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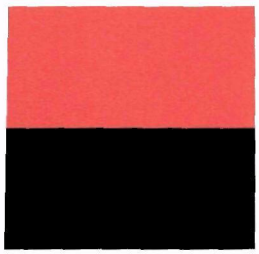

EPRI/EPA 1995 Joint Symposium on Stationary Combustion NO_x Control

Book 3: Thursday, May 18, 1995
Sessions 6A, 6B, 7A, 7B

Sponsored by
Electric Power Research Institute
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U.S. Environmental Protection Agency
Air and Energy Engineering Research Laboratory
Combustion Research Branch

May 16–19, 1995
Hyatt Regency Crown Center
Kansas City, Missouri

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Prepared by
ELECTRIC POWER RESEARCH INSTITUTE

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Session 6A
Coal Combustion

Commercialization of Low NO_x Cell Burner (LNCB®) Technology

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Abstract

Standard, tightly-spaced cell burners were developed by Babcock & Wilcox during the 1960s in response to economic demands for highly efficient burner designs. However, the downside of this 1960s' design is the production of elevated levels of nitrogen oxides (NO_x) emissions which negatively impact the environment.

Cell-fired units have been designated as Phase II, Group II boilers under Title IV, Acid Rain Control, of the Clean Air Act Amendments of 1990 for NO_x control. This paper will discuss one technology developed under the auspices of the U.S. Department of Energy's Clean Coal Technology program for pulverized coal, cell-fired units - namely, the Low NO_x Cell burner (LNCB®) technology.

The body of this paper will describe the development of Low NO_x Cell burner technology and examine six follow-on commercial contracts. The purpose of the paper is to identify similarities and differences in design, fuels, costs and performance results when compared against the Clean Coal Technology prototype.

Introduction

In 1986 Congress established the Department of Energy's Clean Coal Technology (DOE-CCT) program to encourage industry to find new ways to burn coal more efficiently and with significantly less pollution. These 50% cost share programs had a primary goal of accelerating the time table from research and development to commercialization. This paper discusses the 1989 Round III program "Full Scale Demonstration of Low-NO_x Cell Burner Retrofit" (Project DE-FC22-P0P90545) and six follow-on commercial contracts using this technology.

Low NO_x Cell Burner (LNCB®) Technology

Background

The Clean Air Act Amendments (CAAA) were passed by the U.S. Congress in 1990. Title IV called for reduced NO_x levels from all coal-fired utility boilers greater than 25 MW. Title IV has two phases for compliance. The cell-fired units are Phase II and must be in compliance by January 1, 2000. The regulations for Phase II will be promulgated on January 1, 1997. There are 39 originally designed, cell-fired units (27,000 MW_e) which must be retrofitted with low NO_x technology, NO_x averaged with other coal fired units within an owner's/operator's system, or retired from service.

One method for reducing NO_x is to modify the boiler's combustion system. The traditional and more costly approach for standard cell burners is to increase burner spacing with extensive pressure part modifications, and then install conventional, internally-staged low NO_x burners. Increased spacing between burners essentially reconfigures the boiler from cell-fired to wall-fired. This approach can be cost effective if the waterwalls in the burner region of the furnace are approaching end of life and need replacement and if the utility intends to average NO_x across a

number of units.

Goals and Solutions

The objective was to achieve at least a 50% reduction in NO_x emissions without pressure part modifications and without degrading the boiler's performance or life. The new retrofit technology also had to be less expensive and easier to install than the traditional NO_x reduction method involving pressure part reconfiguration.

B&W, in association with EPRI, developed the state-of-the-art Low NO_x Cell burner (LNCB®), designed to limit the formation of both fuel and thermal NO_x. The LNCB design replaces the two circular burners in a standard cell with a single, high-heat input S-type circular burner and a close-coupled overfire air port or NO_x port (see Figure 1). The key design feature is controlled mixing of the fuel and air streams to delay combustion and limit NO_x formation.

The LNCB concept was designed by B&W to fit directly into existing cell burner furnace wall openings (see Figure 2). Only local changes in coal piping at the burner front are required to accommodate the S-burner's increased coal-firing capabilities.

Under the DOE-CCT program, Babcock & Wilcox (B&W) conducted a demonstration on The Dayton Power & Light Company's (DP&L) J. M. Stuart Station Unit 4, a 605 MW_e B&W pulverized coal-fired universal pressure boiler (co-owned by Columbus and Southern Power Company and Cincinnati Gas & Electric Company). This project also was jointly funded by the Ohio Coal Development Office, B&W, the Electric Power Research Institute (EPRI), DP&L, Centerior Energy Corporation, Duke Power Company, New England Power Company, Tennessee Valley Authority and Allegheny Power System.

DOE-CCT Demonstration: Dayton Power & Light

At Stuart Station the burners are arranged in six columns and two rows of two-high LNCB® burners. The heat input per LNCB® burner is 220 x 10^6 Btu/hr at maximum continuous rating (MCR) with six mills in service and 265 x 10^6 Btu/hr at MCR with five mills in service. The scope included provisions for future scanners. New lighters were provided under a separate contract. The fuel burned is an Eastern bituminous coal and has a fixed carbon to volatile matter ratio (FC/VM) of 1.44 and a nitrogen content of 1.23% (see Table 1). During initial implementation of this patented technology, sophisticated three-dimensional computer analysis programs were used to optimize the overall LNCB configuration and combustion performance. This included minimizing CO and H_2S concentrations by inverting a selected number of S-burner and NO_x port couplets (see Figure 3). Field tests confirmed that this final configuration produced the desired retrofit performance.

The result: Stuart Station Unit No. 4's baseline NO_x level of 1.17 lb/10^6 Btu had been reduced by greater than 50% and routinely operates on a system demand basis at less than 0.58 lb/10^6 Btu. A secondary benefit is a cleaner furnace and convection pass which has reduced annual maintenance costs.

One drawback to the design of the deeply-staged system is the elevated potential for corrosion on the furnace walls as a result of localized reducing atmosphere. One commercial remedy for localized corrosion is a field applied chromized coating. Typical costs average $95/ft^2$ and should last a minimum of five years. Many utilities already routinely use commercial coatings to protect furnace walls.

Integration of B&W's LNCB technology into the existing pressure part openings of cell-fired boilers offers many benefits. The demonstration project at Stuart Station achieved NO_x reductions in excess of 50% with no degradation in combustion efficiency and minimal change in carbon monoxide emissions. The twin features of minimum burner piping modifications and no boiler pressure part changes provide benefits of lower cost and shorter schedule than other options.

First Commercial Contract: Allegheny Power System

The first commercial contract for LNCB® technology was awarded to Babcock & Wilcox by Allegheny Power System (APS) for Hatfield's Ferry Unit 2, a 555 MW universal pressure, once-through boiler burning Eastern bituminous coal with a fixed carbon/volatile matter ratio of 1.38 and 1.56% nitrogen in the fuel (see Table 2). This opposed-wall-fired unit consisted of five columns and two rows of two-high LNCB® burners. The original 40 burners were replaced with 20 LNCB® burner sets. Each burner is capable of 252 x 10^6 Btu/hr with all five mills in service and MCR on the unit or 315 x 10^6 Btu/hr at MCR with four mills in service.

Again, mathematical modeling was used to optimize the arrangement of burners and NO_x ports (see Figure 4). The five column arrangement and the angle setting on the spin vanes and louvers are almost identical to the DP&L design. However, at Hatfield, the impeller position is moved back from the front of the nozzle.

Results and Follow-on Contracts: With a NO_x baseline of 1.17 lb/10^6 Btu, the guarantee level was established at 50% reduction or 0.58 lb/10^6 Btu. This condition was met. Based on the performance of Hatfield's Ferry 2, APS contracted with B&W to provide the same design for Units 1 and 3 at this site. These units share the same design, fuel, and burner/NO_x port arrangement as Unit 2.

Unit 1 was installed in October and November 1994. At this writing, Unit 3 is being installed and will start up in April 1995.

Cost Savings: APS enjoyed some significant cost savings on burner auxiliaries by reusing existing drives and scanners. In addition on Unit 3, the original cell burners had been upgraded to convertible S-type burners for mechanical reliability in the mid-80s. Using a large portion of the hardware already in place, B&W was able to field modify 20 S-burners from a 15¼" I.D. coal nozzle to a 21" I.D. nozzle for additional savings to APS.

Second Commercial Customer: Detroit Edison Company

Detroit Edison Company was the second utility to award Babcock & Wilcox a contract for LNCB® technology at Monroe Unit 1. This unit is one of four identical 780 MW universal pressure, once-through boilers with 28 sets of two high burners arranged in eight columns on opposite walls and symmetrically about a furnace division wall.

The fuel burned in the Monroe units is a blend of 45% low sulfur sub-bituminous, 47% mid-sulfur Eastern bituminous and 8% low sulfur Southern bituminous (see Tables 3 and 4).

Design Differences: While Hatfield's Ferry was similar in design and fuel type to Stuart #4, the Monroe #1 unit represented a significant change in furnace design and fuel type. The division wall in the original Monroe furnace design changed the horizontal spacing of the bottom row of burners and eliminated the need for a second set of burners about the two middle columns. Using the existing openings, the mathematical modelers determined that the most effective design to reduce NO_x and minimize CO formation and corrosion along the sidewalls was to invert the burner/NO_x port of each outside cell on the lower row. This variation from the earlier inverted design means that on the lower set of LNCB® burners, burners oppose burners and NO_x ports oppose NO_x ports (see Figures 5 and 6).

The blended fuel with a significant portion (45%) of Western sub-bituminous, low sulfur coal requires a higher fuel flow and consequently a higher air flow to reach MCR on the unit. Therefore, the two original 15¼" I.D. coal pipes were joined with a Y section to form a 23" I.D. pipe and coal nozzle. The burners are rated at 251 x 10^6 Btu/hr with all seven mills in service and 293 x 10^6 Btu/hr with six mills at MCR conditions.

Results and Follow-on Contracts: Baseline NO_x conditions with a blended fuel were 0.93 lb/10^6 Btu at 730 MW. Babcock & Wilcox guaranteed a post-retrofit NO_x level of 0.52 lb/10^6 Btu with CO levels not to exceed 150 ppm. Initial performance tests were analyzed at two loads, seven mills at 750 MW yielding NO_x levels of 0.52 lb/10^6 Btu and at 650 MW, yielding 0.48 lb/10^6 Btu. CO level guarantees were also met. While the % reduction from baseline conditions i.e. 44% was not as large as the demonstration program goal of 50%, the final absolute NO_x numbers were lower i.e. .52 vs. .58 lb/10^6 Btu. The success at Monroe #1 lead to additional orders on Monroe #3 and #4.

Cost Comparison of Design Variations

Variations in scope and material costs will range from a low installed cost of $5.5/kW to $8.5/kW for units with convertible S-burners to a high of $7.0/kW to $10/kW for units where extended scope includes new drives, scanners and lighters as well as field applied coatings for corrosion protection.

NO_x Results Compared Against DP&L Demonstration

Application-specific burner zone heat release rates, furnace configuration, and coal type (e.g., volatility, fixed carbon level, bituminous versus sub-bituminous, nitrogen content, and oxygen content, etc.) will impact expected NO_x results. Boilers firing fuels similar to DP&L's Stuart Station are expected to experience at least a 50% NO_x reduction when retrofitted with LNCB® technology. The results at Hatfield's Ferry confirm this.

Units with higher burner zone heat release rates than J. M. Stuart Station will generally have higher baseline NO_x levels. For these units, LNCB technology has the potential to reduce NO_x emissions by 50% from original design, baseline levels, but not necessarily to absolute levels as low as those attained at DP&L Stuart Station.

Conversely, units like Monroe, with lower baseline NO_x levels, burning sub-bituminous coals or blends may not achieve a 50% NO_x reduction but will have lower absolute NO_x values than similarly configured units burning bituminous fuels (see Table 5).

LNCB Receives R&D 100 Award

The Low NO_x Cell burner was so successful, cost effective and innovative, that it was selected to receive the coveted 1994 R&D 100 Award. This award, sponsored by R&D Magazine, recognizes products or inventions introduced for commercial use during the preceding year that have had a meaningful and profound impact on technology or industry. With that in mind, the panel of judges cited LNCB technology to be "one of the most technologically significant new products of the year."

Conclusion

B&W has installed or received orders for its LNCB® technology on more than 5750 megawatts of capacity in the United States. Regardless of boiler configuration or fuel characteristics, the advantages of the LNCB technology for utilities include:

- No pressure part modifications
- Short outage schedule
- Same number of original lighters
- Elimination of 50% of the original scanners
- Potential reuse of existing drives, lighters and scanners
- Cleaner furnace and convection surface
- NO_x reductions of 50% or absolute values below 0.6 $lb/10^6$ Btu
- Low installed cost

The success of this design on a variety of fuels and boiler configurations makes it a prudent choice for establishing levels of NO_x reduction for Phase II, Group II cell-fired boilers.

Acknowledgements

The authors wish to thank the following government agencies, private entities and utilities whose sponsorship made these projects possible:

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- Ohio Coal Development Office (Ohio Department of Development)
- Electric Power Research Institute
- The Babcock & Wilcox Company
- The Dayton Power & Light Company
- Allegheny Power System
- Centerior Energy Corporation
- Duke Power Company
- New England Power Company
- Tennessee Valley Authority
- Cincinnati Gas & Electric Company
- Columbus and Southern Power Company
- Detroit Edison Company

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TABLE 1 - DAYTON POWER & LIGHT, STUART #4, FUEL			
Ultimate Analysis		**Proximate Analysis**	
(as received)	**%**	**(as received)**	**%**
Carbon	66.52	Moisture	5.55
Hydrogen	4.49	Fixed Carbon	47.61
Oxygen	7.51	Volatile Matter	33.14
Nitrogen	1.23	Ash	13.70
Sulfur	1.00	**Total**	**100.00**
Moisture	5.55	FC/VM = 1.44 HHV = 11,880 Btu/lb	
Ash	13.70		
Total	**100.00**		

TABLE 2 - ALLEGHENY POWER SYSTEM, HATFIELD'S FERRY #2, FUEL*			
Ultimate Analysis		**Proximate Analysis**	
(as received)	**%**	**(as received)**	**%**
Carbon	72.79	Moisture	4.90
Hydrogen	4.85	Fixed Carbon	50.13
Oxygen	4.95	Volatile Matter	36.29
Nitrogen	1.56	Ash	8.68
Chlorine	0.11	**Total**	**100.00**
Sulfur	2.16	FC/VM = 1.38 HHV = 13,083 Btu/lb	
Moisture	4.90		
Ash	8.68		
Total	**100.00**		

* Typical analysis used during guarantee testing.

TABLE 3 - DETROIT EDISON, MONROE #1, BLENDED FUEL				
Ultimate Analysis	**45% Western**	**47% Eastern**	**8% Southern**	**Blend**
Carbon	51.96	72.76	69.86	63.17
Hydrogen	3.71	4.88	4.30	4.31
Oxygen	12.44	6.09	8.60	9.15
Nitrogen	0.70	1.41	1.22	1.07
Chlorine	0.00	0.09	0.09	0.05
Sulfur	0.30	1.30	0.90	0.82
Moisture	26.28	6.38	6.05	15.31
Ash	4.61	7.09	8.98	6.12
Total	**100.00**	**100.00**	**100.00**	**100.00**

TABLE 4 - DETROIT EDISON, MONROE #1, BLENDED FUEL				
Proximate Analysis	**45% Western**	**47% Eastern**	**8% Southern**	**Blend**
	Proximate	**Proximate**	**Proximate**	**Proximate**
Fixed Carbon	36.61	50.46	49.65	44.16
Volatile Matter	32.50	36.07	35.32	34.40
Moisture	26.28	6.3	6.05	15.31
Ash	4.61	7.09	8.98	6.13
FC/VM:	1.13	1.40	1.41	1.28
Total	**100.00**	**100.00**	**100.00**	**100.00**
Btu/lb	9,055	13,090	12,620	11,237

TABLE 5 - COMPARISON OF RESULTS			
Site	**DP&L, Stuart**	**APS, Hatfield's Ferry**	**DECO, Monroe (blended analysis)**
HHV, Btu/lb	11,880	13,083	11,237
FC/VM	1.44	1.38	1.28
N_2	1.23%	1.56%	1.07%
Baseline NO_x	1.17	1.17	0.93
Retrofit NO_x	<0.58	<0.58	<0.52
% Reduction	50%	50%	44%

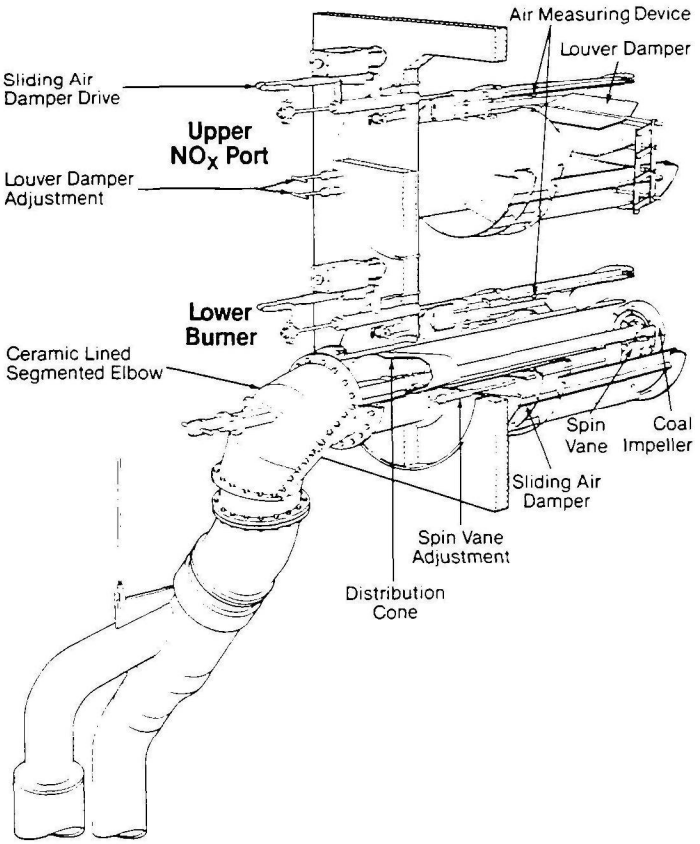

Figure 1 - B&W's Low NO_x Cell Burner.

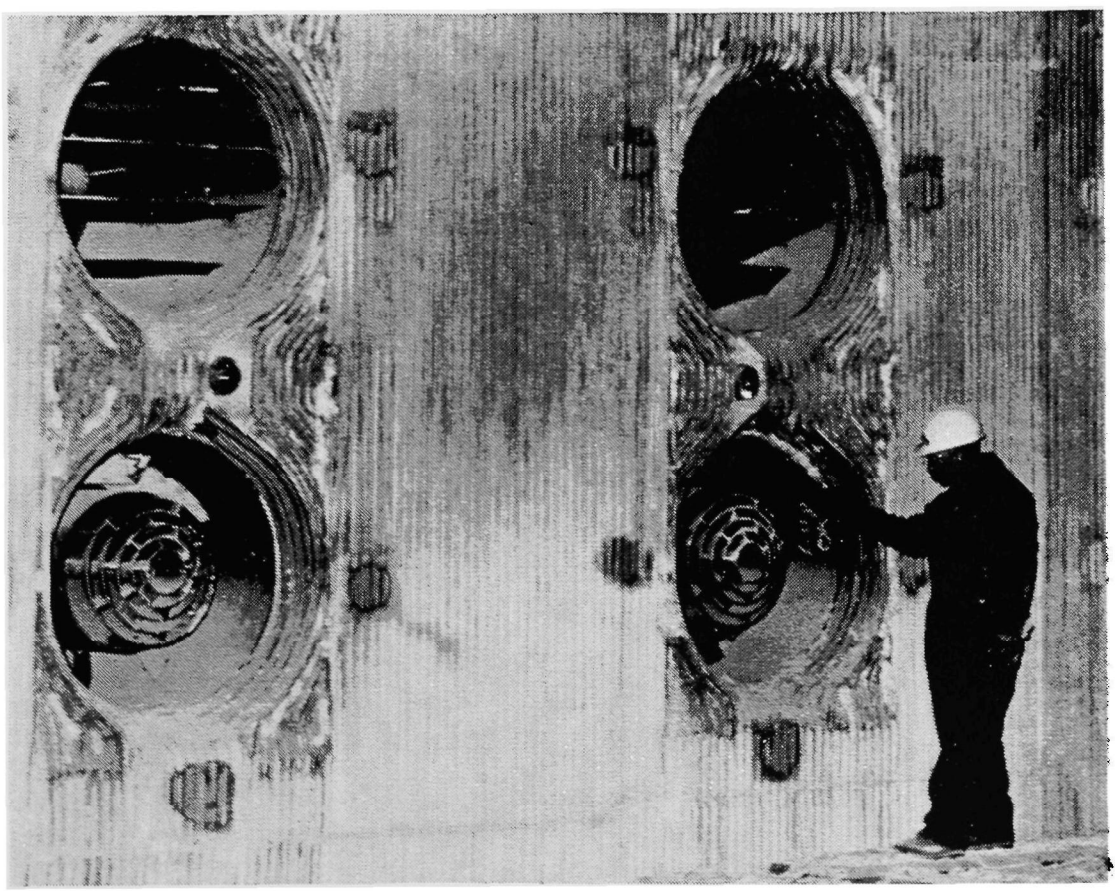

Figure 2 - The LNCB was designed to fit directly into existing cell burner furnace wall openings.

Dayton Power & Light
Stuart #4 605 MW

Scope of Supply:
24 Low NO_X Cell Burners

● **NO_X Port**
○ **Burner**

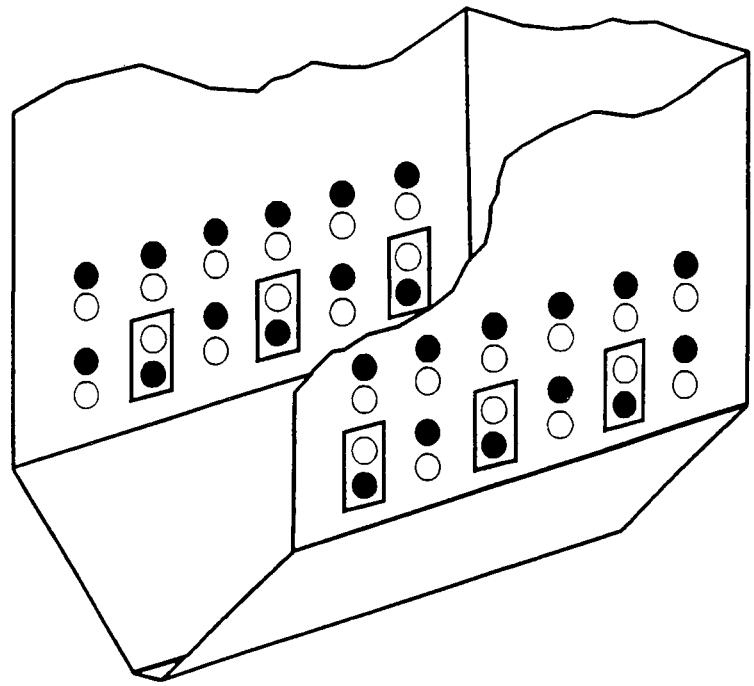

Figure 3 - Demonstration site furnace configuration.

Allegheny Power System
Hatfield's Ferry #2 555 MW

Scope of Supply:
20 Low NO_X Cell Burners
20 Reused existing lighters, scanners and drives

● **NO_X Port**
○ **Burner**

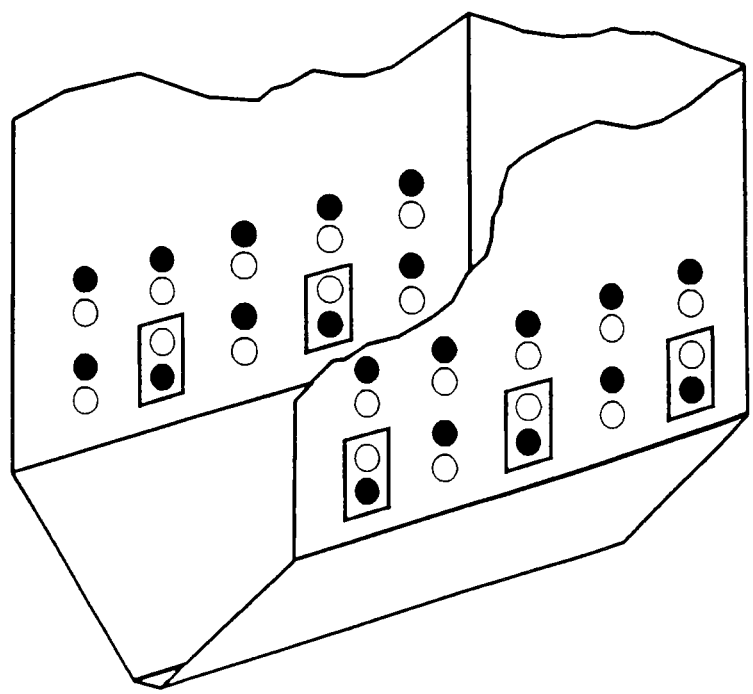

Figure 4 First commercial contract furnace configuration.

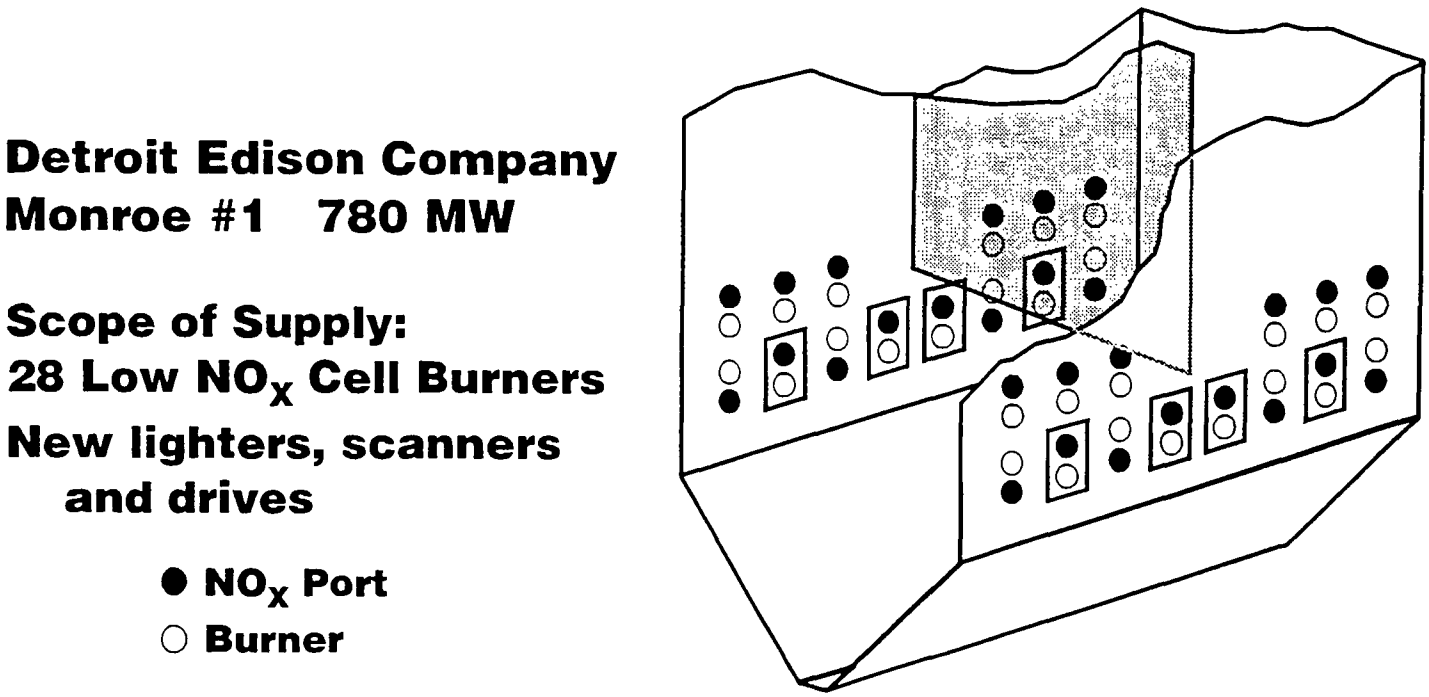

Figure 5 - Second commercial contract furnace configuration.

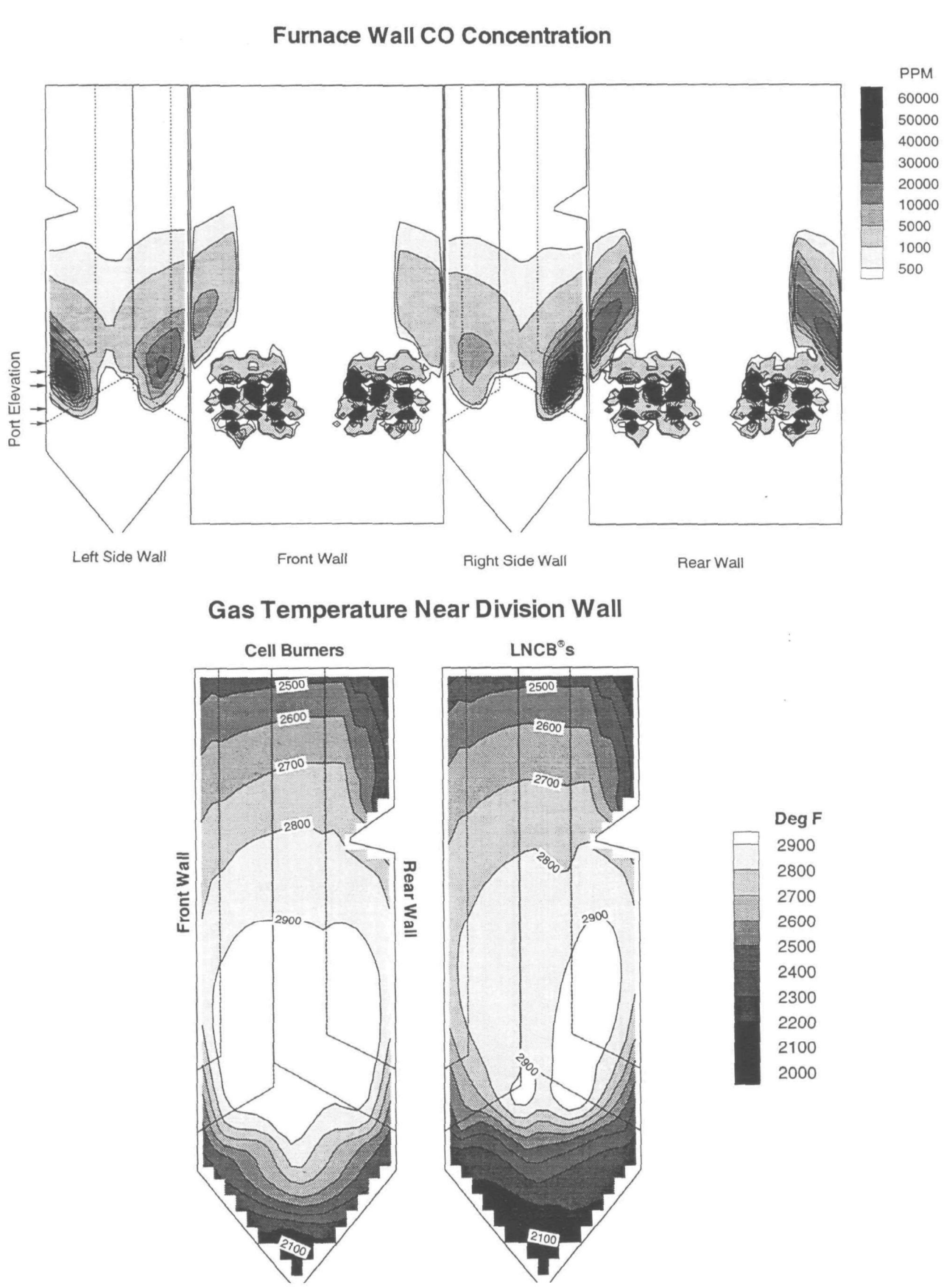

Figure 6 - DECO LCNB® base case.

ALTERNATIVE SOLUTIONS FOR REDUCING NO_x EMISSIONS FROM CELL BURNER BOILERS

by

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Abstract

An economical solution for reducing NO_x emissions from coal fired utility boilers equipped with pre-New Source Performance Standard (pre-NSPS) cell burners is being developed. Since over 12% of the U.S. generating capacity is produced from cell burner boilers, this development is very significant for the utility industry.

Cell burners, manufactured in the 1950's and 1960's, rapidly mix the pulverized coal and combustion air resulting in highly turbulent and efficient combustion. Typical NO_x emissions from cell burner boilers average 1.0 to 1.8 lb/10^6 Btu.

This paper presents the results of retrofitting American Electric Power's Muskingum River Unit 5, a 600 MWe supercritical cell burner boiler, with Riley low NO_x CCV™ burners. Initial results demonstrated the ability to reduce NO_x emissions greater than 50% without the requirement for overfire air, off stoichiometric firing, burner respacing, mill system and coal piping changes or pressure part modifications. Long term operation, which recently included switching coals, has resulted in some deterioration in NO_x performance. Results of the initial optimization testing are presented along with plans for reoptimizing unit operation to restore the low NO_x levels.

Plans for developing Riley's "next generation" low NO_x pulverized coal burner (CCV II) will also be presented. The intent is to potentially utilize this advanced design on other cell burner units in American Electric Power's system for improved NO_x reduction performance.

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Introduction

American Electric Power (AEP) based in Columbus, Ohio has over 24,000 MWe of installed electrical generating capacity. Over 5600 MWe of this electrical capacity comes from six (6) coal fired boilers equipped with pre-NSPS cell burners making AEP one of the largest owners of these boiler types. The six (6) units range in nominal size from 600-1300 MWe generating capacity.

In 1993, AEP implemented a program to find a viable solution for reducing NO_x emissions from cell burner boilers. Specific objectives were to:

- reduce NO_x emissions by at least 50%
- reduce NO_x emissions using low NO_x burners that would not require air staging via overfire air (OFA) or off-stoichiometric firing
- develop a technical and economical low NO_x technology for application to all of AEP's cell burner boilers
- reduce NO_x emissions without degradation in boiler performance

A review of available technology in the industry indicated that NO_x reduction was possible by, (1) installing modified cell burners that replace the upper nozzle with an OFA port or by, (2) respacing and installing low NO_x burners to resemble a more conventional wall fired installation. These options were abandoned because of concerns for lower furnace corrosion, the long term effects of deep staging on overall boiler operation, and the significant cost associated with pressure part modifications to rearrange the burners.

The technical approach desired by AEP was to install low NO_x burners without OFA as a direct replacement for all the cell burners. Riley Stoker Corporation (Riley) was selected by AEP to implement this approach. The significant factors influencing this decision were:

- With the Riley Controlled Combustion Venturi (CCV®) burners we did not anticipate any requirement for overfire air or air staging for reducing NO_x by 50% thus minimizing lower furnace waterwall corrosion due to reducing atmospheres.
- The CCV® burners were simply a "plug in" replacement for the pre-NSPS cell burners and would not require costly pressure part changes.
- The CCV® burners would not require any changes to the existing milling system and coal piping configuration.
- The CCV® burners would not require respacing the burners despite the high furnace heat release rates typical of these types of boilers.

Unit Description

AEP and its subsidiary, Ohio Power, elected to utilize Muskingum River Unit 5 (MR5), shown in Figure 1, as the demonstration site for the Riley burner technology. This unit, originally designed and manufactured by Babcock and Wilcox (B&W) in the early 1960's, is located in Beverly, Ohio. The boiler is a supercritical design producing superheated steam at a rate of 4,035,000 lb/hr, 3800 psig operating pressure and 1000°F operating temperature. The electrical generating capacity is a nominal 600 MWe.

As shown in Figure 2, pulverized coal was previously burned using twenty (20) B&W two-nozzle cell burners and ten (10) standard circular burners arranged in an opposed fired configuration. Figure 3 shows a schematic of the original 2 nozzle cell burner. The unit has five (5) B&W MPS size 89 pulverizers which feed ten (10) coal nozzles each. The furnace dimensions are approximately 63' W x 39' D. MR5 burned a high volatile bituminous coal from Central Ohio. Typically, uncontrolled NO_x levels from cell burner type boilers average from 1.0 to 1.8 lb/10^6 Btu. The uncontrolled NO_x level measured at Muskingum River Unit 5 was 1.2 lb/10^6 Btu. The objective of this project was to reduce NO_x emissions to below 0.6 lb/10^6 Btu with minimal impact on boiler performance and, in particular, flyash unburned carbon (UBC) levels.

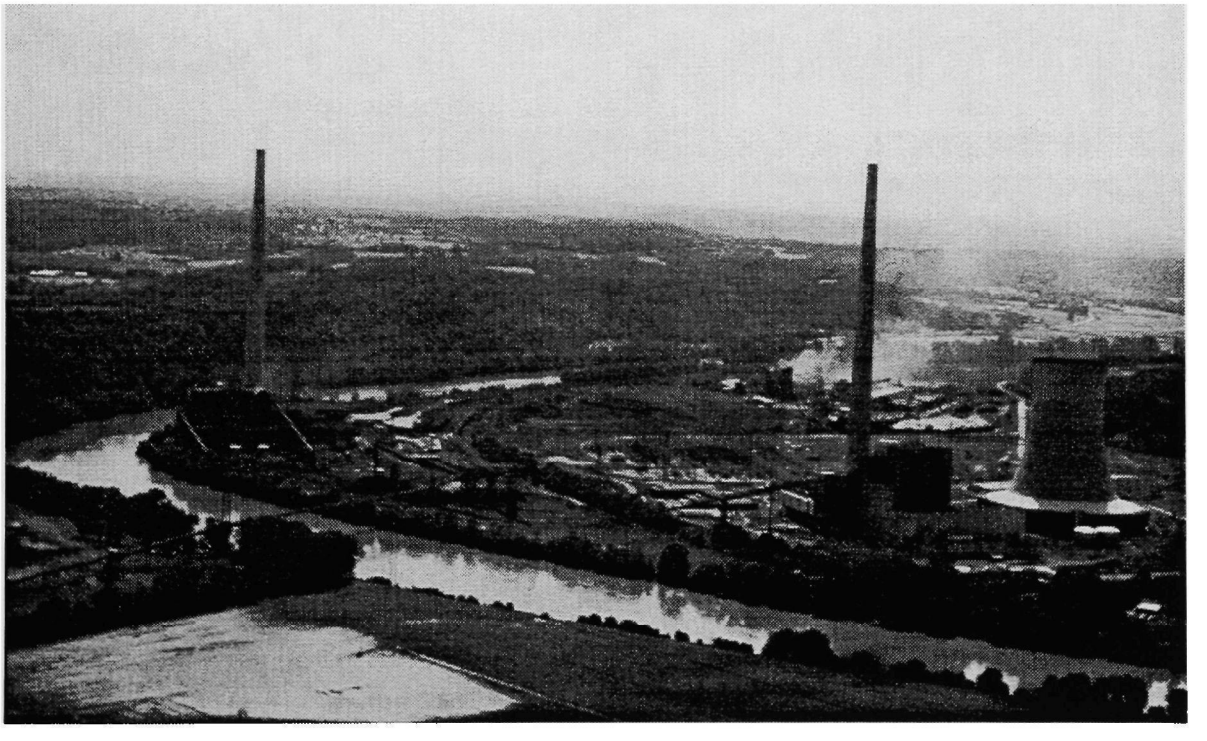

Figure 1
Muskingum River Unit 5 (on right)

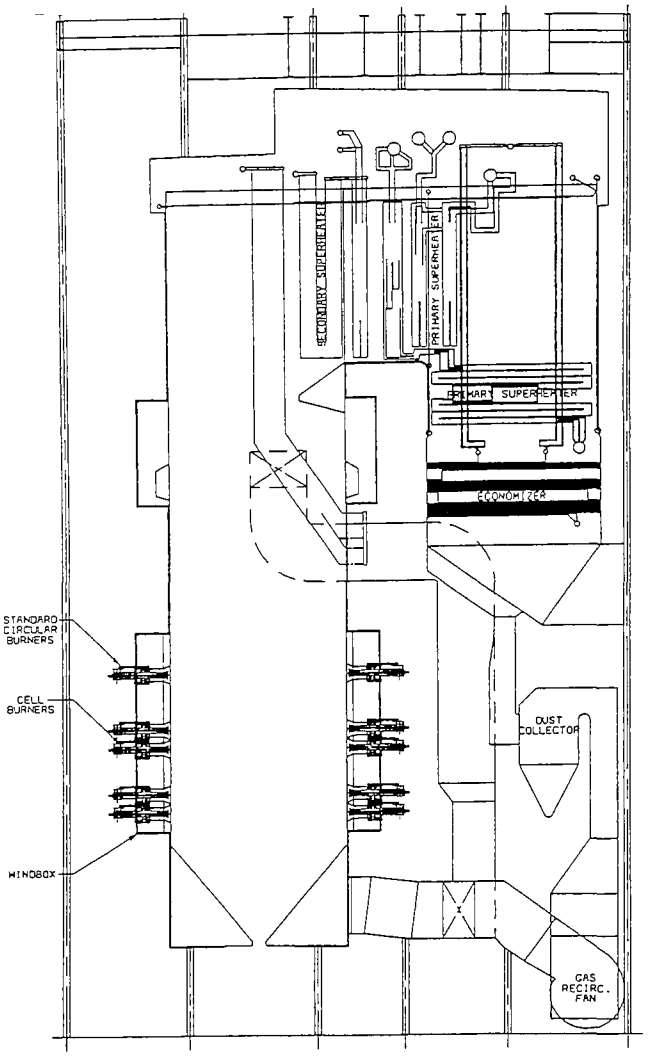

Figure 2
Muskingum River Unit 5

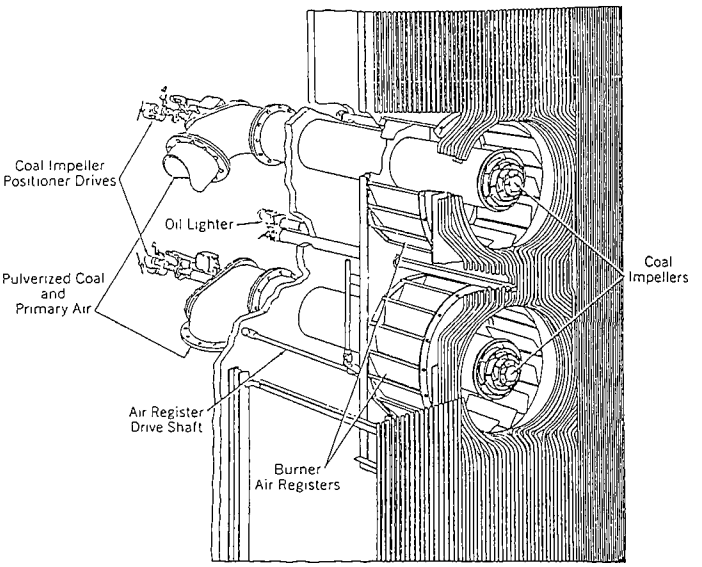

Figure 3
Original 2-Nozzle Cell Burner

Low NO_x Combustion Retrofit

In the spring of 1993, AEP contracted with Riley for the design, fabrication and installation of fifty (50) low NO_x CCV® burners for Ohio Power's Muskingum River, Unit 5. Each burner was rated for 107 MMBtu/hr heat input at MCR conditions. The basic CCV® burner design is shown in Figure 4. The key element of the burner design is the patented (U.S. Patent No. 4,479,442) venturi coal nozzle and low swirl coal spreader located in the center of the burner (1). The venturi nozzle concentrates the fuel and air in the center of the coal nozzle creating a very fuel rich mixture. As this mixture passes over the coal spreader, the blades divide the coal stream into four (4) distinct streams which then enter the furnace in a gradual helical pattern producing very gradual mixing of the coal and secondary air. Secondary air is introduced to the furnace through the air register, supported off the burner front plate, and subsequently through the burner barrel and over the secondary air diverter.

Devolatilization of the coal in the fuel rich mixture occurs at the burner exit in an oxygen lean primary combustion zone, resulting in lower fuel NO_x conversion. Peak flame temperature is reduced by the gradual mixing, thus suppressing the thermal NO_x formation.

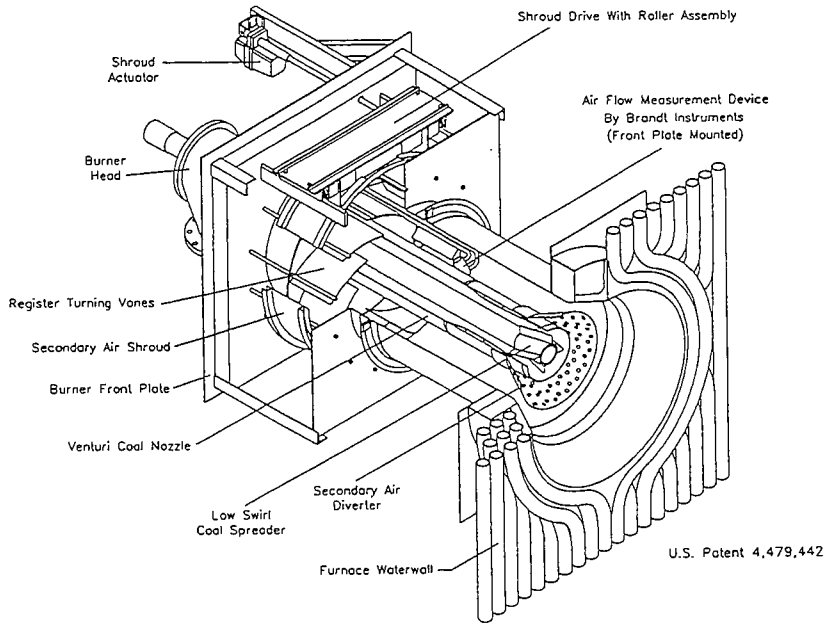

Figure 4
Riley Low NO_x CCV® Burner

The resulting coal flame produced by the CCV® burner is longer, narrower and more tubular in shape than the flames produced by the highly turbulent original B&W cell burners. The CCV flame is also typically well attached to the coal nozzle tip resulting from recirculation eddies formed on the backside of the secondary air diverter. For good NO_x control, it is very important to have a well attached coal flame (3). As discussed later, detached coal flames on the upper row of standard circular burners may be limiting the full NO_x reduction potential of this retrofit.

The furnace depth of 39' is sufficient to accommodate the longer low NO_x coal flames without concern for waterwall flame impingement. However, our most significant concern on this project was that, with the very close burner spacing (4'-2") between the two (2) nozzles in a cell burner (Figure 5), the coal flames would interact with each other causing more turbulent combustion than desired and minimal reduction in NO_x emissions.

The new low NO_x burners were installed in the fall of 1993 during a six (6) week outage. The unit was also equipped with a 33 point sampling grid at the economizer outlet for measuring gas emissions during the post retrofit performance testing. Start-up of MR5 commenced in December, 1993. However, difficulties in trying to utilize the original mechanically atomized oil ignitors, which were shared between the burner nozzles within the cell, forced a delay in testing the CCV® low NO_x burner configuration until June of 1994.

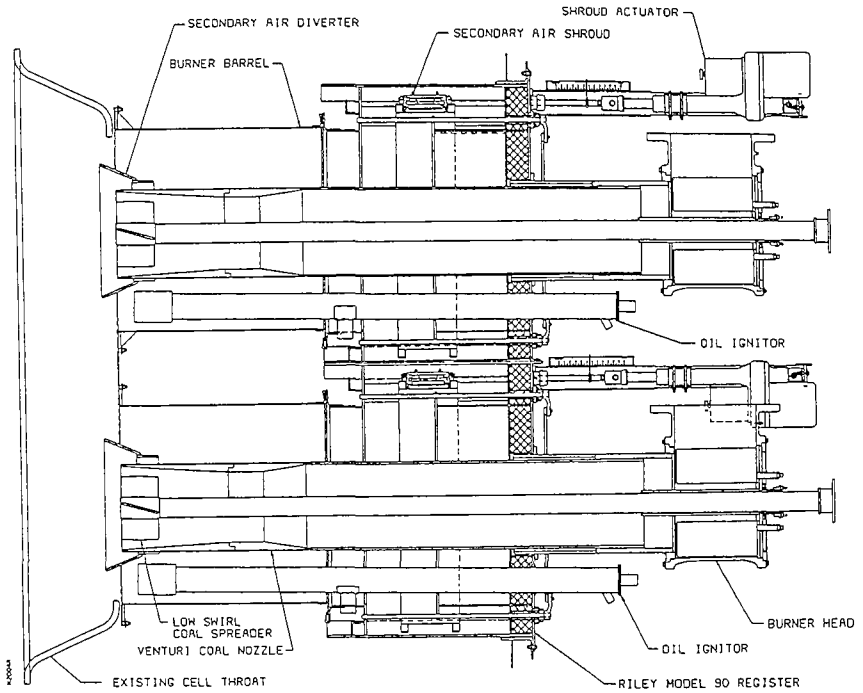

Figure 5
Riley Low NO_x CCV® Cell Burner

Optimization Test Results

Post retrofit optimization testing was performed in early summer, 1994. Our primary concern regarding flame interaction impacting NO_x with the close burner spacing quickly diminished after observing the CCV® cell burner flame shape. The CCV® cell burner flames were all well attached , narrow and tubular in shape. The flames were distinctly independent from each other with little or no flame interaction between adjacent flames.

The upper row of non-cell configured burners also produced a long narrow flame but were detached by various amounts. The detachment was felt to be the result of:

- windbox air maldistribution
- insufficient diverging throat depth for adequate burner recirculation.

Regardless, the optimization testing continued.

As summarized in Figure 6, NO_x was reduced to below 0.6 lb/10^6 Btu for a 52% reduction from pre-retrofit levels. CO emissions remained extremely low and consistent with baseline operation. Flyash LOI averaged 2.23% as compared to 2.7% prior to the retrofit while the actual % of unburned carbon in the ash averaged $< 1\%$ as compared to 1.5% prior to the retrofit. As shown in Figures 6 and 7, NO_x decreased with reduction in load and with normal increases in excess air for maintaining steam temperature.

Parameter	Pre-Retrofit Baseline	Post Retrofit Optimization		
Test No.	2	9	11	10
Date	9/16/93	7/6/94	7/7/94	7/7/94
Gross Generation, MW	600	607	454	368
Excess Air, %	25	22	33	40
Superheat Outlet Temperature, °F	1000	994	996	993
Reheater 1 Outlet Temperature, °F	1023	1010	970	937
Reheater 2 Outlet Temperature, °F	1024	1003	960	917
Econ. Outlet Gas Temperature, °F	690	735	676	648
AH Outlet Gas Temperature, °F	341	353	328	325
Boiler Efficiency, % (Approx.)	87.85	87.63	-	-
NO_x @ Econ. Outlet by CLA[1], 10^6 Btu		0.56	0.53	0.51
NO_x @ Stack by CEMS, lb/10^6 Btu	~ 1.2[2]	0.59		
CO @ Econ. Outlet, ppm	0	1	1	1
Flyash LOI, %	2.7	2.23	1.89	1.70
Carbon in Ash %	1.5	0.78	0.67	0.95

(1) Chemilluminescent Analysis
(2) Measured during previous boiler testing.

Figure 6
Summary of Test Results

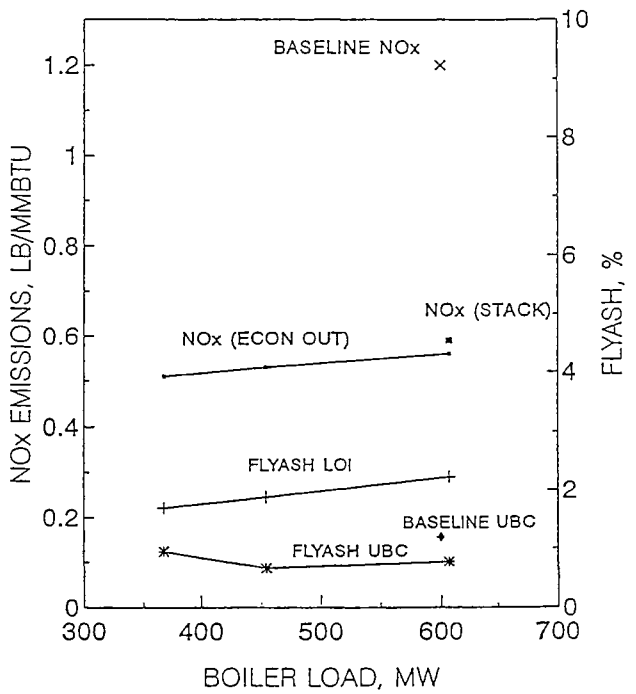

Figure 7
The Effect of Load on NO_x and LOI

While the Muskingum River Unit 5 burner arrangement benefits somewhat in terms of NO_x control (baseline and post-retrofit) from a row of circular burners above the two levels of cell burners (Figure 2), full load is achievable with this row of burners out of service. This firing condition was also tested to simulate the furnace thermal environment of AEP's cell fired 1300 MWe units. The NO_x emissions under this condition increased at full load to 0.63 lbs/10^6 Btu, less than the 0.66 lbs/10^6 Btu predicted, which supports that a nearly 50% reduction in NO_x from the existing 1300 MW units is possible with "plug in" CCV® burners.

It was also observed throughout this testing that there was no significant change in boiler thermal performance as a result of installing low NO_x burners. Figure 8 shows a typical fuel analysis of the coal burned during the optimization testing along with pulverized coal fineness results.

Coal Source	Central Ohio	E. Kentucky
Proximate		
Moisture %	6.6	6.8
Volatile Matter, %	39.1	33.54
Fixed Carbon, %	42.5	47.82
Ash, %	11.8	11.84
Ultimate		
Carbon, %	68.3	74.0
Hydrogen, %	5.0	4.79
Oxygen, %	8.41	6.20
Nitrogen, %	0.99	1.48
Sulfur, %	4.7	0.70
Higher Heating Value, Btu/lb	11,660	12,153
Hardgrove Grindability	50	45
Coal Fineness		
% 50 Mesh	99.5	98.4 99
% 100 Mesh	95.5	
% 200 Mesh	79.1	60 +

Figure 8
Fuel Analysis and Coal Fineness Results

Long Term Operating Results

Long term operation following the optimization testing has resulted in NO_x levels deteriorating. Data collected from the CEMS equipment indicated NO_x emissions had increased to an average 0.69 lb/10^6 Btu up until mid December, 1994. Then, a coal change to a lower sulfur E. Kentucky coal occurred. NO_x emissions increased further to an average 0.75 lb/10^6 Btu and also became more erratic. Figure 8 shows a comparison of the fuel analysis.

The most significant factor affecting the performance is felt to be due to changes in mill operation as a result of changes in grindability, moisture content and heating value. Primary air flow quantity was reduced with the new fuel's higher heating value which in turn affects venturi nozzle velocities, burner mixing and NO_x emissions. Other factors may include boiler seasoning, excess air and burner air distribution.

Plans are to reoptimize the unit operation with particular emphasis on burner air distribution, proper flame attachment and mill operation. The objective is to reduce the NO_x level back down to below 0.6 lb/10^6 Btu. Long term parametric testing is planned which will include evaluating specific variables such as primary air flow, excess air, burner air distribution and various burner adjustments.

In parallel with this reoptimization effort, Riley will be utilizing mathematical modelling (Fluent analysis) to evaluate the flame detachment concern for the upper circular burners. Control of NO_x is difficult with detached coal flames. Potential solutions will be developed.

Advanced Low NO_x Coal Burners

Riley is currently implementing a program to develop its "next generation" low NO_x coal burner design. The program will evaluate the effects of dual secondary air streams, nozzle and spreader configurations on controlling NO_x emissions. Testing, which is scheduled to commence in early summer 1995, will be conducted on prototype burners installed in the 100 MMBtu/hr coal burner test facility at Riley Research. All previous testing of CCV burners have been conducted here for direct comparisons of the test results.

Included in this program will be the installation of DB-Riley's dynamic SLS classifier for studying the effects of various pulverized coal particle size distributions on emissions performance and flyash unburned carbon levels. The dynamic classifier will be integrated with the existing Atrita milling system for direct coal firing. An evaluation of the feasibility of incorporating dynamic classifiers on Atrita milling systems will be made.

The program will also focus on testing three (3) different coals of various volatile contents and reactivity. Our field experience has shown a significant impact of coal quality on combustion efficiency (unburned carbon) and to a lesser extent on NO_x. The effects of various coal qualities on burner performance will be determined during this test program.

Summary

DB-Riley and AEPSC have retrofitted low NO_x burner technology on a coal fired 600 MWe cell burner supercritical unit. Of particular significance is the ability to retrofit with this "plug in" technology without pressure part modifications, burner respacing and off stoichiometric firing using overfire air ports.

Test data demonstrated the ability to achieve 50% NO_x reduction at MR5 (< 0.6 lb/10^6 Btu) without degradation in boiler performance during post retrofit optimization testing. Long term operation has shown some deterioration in NO_x to levels exceeding 0.7 lb/10^6 Btu. Efforts to reoptimize the unit operation as well as an evaluation of design upgrades are continuing.

Advanced burner designs are being tested at Riley's Research facilities for improved NO_x control and combustion efficiency. A total systems approach including the burner, dynamic classifier and various coals of different quality and characteristics is being evaluated to meet future requirements. The feasibility of using dynamic classifiers for improved fineness and combustion efficiency would require an economic analysis on a site specific basis. It is anticipated future requirements will include stricter air quality standards as well as special requirements for ash disposal and utilization (3).

Acknowledgements

Special thanks are extended to Ohio Power and AEPSC for their continued support throughout this low NO_x retrofit project. Also, thanks are extended to Riley's Service Department for their dedicated efforts during the start-up and testing of MR5.

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CO-FIRING COAL-WATER SLURRY IN LOW-NOx BURNERS: Experience at Penelec's Seward Station

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Abstract

The electric utility industry is facing significant challenges for meeting environmental regulations for its coal fired generation units, and at the same time must stay cost competitive in an industry that is being deregulated. An emerging technology that can reduce environmental equipment costs, lower fuel costs, and help to mitigate the environmental hazard of coal fine impoundments is the recovery and use of coal fines (<100 mesh) in the form of a coal-water slurry (CWS). The are a number of issues to be addressed in the commercialization of this technology: coal fines recovery and beneficiation; CWS formulation and transportation; and storage, handling, and firing by the combustion equipment owner. This paper deals primarily with the utilization of coal-water slurry in utility boilers. Experience on a 32MWe wall-fired boiler at the Pennsylvania Electric Company (Penelec) Seward Station indicates that CWS can be stored, handled, and co-fired with pulverized coal with only minimal modifications to the low-NOx coal burners and no modifications to the boiler. There was no degradation in combustion performance and NOx emissions were reduced approximately 20% when firing up to 35% of the total boiler heat input as coal-water slurry. Penelec contracted to purchase 2,500,000 gallons of CWS and are currently co-firing the fuel as it is produced and delivered. CWS fuel made from coal cleaning plant waste fines and from coal fines recovered from existing impoundments is expected to be available for less than $1.00/MMBtu.

Introduction

Emissions from coal-fired central-station power plants continue to play a significant role in the quality of the environment. The preparation and combustion of coal in a pulverized coal fired boiler raises several environmental concerns, namely the release of:

- fine coal particles lost from the preparation process
- sulfur dioxide due to pyritic and organic sulfur in the coal,
- nitrogen oxides due to fuel bound nitrogen and to high combustion temperatures,
- particulate matter due to the ash content of coal and unburned carbon,
- hazardous air pollutants (HAPs) present in the coal.

In general, the higher the ash, sulfur, and nitrogen content of the coal, the higher the emissions per million Btu of coal fired. The recent revisions to the Clean Air Act Amendments mandate reductions in NO_x and SO_x emissions and future legislation is expected to address reduction requirements for both fine particulate and hazardous air pollutants.

In addition to the emissions from the plant sites themselves, there are concerns with the fine coal waste streams from coal processing facilities operated to provide fuel to numerous power plants. Also of significant concern are the two billion tons of coal fines deposited over the years in pond impoundments of various sizes all over the country.

An innovative approach is needed to apply technology that addresses each concern in a cost effective, environmentally acceptable manner that may be implemented synergistically. The approach should be applicable to a variety of coal-fired combustion systems and power cycles. Further, the commercialization of the integrated technologies should consider both domestic and international applications.

Conventional washing removes some impurities from coal and helps in reducing sulfur and particulate emissions. Deep physical cleaning, to remove additional impurities, has not been commercially practiced because it requires energy intensive grinding of the coal to a very fine size so that impurities can be liberated. The resulting fine coal product is difficult to handle and transport. Adding water to these fines to produce a coal water slurry fuel has been considered impractical due to the cost of fine coal preparation, slurry formulation additive cost and "potential" boiler impacts.

It is important to note that the coal-water slurry discussed in this paper is significantly different than the coal-water slurries which were the subject of so much attention in the early 1980's.[1-8] The objective of the work at that time was to develop a coal-based fuel that would displace oil and natural gas in boilers designed to fire only oil or natural gas. Efforts were focused on forulating slurries with very high solids loading (65-75%). A number of issues arose as attempts were made to demonstrate the feasibility of the technology:

- slurry formulation required significant amounts of grinding and expensive additives
- slurry rheology very sensitive to coal properties
- slurry characteristics changed during storage
- flame stability difficult to achieve
- difficulty with numerous impacts on boilers due to the presence of coal ash

Eventually, the technology was abandoned due to the unfavorable economics created by the high costs incurred to address these issues and the relatively low cost differential between coal and the premium oil and natural gas fuels.

Slurry formulation developments, along with the identification of available fine coal streams, have eliminated the need for the grinding and the additives. This approach results in CWS fuels which have lower concentrations of solids (coal), in the range of 50-55%, but are significantly lower in cost than conventional CWS fuels developed in the past.

To encourage the further development of this technology, Penelec is proceeding with a multi-phase program to demonstrate the technical and economic viability of using waste coal fines to reduce operating costs. EER has supplied the technical expertise to store, handle, control, and use the CWS in their Seward Station Unit #14. Recently, EPRI's Upgraded Coal Interest Group (UCIG) has provided financial support for additional studies of fine coal resources, their cleanability, and use. In addition, the UCIG is assisting in the transfer of this technology throughout the utility industry.

Coal Fines Utilization

The history of coal cleaning practices in the United States involving the recovery of fine clean coal has always been conspicuously inadequate. Because any cleaning (demineralization) and inclusion of fines into the main product frequently marginalized or degraded the market value of the coal due to the poor removal of impurities and/or moisture,

there has always been a tendency to discard all or much of the fine clean coal. Frequently, the disposal is in the form of a slurry into waste impoundments either directly or as poorly or partially cleaned refuse. It is not unusual for discarded fine slurry coal to contain more mineral impurities than the original raw coal and, even if dewatered, to contain an unmarketable moisture content in the range of 20 to 40 percent. The problem of excess moisture content is completely mitigated by the direct combustion of the fines in slurry form.

In a commercial coal prep plant, coarse coal, usually plus 28 mesh, is cleaned and sold. The -28 mesh coal may be recovered in conventional flotation cells or as raw coal through thickening and filtration circuits. It can then be mixed with the clean product. If fine coal is to be widely used as a fuel, it must be cleaned and then recovered in a usable form. Coal-water slurry is an obvious solution to the fines handling problems. If CWS is produced, then the fuel consumer must be willing to purchase this coal and to modify combustion equipment to burn the slurry. The incentive for the consumer, primarily the coal burning utility companies, will be a clean fuel at low cost. In some instances, the CWS will be a source of compliance coal or low sulfur coal for blending.

The Pennsylvania Electric Company (Penelec) began looking for alternative methods of utilizing the <100 Mesh coal fines stream from the Homer City Coal Cleaning Facility located adjacent to the Homer City Generating Station in Homer City, Pennsylvania. The current practice involves dewatering the fines stream and further reducing the moisture content in large, coal-fired dryers. The fines are then added to the remainder of the coal stream into the plant. The Station experiences a number of storage and handling difficulties, however, as a result of the presence of those fines. In addition, the coal dryers consume 50,000 tons of coal and are responsible for almost 1500 tons of SO_2 emissions each year.

One of the most promising alternatives appeared to be the use of the waste coal fines stream in its existing form - a coal and water mixture. Subsequent laboratory and pilot scale work indicated that the mixture could be concentrated to approximately 50% coal by weight without additives or additional adjustment to the particle size distribution. A series of small scale combustion tests at Penn State University verified that the mixture could be co-fired with pulverized coal with minimal changes to the flame characteristics.[9] Penelec management felt that there would be minimal risk in proceeding with a larger scale demonstration in a small utility boiler.

Application to Utility Boiler

Seward Station Boiler #14 was selected as a candidate host unit for the demonstration. The unit was already scheduled for a low-NOx burner retrofit and Penelec simply added the CWS co-firing requirement to the low-NOx firing system specification.

Boiler Description

Penelec's Seward Station Boiler #14 is a B&W Sterling type, front wall fired, dry bottom furnace that began operation in 1950. The boiler is normally operated at 360,000 lb/hr steam at 675 psig and 835 °F. Each boiler superheater is a pendant type with steam temperature controlled by a spray attemperator. Figure 1 illustrates a side view of the boiler and shows the principle dimensions. The burners are configured in two elevations of three burners per row. This 2 by 3 burner array is closely spaced with a horizontal spacing of 6 feet and a vertical spacing between the rows of just 5 feet. Pulverized coal is supplied to the burners by two modified B&W "EL" pressurized pulverizers. Each pulverizer has a capacity of 8-10 tons per hour. EER was awarded a contract in September 1993 to supply 6 FlameMastEER™ Low-NOx burners and associated equipment. Three of the burners can supply up to 40% of the burner heat input as coal-water slurry.

Burner Description

EER designed, fabricated, and delivered 6 Model 4AF FlameMastEER Low-NOx burners in April 1994. Each burner is rated at 91 MMBtu/hr which enables the unit to achieve full load operation with one burner out-of-service. The burner throats were increased from 31 to 34 inches in diameter with new refractory throat tiles and tube openings. Each burner was equipped with a #2 oil lighter. The lower three burners were supplied with air atomized CWS atomizers that can provide up to 40% heat input to each burner.

A sectional assembly view of the FlameMastEER™ Burner is shown in Figure 2. The key functional features of the design include:

- Combustion air supply through separate secondary (SA) and tertiary (TA) passages
- Variable SA/TA flow distribution

- Variable SA and TA swirl
- Low pulverized coal velocity
- Flameholder attached to the coal nozzle

The air distribution between the secondary and tertiary zones, along with the adjustment of swirl of both flows, represents the operating variables for control of NO_x emissions. The low primary air/coal velocity and flameholder are designed to provide good flame stability and acceptable flame characteristics for a wide range of operating conditions and fuel characteristics. The flameholder establishes local recirculation zones and promotes local mixing between the coal and the secondary air. This leads to a rapid devolatilization of the coal and liberation of fuel nitrogen in a low excess air environment resulting in reduced NO_x formation.

Mechanically, the burner has been designed to minimize the number of moving parts. Those parts which do move, slide axially, eliminating complex linkages and gears. The secondary and tertiary swirl control vanes move within conical passages of the burner. As the swirlers moved toward the narrow end of the cone more air passes through the vanes increasing the amount of swirl. As the swirler is moved in the opposite direction, the air follows the path of least resistance and by-passes the vanes, resulting in less swirl. The amount of combustion air entering each burner is controlled by a sliding ring damper. Similarly, the split between secondary and tertiary air is controlled by a second ring damper. During normal operation, the tertiary air ring damper is actuated automatically while the ring damper controlling the SA/TA flow split is set during commissioning and is manually operable if necessary. These moving parts are located farthest away from the high heat flux environment in order to reduce warpage and binding. The parts of the burner that are subjected to a high heat flux are fabricated from a high strength heat resistant alloy.

The three CWS-capable burners were supplied with air-atomized slurry atomizers installed through the secondary air zone. The VEER-jet^R was developed by EER for a much smaller CWS burner designed for retrofit to fire-tube boilers. A non-plugging spray tip that would provide effective atomization with a low air-to-fuel ratio was required. The tip has been demonstrated to be very effective during the Seward operation. Not once did the tips require cleaning during the initial slurry testing, even though the quality of the slurry was sometimes poor.

Coal fineness and distribution to the burners also affects low-NO_x burner performance. Due to the low volatile content of the fuel normally fired at the Seward Station, EER specified the coal fineness to be no more than 0.5% >50 mesh and at least 75% <200 mesh. EER also recommends that, within multiple burner arrays, each burner should be supplied with equal amounts of fuel and air. EER included an adjustable orifice coal flow balancing damper, marketed as the PC FlowmastEER™, with each burner. This device, shown in Figure 3, allows for the adjustment of pulverized coal flow to within +/- 5% between burners. The FlowmastEER™ dampers were installed just above the pulverizer due to limited space along the coal piping.

Coal-Water Slurry System Description

The first CWS handling and feed system installed at Seward, shown Figure 4, consisted of an air operated diaphragm pump, a transfer line, a 4,000 gallon stationary tanker trailer, a slurry recirculation line fed from another diaphragm pump, and a high pressure slurry positive displacement Moyno pump. The system was designed and constructed to be a temporary installation to enable the CWS test program to be conducted. Due to the success of the CWS demonstration, Penelec later installed a permanent coal water slurry system that is similar to the first system, but now utilizes two 9,000 gallon tanks with agitators. This system is being used to co-fire 2,500,000 gallons of slurry over a six-month period which began in March of this year.

Instrumentation in the piping system consists of pressure gauges before and after the strainers, a pressure switch to protect the slurry pump from overpressure, local pressure gauges at the atomizers, low pressure switches on the atomizing air, a main fuel trip valve, automatic shutoff valves for the slurry and water purge system, a pressure transmitter for the air header, and an orifice plate and transmitter for the atomizing air flow. The slurry flow was determined from a mass flow meter that also provides totalization, density, and temperature.

A Bailey INFI-90 control system, installed simultaneously with the new burners, includes both Burner Management and Combustion Control system (CCS) functions. The INFI-90 is a distributed control system that provides integrated modulating control, sequential control and the capability of the operator to monitor and interface with the various instruments and devices associated with the Low NO_x Burner and coal-water slurry systems.

The control of most plant systems remains unchanged when switching to CWS co-firing. The CCS calculates a Btu-corrected slurry flow to compensate for changes in slurry density during firing. The approximate heating value is then subtracted from the Fuel Master demand signal controlling the pulverizer coal feeder, reducing the pulverized coal heat input by an equal amount. This control scheme prevents over-firing of the boiler when co-firing CWS and, at the same time, uses the slurry pump to maintain a constant heat input instead of modulating the pulverizer feeder to make up for changes in slurry heating value. This resulted in a more stable boiler operation due to a more constant fuel input. Primary air flow remains unchanged during CWS co-firing.

The slurry atomizing air system operates at a constant pressure and does not have a modulating control loop. Atomizing air flow and pressure are measured and transmitted to the CCS for display on the operator screen. Control of the various slurry valves and flush water valves are handled by the Burner Management System, to ensure proper sequential operation. Typical operation of the atomizers is with 0.15-0.25 lb atomizing air/1.0 lb of slurry.

Operating Results

The testing conducted at the Seward Station was focused on identifying any potential operating problems which could be attributed to the low-NOx burners, due to their CWS firing capability, or to the slurry fuel itself.

Low-NOx Burner Performance

Prior to any CWS co-firing, the low-NOx burners were optimized and guarantee performance tests were completed. The coal flow to each burner was balanced using the International Standards Organization (ISO) technique using a Rotorprobe™ sampling instrument. Initial measurements on Boiler #14 "A" pulverizer (lower burner elevation) indicated that one burner was receiving 17% more coal than the other two. After 3 iterations of FlowmastEER™ damper adjustments, the coal flow balance was measured to be within 3%. EER also rechecked the "A" pulverizer two months later and the coal flow balance was within 4%.

Prior to the burner retrofit, the average Boiler #14 baseline NO_x emissions were 0.90 lb/MMBtu with coal volatile and nitrogen contents of 25% and 1.3%, respectively. The FlameMastEER™ Burner emission guarantees were 0.5 lb/MMBtu NO_x, less than 100 ppm CO with 15-20% excess air in the furnace. At the same time, carbon in the fly ash was not to increase above baseline levels. Requirements of the guarantee were that the coal volatile content would not be less than 24% and that the nitrogen content would not exceed 1.3%. Results of the guarantee tests indicated that all guarantees were satisfied with a post-retrofit NO_x level of 0.47 lb/MMBtu.

CWS Co-firing Performance

The primary objective of the CWS test program was to evaluate the feasibility of co-firing low solids (40-50% by weight) CWS with pulverized coal in the Seward boiler. The results of the study were necessary to determine if a large scale CWS co-firing demonstration project is feasible. Penelec's ultimate objective is to utilize the coal fines stream from the cleaning plant in Units #1 and #2 (660 MWe wall-fired boilers) at the Homer City Station.

At the completion of the initial test program approximately 125,000 gallons of CWS were produced at the Homer City cleaning facility, trucked to Seward Station, and burned in Boiler #14. A total of approximately 150 hours of operation were accumulated on the system components.

The test program was conducted in such a way that direct comparisons between boiler operation, with and without slurry, were possible. A typical test burn involved three hours of data collection firing coal under steady operating conditions to determine the baseline performance. That was followed by three hours of testing with slurry co-firing. This procedure was designed to compensate for the normal variability which occurs in the operation of Boiler #14. Bunkered coal was used as the base fuel and consisted of the normal Seward coal supply.

The boiler control system measures and records all pertinent boiler operating data which is then down loaded onto a diskette at the end of each test. Continuous emission monitors were utilized as the primary data collection method for SO_2, NO_x, CO, and O_2. Flue gas samples were extracted just upstream of the air heater. At selected times, manual isokinetic samples were taken for particulate, SO_2, and NO_x using standard EPA methods.

Initial testing evaluated the CWS firing system with respect to materials handling and CWS flow characteristics. The boiler functioned in a normal manner without any boiler trips, mill trips, opacity excursions, or loss of flame. Although a very cautious test plan had been proposed, within one hour of beginning the first co-firing test, CWS was successfully burned in all three burners.

After the initial demonstration that the CWS fuel could be fired safely and without causing boiler upsets, parametric tests were conducted to evaluate the effect of slurry input, excess air level, atomizer operating conditions, and burner variables. With the majority of the testing conducted with less than 35% heat input from CWS, the burner and boiler performance is still dominated by the pulverized coal combustion.

One noticeable impact was the reduction of NO_x emissions during CWS co-firing. Figure 5 illustrates the emissions during boiler operation with and without slurry co-firing. The data points shown include a range of CWS heat input conditions from 10-35%. It was apparent that the NO_x reduction was not limited to reductions caused by the water in the slurry stream. At the conclusion of each test, the slurry tank was cleaned with water which was subsequently injected into the boiler. For an equivalent amount of water, the NO_x emission level was always lower during slurry co-firing. The reader should be aware that EER has demonstrated on pilot scale facilities that CWS can be a very effective reburn fuel. For equivalent levels of CWS firing demonstrated at Penelec, significantly higher reductions (>50%) of NOx emissions could be achieved by utilizing the CWS as a reburn fuel.[10]

The only operating problem encountered during the test program was with the slurry stability. Several slurry batches were delivered with solids contents less than 40%. It was discovered that solids were settling in the storage tank at the production facility. Agitation in the storage tank solved the problem. Mining labor negotiations resulted in coal supplies with significant variability during the test period also causing some problems in the production process. In spite of the low solids content and variability of the CWS quality, the slurry was burned at Seward without difficulty.

Throughout the testing, boiler performance and operations were unaffected by co-firing CWS. Steam temperature, flow, and pressure, opacity, attemperation flow, and other performance indicators were essentially constant. Boiler efficiency dropped by 1-2% as expected from the losses caused by the water being injected. Carbon loss in the flyash increased by a small amount when co-firing the slurry. It appears that this is due to the coarser size distribution of the CWS compared to the pulverized coal that it displaces. The

CWS fired was typically 3-5% >50 mesh material. This issue could be addressed in the production process as necessary.

Penelec determined from the results of this performance testing that, at least for relatively short-term tests, coal-water slurry co-firing was a viable alternative for reducing their cost of power generation. The next step in the demonstration program is to verify that same conclusion when CWS fuel is co-fired on a continuous basis for periods of at least one week. Penelec contracted EER to modify the remaining three low-NOx burners on Unit #14 for CWS co-firing capability. Penelec has also awarded a CWS supply contract to Washington Energy Processing, Coraopolis, PA to deliver up to 2,500,000 gallons of coal slurry to the Seward Station over a three to six-month period. The fuel is currently being fired in all burners on Unit #14. The results of the long-term testing will be reported at future symposia as they become available.

Economics

The main reason that coal is the primary fuel used for production of electricity worldwide is its low cost. Any process for removal of pollutants, whether coal cleaning or post combustion scrubbing, will add to the cost of producing electricity. This extra cost, which will be passed on to consumers, has to be justified based on the environmental regulation requirements. The pollution control steps or coal cleaning processes will usually increase the cost of coal. Assuming that coal would still be used, and the choice has to be made between the coal cleaning and post combustion scrubbing, the pre-combustion cleaning, in the long term, may provide a more economical alternative.

EER has performed several preliminary evaluations of applying CWS fuel to utility boilers. In addition, both Penelec and EER have worked closely with Mr. Ed Zawadzki, a fuel technology consultant, and the EPRI UCIG to evaluate the potential for recovering and utilizing coal fines. Based on this experience, the authors believe that there is considerable potential for coal-water slurry fuels to be produced in sufficient quantities for utility application at less than $1.00/MMBtu.

The fixed cost to retrofit the slurry co-firing equipment on the very small Penelec unit was approximately $25/kw. EER has estimated the cost for CWS co-firing capability on larger units (500-700MWe) to be in the range of $5/kw.

Variable costs are dominated by the fuel cost differential between the CWS fuel and the coal it displaces. The costs to operate the atomizing air compressors and the reduction in boiler efficiency are the only significant operating costs to consider. No additional operations personnel are required. Normal maintenance of the slurry handling equipment is required.

EER also completed another evaluation in the past year looking at the benefit of combining advanced cleaning methods with the coal slurry technology. Based on a cursory evaluation of the individual characteristics of the coal cleaning techniques and the slurry co-firing/reburning methods to be investigated, the reduction of pollutants; NO_x, SO_x, HAPs, and PM_{10} were estimated. Table 1 summarizes the technical and economic results of this evaluation. With the use of advanced physical cleaning of coal to make a slurry for use as a co-firing fuel, there would be a net savings rather than a cost for NO_x reduction. Also, the chemical cleaning of coal may look relatively expensive, but under the current Federal government tax codes, IRS §29, a tax credit of $0.97/MM Btu may be obtained, which when applied also yields a payback for the chemical cleaning of coal, not a cost.

TABLE 1. ESTIMATED POLLUTANT REDUCTIONS AND COST

Case:	Percent of Baseline				Cost, $/ton		
Pollutant:	NO_x	SO_x	HAPs	PM_{10}	NO_x	SO_x	Total
Baseline	100	100	100	100	-		-
Physically Clean CWS Co-Firing	70	100	100	100	(420)	-	(420)
Physically Clean CWS Reburning	40	100	100	100	52	-	52
Adv Phys Clean CWS Reburning	40	88	88	88	52	326	378
Adv Phys + Chem Clean CWS Reburn	30	81	80	85	44	531	575
Adv Phys + Chem Clean CWS Reburn (w/IRS §29 tax credit)	30	81	80	85	44	(217)	(173)

Note: The analysis is based on applying the technologies to a 700 MWe wall-fired boiler that yields a baseline emission rate of 0.8 lb NO_x/MM Btu of coal fired. The baseline coal for the slurry is assumed to contain 10% ash, 4% sulfur with a 50-50 mix of pyritic and organic sulfur, and a heat input of coal water slurry of 25%. HAPs reduction is shown as mercury; other heavy metals such as selenium, cobalt, arsenic etc. will also be reduced. The physical and chemical cleaning applies only to the CWS used for reburning. In the physical plus chemical cleaning case, all of the added cost was placed as a cost of SO_2 reduction.

Conclusions

There has been a considerable amount of work completed over the past several years to evaluate the potential for utilizing coal fines as a low cost boiler fuel. Significant quantities of fines are produced and disposed of every day by operating coal cleaning plants. In addition, large quantities of fines are stored in impoundments throughout the country. Coal fines can be recovered economically from both sources, converted to a coal-water slurry fuel, and can be made available at a very attractive price for supplementing existing fuel supplies.

Pennsylvania Electric Company has proceeded with a demonstration of the utilization of a relatively low solids content (50%) coal-water slurry fuel on a small utility boiler located at their Seward Station. Low-NOx burners, some with CWS firing capability, were retrofit and a slurry fuel handling system installed to evaluate the technology. The 32 MWe boiler has fired several hundred thousand gallons of CWS fuel without any detrimental boiler impacts whatsoever. In addition, during CWS co-firing tests, NO_x emissions have been reduced approximately 20%. Penelec has contracted for an additional 2,500,000 gallons of coal-water slurry and are continuing their evaluation with long-term slurry firing currently in progress.

EER is prepared to offer complete coal-water slurry firing systems, with performance guarantees, for retrofit to most utility and large industrial boilers.

Acknowledgements

The authors would like to acknowledge the Penelec Seward Station operations and maintenance staff and Mr. John Kline, Station Engineer, whose cooperation and support made the introduction of this new fuel a relatively easy task.

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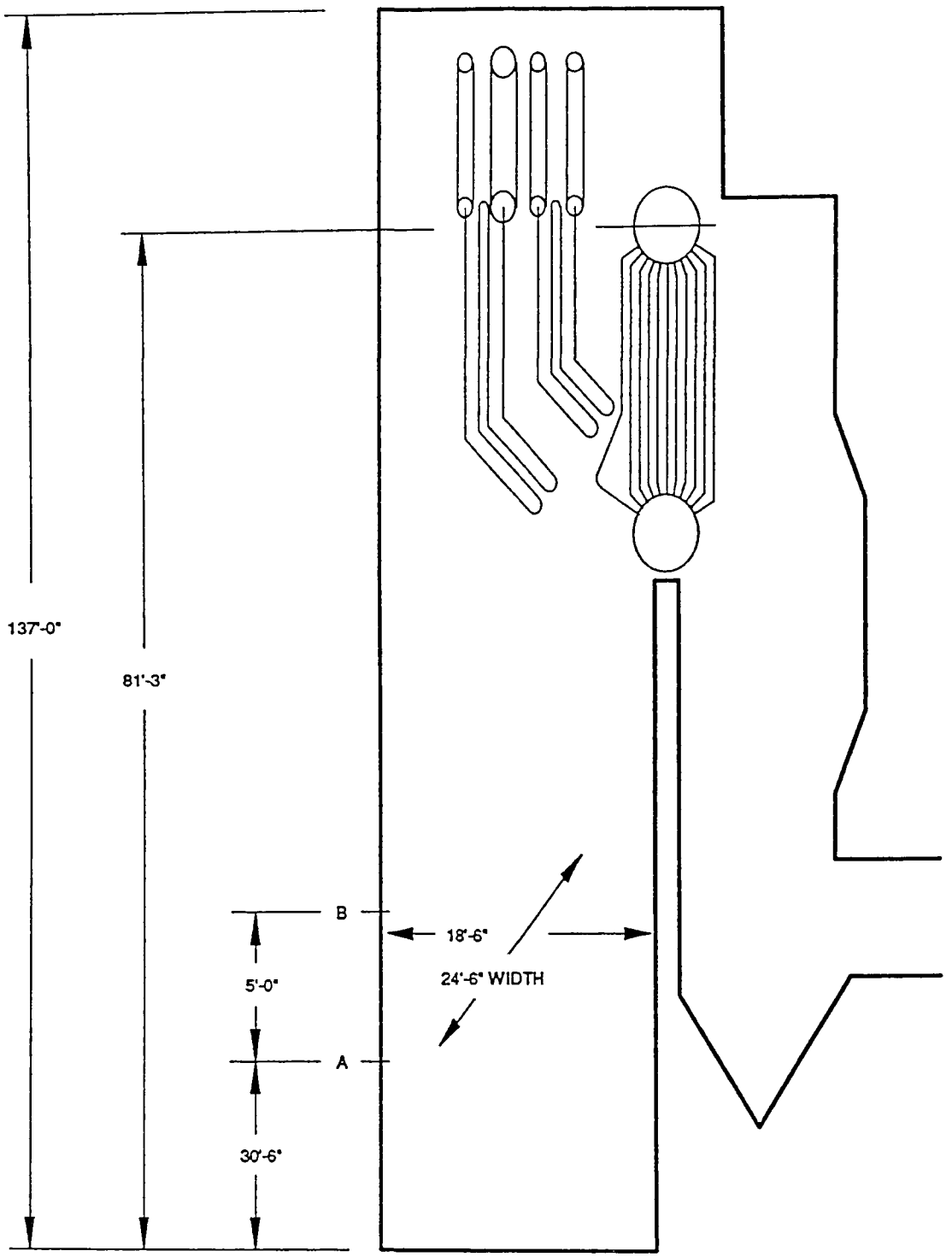


Figure 1. Seward Boiler 14 dimensions

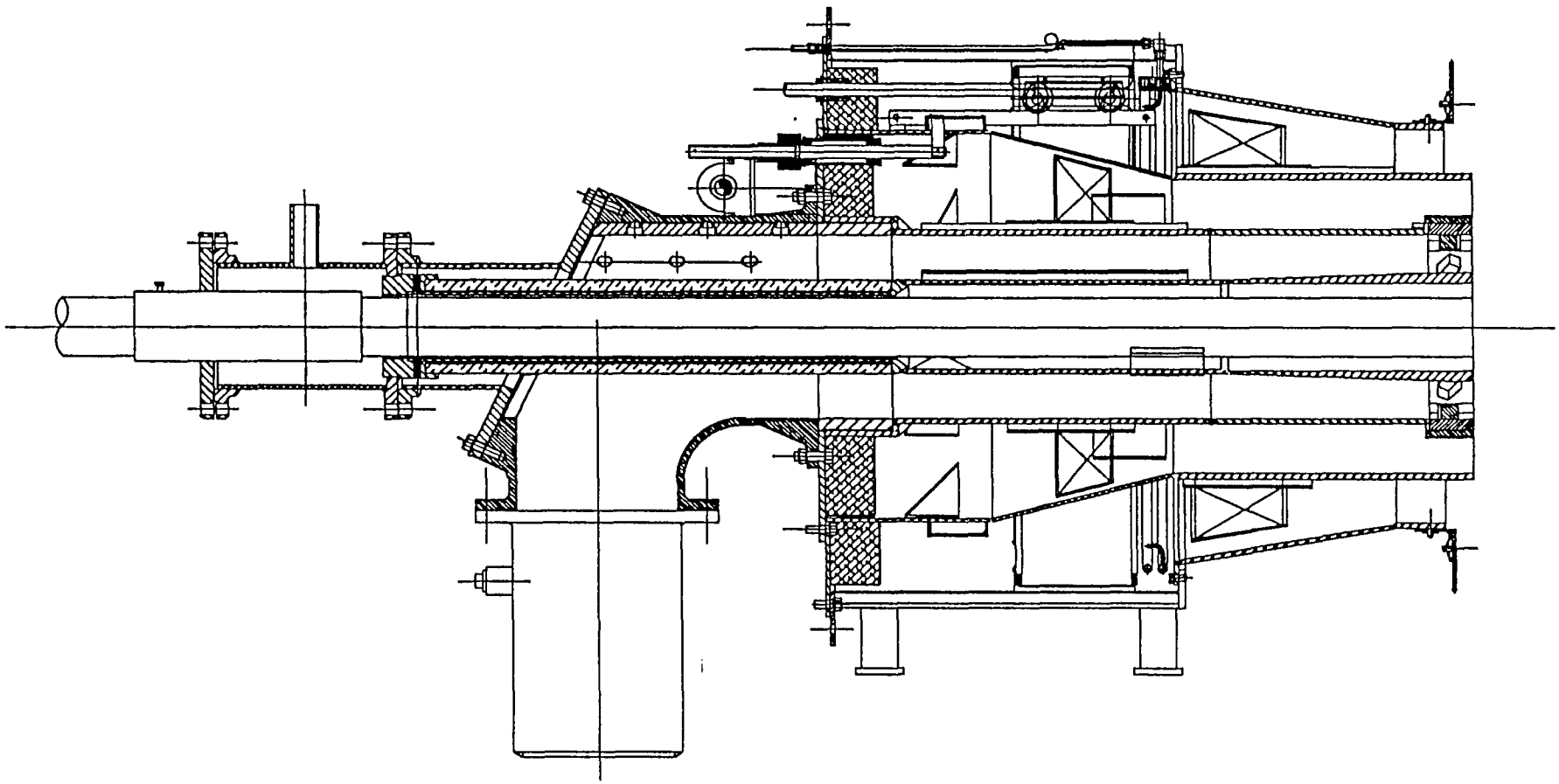
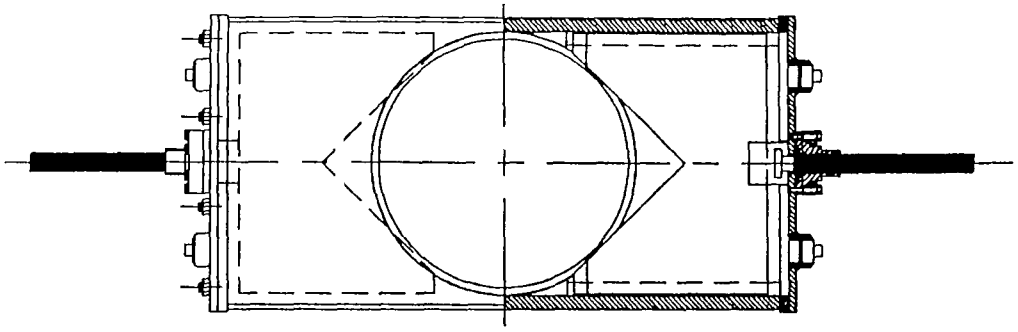
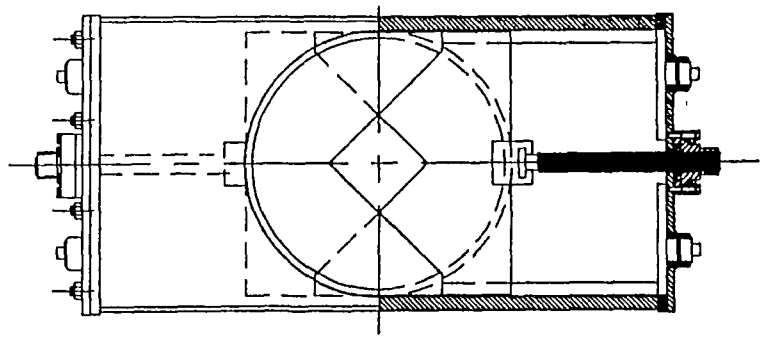


Figure 2. FlamemastEER™ Burner



FULLY OPEN
100% FLOW AREA-NO RESTRICTION



FULLY CLOSED
10% FLOW AREA-MAXIMUM CONTROL

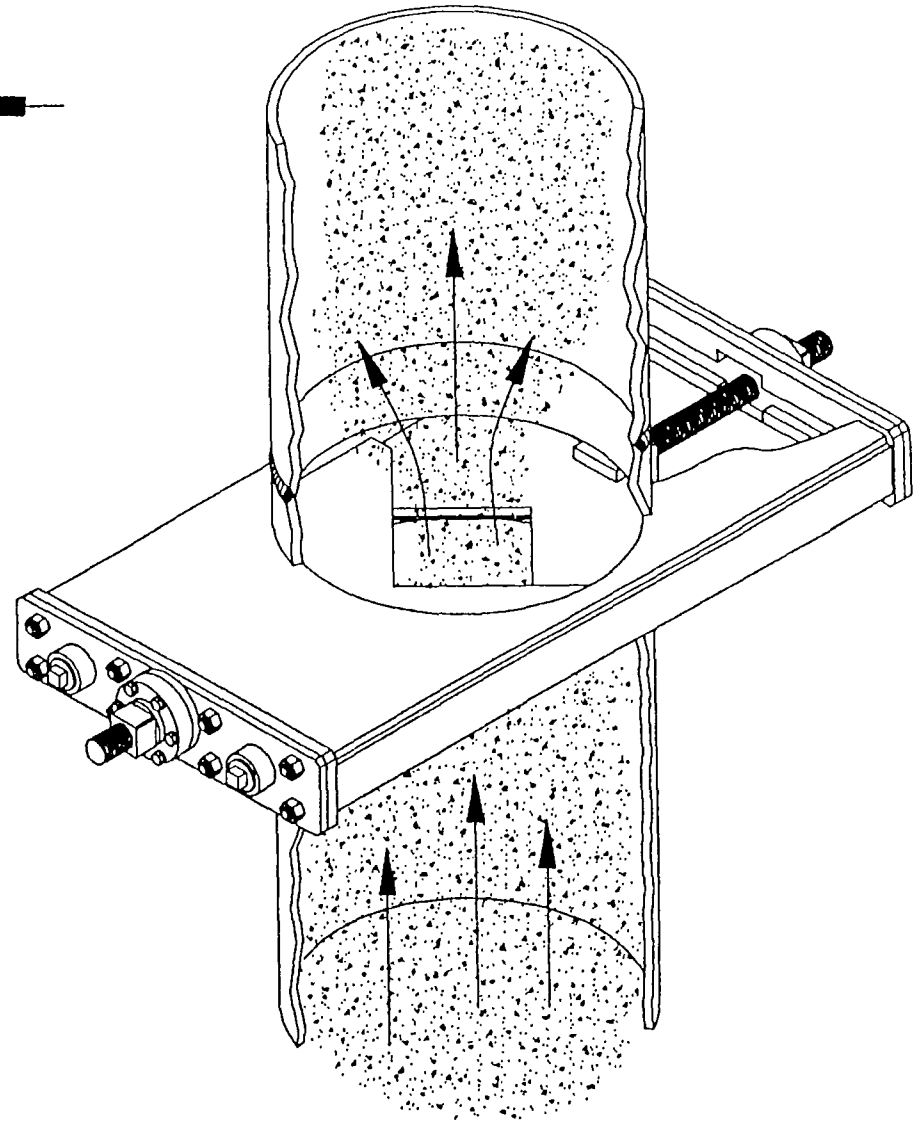


Figure 3. PC FlowmastEER™.

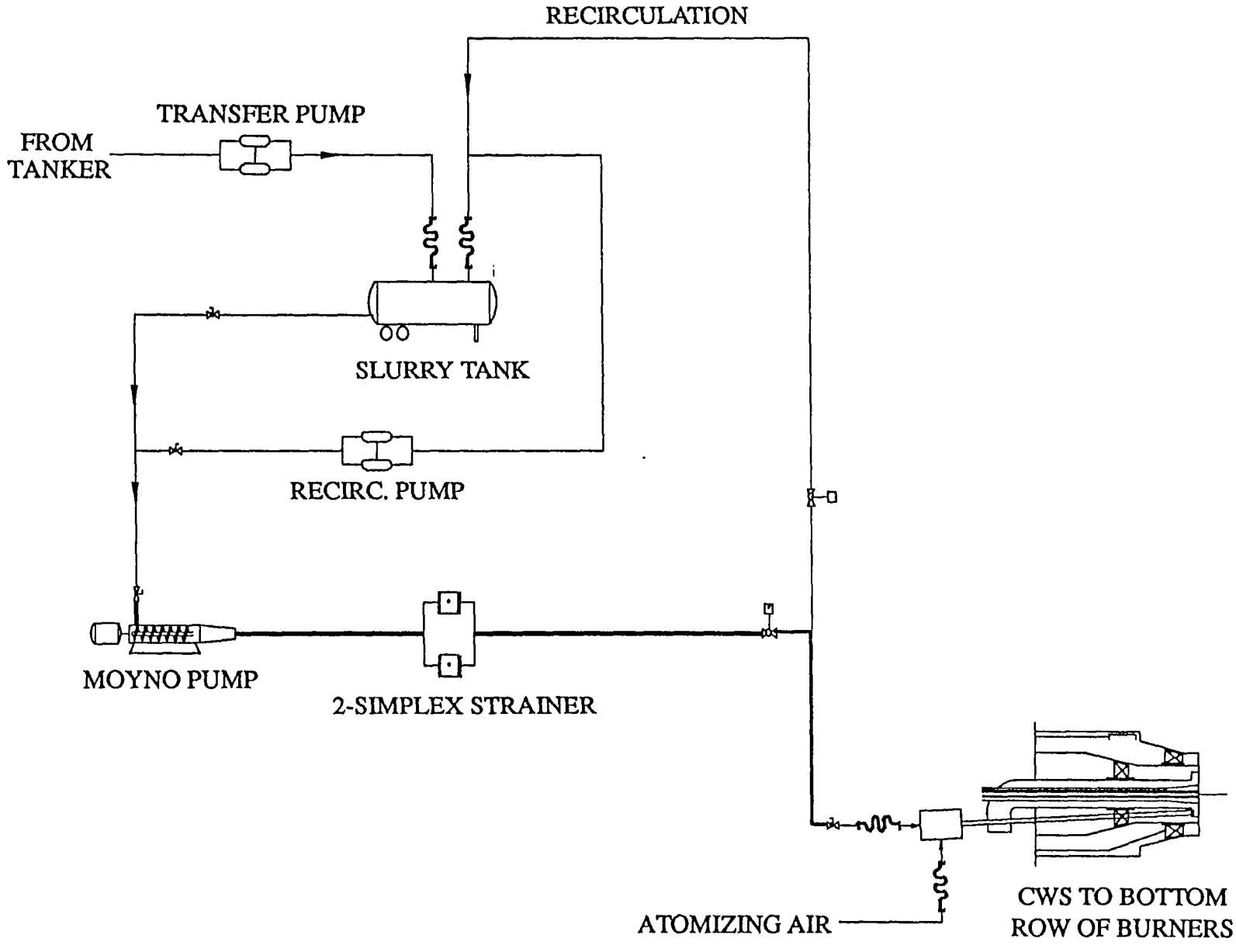


Figure 4. Seward Coal Water Slurry System

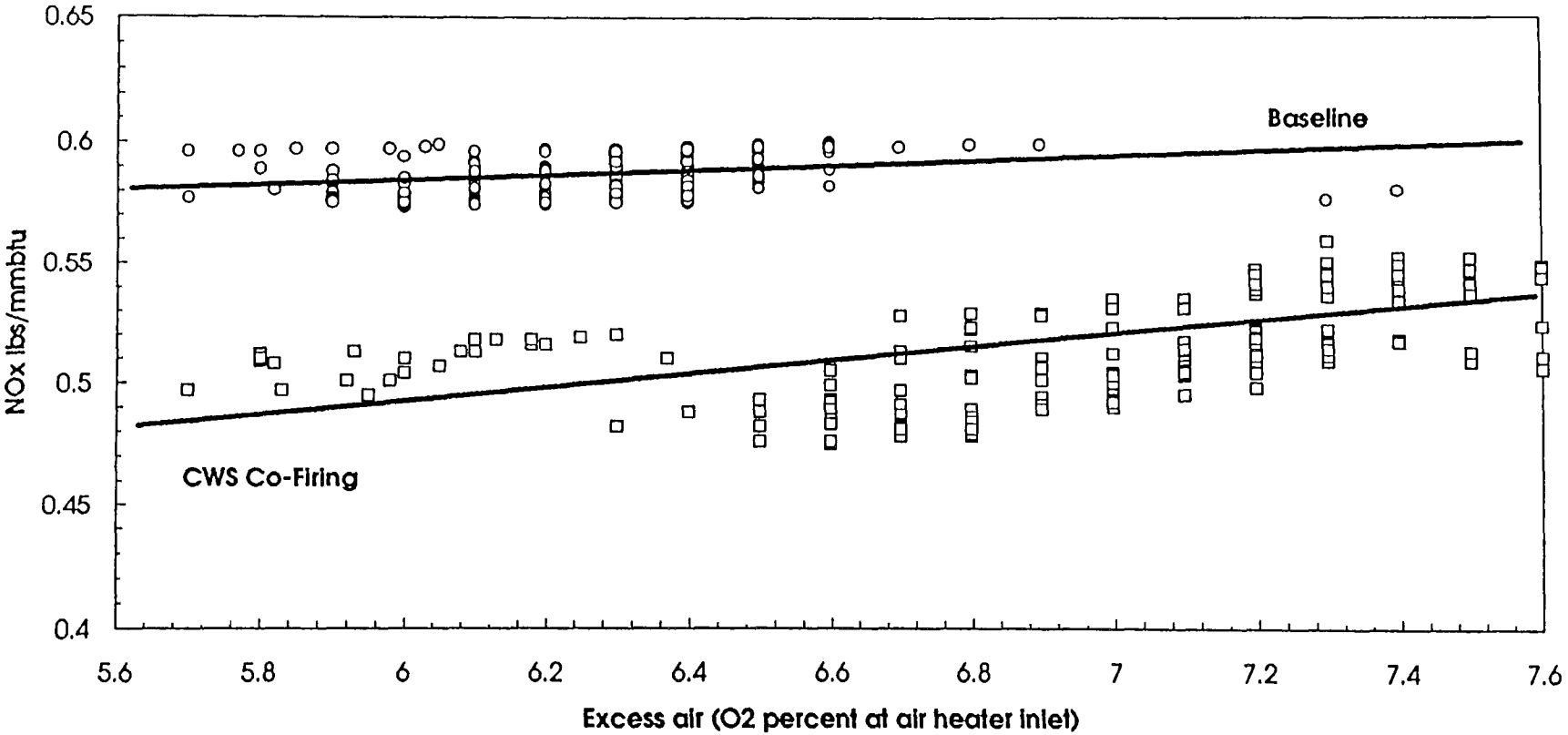


Figure 5. NOx Emissions Impact

IN-FURNACE, RETROFIT ULTRA-LOW NO_x CONTROL TECHNOLOGY FOR TANGENTIAL, COAL-FIRED BOILERS: THE ABB C-E SERVICES TFS 2000™R SYSTEM

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United Illuminating

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ABB C-E Services, Inc.

Abstract

United Illuminating and ABB C-E Services, Inc. report the first commercial retrofit installation and performance results from a TFS2000™R firing system. Pre-retrofit and post-retrofit field trials were conducted to evaluate the impact of the retrofit design on the boiler emissions and thermal performance. During testing, the retrofitted 390-MW_e utility boiler demonstrated NOx emissions on the order of 0.25 lb/10^6 Btu, while firing Eastern bituminous coal over the entire load range, without increase in unburned carbon (UBC). A potential minimum NOx emission level of 0.16 lb/10^6 Btu was achieved in parametric testing. The effects of the retrofit on boiler emissions, thermal performance and operating experience are reported.

Introduction

United Illuminating (UI) provides electricity to south-central Connecticut. In 1984, the electricity produced in the UI system came from an energy mix that was 94% fuel oil and 6% nuclear. To diversify its fuel base, in that year UI reconverted the Bridgeport Harbor Station Unit 3 (Figure 1) for coal firing. By 1985, the contribution of oil to UI's energy mix was reduced to 53%; nuclear was 9%, and coal had provided 37%. Continuing with its strategy of utilizing diverse fuels, UI shifted its energy mix to 1% natural gas, 5% hydro, 8% trash-to-energy, 17% oil, 35 % nuclear, and 34% coal by 1992.[1]

The city of Bridgeport is located in a "Severe" ozone nonattainment area under the 1990 Clean Air Act Amendments (CAAA) Title I. Bridgeport Harbor Station Unit 3 (BHS Unit 3) is a Phase II unit under CAAA Title IV. The State of Connecticut's Reasonably Achievable Control Technology (RACT) NOx limitation is 0.38 lb/10^6 Btu for tangential coal-fired boilers. With UI's fuel strategy in place, the utility decided to retrofit BHS Unit 3, its only coal-burning unit, with an aggressive low NOx firing system.

ABB C-E Services invited UI to participate in a research and development project in which BHS Unit 3 would serve as the first commercial field demonstration of TFS

2000™R technology. Similar technology had previously demonstrated ultra-low NOx emissions at the laboratory scale.[2]

Unit Description

BHS Unit 3 is a Combustion Engineering, Inc., Controlled Circulation® steam generator with radiant reheat cycle and a pressurized furnace (Figure 2). It was designed in 1965 and commissioned in 1968. The steam generator is rated at 2,700,000 lb/hr primary steam flow at maximum continuous rating (MCR), with a corresponding reheat flow of 2,387,000 lb/hr. The MCR design superheat and reheat outlet steam temperatures are 1005 F. Operating pressure at the superheater outlet is 2629 psig.

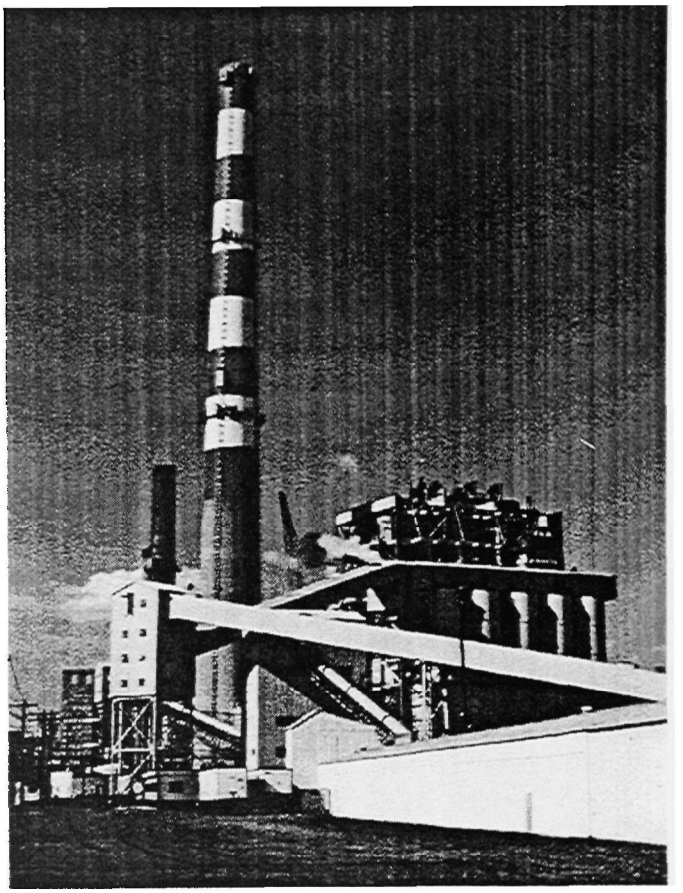

Figure 1: United Illuminating's Bridgeport Harbor Station

Nominally rated at 390 MWe, the unit was equipped with a Tilting Tangential Firing System for firing pulverized coal from five elevations and oil from four elevations. During the reconversion to coal firing in 1984, close-coupled overfire air was added. BHS Unit 3 operates with Eastern U.S. bituminous coals from sources in Kentucky. The coal composition is relatively uniform, with a low sulfur content and low slagging/fouling potential. Table 1 shows a typical coal analysis for BHS Unit 3.

BHS Unit 3 is typically operated on automatic load dispatch, generating steam at MCR on weekdays and at control load or lower on nights and weekends. Pre-retrofit NOx emissions under normal operating conditions were in the range of 0.55-0.60 lb NOx/10^6 Btu. The unit had no history of significant slagging or fouling, and no history of pressure part failures related to the coal properties.

Moisture	5.4%
Volatile Matter	30.1%
Fixed Carbon	57.7%
Ash	6.8%
Nitrogen	1.4%
Sulfur	0.7%
FC/VM	1.92
HHV (Btu/lb)	13,400
Hardgrove Index	45

Table 1: Typical Coal Analysis

TFS 2000™R SYSTEM DESIGN

The TFS 2000™R System at BHS Unit 3 is an integrated retrofit design based on the successful laboratory development of Combustion Engineering, Inc.'s (ABB C-E) TFS

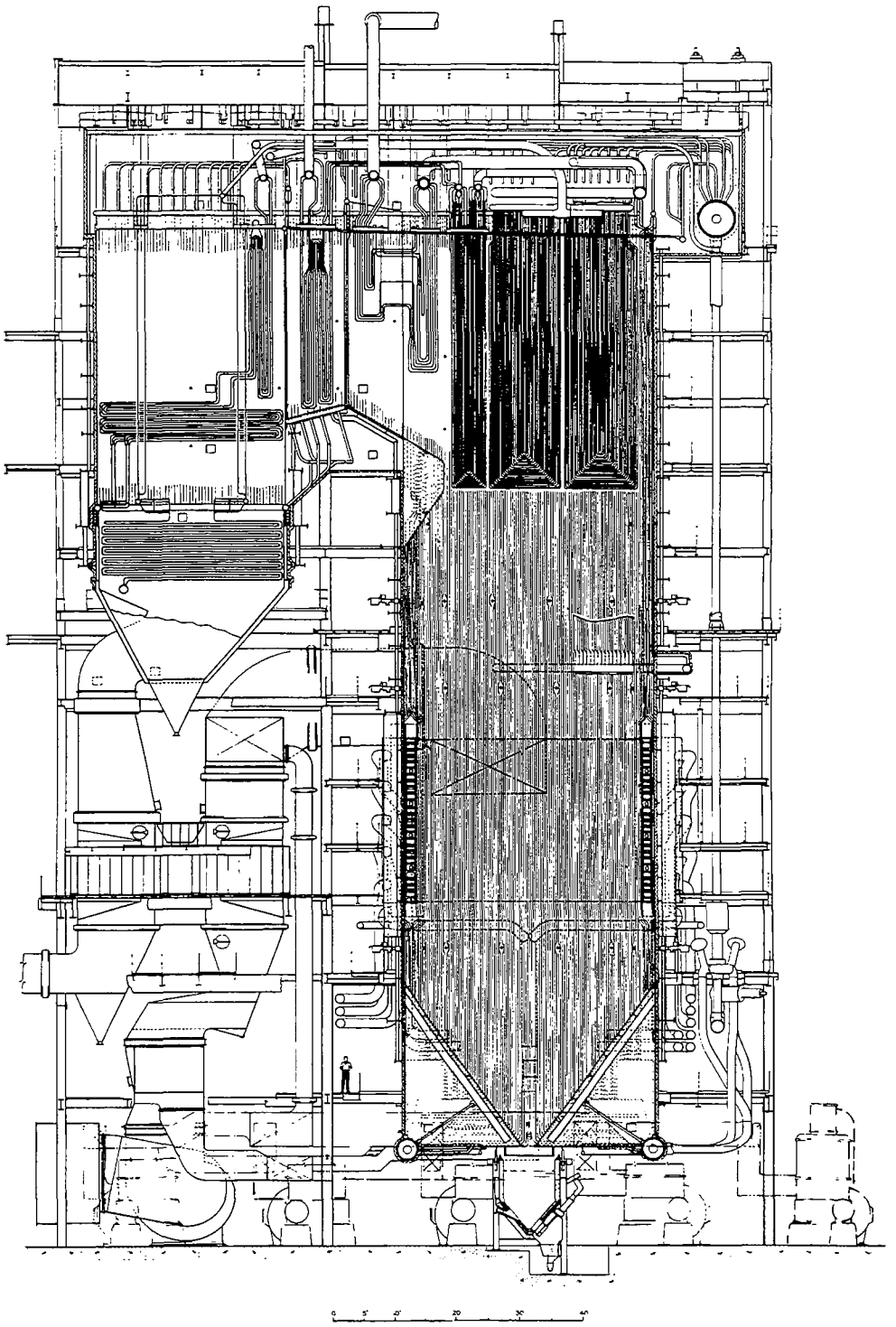

Figure 2: Bridgeport Harbor Station Unit 3, Pre-Retrofit Side Elevation

2000™ system for new boilers.[2] The challenge is to provide the most aggressive control of NOx emissions possible within the constraints of a fixed furnace geometry, without introducing any radical or negative departures from either design or operating

practices. Previous research and development efforts suggested that the laboratory results for absolute NOx emissions, and trends for carbon monoxide and unburned carbon, were consistent with a utility boiler.[3] Therefore, the next step in the commercialization of the TFS 2000™R technology was a field demonstration on a large utility boiler.

The basic design philosophy of the TFS 2000™R firing system is based on the integration of four major principles:

1. Firing zone stoichiometry control
2. Pulverized coal fineness control
3. Initial combustion process control
4. Concentric firing

Laboratory testing has indicated that there is an optimum main firing zone stoichiometry for minimizing NOx emissions.[2] However, achieving this level of stoichiometry can result in high levels of CO and UBC. The TFS 2000™R system (Figure 3) controls the process of NOx formation and destruction in distinct regions of the furnace by "staging" the introduction of air through flame attachment coal nozzle tips and multiple levels of separated overfire air (SOFA) and close-coupled overfire air (CCOFA). The TFS 2000™R system thereby optimizes the entire stoichiometry history of the coal particles, to minimize NOx emissions.

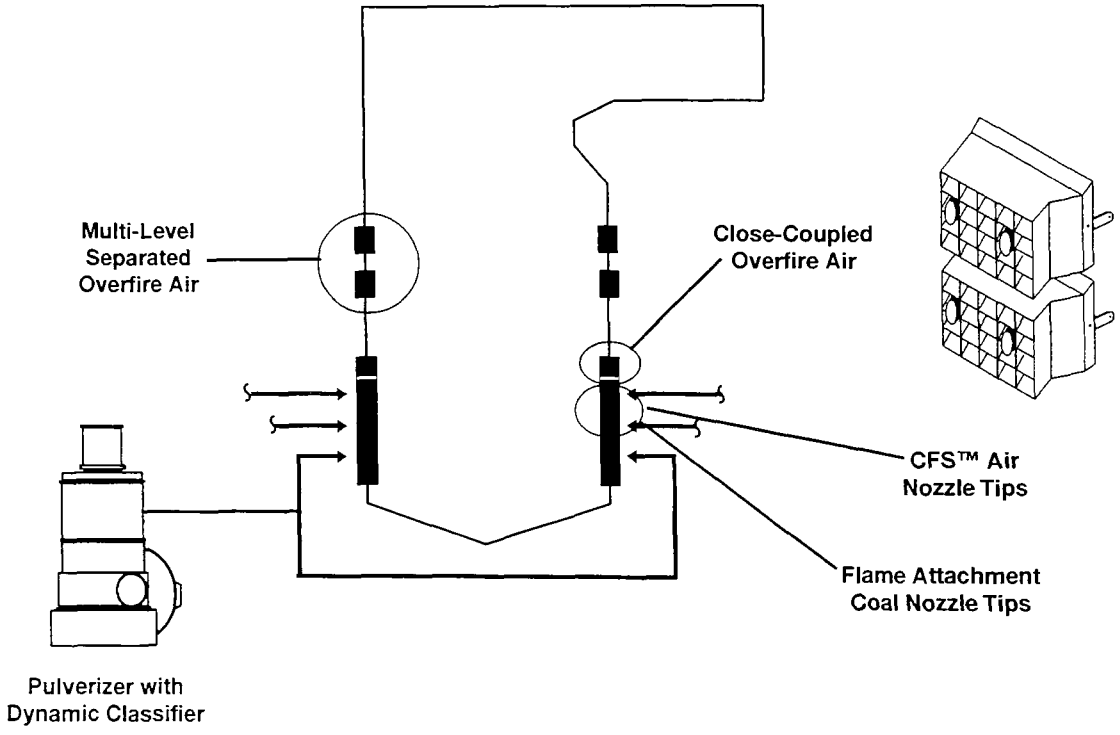

Figure 3: Schematic Diagram of a TFS 2000R Firing System

Pulverized coal fineness is controlled by use of a Dynamic™ classifier. The rotating classifier vanes more effectively prevent larger coal particles from exiting the pulverizer, and this helps decrease the UBC levels in the flyash. Finer coal particles can also enhance fuel-bound nitrogen conversion and its subsequent reduction to molecular nitrogen under staged firing conditions by allowing rapid ignition near the coal nozzle tip.

Flame attachment coal nozzle tips are incorporated in the TFS 2000™R system design to provide early fuel devolatilization within an oxygen-deficient zone. With conventional firing systems, coal is devolatilized in an oxygen-rich environment, and the fuel nitrogen released can readily react with the available oxygen to form nitrogen oxide compounds. With the flame attachment coal nozzle tip, rapid coal devolatilization is accomplished by establishing a flame front near the exit of the tip. The coal nozzle tip design is based on existing flame characteristics, coal constituents, and fuel line transport conditions. Besides the NOx emissions control benefits, establishing coal ignition early in the combustion process improves flame stability and minimizes increases in unburned coal levels.

ABB's patented CFS™ concentric firing system air nozzle tips direct some of the secondary air in the main firing zone away from the fuel streams. Offsetting the air decreases the local firing zone stoichiometry during the initial combustion stages.

Concentric firing also creates an oxidizing environment near the furnace waterwalls in and above the main firing zone. This reduces ash deposition quantity and tenacity. Increased oxygen levels along the waterwalls also decreases the potential for corrosion, especially with coals having high concentrations of sulfur, iron, or alkali metals.

The specific equipment components selected to achieve these elements of combustion will vary for different retrofit installations, depending on the design and maintenance condition of the installed equipment, and on the constructability constraints at the site.

TFS 2000™R System Implementation

The retrofit equipment described below for the field demonstration of TFS 2000™R technology at BHS Unit 3 was installed in the Fall of 1993. The installation coincided with a scheduled maintenance outage for the turbine-generator. The outage duration was 8.5 weeks.

Windboxes

Because the existing main windboxes at BHS Unit 3 were in a deteriorated condition and the planned outage duration was short, the main windboxes were completely replaced with new, pre-assembled units. Each new main windbox (Figure 4) contains one bottom air compartment, four elevations of air/oil compartments with CFS™ air nozzle tips above and below the oil gun tips, two elevations of CCOFA compartments, and five elevations of coal compartments with flame attachment coal nozzle tips. New

tilt mechanisms were provided at the compartments, re-using existing tilt drives. Secondary air flow to the windbox air registers is controlled by means of louver dampers equipped with self-lubricating damper bearing assemblies.

With ABB's flame attachment coal nozzle tips, the ignition point of the coal occurs closer to the nozzle tip than it does for conventional coal nozzle tips. The rapid fuel ignition produces a stable volatile matter flame and minimizes NOx production in the fuel-rich stream.

The CFS™ air nozzle tips supplied at BHS Unit 3 are equipped with manually-adjustable horizontal yaw mechanisms. The yaw adjustment is set so that a portion of the secondary air is directed away from the fuel streams toward an imaginary circle that is concentric with the main firing circle. The yaw angle is set during commissioning and is not changed during normal operation of the boiler.

The CCOFA elevation air registers direct a portion of the secondary air into the furnace at the top of the main windboxes. Each CCOFA compartment is equipped with ABB's patented horizontal yaw adjustment mechanism. The manual yaw adjustment enables each CCOFA air jet to be independently directed for effective mixing.

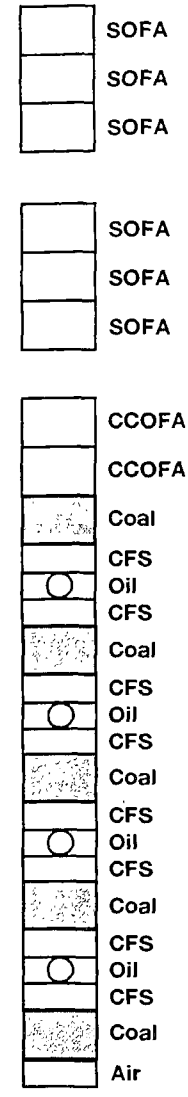

Figure 4: Schematic Diagram of TFS 2000R Windboxes at BHS Unit 3

Two new SOFA registers were added above each of the new main windboxes. Each SOFA register contains three air compartments with adjustable horizontal yaw and vertical tilt mechanisms (Figure 5). During commissioning, the yaw angle is set to minimize carbon monoxide and UBC emissions. This is a manual adjustment that is not intended to be varied during operation.

To measure the SOFA air flow, an annular venturi (Figure 6) was installed in each SOFA air supply duct. ABB's patented annular venturi design requires only about two-thirds the length of a standard venturi and measures air flow with an accuracy of ±5 percent. It has a signal-to-noise ratio of approximately 10. Annular venturi are not required components for a TFS 2000™R system retrofit.

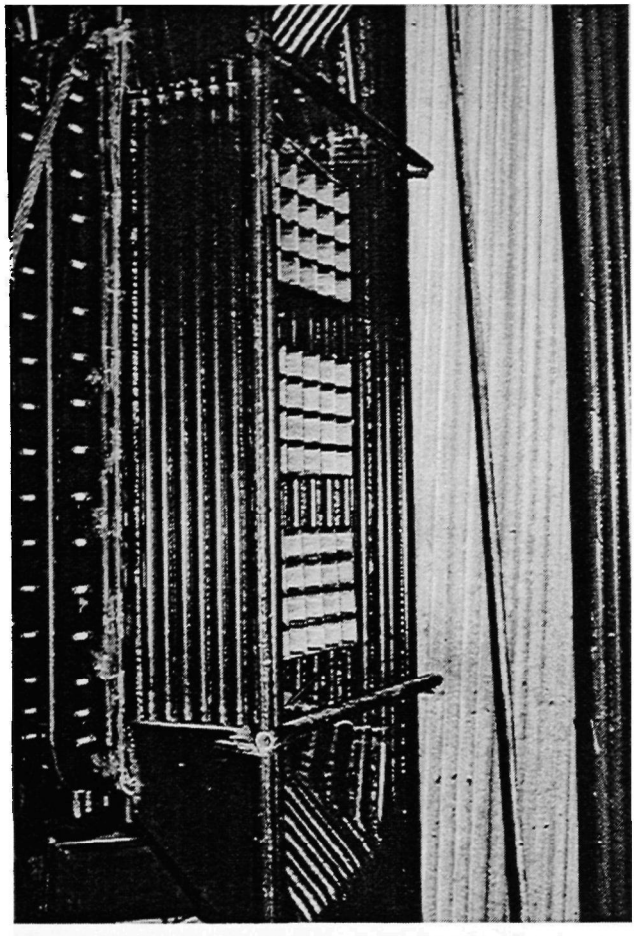

Figure 5: New SOFA Register During Installation

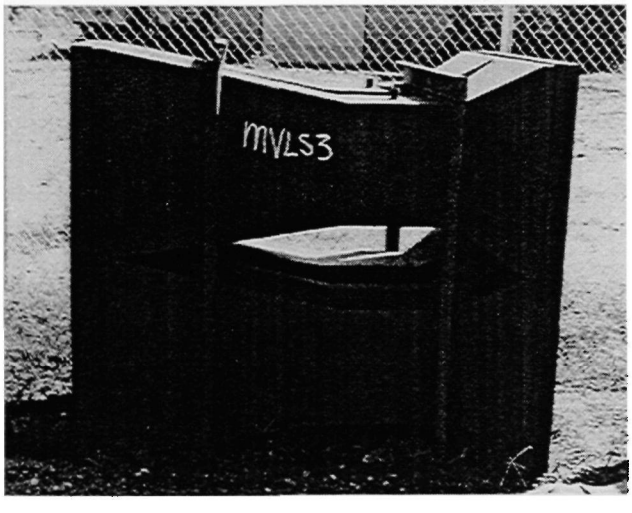

Figure 6: Annular Venturi for SOFA Ductwork in Laydown Area

Pulverizer Modifications

Pulverizer modifications to implement TFS 2000™R technology are also site-specific, and depend greatly on the condition of the existing pulverizers, as well as the coal to be fired after the retrofit. BHS Unit 3's five pulverizers were well-maintained and in good operating condition prior to the retrofit. The pulverizers were upgraded to permit operation at higher fineness levels without coal flow de-rating. The existing "spider" fan wheels were replaced by new high efficiency fans (HEF) utilizing the existing exhauster casings. In addition, the existing 600-Hp pulverizer motors were replaced with new 700-Hp motors. Figure 7 shows one of the new HEF wheels.

In each pulverizer, a new Dynamic™ classifier replaced the existing static classifier. The Dynamic™ classifier has a vaned rotor that is supported by two bearings. It is driven by a 40-Hp motor, and the speed of rotation is controlled through an ac variable-speed controller. Figure 8 is a photograph of one of the pulverizers during the installation of the Dynamic™ classifier. The Dynamic™ classifier effectively eliminates large coal particles (+50-mesh or +70-mesh) and minimizes the fraction of +100-mesh coal particles. It allows extensive operational flexibility, and can be used to compensate for the effects of pulverizer wear, load changes, and changes in coal type or grindability.

Additional Work

Pressure part replacements requiring four main windbox tube panels and four SOFA tube panels accompanied the new windboxes and SOFA registers. Additional pressure part modifications were made at BHS Unit 3 to eliminate interferences with the SOFA register installation.

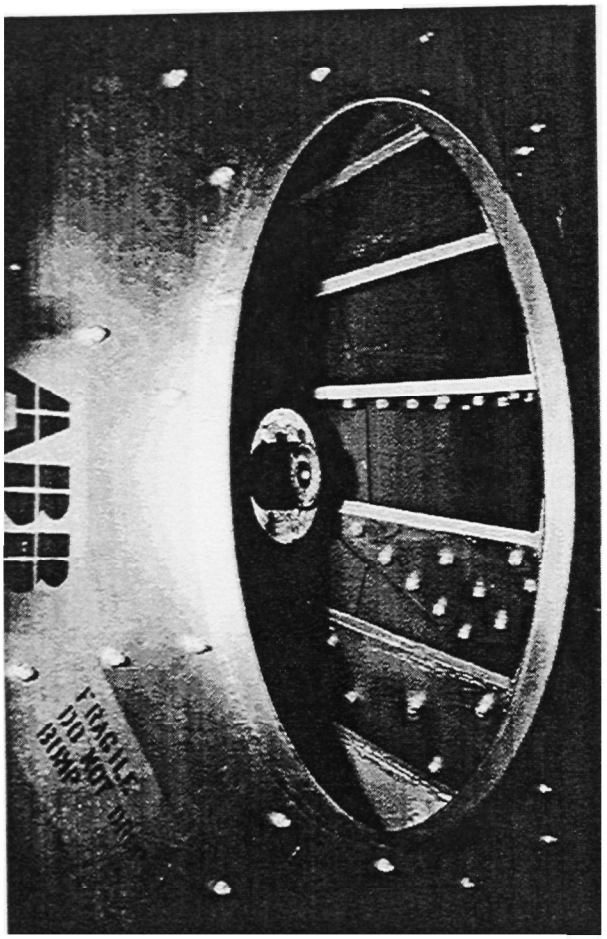

Figure 7: New HEF Wheel in the Existing Exhauster Casing

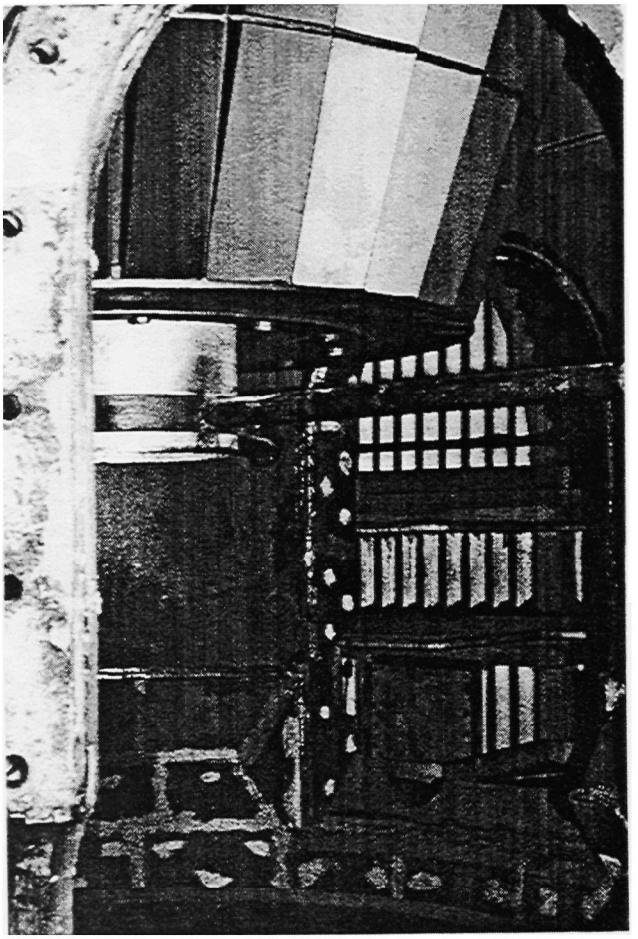

Figure 8: New Dynamic™ Classifier During Installation

As part of the research and development project, 39 waterwall chordal thermocouples and 135 convective section thermocouples were installed to provide accurate and convenient measurements of the boiler's thermal performance under load. In addition, six waterwall test panels were installed to investigate industry concerns regarding long-term waterwall tube wastage under substoichiometric firing conditions. These panels were fabricated of new waterwall tubing and were subjected to ultrasonic thickness measurement prior to installation. Tubing thickness will be regularly monitored during future maintenance outages. Figure 9 shows the approximate locations of this test equipment.

Control system inputs/outputs and logic were added for operation of SOFA dampers and Dynamic™ classifiers, and to expand the operational flexibility of all windbox dampers. In addition, UI elected to perform additional back pass modifications, to upgrade the DCS control system and to add continuous stack emissions monitors and stack elevator during the outage. These modifications were not required for the new firing system.

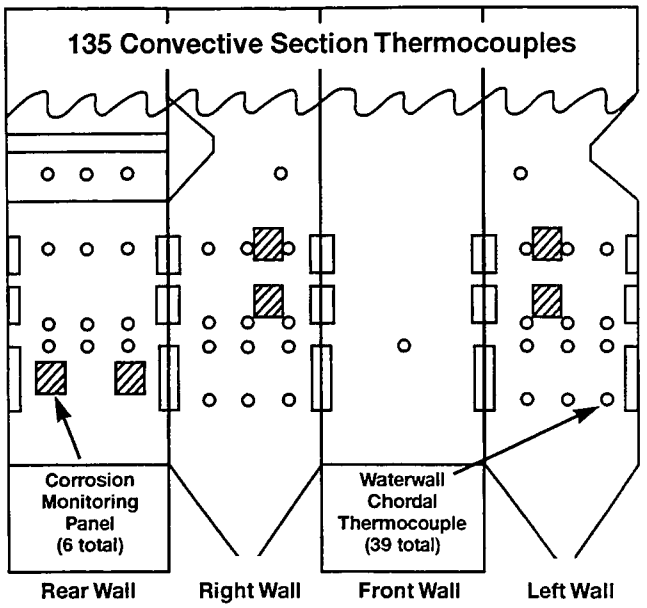

Figure 9: Locations of Test Thermocouples and Test Panels

TFS 2000™R System Performance Evaluation

Pre-retrofit and post-retrofit field trials were conducted to evaluate the impact of the new design on the boiler emissions and thermal performance. The focus of the field trials was to quantify the impact of the new firing system over the full operating range of the boiler.

Boiler Emissions Performance

The boiler emissions performance was characterized through a series of parametric tests during which certain operational parameters were varied in a systematic fashion for several scenarios of boiler load, staged firing, and secondary air biasing.

NOx Emissions

Except where noted, all NOx measurements in this paper were determined via EPA Method 7E, using a chemiluminescent NOx analyzer, and are reported in units of lb NOx/10^6 Btu. Figure 10 shows the relationship of the measured NOx emissions from BHS Unit 3 to the calculated stoichiometry at the top coal elevation for both the pre-retrofit and post-retrofit configurations of the boiler. All measurements were taken at MCR. The characteristic decrease in NOx emissions with decreasing stoichiometry is evident. Pre-retrofit NOx testing with the use of CCOFA showed NOx levels in the range of 0.46 - 0.58 lb NOx/10^6 Btu.

Sixty-six post-retrofit tests were conducted while varying the coal fineness and the degree of staging and mixing, along with a number of operating variables such as excess air. Post-retrofit NOx emissions as low as 0.20 lb NOx/10^6 Btu were achieved with no increase in the UBC in the flyash.

The two data points labeled "Potential Minimum NOx" (0.18 and 0.16 lb NOx/10^6 Btu) represent short-term (approximately 3 hours) test results. These results were achieved with carbon monoxide emissions less than 200 ppm and only a two-percentage point increase in UBC emissions over the pre-retrofit level. *It is significant that the potential minimum NOx results were achieved at a higher stoichiometry than many of the higher post-retrofit testing results, demonstrating that stoichiometry is not the only variable affecting NOx emissions.*

The post-retrofit test NOx emissions as a function of boiler load are shown in Figure 11. The secondary air dampers and tilts were controlled to operate the boiler with NOx

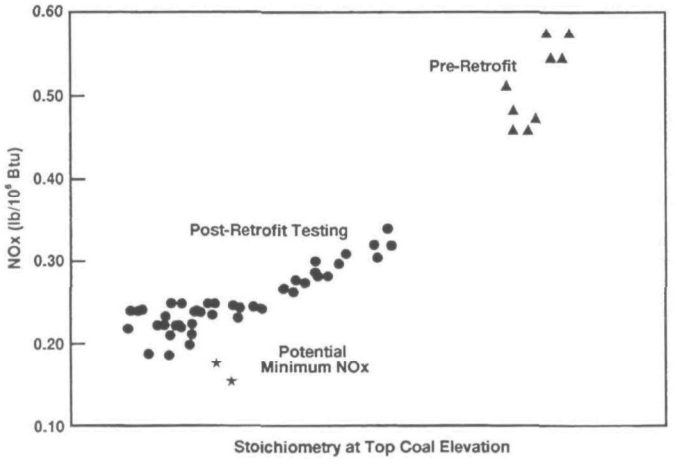

Figure 10: NOx Emissions vs. Stoichiometry at MCR

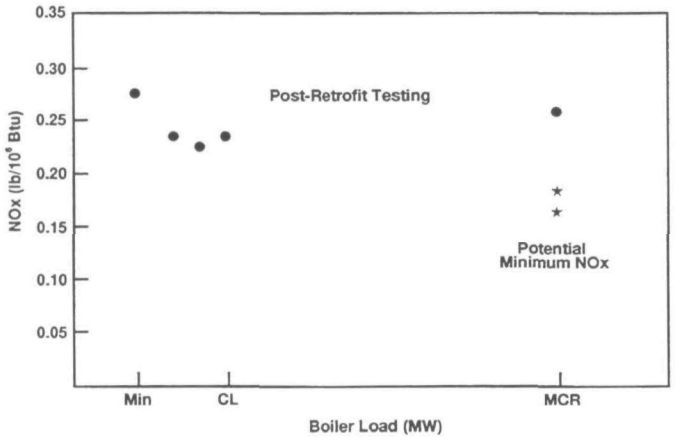

Figure 11: NOx Emissions vs. Boiler Load

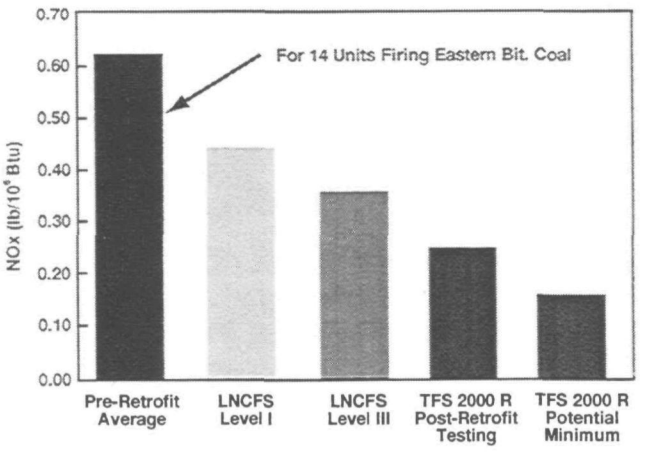

Figure 12: Comparison of ABB Retrofit Results for NOx Emissions

emissions on the order of 0.25 lb NOx/10^6 Btu from MCR through control load (CL), to minimum load, with no increase in UBC in the flyash. Although it is typically expected that NOx levels will increase dramatically at low boiler loads because of the required increase in excess air, at BHS Unit 3, the post-retrofit NOx emission at minimum load can be controlled to less than 0.30 lb/10^6 Btu.

Figure 12 compares the BHS Unit 3 post-retrofit testing for NOx emissions to other low NOx retrofit results for similar coals in tangentially-fired boilers. The pre-retrofit average NOx emissions of 0.62 lb/10^6 Btu for 14 other units firing Eastern bituminous coals is shown in the first (left) bar. ABB C-E Services' LNCFS™ firing systems were applied in these units.[4] As shown in Figure 12, LNCFS™ system field results reached a lower limit for NOx emissions at an average of 0.36 lb/10^6 Btu. The BHS Unit 3 field demonstration test results for NOx emissions are significantly lower.

Limited testing was performed to evaluate firing system performance using No. 6 fuel oil, currently used as an emergency backup fuel to coal at BHS Unit 3. These brief tests, under non-optimized conditions, indicated that NOx levels on the order of 0.12 lb/10^6 Btu to 0.15 lb/10^6 Btu can be obtained from control load through 365 MW, with opacity in the range of 2-7 percent.

Carbon Monoxide Emissions

All carbon monoxide (CO) measurements reported in this paper are given in units of parts per million

(ppm) of gas and are corrected to 3% oxygen in the flue gas. The test protocols used are in accordance with EPA Method 10. Pre-retrofit CO emissions were less than 50 ppm. During the post-retrofit testing the SOFA yaw angles were varied to demonstrate the variation of CO emissions with NOx. During the tests documented in Figure 10, at full load, CO levels of 44 ppm were obtained at NOx emissions of 0.34 lb/10^6 Btu; CO emissions of 22 ppm occurred with NOx emissions of 0.24 lb/10^6 Btu; and CO emissions of 178 ppm were found with NOx emissions of 0.16 lb/10^6 Btu.

Opacity

Opacity measurements were taken with the plant instrumentation. At BHS Unit 3, the regulated opacity limit is 20%. The pre-retrofit opacity averaged less than 10%. During the post-retrofit testing, the opacity remained less than 10% for most tests, and below the regulated limit under all test conditions. Isokinetic sampling of the flue gas entering the unit's electrostatic precipitator (ESP) confirmed that there was no significant change in the flyash (dust) loading entering the ESP. No significant change in the mass ratio of flyash-to-bottom ash was observed.

ESP collection efficiency prior to TFS 2000™R retrofit was 99.3 percent. During the outage for the TFS 2000™R retrofit, routine maintenance was performed on the ESP. Measured post-retrofit ESP efficiency is on the order of 99.2-99.5 percent.

Boiler Operational Performance

During post-retrofit testing on the BHS Unit 3 boiler, multiple aspects of boiler operation were investigated to ensure that there were no adverse impacts on boiler operation related to the changes in the firing system.

Ash and Slag Deposition Patterns

A long-term change in the ash and slag deposition during operation was noted. Post-retrofit ash deposition has increased in the superheater sections closest to the furnace outlet, the superheater division panels and superheater platen assemblies (Figure 2). These ash deposits are friable and easily removed. No other significant changes in ash accumulation have been observed in the convective sections of the boiler. Slagging has decreased on about one-third of the furnace wall, in the areas near the CFS™ air elevations. Although the ash and slag deposition patterns have changed, they are controllable with the existing sootblowers and wall blowers on the boiler.

The boiler had no history of waterwall corrosion before the retrofit. After approximately 14 months of post-retrofit operation, no evidence of accelerated waterwall wastage has been observed.

Coal Fineness

Calibration runs for the Dynamic™ classifier with the "B" pulverizer established the relationships among coal feed rate, fineness, and classifier rotation speed. Generally, a

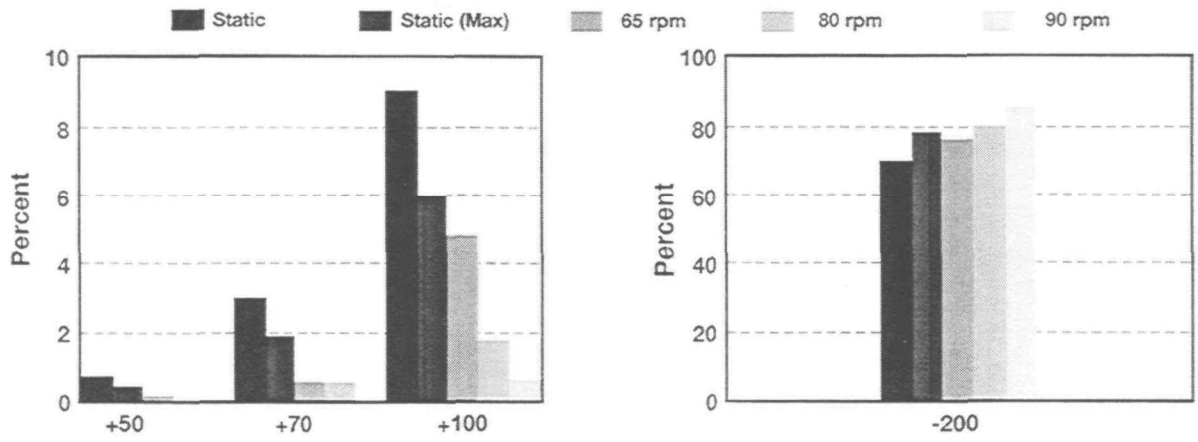

Figure 13: Comparison of Static and Dynamic Classifier Fineness Results

higher classifier rpm produces greater fineness, and rpm can be decreased as coal feed rates are decreased. At all coal feed rates, the coal fineness achievable with the Dynamic™ classifier is finer than with the static classifier, particularly in terms of decreasing or eliminating the largest +50 and +70-mesh particles. Coal particles in these size ranges have significant impact on UBC. Figure 13 compares the performance of the static classifier and the Dynamic™ classifier at BHS Unit 3 with five pulverizers, each in service at 55,000 lb coal/h.

Pulverizer performance has met expectations, with the exception of a "rumble" condition that occurred during testing at high classifier rotation speeds. High fineness "rumble" can occur with either dynamic or static classifiers on a high-fineness setting. High fineness "rumble" is an instability, leading to vibrations, that is caused by an increase in recirculation of fine particles. At BHS Unit 3, the Dynamic™ classifier rotational speed is currently limited to avoid high fineness "rumble". A study is in progress at the ABB Power Plant Laboratories Pulverizer Development Facility in Windsor, Conn., to develop a methodology for predicting/preventing the onset of high fineness "rumble".[5]

Furnace Oxygen Imbalance

The oxygen concentration in the flue gas was measured at the economizer outlet in accordance with EPA Method 3A. Post-retrofit left/right oxygen imbalance is less than or equal to the pre-retrofit performance.

Boiler Thermal Performance

Boiler Efficiency

The installation of the TFS 2000™R firing system did not affect the boiler thermal efficiency (ASME Performance Test Code 4.1). Pre-retrofit and post-retrofit boiler

efficiencies were calculated at MCR and at control load, and the efficiency remained at 91.4 - 91.7 percent, regardless of the NOx emissions level.

Steam Temperature/Flow Control

All post-retrofit operation of the boiler confirms that the superheater and reheater design outlet steam temperatures can be maintained at loads from MCR through control load. In addition, the superheater and reheater design pressures and mass flow rates are maintained at all loads from MCR through control load.

Steam temperature control is accomplished through the use of the adjustable tilts and the interstage desuperheaters. The windbox tilts continue to operate within their normal range.

At both the maximum and potential minimum NOx emissions levels, the post-retrofit reheater desuperheater spray water flows were about the same as the pre-retrofit levels. Thus, the implementation of TFS 2000™R technology does not adversely impact the unit's heat rate.

Element Steam Temperature Imbalance

Eight pre-retrofit tests and two post-retrofit tests were analyzed. Two of the pre-retrofit tests were for normal operation, three were for operation with the top secondary air dampers closed, and three were for operation with three tilt positions. One post-retrofit test was conducted with maximum SOFA and acceptable boiler operation, and the other was at the minimum NOx emission. The (low temperature) superheater rear pendant outlet steam temperatures, (high temperature) superheater finishing pendant outlet temperatures, and the high temperature reheater outlet temperatures were measured and analyzed. As compared to the initial operation of the unit, firing oil, in 1968, there was no significant difference in the element steam temperature profiles caused by the TFS 2000™R system.

Maximum Local Heat Absorption Rates

The peak waterwall heat absorption rates calculated from readings with the chordal thermocouples installed in the furnace walls were well below the design values and confirm that the post-retrofit departure from nucleate boiling (DNB) margin for the boiler remains within ABB C-E design standards.

Vertical Heat Absorption Profile

The vertical heat absorption profile, as measured through the chordal waterwall thermocouples is similar under all post-retrofit operating conditions. There is a slight shift in the furnace vertical heat absorption profile towards the upper furnace under potential minimum NOx conditions. This shift did not adversely affect boiler waterwall circulation.

UBC as a Function of NOx Emissions

Significant increases in UBC levels in the flyash have been documented for boilers retrofitted with earlier low NOx firing systems.[4] Pre-retrofit UBC levels at BHS Unit 3 were in the range of 5.8 - 8.0 percent carbon. For a tangentially-fired boiler with an Eastern bituminous coal, this range is about average.

Except where noted, the flyash samples for both the pre-retrofit and post-retrofit UBC results were obtained in accordance with EPA Method 17. Carbon content was determined directly, not by loss of ignition (LOI).

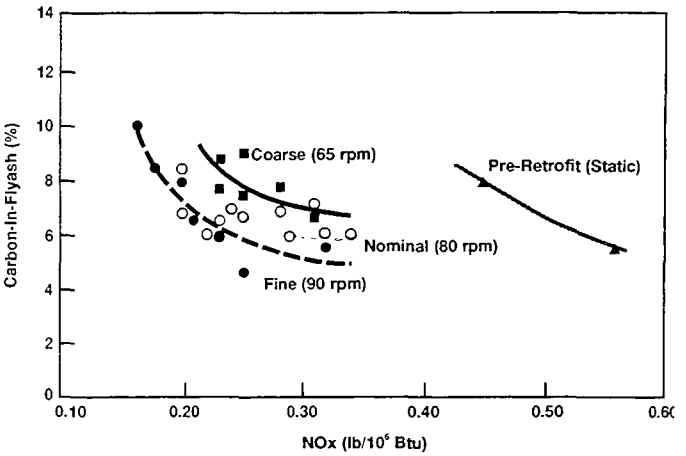

Figure 14: UBC in Flyash vs. NOx Emissions at MCR

UBC levels for post-retrofit operation at BHS Unit 3 with three different fineness levels are given in Figure 14. For this comparison, boiler load was held constant at MCR. The trend of increasing UBC with decreasing NOx emissions is evident for the three post-retrofit data sets. The trends also illustrate that UBC control is dependent upon the particle size of the coal. NOx emissions as low as 0.20 lb/10^6 Btu were obtained with no increase above pre-retrofit levels of UBC in the flyash.

Commercial Operating Experience

The unit has been operating commercially, post-retrofit, firing coal for over 14 months. The unit operates under load dispatch at MCR on weekdays from about 8:00 am to 11:00 pm. At night and on weekends, the unit load is decreased to as low as 140 MW. Operators report no significant operational problems, and no indication of accelerated waterwall wastage or corrosion has been observed.

Daily Average Value for Megawatts, NOx , and UBC

Figure 15 is a plot of the megawatt load and NOx emissions during the period from January 3, 1995 to February 27, 1995. The values given for NOx emissions in lb/10^6 Btu are as measured on the Continuous Emissions Monitor (CEM) at the stack. The CEM is RATA certified. NOx emissions during this period of normal operation averaged 0.23 lb/10^6 Btu.

During the pre-retrofit period from January 2, 1993 to September 25, 1993, a survey of weekly ash truck samples (combining flyash and bottom ash) showed an average of 7.6 percent carbon in the ash. For the post-retrofit period from January 2, 1994 to February

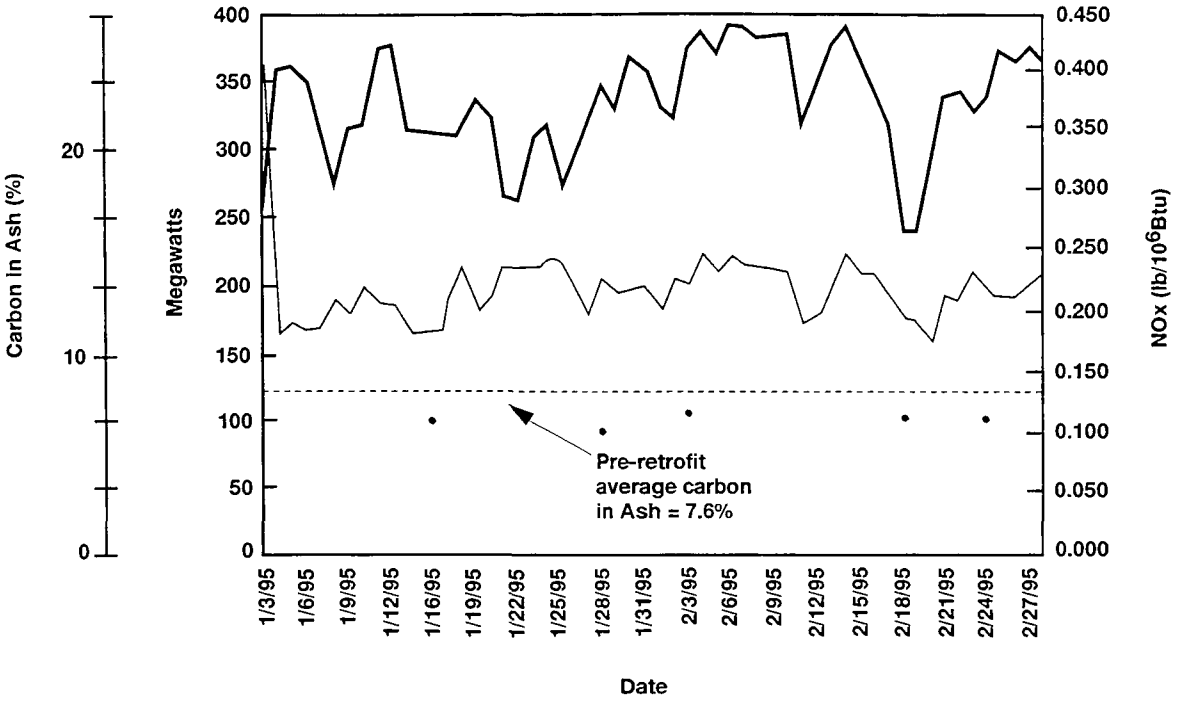

Figure 15: BHS Unit 3 NOx Emissions vs Megawatt Load and Unburned Carbon

25, 1995, the weekly survey of the ash truck samples averaged 5.8 percent. Figure 15 also shows the weekly ash truck sample carbon content.

Unit Availability and Heat Rate

Table 2 shows the five leading causes per year for lost generation at BHS Unit 3 from 1990 through 1994. Comparisons of the equivalent forced outage rates, which take into account any capacity reduction as well as shutdown, show that the unit's overall post-retrofit availability is consistent with its historical performance. Post-retrofit availability losses caused by economizer leaks and coal pulverizers are higher than the historical levels. The economizer leaks were definitely not related to the TFS 2000™R installation, and problems have been corrected. The increased availability loss caused by the coal pulverizers is attributable to problems with the initial design of the seals for the Dynamic™ classifiers, which have since been corrected. No ongoing loss of availability attributable to the Dynamic™ classifiers is expected.

Figure 16 shows the year-to-date effective availability factor (EAF), calculated both including overhauls and excluding overhauls, for the years 1990 through 1994, and for January through February of 1995. For 1993, the year-to-date figure is taken from the pre-retrofit period of January through September. For 1994, the year-to-date figure is entirely post-retrofit experience. The data in Figure 16 confirm that the post-retrofit availability is consistent with the pre-retrofit experience.

	1994	1993	1992	1991	1990
Annual Overhaul	4.4	16.5	4.4	6.2	4.8
Superheater Leak					1.4
Reheater Tube Leak	1.9			1.8	1.2
Economizer Tube Leak	2.6				
Precipitator Repair					1.2
Waterwall Leak	1.5	3.8	5.5	1.6	0.7
Coal Mill	1.5				
Main Boiler Feed Pump (MBFP)			1.5		
Casing Leak				1.5	
Condenser Cleaning		0.9			
Boiler Circulating Pumps				1.5	
Lube Oil System			1.9		
Boiler Circ. PP			1.0		
Boiler Water Chemistry		1.6			
Turbine Vibration		1.6			
Total of 5 Causes	11.9%	24.4%	14.3%	12.6%	9.3%
Total Lost for Year	17.8%	27.8%	17.9%	15.6%	11.4%
Equivalent Forced Outage Rate (EFOR)	13.4%	11.3%	13.5%	9.4%	6.8%

Table 2: BHS Unit 3 – Five Leading Causes Per Year of Lost Generation (Retrofit Work Done October – November, 1993)

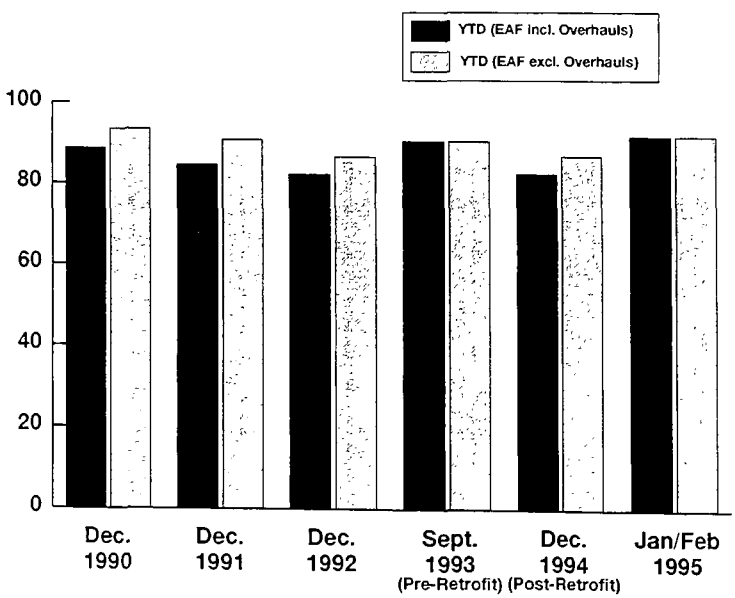

Figure 16: BHS Unit 3 Equivalent Availability Factor Including and Excluding Overhauls (Post-retrofit start-up was December, 1993)

It is also apparent that the heat rate for BHS Unit 3 has not deteriorated as a result of the TFS 2000™R retrofit. Figure 17 is a graph of the year-to-date heat rate in Btu/kW-h for the unit from December of 1987 through December of 1994. The heat rate post-retrofit (1994) is well within the historical trend of heat rate for the unit.

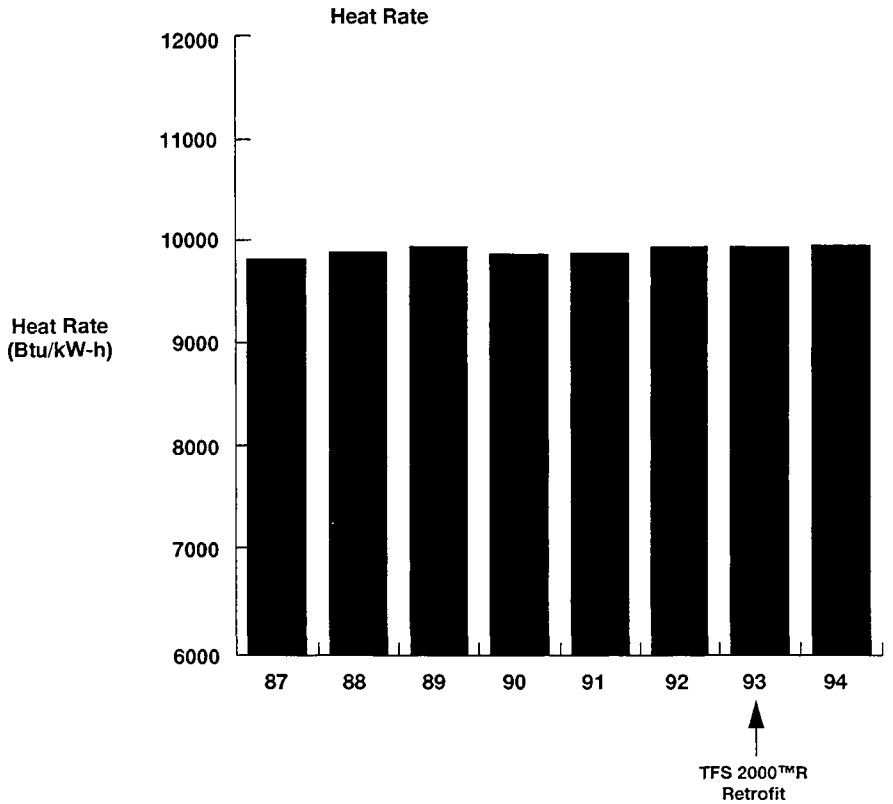

Figure 17: BHS Unit 3 Heat Rate Performance History

Equipment Operational Experience and Inspection

During a scheduled maintenance outage inspection conducted in the fall of 1994, the new firing system condition was assessed as excellent. Tilts were stroked with no evidence of binding or erratic operation. Eight of twenty coal nozzle tips had slight warpage in splitter plates that was not affecting performance, and repair was not required. One oil nozzle tip bluff body diffuser was slightly cracked, and repair was not required. SOFA-related components were in "as new" condition. The ignitor system was also in "as new" condition.

Detailed visual inspections were made of the economizer, steam drum, lower waterwall drums, waterwalls, bottom ash hopper, superheater steam cooled roof and wall headers, backpass steam cooled outlet headers, low temperature superheater, superheater division wall panels and pendant platen assemblies, and reheater

assemblies, and all were found to be in a condition typical of pre-retrofit experience. Service recommendations in the outage report were consistent with pre-retrofit experience on this unit.

A visual inspection of the Dynamic™ classifer during the 1994 fall outage showed no visible wear on the rotor assembly. An exhauster inspection indicated no required maintenance after nine months (approximately 8500 hours) of service.

Conclusions

United Illuminating and ABB C-E Services consider the retrofit of Bridgeport Harbor Station's Unit 3 to be a commercially and technically successful full-scale demonstration of TFS 2000™R technology. The boiler thermal performance and efficiency are unchanged from the pre-retrofit conditions. Although the slagging/fouling patterns have changed slightly from pre-retrofit, the existing sootblowers and wall blowers are capable of controlling them.

During testing, the boiler consistently demonstrated NOx emissions on the order of 0.25 lb/10^6 Btu over the entire load range, with no increase in unburned carbon in the flyash. The lowest NOx emissions measured for this boiler during post-retrofit parametric testing is 0.16 lb/10^6 Btu. The potential for long-term operation of the boiler at this level has not been thoroughly investigated. In over 14 months of commercial operation, operation of the boiler with the TFS 2000™R technology has caused no significant adverse impact on boiler operation or availability.

Acknowledgements

The authors acknowledge and appreciate the efforts and expertise of all the individuals from United Illuminating and ABB who were involved in the success of this field demonstration project. The contributions of D. Gillespie, P. Olson, A. Cortiglio, R. Diotalevi, D. Cahill, J. Buffa, P. Acimovic, T. Dorazio, W. Derech, V. Piras, and R. Collette are especially noted. Thanks also to R. Lewis, G. Strich, D. Choi, P. Stanwicks, T. Kelly, C. Boyle, B. Walsh, D. Gelbar, D. Barlow, J. Dallen, and C. Doherty for their valuable contributions.

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ESTIMATION OF NO_x EMISSIONS FROM PULVERIZED COAL-FIRED UTILITY BOILERS

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Abstract

The formation of nitrogen oxides (NO_x) during pulverized-coal combustion in utility boilers is governed by many factors, including the boiler's design characteristics and operating conditions, and coal properties. Presently, no simple, reliable method is publicly available to estimate NO_x emissions from any coal-fired boiler. A neural network back-propagation algorithm was previously developed using a small data set of boiler design characteristics and operating conditions, and coal properties for tangentially fired boilers. This initial effort yielded sufficient confidence in the use of neural network data analysis techniques to expand the data base to other boiler firing modes. A new neural network-based algorithm has been developed for all major pulverized coal-firing modes (wall, opposed-wall, cell, and tangential) that accurately predicts NO_x emissions using eleven readily available data inputs. A sensitivity study was completed for all major input parameters, which yielded results that agree with conventional wisdom and practical experience. This new algorithm is being used by others, including the Electric Power Research Institute who has included it in its new software for making emissions compliance decisions, the Clean Air Technology Workstation.

Introduction

The formation of nitrogen oxides (NO_x) during pulverized-coal combustion in utility boilers is governed by many factors, including the boiler's design characteristics and operating conditions, and coal properties. Presently, no simple, reliable method is publicly available to estimate NO_x emissions from any coal-fired boiler. The authors previously showed that existing empirical methods were inadequate for predicting NO_x emissions from even a small data set of tangentially fired utility boilers (1). To provide a simple, short-term, multi-purpose means to predict NO_x emissions from pulverized coal-fired utility boilers, a neural network back-propagation algorithm was previously applied to a data set for nine tangentially fired boilers (1). The purpose of this initial study was to assess the ability of a neural network to reasonably predict NO_x emissions for a qualified data set. As previously reported (1), the algorithm developed by the Pittsburgh Energy Technology Center (PETC) generally

predicted NO_x emissions to within ±50 ppm of actual values for the tangentially fired boilers in the limited initial data set.

The Electric Power Research Institute (EPRI) is developing a variety of software tools to assist its member utilities with making decisions on how to comply with the Clean Air Act Amendments of 1990 and the changing power generation marketplace. Sargent & Lundy Engineers has developed the Clean Air Technology (CAT) Workstation as the primary compliance decision tool for EPRI's member utilities. The CAT Workstation includes modules to comparatively evaluate various emissions control technologies. After the 1993 Joint Symposium on Stationary Combustion NO_x Control, where the previous work was presented, EPRI expressed an interest in using PETC's neural network-based algorithm within the CAT Workstation. However, because the data base upon which the initial algorithm was developed was limited to tangentially fired boilers, PETC recommended collection of additional data and development of a new NO_x algorithm. Subsequently, a Cooperative Research and Development Agreement (CRADA) was implemented between PETC and EPRI to collect additional power plant data, develop a new algorithm, and provide a copy of the coded algorithm to Sargent & Lundy for inclusion in the CAT Workstation. This effort was initiated in early 1994 and concluded by mid-year 1994 in order to meet the deadline for release of the CAT Workstation to EPRI's member utilities.

Technology-based standards, such as Reasonably Available Control Technology (RACT), that require the installation of NO_x control technologies on a boiler-by-boiler basis, are driving the emissions compliance strategies of U.S. utilities. Carnegie Mellon University's (CMU) Department of Engineering and Public Policy is examining the potential cost savings achievable through NO_x emissions trading, an alternative market-based strategy to achieve federal and state goals for NO_x emission reduction (2). To estimate the potential cost savings achievable through inter-utility NO_x trading, CMU is using a combinatorial optimization approach to identify boiler retrofits and operating parameters that yield the most cost-effective means for NO_x abatement (2). PETC's new neural network-based algorithm for predicting NO_x emissions was utilized by CMU to predict emissions from pulverized coal-fired boilers in the Pittsburgh, Pennsylvania area as part of their efforts. CMU is continuing development of their trading analysis software for NO_x emissions using PETC's algorithm to predict NO_x emissions.

This paper summarizes the development of the new neural network-based algorithm, including a review of the data base used in the effort and a sensitivity study for the major input parameters. Planned efforts to significantly expand the existing data base using data and information being collected by the U.S. Environmental Protection Agency (EPA) from U.S. utilities, as required by the Clean Air Act Amendments of 1990, is briefly discussed.

Neural Network Approach for Estimating NO_x Emissions

Neural networks became popular in the 1980s for analyzing complex, inter-related data sets. Neural networks do not assume that a relationship is known between process input and output but rather try to determine the relationship by analyzing data sets of input and output data.

Neural networks, which can be viewed as a non-linear data analysis technique, are computational systems that use the organizational principles of biological nervous systems. Computer scientists imitate these principles because biological systems easily outperform all current approaches for pattern recognition. The primary requirement for developing a neural network is a complete data set. Also, the proper representation of the input (e.g., a ratio of two variables may be the best representation) hastens the training process.

Neural networks consist of several simple, highly interconnected data-processing units. These units approximate the complex system of neurons and electrochemical signals that are used by the human brain to process information (3). Each of the neurons in the human brain functions like a tiny computer with limited capabilities. Connected together, these cells form the most intelligent system known. Neural networks are a relatively new class of computing systems formed from hundreds of simulated neurons interconnected like the brain's neurons. The principal aim of the technology is to mimic nature's approach for processing data and information. The back-propagation algorithm developed independently by at least three groups (4-6) is the most widely used neural network approach.

Neural networks are being used to solve real problems. For example, they have been shown to be effective in estimating the fatigue life of mechanical parts (7), performing fault detection in chemical plants (8), diagnosing automobile malfunctions (9), and recognizing human speech (10). Neural networks are also being widely applied within the power industry (3). During the last five years, more than 200 technical papers discussing artificial intelligence applications in the power industry have been published; more than one-half of these applications involved the back-propagation algorithm (11). All of the above applications used noisy data, where the relationships between input-output pairs were complicated and poorly understood. Neural networks generally excel over traditional data analysis techniques, such multiple linear regression techniques, in analyzing noisy data sets.

Commercial neural network software, Brainmaker Professional (Version 2.03) from California Scientific Software, was used in this study. During the network development process, the user defines the problem and collects the input-output pairs. If the user neglects key input parameters for a particular application, the predictive capabilities of the network will suffer. A neural network is not programmed with rules like an expert system, but rather it "learns" in much the same way that people do, i.e., by example and repetition.

After the input-output pairs are collected, the learning stage commences with the network associating output data with input data. Each time an input is presented, the network sends back an answer of what it thinks the output should be. When it is wrong, the network corrects itself by changing the weight matrices that are used to adjust the input-output relationships. Brainmaker Professional uses a back-propagation algorithm to adjust these weight matrices. The training process is repeated until the network derives answers that are within a user-specified tolerance for all inputs. This learning process can take considerable time, possibly up to several days or weeks, to complete depending on the complexity of the problem and the processing speed of the computer used. The final weight matrices are saved to file and can be used external to Brainmaker Professional to predict outputs for other sets of input parameters.

Key to the training process in a neural network is the transfer function. Output is determined by the network's nodes, which combine signals sent to them by lower-layered nodes and transfer these adjusted signals to higher-layered nodes. Neural network nodes are designed to process signals with values between 0 and 1; therefore, all input variables must be scaled to the unit interval. Non-linear transfer functions send signals from node to node within the network. The default transfer function used by Brainmaker Professional is the sigmoid function:

$$f(x) = \frac{1}{1 + \exp(-x)}$$

Figure 1 shows that the sigmoid function maps real numbers into the unit interval. It is often called a "squashing function" because very large positive values are asymptotically mapped to 1 and very large negative values are asymptotically mapped to zero. The squashing function forces the network to focus on the bulk of the data and to place less emphasis on the data at the extreme lower and upper ends of a data set. Hence, network predictions are generally less accurate at the data extremes.

A network stops training when its answers are within a user-specified tolerance. This tolerance specification is applied to the data after it has been subjected to the squashing function. The non-linear nature of the squashing function often causes extreme data values to deviate more from actual values than non-extreme data. Typically, training is conducted at progressively smaller tolerance levels until the deviation between predicted and actual values is acceptable by the user. The value of acceptable deviation between the network answers and the actual data depends upon the accuracy of the available input-output pairs. For example, if inherent errors in the available input-output pairs limit their repeatability to 5%, then training to a tolerance less than 0.1 is unwise. Although decreasing the tolerance will force the network answers to more closely approximate the training data, the network will not necessarily perform better on new data sets. Usually, continued training at progressively smaller tolerances improves the network's performance to a point beyond which it deteriorates. Once confidence is obtained in the predictive capabilities of the neural network, it is used to generate answers for input sets with unknown outputs.

Definition and Scope of Study

To expand the initial data base, additional data were collected from a variety of sources, including several electric utilities, the Utility Air Regulatory Group (UARG), and several projects within USDOE's Clean Coal Technology Program. A total of 384 data sets from 36 pulverized coal-fired boilers were collected and used in this study. Thousands of other data sets were also collected that were not usable because they lacked key data inputs. Also, some data was collected from cyclone-fired utility boilers but was not analyzed because this effort was limited to pulverized coal-fired boilers. Multiple data sets were available for some boilers because of variations in operating conditions and coal type. Some boilers had emissions data from the burners that were originally installed when the unit was built, which

were either conventional or low-NO_x burners, and low-NO_x burners that were retrofitted at a later date. Both pre- and post-NSPS (New Source Performance Standards) boilers were included in the data set. Table 1 summarizes the characteristics of the boilers in the available data set as a function of firing mode (e.g., wall, opposed-wall, cell, tangential). No data were available from wall-fired units with low-NO_x burners, largely as a result of the limited time available to produce an algorithm for inclusion in EPRI's CAT Workstation. The new tangentially fired boiler data consisted of a few data sets from many boilers whereas the data for the other boiler types consisted of many data sets from a few boilers. The capacity ranges of the wall-fired boilers, opposed wall-fired boilers, and cell burner-equipped boilers were limited compared to that of the tangentially-fired boilers because attempts to acquire additional data over a broader capacity range were unsuccessful during the limited available time.

A number of possible input parameters, based on boiler design characteristics and operating conditions, and fuel properties, were considered during development of the new NO_x algorithm. The objective was to use a limited number of simple, readily available parameters as long as the predicted NO_x values agreed reasonably with actual emission values for a range of boiler types, boiler operating conditions, and coal properties. The starting point for selection of the new input parameters was the list of parameters used for the original NO_x algorithm. However, data availability precluded use of some of the original inputs and extension of the data set to additional boiler types required inclusion of new parameters to adequately describe the various boiler designs.

Table 2 summarizes the parameters that were eventually selected as input to the new neural network. Furnace height, which is used to calculate the furnace volumetric heat release rate, was defined in this effort as the distance from the bottom of the ash hopper to the furnace roof (i.e., the overall boiler height). The more standard definition for boiler height, which is the distance from the furnace hopper knuckle to the furnace nose, was not used because detailed furnace dimensions were not available for all units and time limitations did not permit collection of this data from individual power stations. The year that the boiler first became operational was used because recent boiler designs attempt to limit NO_x formation in conjunction with low-NO_x burners. The year that the current burners were installed was used as an indicator of the developmental status or maturity of low-NO_x burner technology. It would be too difficult to use individual burner design specifications for different manufacturers in this simplistic approach. The ratio of boiler load to the boiler's maximum continuous rating (MCR), along with the plan area and volumetric heat release rates, are used to reflect a boiler's overall design and operating characteristics. Most utilities provided the NO_x and excess oxygen values at the economizer outlet; therefore, this emissions sampling location was selected. Other utilities provided NO_x and O_2 data at the boiler exit or stack. However, there were fewer data sets with this information. Also, because leakage of air into the exhaust system of a plant can not be distinguished from operating conditions with higher excess air, stack emissions data are difficult to use. Therefore, only data sets with the oxygen content in flue gas at the economizer outlet were used in this study. No input parameter was used to address the impact of over-fire air usage or mill firing pattern on NO_x emissions because sufficient data were not available for network training. While omission of these parameters is recognized as a significant weakness because they are known to have a major

impact on NO_x emissions levels, the previous effort yielded reasonable NO_x emissions estimates for a wide variety of boilers, boiler operating conditions, and coal types without inclusion of this information.

Neural Network Results

Figure 2 compares the actual and estimated NO_x emission values for the 346 data sets used to train the network. The network was trained at a tolerance of 0.10, i.e., training was complete when all of the scaled predicted values (i.e., from the squashing function) agreed to within 10% of scaled actual values. As previously stated, the non-linear nature of the sigmoid function causes large deviations between actual and estimated values at the extremes of the data. The greatest difference between an actual NO_x value and its corresponding estimated value is 263 ppm, which occurred for a measured NO_x value of 1350 ppm (0% O_2, dry volume basis). The standard error of estimate for the training data was 67 ppm. Training at a smaller tolerance reduced the maximum deviation and the standard error of estimate but the network performance degraded on the data that was reserved for testing. As shown in Figure 2, the neural network's maximum estimate is 1150 ppm (0% O_2, dry volume basis) Increasing this value would improve the fit for extremely high NO_x emission values but would degrade the fit of the non-extreme data. It was felt that minimizing the deviations between actual and estimated NO_x emissions for the bulk of the (i.e., non-extreme) data was more important than minimizing deviations for the extremely high NO_x emission values. The standard error of estimate decreased from 67 ppm to 52 ppm if only the data sets that had an actual NO_x values less than 1150 ppm (0% O_2, dry volume basis) were considered.

Figure 3 compares actual and estimated NO_x emissions for the 38 data sets reserved for testing by the neural network. These data sets represent the balance of data sets in the master data set. The standard error of estimate for the test data was 64 ppm. The greatest difference between a measured NO_x value and its corresponding estimated value was 156 ppm, which was again for a data set with an extremely high actual NO_x emission value.

Sensitivity of Input Variables

Although the prior study yielded a neural network that was very good at estimating NO_x emission values using readily available input data, little effort was placed on assessing the reasonableness of the data trends predicted using the network algorithm. To assess the sensitivity of the selected input parameters on NO_x emission trends, a series of numerical experiments were performed using the new algorithm. While these experiments do not fully evaluate the accuracy of the algorithm because actual data were not available for comparison in all cases, they do reveal trends that agree with conventional wisdom and practical experience. Also, they demonstrate how the algorithm can be used in various utility scenarios, such as low-NO_x burner retrofits and boiler operational changes. These sensitivity analyses are not meant to describe trends that are universally applicable to all pulverized coal-fired boilers but rather to show that, while the trends are generally reasonable, specific analyses can yield results that require further explanation. To simplify these analyses, input

parameters were varied individually except when fuel analysis effects were being assessed. As detailed in Table 3, four different scenarios were analyzed:

1) a pre-NSPS 200 MW_e wall-fired boiler built in 1955 that is equipped with the original burners
2) a pre-NSPS 200 MW_e wall-fired boiler built in 1955 that was hypothetically retrofitted with low NO_x burners in 1985
3) a pre-NSPS 180 MW_e tangentially fired boiler built in 1967 that is equipped with the original burners
4) a post-NSPS 250 MW_e tangentially fired boiler built in 1984 that is equipped with the original burners

At 5-year increments from 1975 until 1990, the effect of retrofitting low-NO_x burners on NO_x emissions is shown in Figure 4 for the pre-NSPS 200 MW_e wall-fired boiler built in 1955 and the pre-NSPS 180 MW_e tangentially fired boiler built in 1967. In this figure and all subsequent figures, the two dates in parentheses indicate the year that the unit first commercially operated and the year that the current (i.e., either original or retrofit burners) were installed. In this analysis, it was assumed that other boiler improvements were implemented when the burners were retrofitted to reduce the flue gas oxygen content at the economizer exit from 8.4% to 5.4% for the wall-fired unit and from 5.5% to 3.5% for the tangentially fired unit. No other input parameter was varied. Figure 4 and subsequent figures (where data were available) shows the excellent agreement between actual and predicted NO_x emission values. Retrofitting low-NO_x burners in 1975 reduces the emissions from the wall-fired unit and the tangentially fired unit by about 28% and 35% and in 1990 by about 38% and 38%, respectively. Interestingly, the same 38% reduction was predicted for both units when low-NO_x burners are retrofitted in 1990. It is not known whether this level of NO_x reduction represents a limit imposed by the limited available data set; additional sensitivity analyses need to be performed for other units to investigate this issue. However, these trends generally do agree with the observed ability of low-NO_x burners to reduce emissions. Also, essentially no improvement in the ability of low-NO_x burners to reduce emissions from the tangentially fired boiler was predicted over the period from 1975 to 1990; other plants may show significantly different results.

One of the simplest, most cost-effective means to control NO_x emissions from pulverized coal-fired utility boilers is reducing the operating excess air level. The effect of excess air level on NO_x emissions from the above four units is shown in Figure 5. In these numerical experiments, the excess air level was varied 2% higher and lower than the actual reported value at the economizer exit of each boiler. Significant changes in NO_x emissions are predicted as the excess oxygen content is varied for all units except the post-NSPS 250 MW_e tangentially fired unit, which has an actual NO_x emission level of only 270 ppm (0% O_2, dry volume basis). These trends generally agree with actual experience in these units and other field measurements.

Fuel properties can have a significant impact on NO_x emissions from pulverized-coal combustion. Substantial differences in NO_x emissions have been observed when the coal being fired is changed or blended with other coals. The two coals that were fired in the pre-

NSPS 200 MW_e wall-fired unit and the pre-NSPS 180 MW_e tangentially fired unit were selected for a numerical blending simulation. As shown in Table 3, the coal burned in the wall-fired unit had a fixed carbon-to-volatile matter (FC/VM) ratio of 1.42 and the coal burned in the tangentially fired unit had FC/VM of 1.02. These two coals were numerically blended at ratios of 1/3, 1/1, and 3/1 to produce simulated coals with compositions between the two actual coals. The effect of FC/VM ratio on NO_x emissions is shown in Figure 6. Significant changes in NO_x emissions are predicted for the two pre-NSPS units whereas the two units with low-NO_x burners are relatively insensitive to fuel properties. Some low-NO_x burner manufacturers claim that their burners are not sensitive to some fuel properties, such as FC/VM ration, while others report significant sensitivity. Again, other scenarios may show substantially different results.

Boiler design obviously has a major impact on NO_x emissions. Generally, lower furnace heat release rates will yield lower NO_x emissions as a result of lower flame temperatures and changes in burner/furnace aerodynamics. Figure 7 shows the effect of plan area heat release rate on NO_x emissions. In this analysis, incremental changes, both higher and lower, in the plan area heat release rate of 5%, 10%, 15%, and 20% from the actual design point for these boilers were evaluated. Also, the volumetric heat release rate and other input parameters were held constant. Because volumetric heat release rate was held constant, these analyses reflect a change in boiler design (e.g., to achieve lower plan area heat release rates, a unit will be made wider and/or deeper while the unit height is decreased) not a change in boiler firing rate (i.e., load). In general, the effect of plan area heat release rate was minimal for all four units. This result seems contrary to practical experience but can be partially explained by the fact it is unlikely that only plan area heat release rate would be changed if a boiler design was being changed to achieve lower NO_x emissions. This is supported by Figure 8, which shows a very strong effect of volumetric heat release rate for all units except the post-NSPS 250 MW_e tangentially fired unit, which has very low actual NO_x emissions. In this analysis, NO_x emissions were determined for incremental changes in furnace volumetric heat release rate of 5%, 10%, 15%, and 20% higher and lower than the actual design point for each boiler. Because the two input parameters of plan area and volumetric heat release rate are so strongly related, it possible that only one of these inputs is required to adequately describe the combined impact of firing rate and boiler design on NO_x emissions.

Future Work

The Clean Air Act Amendments of 1990 require most U.S. coal-fired power plants to install continuous emissions monitoring (CEM) systems and to electronically report the collected data to the U.S. Environmental Protection Agency on a quarterly basis. EPA has collected up to a full year of data from some utilities and several quarters of data from the remaining utilities. PETC has worked with EPA on various NO_x issues, including assessing the performance and cost of various combustion and post-combustion control technologies. Discussions have been held with EPA on acquiring the first year's emissions data from all U.S. coal-fired power plants equipped with CEMs to expand the data base for development of a new NO_x prediction algorithm. While this emissions data base would be unequalled by any other, EPA is not authorized to collect all the related information that is required to predict

NO_x emissions accurately. For example, EPA is not collecting all the boiler design and coal property data that is used in the current NO_x algorithm. Collection of this data would require an effort that is beyond current resources. However, several options to perform this task are still being considered.

PETC has also discussed similar efforts with EPRI to develop a neural network-based NO_x algorithm for gas- and oil-fired utility boilers. The Gas Research Institute (GRI) has been using EPRI's CAT Workstation, which includes PETC's NO_x algorithm, as part of their market projections for natural utilization in coal-fired boilers (12). GRI has expressed an interest an algorithm to predict NO_x emissions from coal-fired boilers using natural gas co-firing or reburning.

PETC will continue to expand its current data base with a focus on including more data from power plants that have retrofitted low-NO_x burners. When sufficient data are available, a new neural network-based NO_x algorithm will be developed and supplied to existing and new users. PETC hopes to initiate some of the above activities later this year, which will likely result in an improved algorithm for all coal-fired boiler types.

Summary

A neural network has been developed to estimate NO_x emissions for wall-fired boilers, opposed wall-fired boilers, cell burner-equipped boilers, and tangentially fired boilers. Eleven simple, readily available parameters were selected as input parameters. The data base used in this effort consisted of boiler design characteristics and operating conditions, and coal property data for 36 pulverized coal-fired utility boilers. Some of these boilers were either originally equipped with low-NO_x burners or had been retrofitted with low-NO_x burners. A sensitivity analysis of the selected input parameters yielded trends that agree with conventional wisdom and practical experience. Some of the current and future applications for PETC's neural network-based NO_x algorithm were highlighted; the most notable of the current applications is the algorithm's inclusion in the Electric Power Research Institute's Clean Air Technology Workstation, which is the primary tool for making emissions compliance decisions tool for its member utilities.

Disclaimer

Reference herein to any specific commercial product by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof.

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Table 1

Boiler Database Information as a Function of Firing Mode

Firing Mode/ Boiler Information	Wall	Opposed Wall	Cell	Tangential
Number of Boilers	6	6	3	21
Number of Units with Low-NO_x Burners	0	3	1	4
Number of Data Sets	77	96	55	156
Number of Units with 2 Furnaces	1	0	0	9
Range of Boiler Capacity (Gross MW_e)	175-200	265-688	610-1300	85-936

Table 2

Input Parameters Selected for NO_x Estimation Study

Parameter	Range of Values	Units
Plan Area Heat Release Rate	0.4-3.3	$MMBtu/ft^2$
Furnace Volume Heat Release Rate	3.4-29.7	$KBtu/ft^3$
Firing Mode (1=tangential; 2=wall; 3=opposed wall; 4=cell)	1-4	--
Number of Furnaces	1 or 2	--
Year Unit First Commercially Operated	1951-1986	year
Year Burners Installed (Original or Retrofit Burners)	1951-1990	year
Flue Gas Oxygen Content at Economizer Outlet	1.8-12.1	vol%, dry basis
Load/Maximum Continuous Rating	0.53-1.12	MW_e/MW_e
Coal's Carbon-to-Hydrogen Ratio (C/H)	13.3-21.3	wt%/wt%
Coal's Nitrogen Content	0.8-1.8	wt%, dry basis
Coal's Fixed Carbon-to-Volatile Matter Ratio (FC/VM)	1.0-3.3	wt%/wt%

Table 3

Base Input Parameters for Sensitivity Study Cases

Parameter	Units	Unit #1	Unit #2	Unit #3	Unit #4
Plan Area Heat Release Rate	$MMBtu/ft^2$	1.57	1.57	1.71	1.80
Furnace Volume Heat Release Rate	$KBtu/ft^3$	16.53	16.53	18.03	14.55
Firing Mode (1=tangential; 2=wall; 3=opposed wall; 4=cell)	--	2	2	1	1
Number of Furnaces	--	1	1	1	1
Year Unit First Commercially Operated	year	1955	1955	1967	1984
Year Burners Installed (Original or Retrofit Burners)	year	1955	1985	1967	1984
Flue Gas Oxygen Content at Economizer Outlet	vol%, dry basis	8.40	5.40	5.45	2.80
Load/Maximum Continuous Rating	MW_e/MW_e	1.00	1.00	1.00	1.03
Coal's Carbon-to-Hydrogen Ratio (C/H)	wt%/wt%	14.98	14.98	16.67	13.37
Coal's Nitrogen Content	wt%, dry basis	1.70	1.70	1.49	1.09
Coal's Fixed Carbon-to-Volatile Matter Ratio (FC/VM)	wt%/wt%	1.42	1.42	1.29	1.02

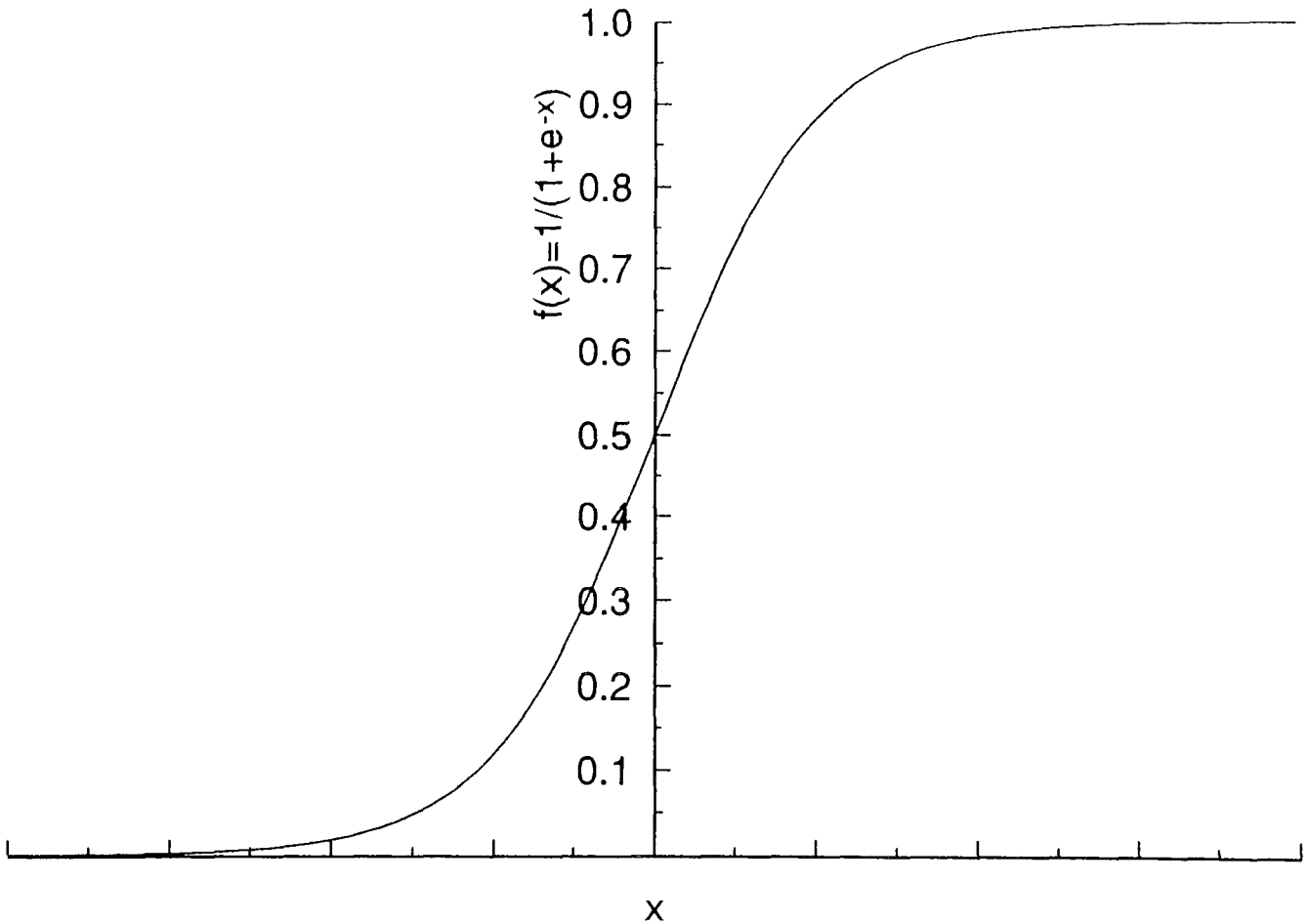

Figure 1. Sigmoid "squashing" function

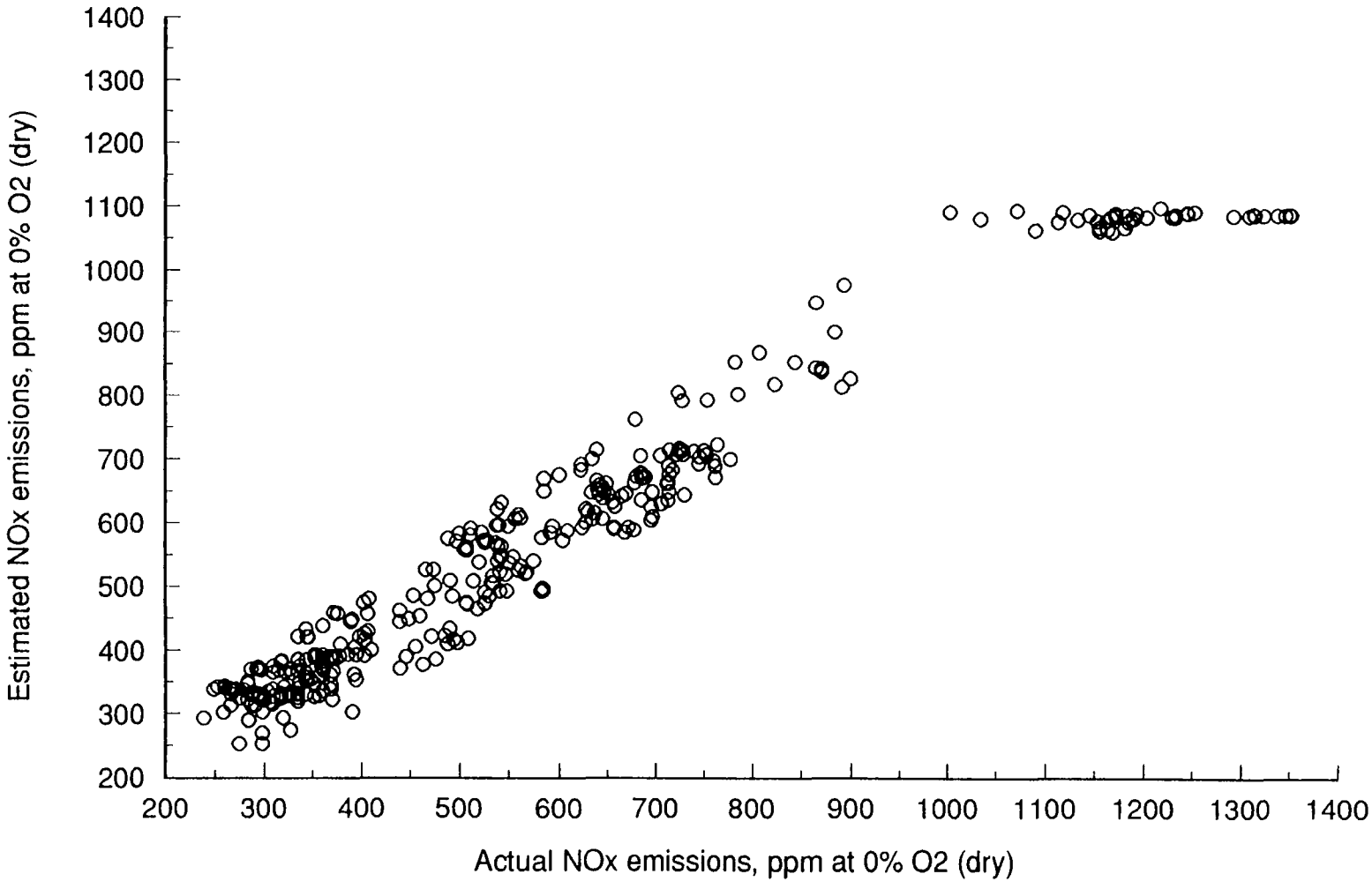

Figure 2. Actual and estimated NOx emissions (training data)

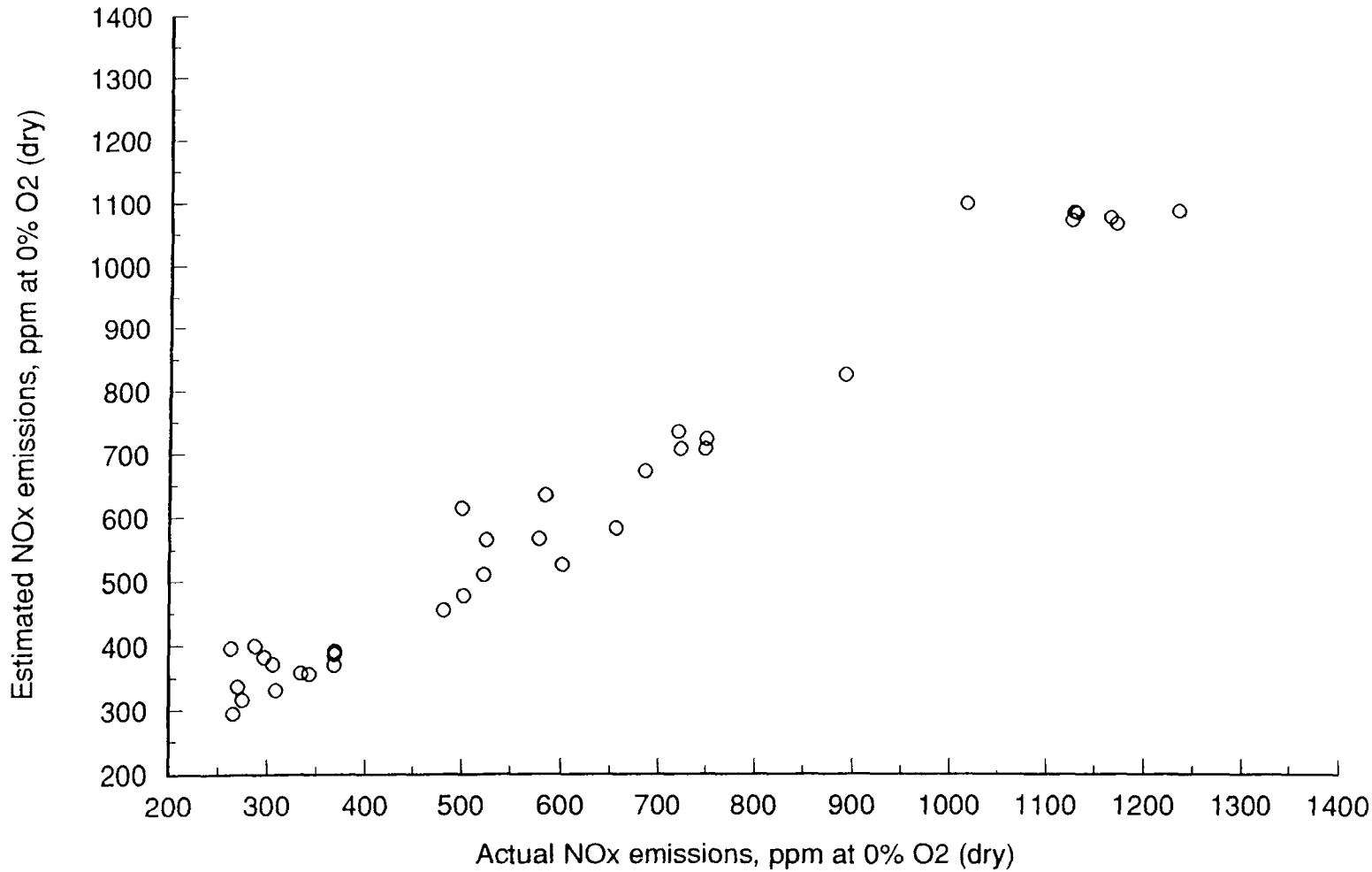

Figure 3. Actual and estimated NOx emissions (test data)

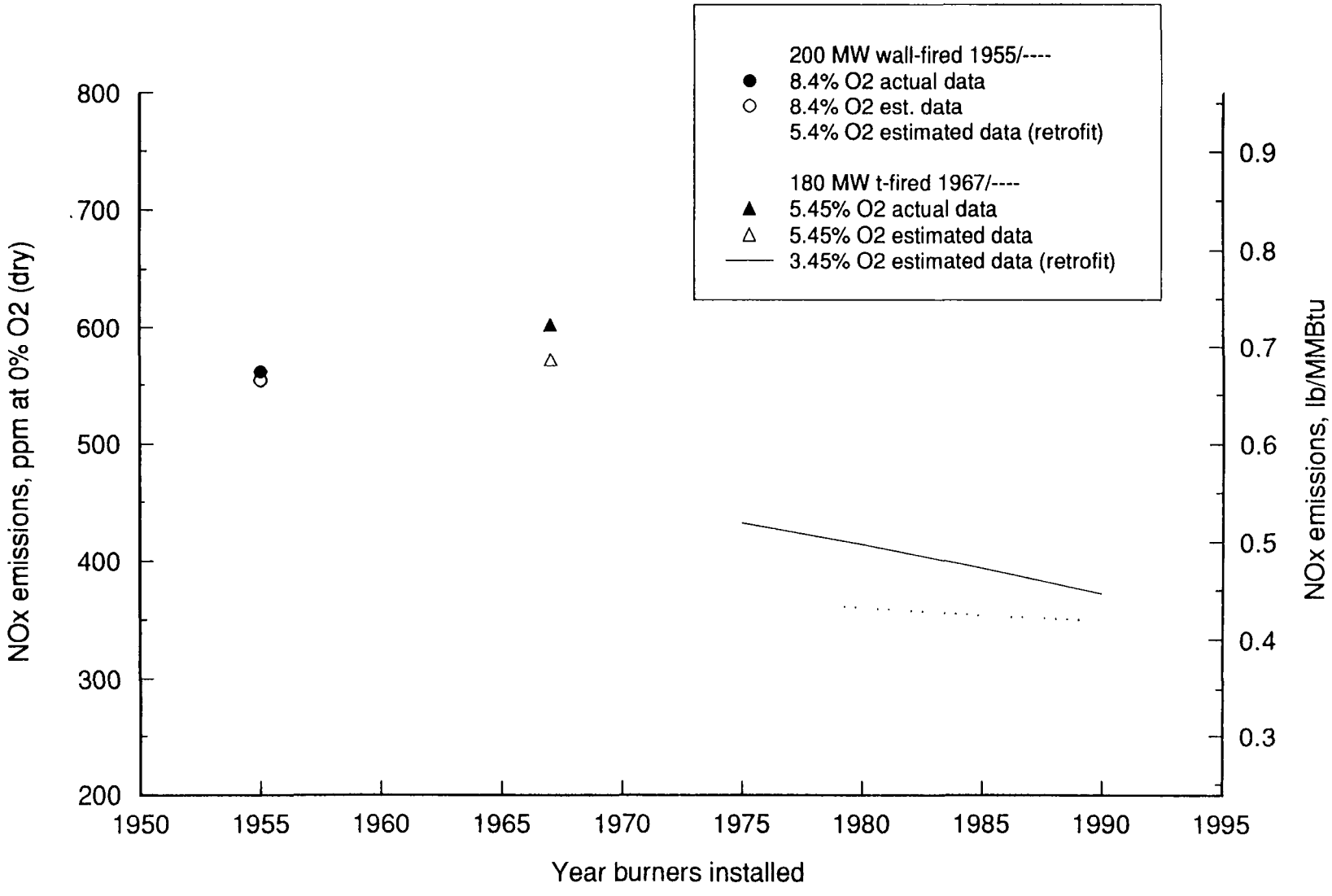

Figure 4. Effect of retrofit burners on NOx emissions

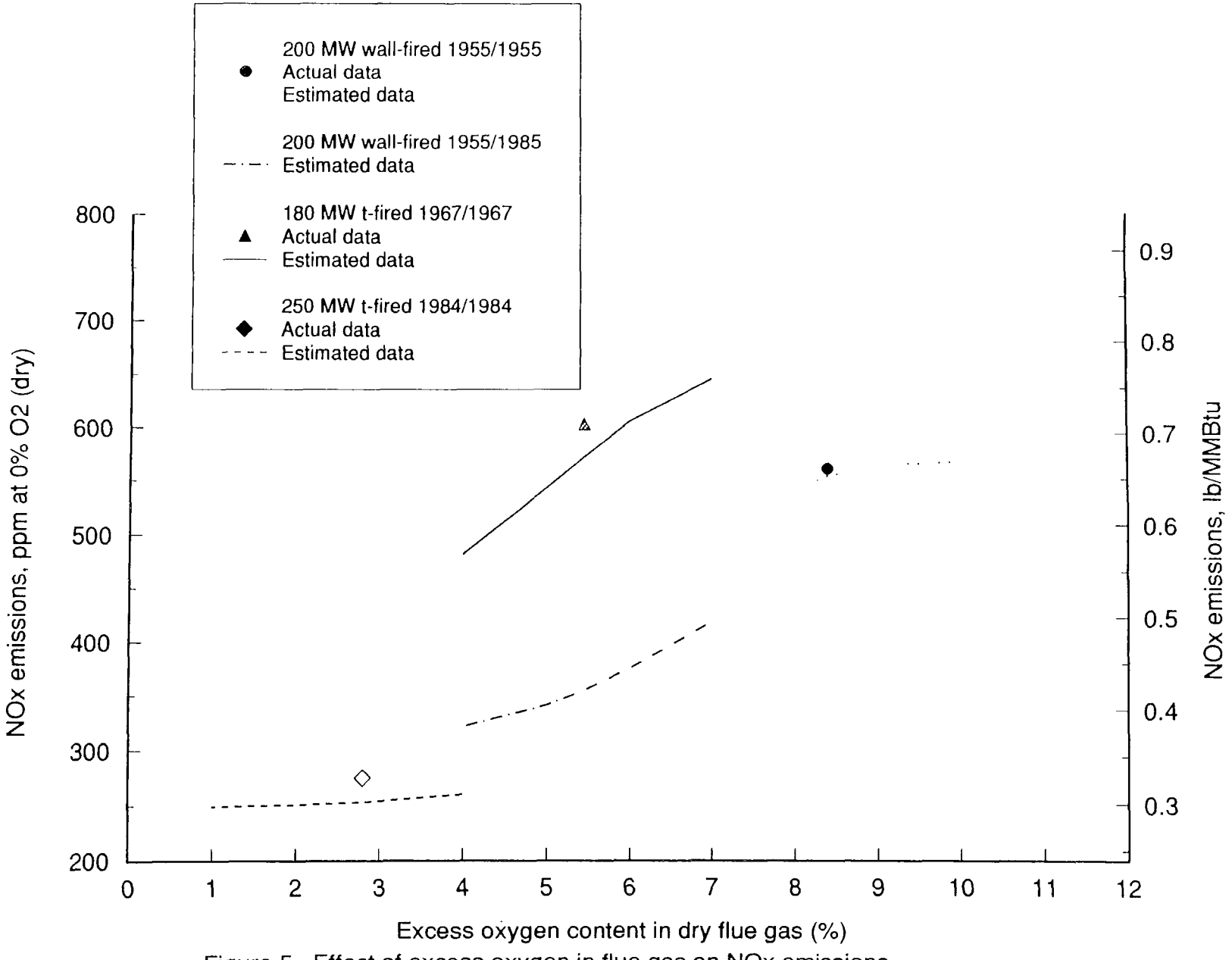

Figure 5. Effect of excess oxygen in flue gas on NOx emissions

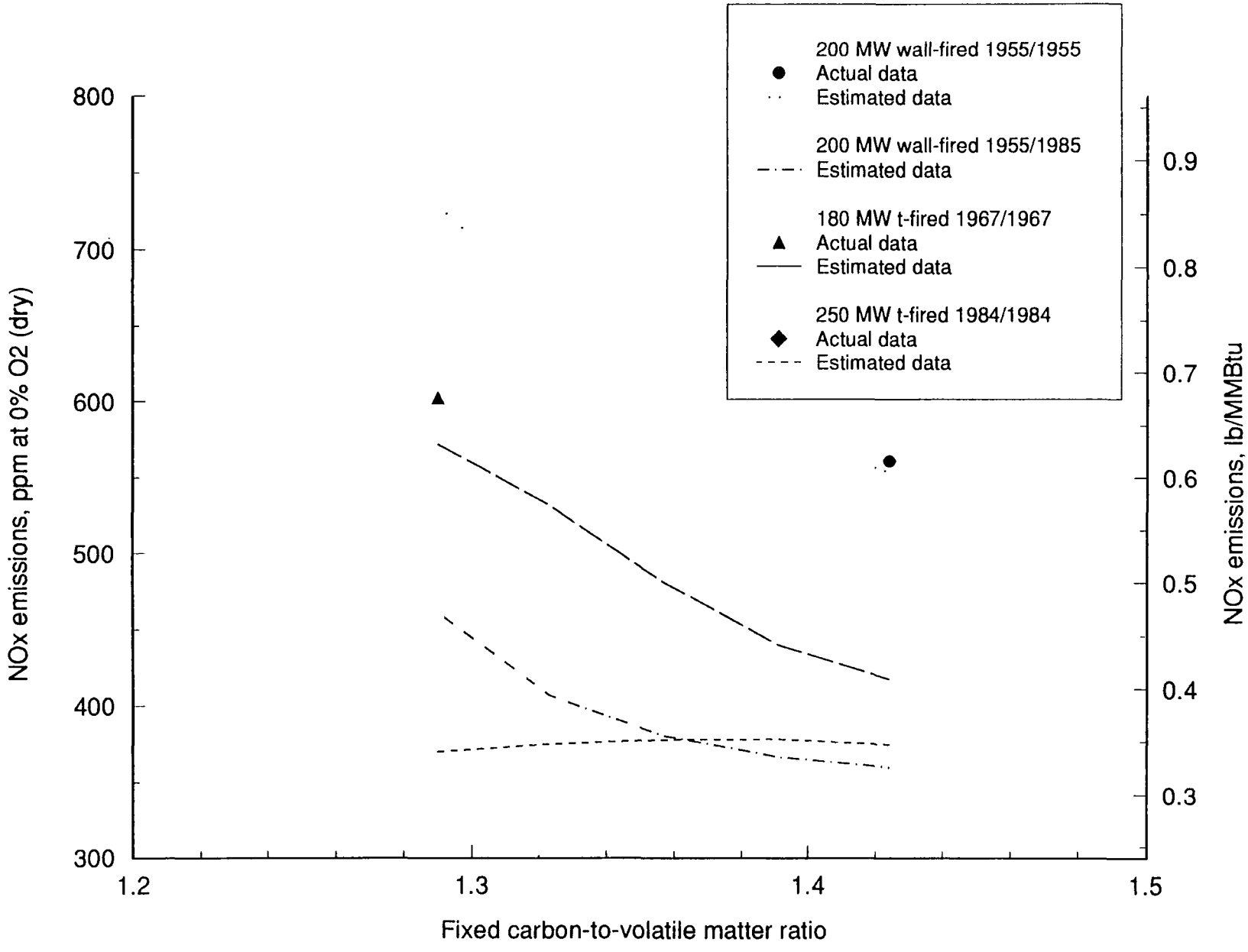

Figure 6. Effect of coal composition on NOx emissions

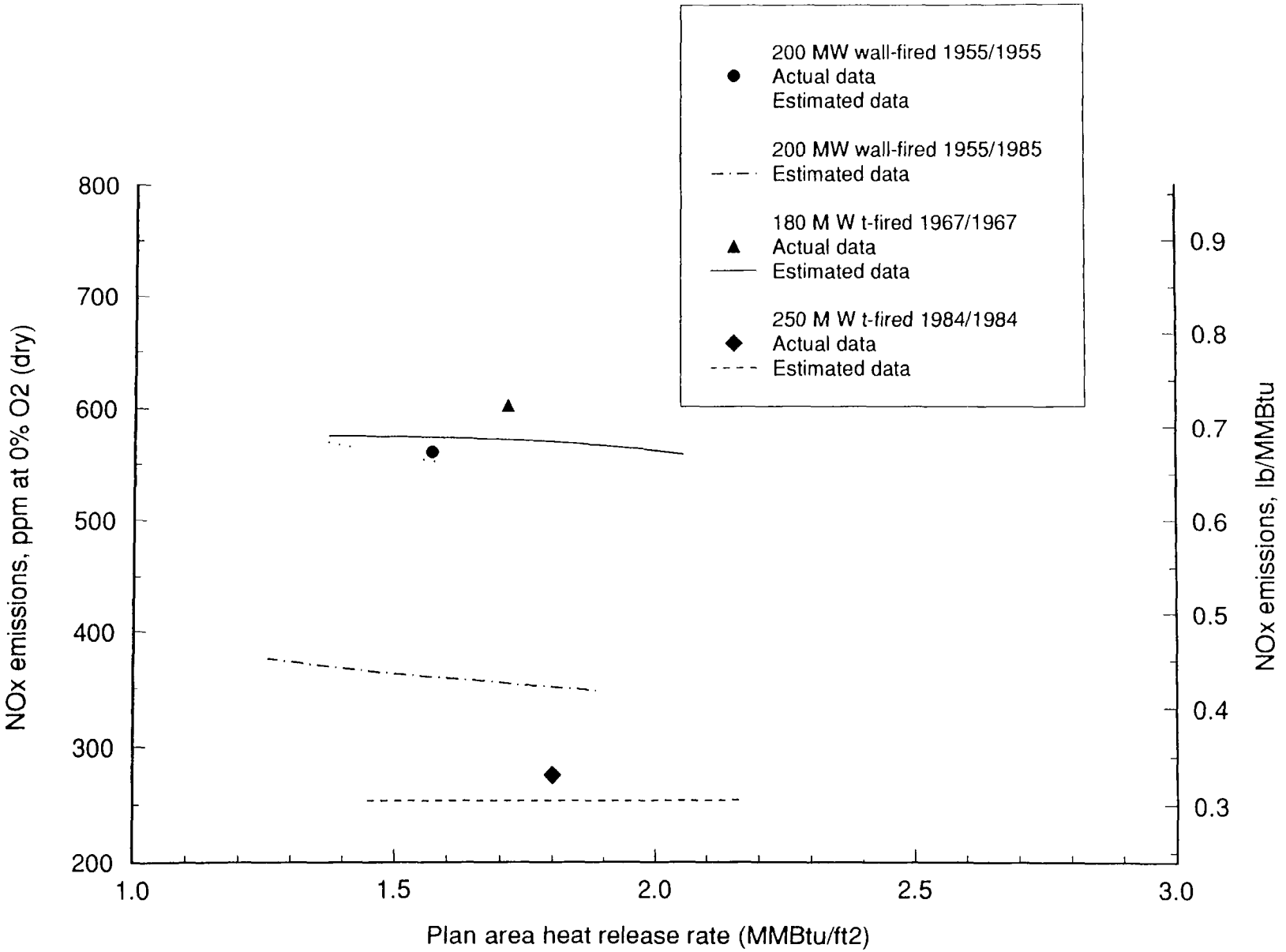

Figure 7. Effect of plan area heat release rate on NOx emissions

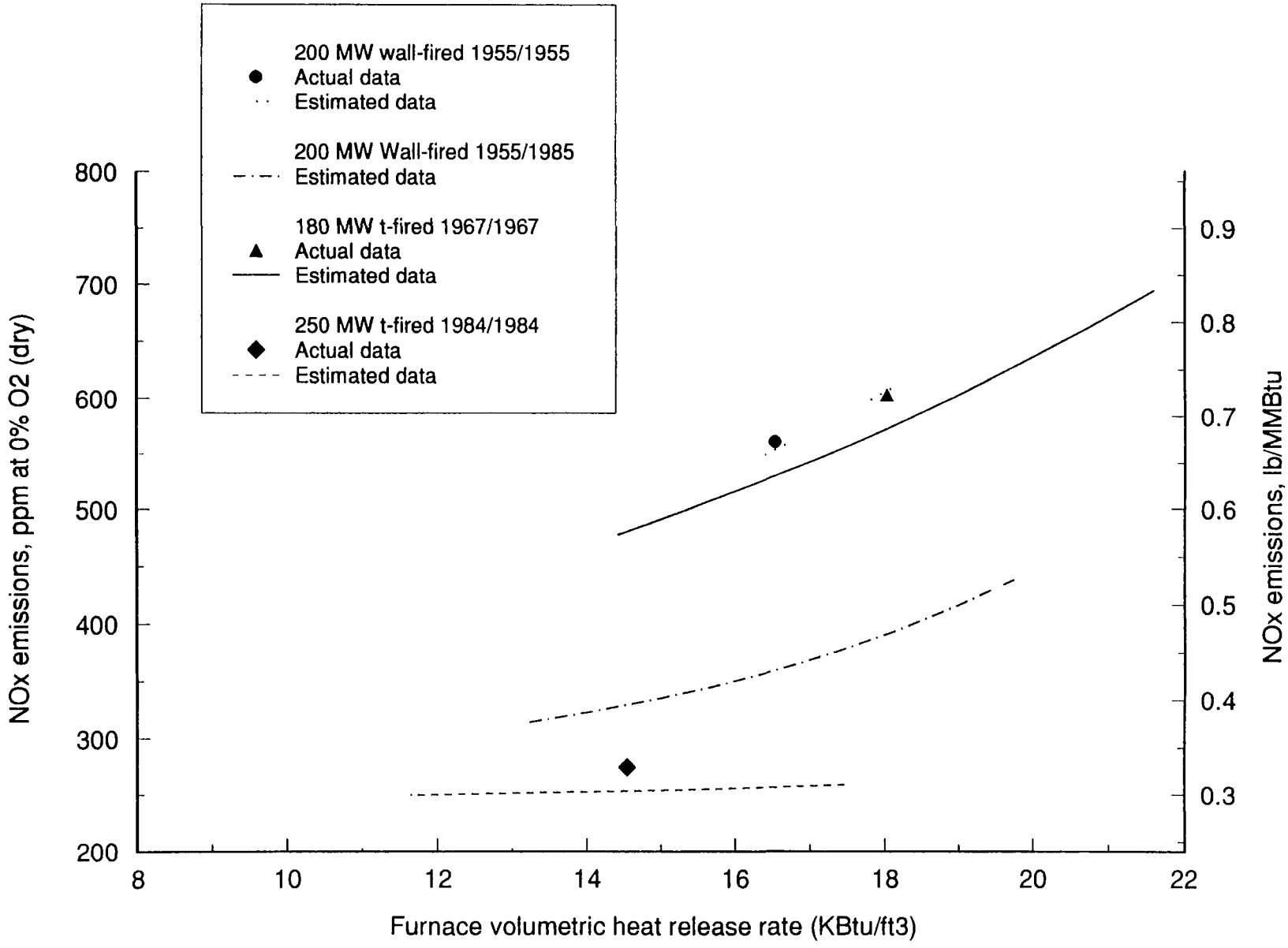

Figure 8. Effect of furnace volumetric heat release rate on NOx emissions

Separation and Recycling of Carbon in Ash

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Abstract

Separation Technologies has developed a new electrostatic process for separating the unburned carbon from utility flyash. As NOx emission levels are reduced through burner modifications and as harder and lower ash coals are burned, the carbon levels in the fly ash increase above the maximum levels for use in concrete. This paper details the results of a production test program carried out in mid-1994 to produce low carbon ash for use in concrete. 3000 tons of low carbon ash were produced at rates of 15-25 tons per hour. The average loss-on ingnition of the fly ash was reduced from 8.2% down to 2.3% with 83% yield.

Introduction

The carbon level in flyash from utility boilers has been increasing in the recent past as a result of a number of factors. Not the least of which is the requirement for many utilities to reduce the amount of NOx emitted. This is usually done through retrofit of special burners and air controls that have the effect of reducing the excess air and reducing the flame temperature. Both of these changes increase the level of unburned carbon in the ash.

Additional factors which compound the problem include increased competition among coal suppliers. With reduced demand for coke, coking coals are now available for combustion. Sulfur content for SO2 emissions and CaO content for precipitator performance are also becoming more important. The net result is that many boilers are burning coals with specifications different than what the boilers were designed for. Two of the more important specifications are Hardgrove grindability index and volatile content. Harder to grind coals with lower volatile do not burn as rapidly and require a longer residence time for complete combustion. This time is not available and so the carbon content of the ash increases. Lower ash coals can exacerbate the problem because with less ash, the concentration of carbon in the ash can increase while the total amount stays the same.

Increased carbon in flyash presents a number of problems to the utility. The unburned carbon represents a loss of fuel value which can amount to a few percent of the total fuel consumption of the boiler (at 1% loss and $40/t this represents $400,000 for every million tons of coal). The

high carbon ash is also more difficult, and more expensive to dispose of. High carbon ash has a lower bulk density, so more trucks are needed to carry the same tonnage off site, and at the same time there is more ash to dispose of, and less fits in the land fill (land fill disposal has a volumetric limit, not a weight bearing limit). Productive uses of high carbon flyash are quite limited.

The largest potential use for flyash is as a pozzolanic additive in concrete where the concrete properties can be improved. When cement hydrates half the calcium is liberated as free lime. This makes conventional concrete quite alkaline, and susceptible to sulfate degradation if dissolved sulfate percolates through the concrete, reacts with the lime and precipitates as gypsum causing swelling and eventually cracking. Flyash in concrete reacts with the free lime to form cementacious products which produce additional strength in the cured concrete. This property of pozzolanic materials has been known for many years, and some building codes allow the substitution of flyash (of certain grades) for a portion of the cement. Other improved concrete properties include lower water content, lower heat of hydration, lower cost, easier flowability, and lower permeability.

High carbon flyash is not usable in concrete. The standard ASTM C-618 specifications for pozzolanic flyash call for less than 6 % loss on ignition. Many specific projects use much lower specifications. For example the Boston Harbor Central Artery Project has a specification of less than 3 % loss on ignition.

The proportion of carbon and the properties of the ash result from the composition of the coal and the conditions it experiences during combustion. Coal is a very inhomogeneous material. It is formed from assorted vegetation that accumulated in bogs and was protected from decay by anaerobic, acidic water. The minerals in coal are present in several forms. The organic material from the original vegetation contains trace minerals as well as carboxyl groups that in low-rank coal act as cation exchange sites. These sites collect cations from water flowing through the coal deposit. Some low rank coals have high levels of these ion exchange sites and have trapped large amounts of alkali metals. As the coal matures and increases in rank, these carboxyl groups decompose and produce CO_2 which escapes. This is accompanied by collapse of pores in the coal structure and a trapping of those elements with high organic affinities and many other generally soluble elements. The other major source of ash in coal is from mineral grains that have been mixed with the coal by the actions of water, wind and earth movements over geologic time, such as sand and clay.

The dominant method of coal combustion is suspension burning, there are multiple types and sizes of burners and points of introduction into the furnace. There is also wide and increasing use of two-stage combustion with sub-stoichiometric firing at the burner with introduction of additional air to the upper furnace for completion of combustion. All these variations together with furnace size and configuration for internal recirculation as well as the diversity of coals being pulverized and burned gives a broad spectrum of range of carbon in the fly ash.

When coal is mined and processed for combustion, the minerals associated with it, along with extraneous rock and seam partings, are all pulverized down to a typical specification of about

70% less than 74 microns. Pulverization liberates many of the mineral particles. The design, state of repair, and operating adjustment of a stations pulverizers affects the actual degree of pulverization achieved. In suspension burning the pulverized coal is dispersed in air and then exposed to thermal radiation from the flame which thermally cracks and vaporizes organic constituents in the coal particle leaving char which burns more slowly. It is this char, particularly from large coal particles, which exits the furnace before completion of combustion. The mineral grains, exposed to the heat from the flame, melt and form spheres. Pyrite is oxidized to magnetite.

The residence time in the furnace is too short for most particles to collide and mix. Mineral particles therefore retain their original oxide content except for being depleted in volatiles. The liquid silicate spheres are quenched as the combustion products pass through the heat exchange surfaces and remain in the flyash as glassy spheres.

Lime produced during the hydration of concrete can dissolve some of the amorphous silicate particles and forms cementacious material that adds strength. This reduces the lime which can otherwise form calcium sulfate compounds which swell and crack the concrete.

Crystalline material is not soluble in the alkali liberated during the hydration of concrete and so merely acts as an inert filler. Ash from some combustion processes cannot be used in concrete. Fluid bed ash is not suitable because of the alkali added for SO_2 control, but also fluid bed combustion occurs at a lower temperature so the particles are not glassy, but instead are crystalline. The thermal history of the ash is important in achieving ash that is pozzolanically active.

For concrete admixture applications, the glassy silicate spheres are the important component. Unburned carbon adsorbs the air-entraining agents which are added for freeze thaw resistance and is undesirable. Water, sulfate and alkali are also deleterious and may not exceed 3.0, 5.0 and 1.5% expressed as H_2O, SO_3 and Na_2O (ASTM C618). The low water content needed precludes a wet process because of the costs of drying and also the potential for pozzolanic reactions to occur during the wet separation process which would cement particles together, decrease subsequent pozzolanic activity and tie up alkali with the silicate particles.

Electrostatic Separation and Contact Charging Theory

Electrostatic separators that utilize contact potential differences between materials are also known as triboelectric separators. Triboelectric (from the Greek tribein, to rub) is a misnomer because the electrostatic charging occurs as a result of surface contact and rubbing and the resultant friction is unimportant except to increase the degree and area of surface contact.

All solids are held together by chemical bonds. These bonds result from the energy structure of electrons in the material. Some electrons are held tightly in bonds, some electrons are free to move, and some can only move if a sufficient activation energy is provided. The magnitude of the activation energy for electron migration determines if a material is a metallic conductor, an

insulator, or a semiconductor. The semiconductor industry is based on the electrical properties that originate at the junction between different material phases.

When two materials are in contact, electrons move (or tunnel) until the energy of electrons in each material at the interface is equalized. The material with a higher affinity for electrons gains electrons and charges negative, while the material with the lower affinity loses electrons and charges positively. A measure of the relative affinity for electrons is the work function, which is the energy to move an electron from the surface to infinity (99 % of which amounts to 10^{-8} meters). The work function is dependent on the chemical composition of the surface. Two surfaces of different work function in contact will exchange electrons until there is a voltage between them equal to the difference in their work functions.

Charge transfer continues between the two materials until the developing electric field at the interface equalizes the energy of electrons at the surfaces of the two materials. As the materials are then moved apart, charge can no longer transfer across the intervening gap, and the two particles now possess equal but opposite sign charges

The contact exchange of charge is universally observed for all materials, and causes the electrostatic nuisances that are a problem in some industries. As a practical matter, two surfaces that are ostensibly the same are still enough different that contact charging will occur. The extreme sensitivity of the surface potential to surface differences explains why in the semiconductor industry dopants at parts per billion can have large effects. Single crystal pure elements can have differences in work function between different crystal faces exceeding one volt (10). Contact charging of a surface is necessarily an averaging of charge exchange at many points. A particle exchanges charge with every particle it contacts, and so its net charge fluctuates around an average which depends upon its neighbors.

The work function of a material is the energy required to remove an electron, and is dependent upon how tightly electrons are held in chemical bonds. A measure of an element's tendency to attract electrons is termed electronegativity. Elements with high electronegativity, such as fluorine, oxygen and sulfur, strongly attract electrons. It has been found that the work function of a material is related to its electronegativity, and that the work function of a compound can be estimated from the electronegativity of its atomic constituents and the degree of electron sharing occurring in the chemical bonds, which allows prediction of work function and allows estimation of separation characteristics of minerals based on their composition. A Lewis acid is defined as an electron receptor, and a Lewis base is an electron donor. Observed separations of minerals are consistent with work function values estimated from electronegativity and bonding. Oxidized materials tend to charge negatively, as do acidic materials. Coal, being a hydrocarbon, charges positively with respect to most metal oxides. Oxides such as silica and sulfides such as pyrite charge negatively with respect to coal, and are readily separable electrostaticly.

Process Description

Separation Technologies, Inc. (STI) has developed a triboelectric separation process with a new patented separator geometry that solves scale-up problems and improves electrostatic separation by 4 or more orders of magnitude, compared to prior technologies. This dramatic improvement has now allowed the economical dry separation of materials inseparable by any other means.

In the STI separator the material is fed into the thin gap between two parallel plane electrodes (see Figure 1). The particles are then swept up by a moving open mesh belt and conveyed in opposite directions. The moving belt sets up a counter current plane couette flow field independent of the electric field. The particles charge by contact with their neighbors. The electric field moves opposite charged particles either up or down depending on their sign of charge. The field need only move the particles a tiny fraction of an inch to move a particle from a left-moving to a right-moving stream. The belt moves the particles adjacent each electrode toward opposite ends of the separator. The counter current flow of the separating particles provides for multiple opportunities for separation and results in excellent separation and recovery.

The separator is very simple. The belt and the associated rolls are the only moving parts. The electrodes are stationary and are composed of an appropriately hard material. The belt is an open mesh polymer. The over all separator length depends on the specific application, but is in the range of 20 feet plus room for rolls and hoppers. The belt speed is the same order as that of pneumatic conveying systems. The power consumption of the electrostatic power supplies is negligible. The power consumption of the belt drive is on the order of 1 kW hr per ton processed.

The short gap, the very high field, the counter current flow, the vigorous particle-particle agitation and the self-cleaning action of the belt on the electrodes are the critical features of the STI separator that give it such good performance. The overall separation performance of the STI separator is superior to all previous contact-charging separators that have been considered for commercial use in terms of:

1. Broadness of feed particle size distribution:
   The STI separator can process material with a particle size from greater than 250 micron to less than 5 micron with no degradation in performance.

2. Electrical conductivity of feed particles:
   The STI separator can contact charge and triboelectrically separate materials irrespective of their electrical conductivity, from good insulators (dielectrics) to good conductors (metals).

3. Low power consumption:
   The STI separator has demonstrated at full scale (25 tons per hour) a power consumption of less than 1 kilowatt hour per ton of feed.

4. Scalability:

   The capacity of the STI separator is a linear function of separator width. A separator has been operated for thousands of tons at rates up to 25 tons per hour.

Laboratory Results

Results from a large number of tests using a several ton per hour laboratory prototype are summarized in Figure 2 where the carbon content of the two product streams is plotted verses the feed carbon content. Many samples of ash from many different sources has been tested with very good results. This includes ash derived from coal from North America, South America, Europe, Africa, Asia and Australia

Commercial Results:

Results from a 3000 ton production test run in summer of 1994 are shown in Figure 3. This figure shows the average LOI for the feed and the two products for each truck load of low-carbon ash produced. The average of the feed, low-carbon and high-carbon over the 3000 tons of ash product were 8.19%, 2.26%, and 35.9% LOI respectively. This represented an ash yield of 83%. All the low carbon ash that was produced was used in concrete projects in the Greater Boston area.

In other tests, thousands of tons of ash from 6 different coal fired units burning several types of coal from North and South America has been processed on the commercial sized unit with good results.

The balance of plant bringing ash to the separator and taking the two products away has recently been upgraded. There is now provision for filling 3 trailers with high and low carbon ash before moving trailers. Production rate has been increased with over 1000 tons being processed in the past 10 days. We expect to get to over 1000 tons of production per week in the next few months as further modifications are done to the balance of plant to increase the feeding capacity.

Combustion tests of the high carbon material have been done, and preliminary indications do not show any difficulty with reburning the high carbon material.

Implications and Conclusions:

1. Carbon level in flyash can be controlled independent of furnace operations. This means that:
- Operators have greater latitude in controlling boiler operations including excess air, heat rate, firing rate, turndown, and supplemental firing with out incurring carbon in ash penalty.
- Harder coals with lower volatile content can be used, allowing greater flexibility of fuel source.
- Reduced need for post combustion NOx treatment.

- Boilers can be operated to reduce NOx emissions still more, generating credits (if applicable).

2. Carbon level in flyash can be reduced with only post combustion processing. This means:
- No plant downtime to install retrofits.
- No boiler modifications.
- No need to derate.
- No need to switch fuels.
- No pulverize retrofit.
- No loss in efficiency.

3. Flyash can be utilized in concrete instead of being placed in landfills. This means:
- No landfill cost.
- Revenues from flyash sales.
- No potential long term liability associated with a landfill.

4. Utility has a "green" environmentally beneficial solution to a current problem.
- Flyash byproduct is recycled, replacing cement, while producing more durable, longer life concrete.
- Reduced cement production means less minerals mined, less CO_2 emitted.
- Per ton of ash used to replace cement, 2 tons less minerals mined, and 1 ton less CO_2 emitted.
- Utility can become one of the largest, if not the largest recycler (on a weight, and percentage basis) in the community.
- Implementing a non-mandated recycling program can promote community good will.

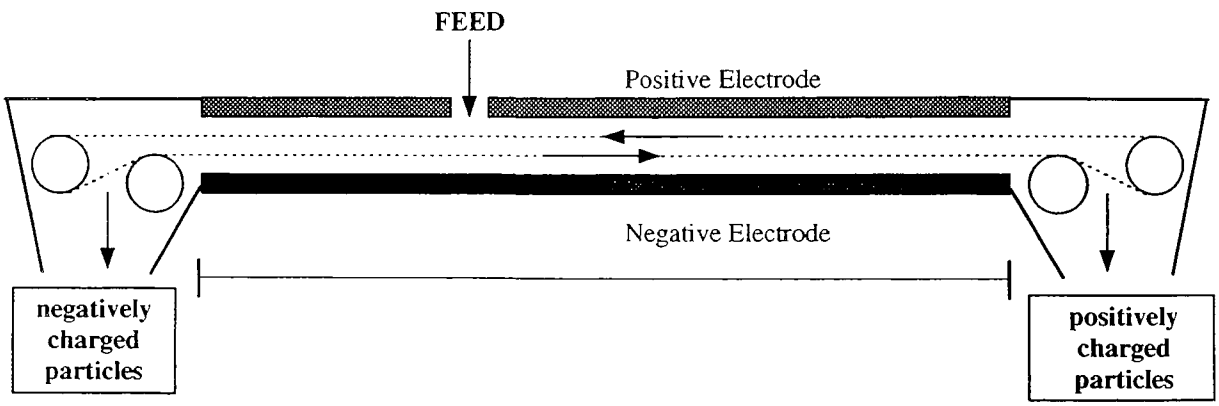
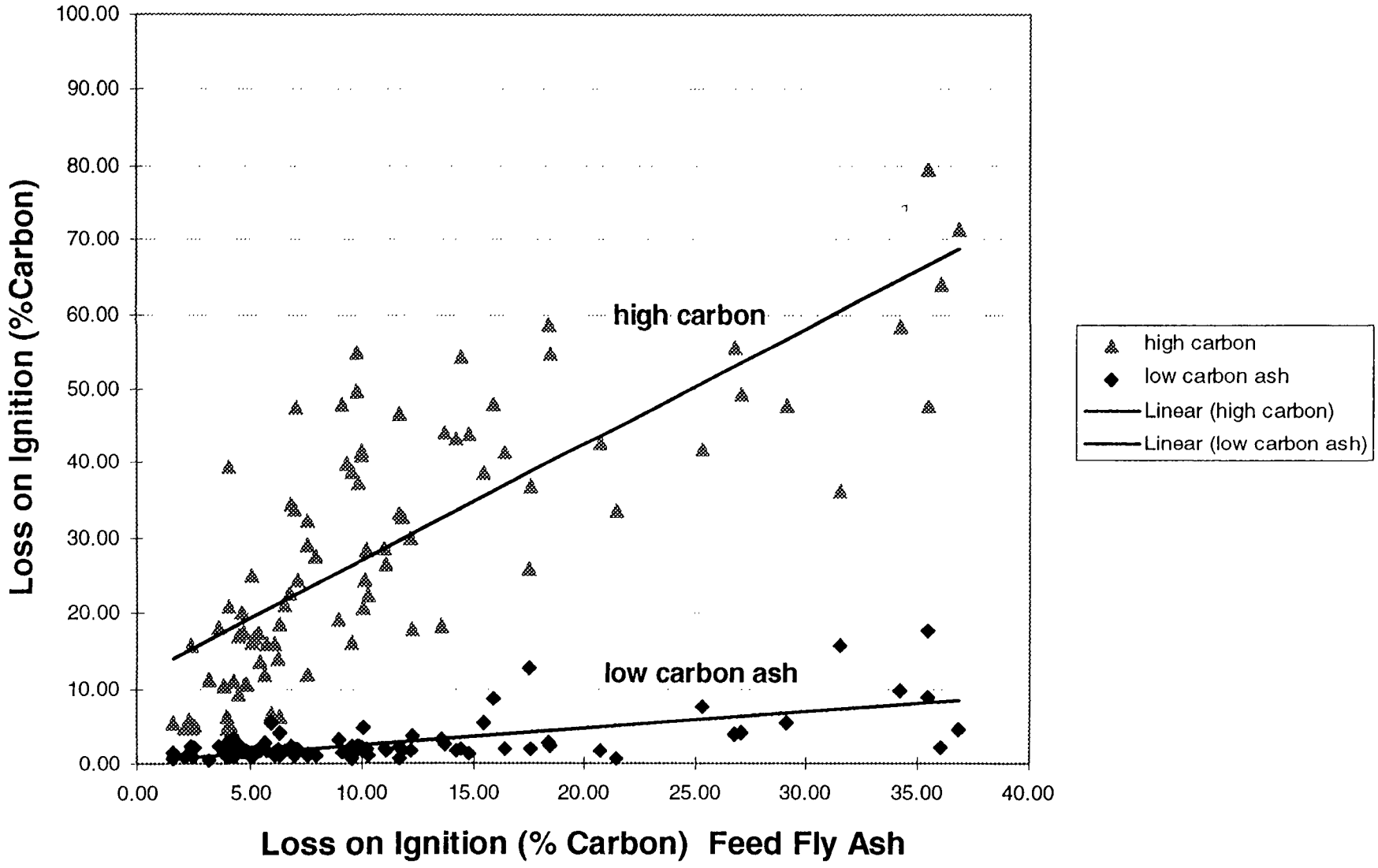


Figure 1
STI Separator

Figure 2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Loss on Ignition (%Carbon)
Loss on Ignition (% Carbon) Feed Fly Ash
high carbon
low carbon ash
high carbon
low carbon ash
Linear (high carbon)
Linear (low carbon ash)

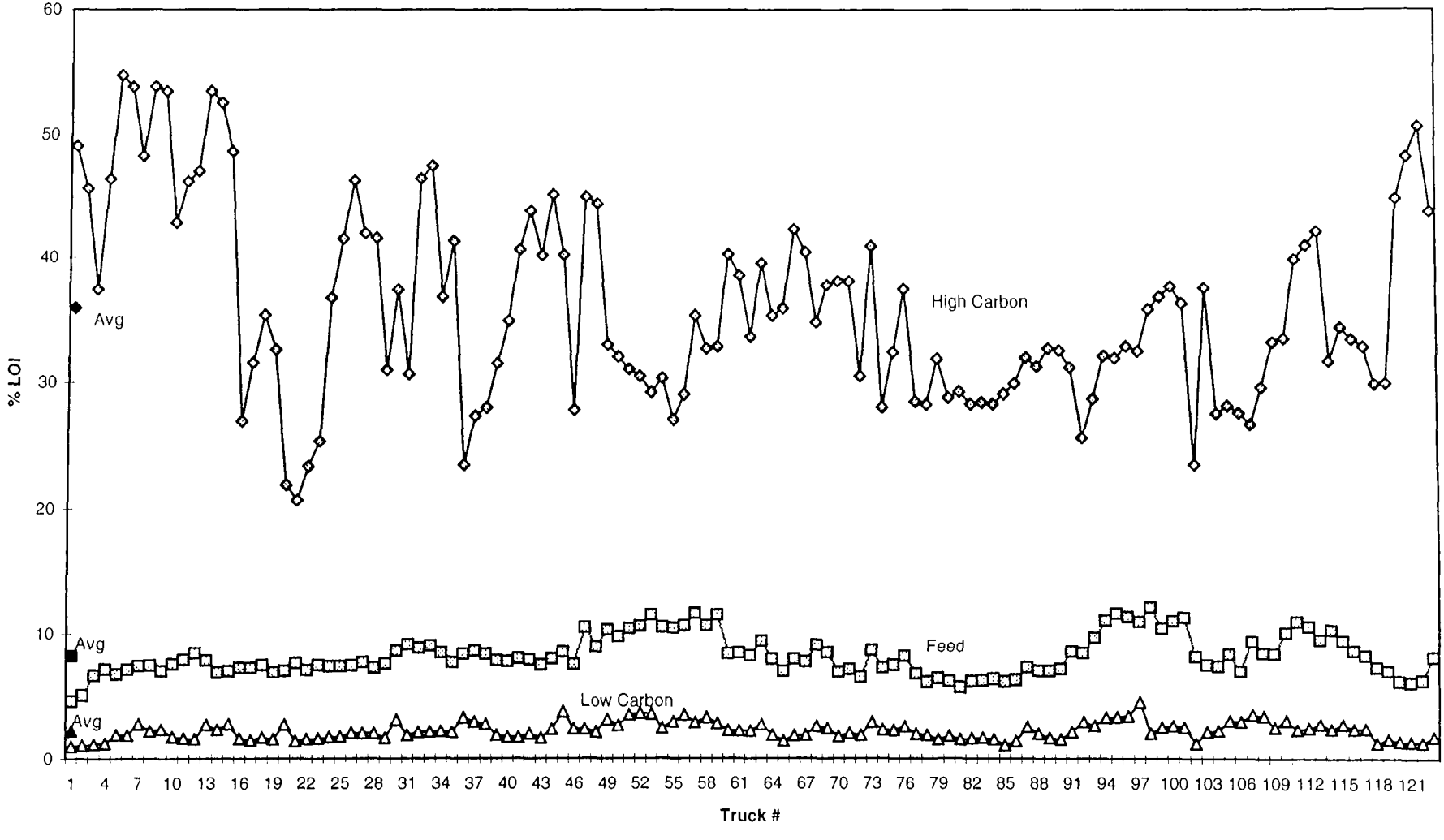

Figure 3
3000 Ton Production Test

NO_x ADVISOR: INTELLIGENT SOFTWARE FOR COMBUSTION OPTIMIZATION

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Abstract

Under a Tailored Collaboration Project funded by PEPCO and EPRI, Lehigh University's Energy Research Center has been developing software for use by plant personnel to optimize a pulverized coal-fired boiler to reduce NO_x emissions and minimize heat rate. Based on expert systems, neural networks and a mathematical optimization algorithm, the NO_x Advisor uses the expert system to safely guide the plant engineer through a series of parametric boiler tests, gathering a database to characterize the operation of the boiler over a wide range of conditions. The neural network develops non-linear mappings between NO_x, heat rate and the controllable parameters. These are then used by the mathematical optimization algorithm to identify optimal operating conditions. The Advisor can be used to adjust a boiler to reach minimum NO_x or identify the conditions that give minimum heat rate subject to a target NO_x value. The Advisor operates off-line on a PC platform, gathering data automatically from the plant data highway and/or manually as needed. Results of a field trial at PEPCO's Potomac River Station are described.

Introduction

Studies performed over the last few years on pulverized coal boilers show that NO_x emission levels can usually be reduced rather substantially through adjustments to the boiler operating parameters. For example, in the case of a corner-fired boiler with conventional burners, NO_x is sensitive to furnace O_2 level, mill loading patterns, burner tilt angle, primary air velocity, and fuel-air and auxiliary-air damper settings. There can be undesirable side effects of reducing NO_x. Generally as the NO_x is reduced, unburned carbon increases, and for some boiler designs steam temperatures drop. This results in increases in heat rate. For units with electrostatic precipitators, increases in unburned carbon can lead to opacity problems. Furthermore, changes in burner settings must be made carefully to avoid unsafe operating conditions or conditions that lead to maintenance problems. Because of the complex relationships between the controllable parameters and the effects of these on emissions, plant performance, plant safety and maintenance, the task of identifying the best set of operating conditions at any particular load level can be difficult. With funding from the Electric Power Research Institute (EPRI) and the Potomac Electric Power Company (PEPCO), Lehigh University's Energy Research Center is developing software, referred to as the NO_x Advisor, for assisting plant personnel in optimizing the combustion of pulverized coal boilers to safely reduce NO_x emissions and optimize heat rate.

The NO_x Advisor combines an expert system, neural networks and an optimization algorithm along with a graphical user interface into a single program. It guides the plant engineer safely through a series of parametric boiler tests, collects and models the data, and then determines the optimal operating conditions. The program can be used to find either the minimum NO_x emissions rate, or the minimum heat rate subject to a target NO_x level.

The NO_x Advisor's operating strategy is based on experience gained from extensive testing at PEPCO's power plants. For example, in the case of Potomac River Unit 4, a corner-fired unit with conventional burners, Lehigh and PEPCO investigators reduced NO_x emissions by close to 40 percent using the combustion optimization approach (1). Full load testing at Dickerson Unit 3, which is also a corner-fired unit with conventional burners, showed that NO_x reductions of close to 30 percent were attainable (2). Using a similar approach, the same investigators have been optimizing combustion at Morgantown Unit 2, which is equipped with an ABB-CE LNCFS III low NO_x firing system. The results from the Morgantown project are reported in a separate paper at this conference (3).

The present paper describes a version of the NO_x Advisor that was developed to optimize boilers similar in configuration to those at the Potomac River and Dickerson Generating Stations. These units have tangentially-fired pulverized coal boilers with conventional burners and no overfire air. A more general version of the NO_x Advisor, which can handle corner-fired boilers with either conventional or low NO_x burners, is under development and will be described in a future paper.

Software Description

The NO_x Advisor is written in C++ and runs under Microsoft Windows on an IBM-compatible PC. The main features of the program are an expert system to direct the parametric boiler testing,

neural networks to model the data, an optimization algorithm to determine the best operating conditions, and a graphical user interface. These components are shown schematically in Figure 1.

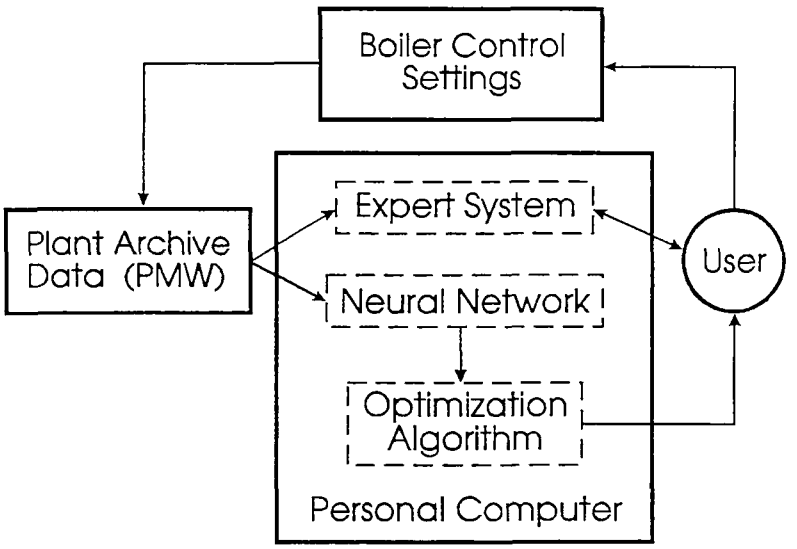

Figure 1
Schematic of the NO_x Advisor.

Expert System

The expert system is used to guide the plant engineer safely through a series of parametric boiler tests. Expert systems are rule-based, incorporating engineering knowledge gained from past experience. An example of a rule is "to reduce NO_x, reduce economizer O_2." Such a system allows the program to react to changing conditions in much the same way as would an engineer experienced in combustion optimization testing. The results of each test are analyzed and used to recommend the settings for the next test. By following this procedure, the expert system is able to minimize the total number of tests while still accumulating sufficient data for the neural network/optimization routines. The expert system was programmed using CLIPS version 6, an expert system development tool created by the NASA Software Technology Branch at the Lyndon B. Johnson Space Center (4).

Neural Networks and Optimization

After the testing is complete, the software analyzes the data using neural networks. The resulting model is then passed to an optimization algorithm that finds the best operating conditions according to the user's desired objective.

Neural networks are a class of mathematical algorithms that simulate the operation of biological neurons. The neural network learns the relationships between the operating conditions and emission and performance parameters by processing the test data. In the training process, a neural network develops a complex non-linear function that maps the system inputs to the corresponding outputs. This function is then passed to a mathematical minimization algorithm that finds the optimal operating conditions.

The neural networks are created and trained using software referred to as Designer Pack by NeuralWare (5). The optimization algorithm is based on the Nelder and Meade Downhill Simplex Method (6).

User Interface

The user interface ties together the expert system, neural networks, optimization algorithm, and data communications into a single program. The expert system and neural network software are compiled as dynamic link libraries that are used as needed. All communication between the user and the program is provided for by the interface through a series of self-explanatory dialog screens. The key advantage of this approach to software development is that the expert system, neural networks and optimization algorithm are invisible to the user. Therefore, the user needs no special knowledge to use these advanced technologies.

The interface also provides the NO_x Advisor with the ability to acquire data automatically from the plant's data archives. The Potomac River Generating Station uses EPRI's Plant Monitoring Workstation (PMW) on a micro-VAX computer to record the plant data. To enable the NO_x Advisor to be tested at Potomac River with automatic data retrieval, a data link was developed specifically for communicating with PMW through a modem. Similar data links will be created in the future to allow the NO_x Advisor to access other types of data archiving systems. The interface also allows for the manual input of test data for those parameters not available on the data highway.

The data collected from each test are stored in a database that is accessible to the user for viewing tests results and is also used to provide a graphing capability. The user is able to plot the test results at any time during the testing process. Single or multiple parameters can be displayed as a function of the test number or any other parameter.

Program Operation

When initiating a test program with the NO_x Advisor, the user is asked a series of questions concerning unit characteristics and the nature of the planned tests. For example, the user will need to specify the load range to be tested and the objective of the optimization: to minimize the NO_x emissions rate or to minimize heat rate subject to a target NO_x emissions rate.

The NO_x Advisor then proceeds through a series of parametric tests, recommending operating conditions important to NO_x reduction and unit performance. Based on previous testing experience with conventional corner-fired boilers, these parameters for full-load are economizer

O_2 level, burner tilt angle, mill loading pattern, and secondary air damper positions. At the end of each portion of the parametric test series, the expert system estimates the best setting for the particular parameter being tested. This new setting is then used as the starting point for the next test series. Using this method, the NO_x emissions rate is gradually reduced as the control settings are moved towards an optimum. This approach also leads to the collection of more test data near the optimal conditions, allowing for better predictions by the neural networks.

Upon completion of parametric testing, the test data are passed to the neural network routine for modeling. Various neural network configurations are evaluated and the most accurate models are saved. The functions generated by these networks are then used by the optimization algorithm to calculate the optimal operating conditions. The solution is checked to ensure it is realistic and that none of the boiler safety limits would be violated. Once the solution is verified, the final set of optimal conditions are displayed for the user.

Results of Field Trial

The NO_x Advisor was tested at PEPCO's Potomac River Unit 5, a 108 MW, subcritical, CE tangentially-fired four-corner boiler, with a GE single-reheat turbine. All of the boilers at the Potomac River Station have conventional burners. The testing, conducted at the upper end of the load range, with four mills in operation, was intended to find the minimum heat rate subject to a target NO_x of 0.42 lb/MBtu. A total of 56 tests was run over the course of seven days of testing. For Unit 5, the test parameters were varied in the following sequence:

- Economizer O_2.
- Burner tilt angle.
- Combined economizer O_2 and burner tilt angle.
- Mill loading pattern.
- Auxiliary air damper positions.
- Fuel air damper positions.
- Combined auxiliary air damper positions and burner tilt angle.

For this unit, the NO_x Advisor required 39 input parameters. Half of these parameters were available on-line from PMW and included unit load, main and hot reheat steam temperatures, burner tilt angle, and various mill parameters. The remaining parameters were recorded from instruments in the control room and CEM system. Unburned carbon was entered into the database after a laboratory analysis of the fly ash for loss on ignition (LOI). The net unit heat rate was entered manually into the database, after being calculated with EPRI's HEATRT code (6).

An overall progression of the test procedure is shown in Figure 2, where NO_x and heat rate are plotted as a function of test number. This figure illustrates a key result of the optimization scheme: NO_x decreases as testing proceeds. Vertical dividing lines on the graph indicate when each control parameter was being varied. This is shown as a different way in Figure 3. First, economizer O_2 was varied while the other parameters were held constant. The same process was repeated for burner tilt angle and the other parameters, according to the previously listed testing sequence.

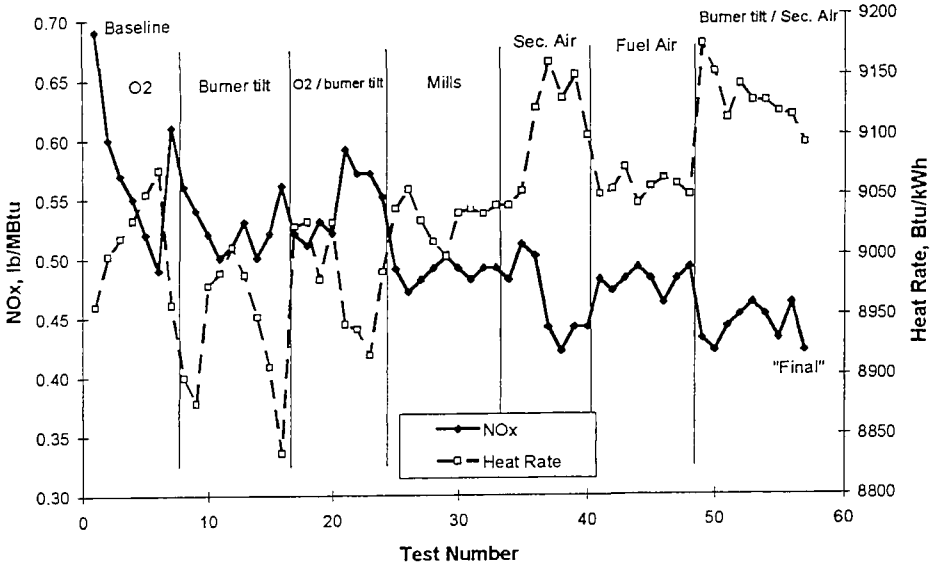

Figure 2
NO_x emissions and heat rate as a function of test number
for Unit 5. Tests were conducted at constant main steam flow rate.

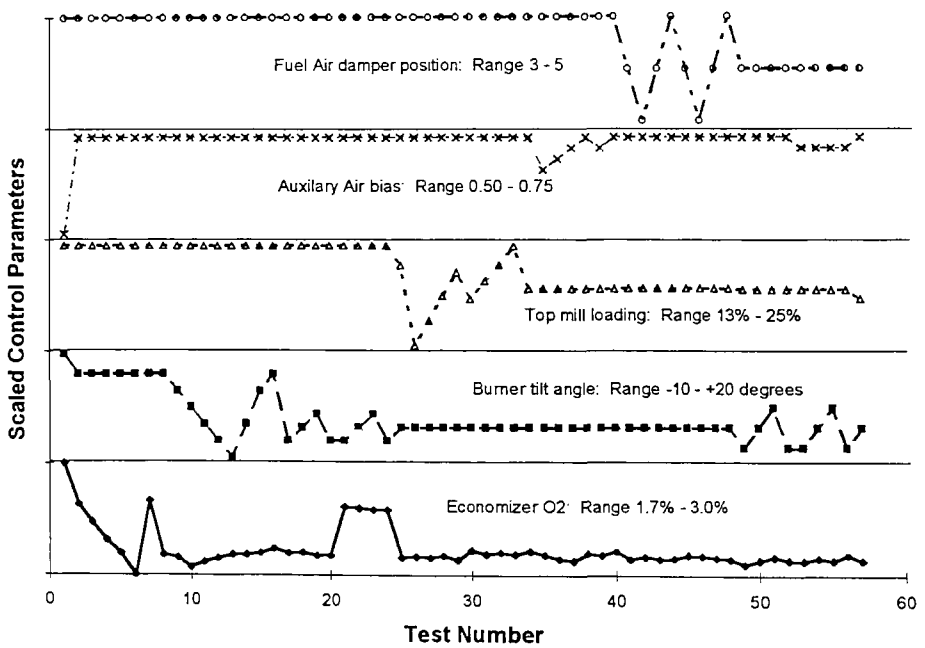

Figure 3
Systematic parametric testing of the control parameters for Unit 5.
Each parameter is scaled for the test range shown on the graph.

In Figures 4 to 7, the detailed variations of each control parameter are shown along with their impact on NO_x and heat rate. A reduction of approximately 1 percent in economizer O_2 resulted in a 0.12 lb/MBtu reduction in NO_x and a 110 Btu/kWh increase in heat rate (Figure 4). Figure 5 shows the results of the parametric tests on burner tilt angle. NO_x displays a parabolic shape with respect to burner tilt angle with a minimum occurring at approximately 0°, while heat rate shows a linear increase as tilt decreases. The increase in heat rate with decreasing tilt angle was due to decreases in main steam and hot reheat steam temperatures. The effect of reducing coal flow rate through the top mill is depicted in Figure 6, where the maximum achievable reduction in flow yielded a 0.02 lb/MBtu reduction in NO_x. Results are given in terms of percentage of total coal flow through the top mill. As the flow is reduced through the top mill, the remainder is evenly distributed among the three lower mills. A value of 25 percent corresponds to even loading of all four mills.

Figure 7 shows the effects of the vertical distribution of air to the auxiliary air registers. The auxiliary air damper bias parameter, a measure of how the air is distributed about the centerline of the boiler, can range in value from -1 to 1. Negative values indicate more air flow to the bottom half of the boiler, 0 is for even distribution, and positive values indicate more air flow to the top half of the boiler. In the range tested, the air flow was biased upwards, producing an overfire air effect, which reduces NO_x.

The data in Figures 4 to 7 illustrate how NO_x and heat rate changed as the testing proceeded, with the control parameters being varied one at a time. However, to use these data to determine the combination of control settings which yields the lowest NO_x or the best heat rate, relationships must be established between NO_x and heat rate and all of the control settings. This is done within the NO_x Advisor using neural networks. The relationships developed by the neural networks are then used to determine the optimal control settings. Figure 8 was created by running the neural network/optimization routine for different values of target NO_x and is a plot of predicted minimum heat rate as a function of target NO_x. This figure can be used to analyze the economic trade-offs associated with operating at a given target NO_x by showing the related heat rate penalty.

Figures 9 to 12 show the recommended control settings required to achieve a specified target NO_x. These figures show that for this boiler, as the target NO_x is reduced, the burner tilt angle is decreased, the top mill is unloaded more, the auxiliary air flow is biased to the upper half of the furnace, and air flow to fuel air dampers is reduced. In all cases the economizer O_2 is driven to the lowest acceptable value, which was found to be 1.9 percent.

Table 1 lists the control settings, NO_x and heat rate for the baseline test, the final test and the neural network/optimization routine predictions. The baseline test was a single test conducted at the normal operating conditions for full load. The settings for the "final test" were selected by the engineer at the end of the testing sequence based on the test data and his institution of what

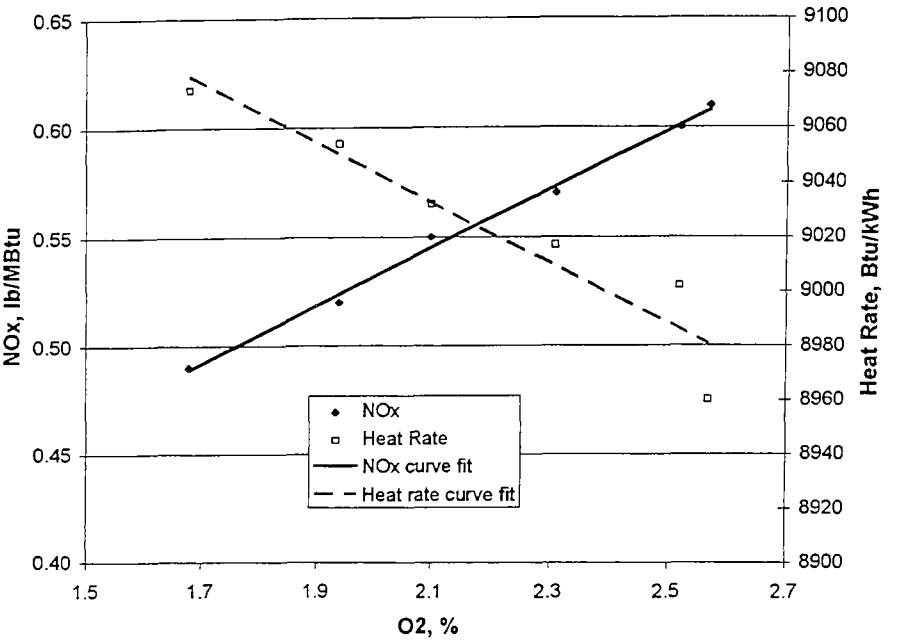

Figure 4
Full-load parametric testing of the economizer O_2 at Unit 5.

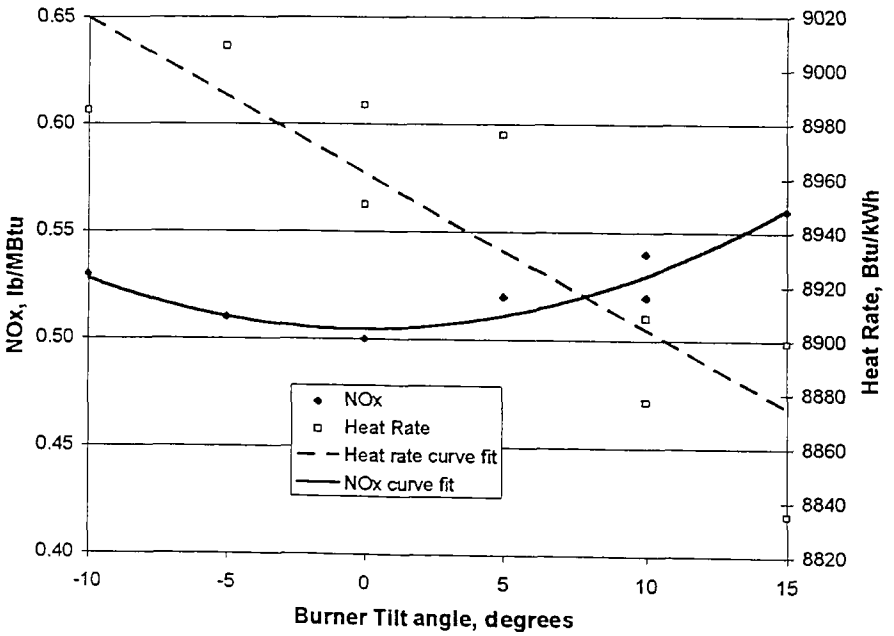

Figure 5
Full-load parametric testing of the burner tilt angle at Unit 5.

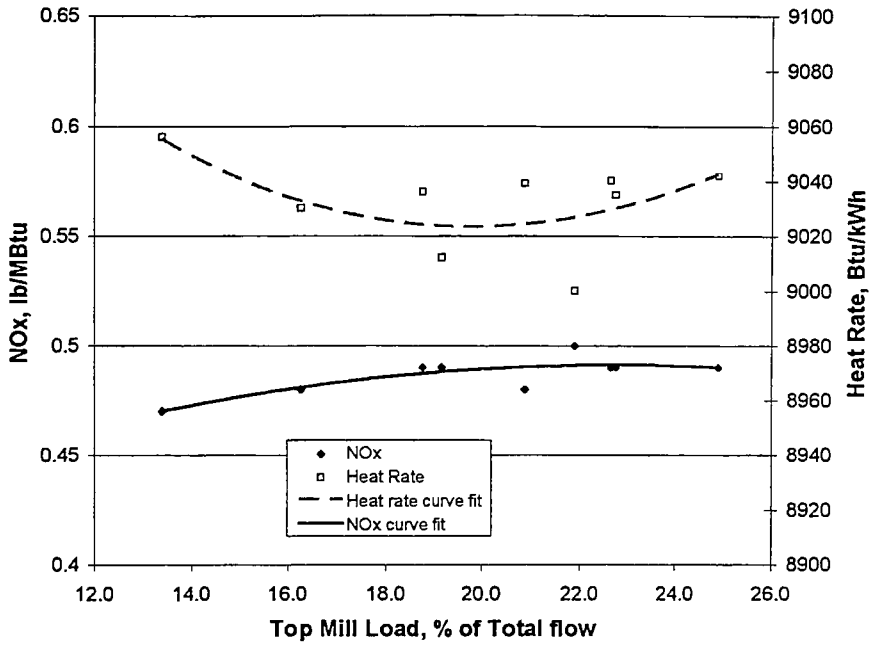

Figure 6
Full-load parametric testing of the mill loading pattern for Unit 5.

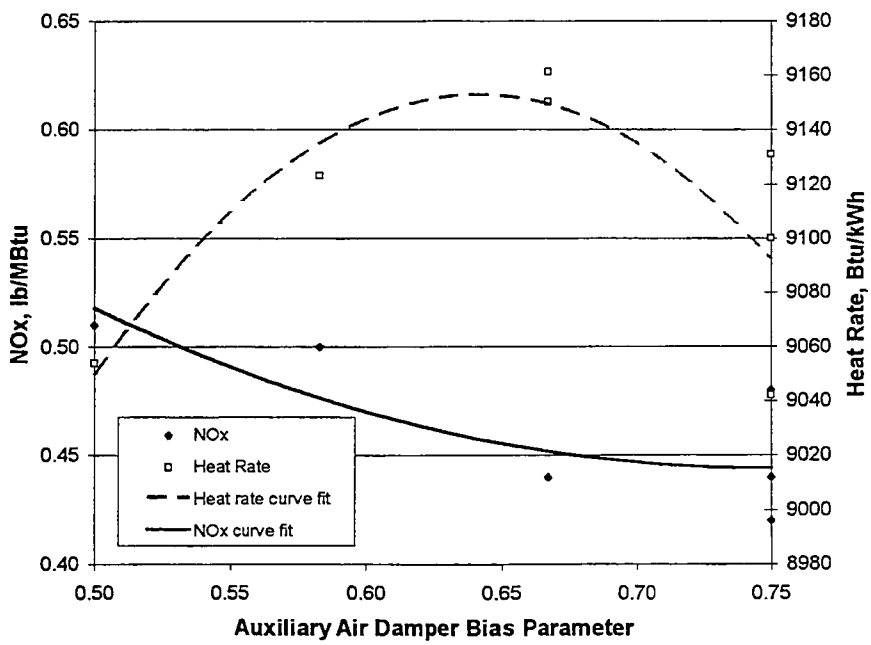

Figure 7
Full-load parametric testing of the vertical distribution of auxiliary air.

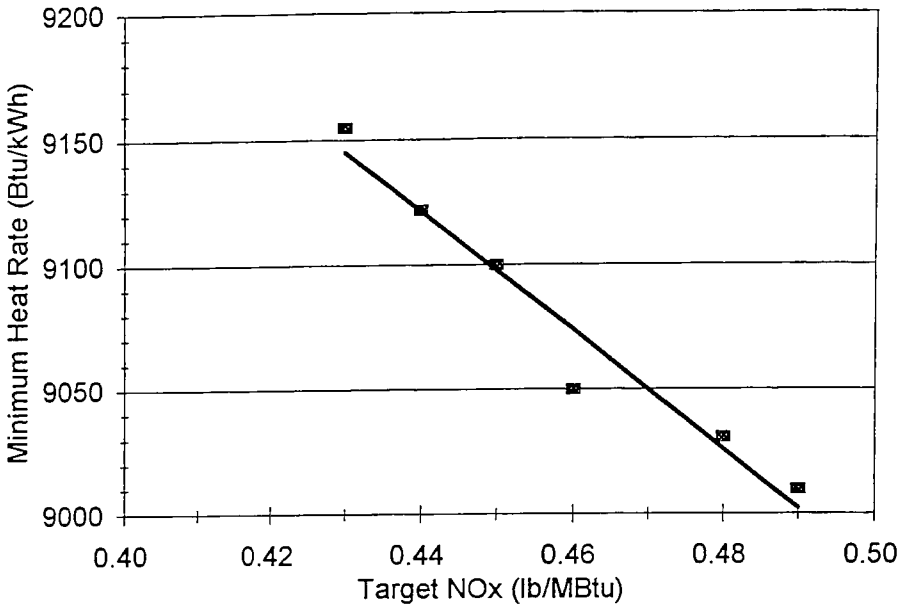

Figure 8
Predicted minimum heat rate as a function of target NO_x.

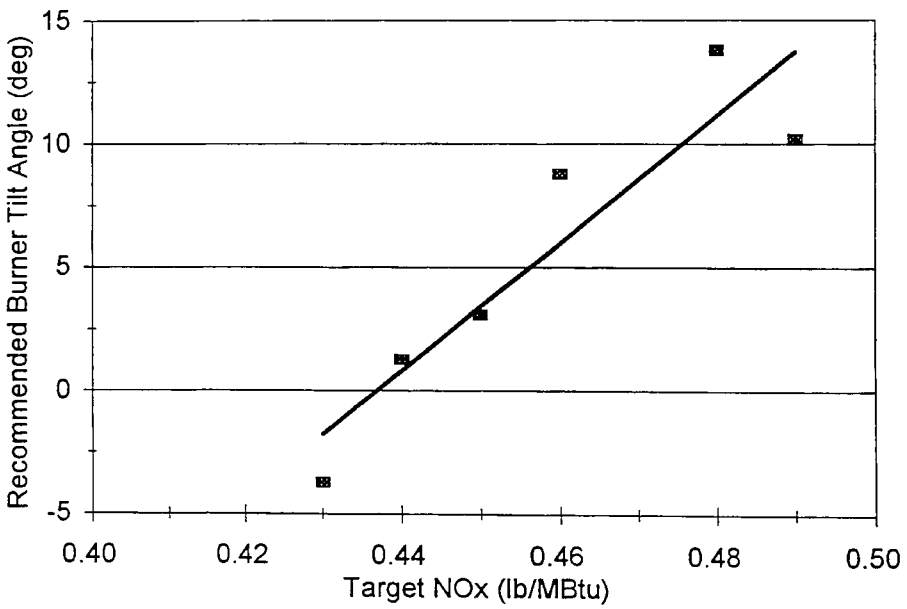

Figure 9
Recommended burner tilt angle to achieve a target NO_x.

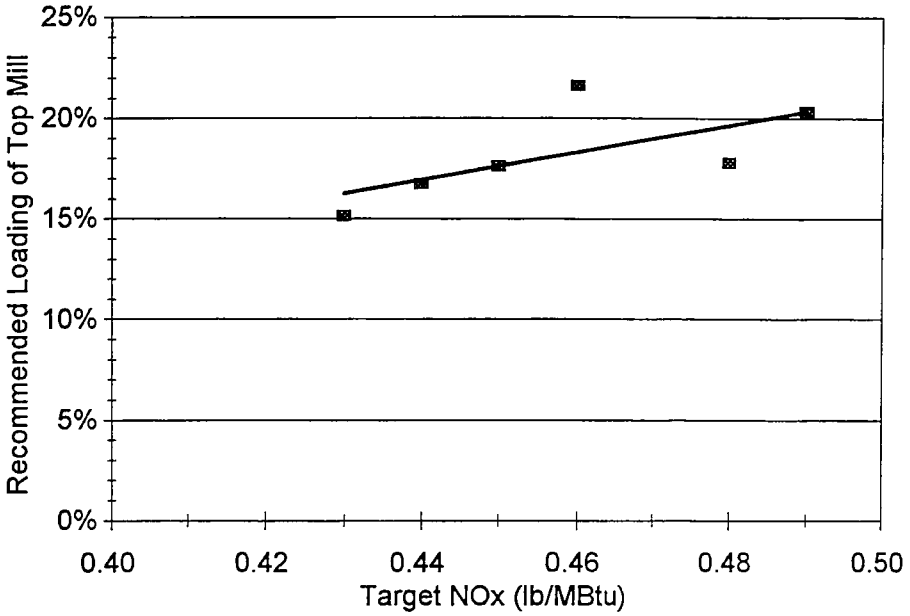

Figure 10
Recommended top mill loading to achieve a target NO_x.

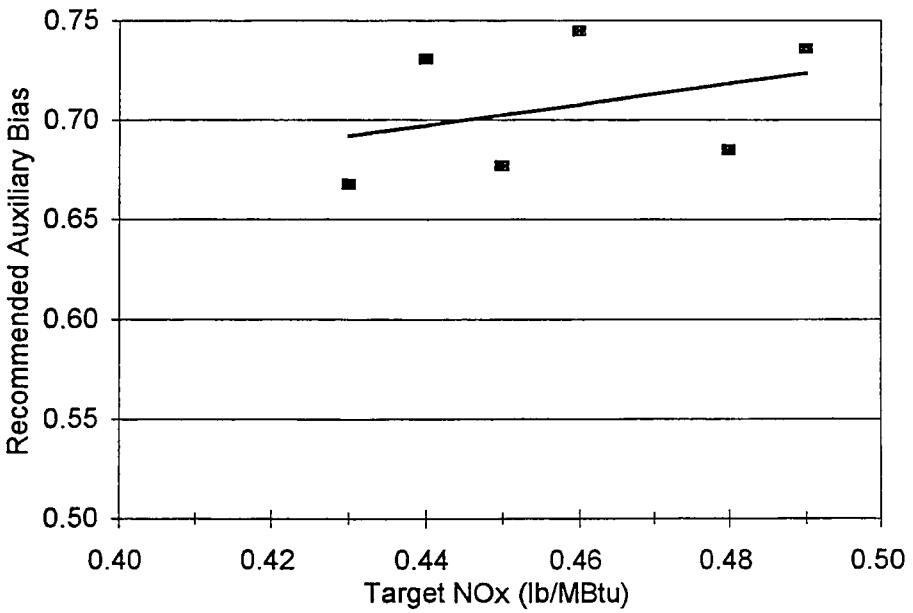

Figure 11
Recommended auxiliary air bias to achieve a target NO_x.

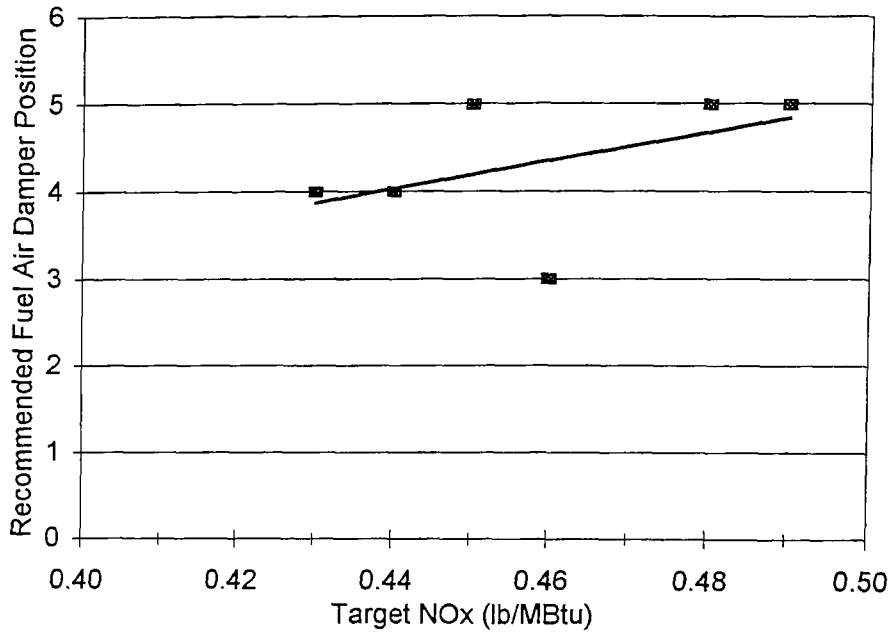

Figure 12
Recommended fuel air damper position to achieve a target NO_x.

the control settings should be for minimum NO_x. The neural network/optimization routine predictions are for the cases of minimum NO_x and minimum heat rate subject to a target NO_x of 0.45 lb/MBtu.

The predicted settings for minimum NO_x are similar to those of the "final test" based on engineering intuition. The NO_x Advisor predicted a 0.27 lb/MBtu reduction from the baseline in the NO_x emissions rate. The unit heat rate was predicted to increase 203 Btu/kWh, due in large part to a drop in steam temperatures. With the testing being performed at constant main steam flow rate, the decrease in steam temperatures also resulted in a 5 MW load reduction.

Comparisons between these predictions and test results obtained at an identical unit (Potomac River Unit 4) show close agreement (see Tables 1 and 2). Further testing is planned at Unit 5 to determine if the recommended control settings shown in Table 1 actually result in the predicted NO_x and heat rate values.

Benefits of the Parametric Testing Approach

The use of a parametric testing approach yields additional benefits besides gathering the test data needed for the neural network. As these tests are conducted, the test engineer can view and plot the data in a variety of ways. This will improve the engineer's understanding of how NO_x and performance vary with boiler settings. By doing this the engineer develops the physical insight needed to overcome constraints on equipment that limit the ability to reduce NO_x or optimize heat

Table 1

Control settings, NO_x Emissions, and Heat Rate
for Baseline and Final Test, and NO_x Advisor Predictions

	Baseline	Final Test	Predicted Min. NO_x	Predicted Min. Heat Rate, NO_x = 0.45 lb/MBtu
Economizer O_2, %	3.02	1.86	1.9	1.9
Burner Tilt Angle, °	20	-1	-3	+3
Top mill load, % of total coal flow*	25	19.0	14.6	17.7
Auxiliary Air**	5-5-5-5-5	5-5-2-2-2	5-5-2-2-2	5-4-4-2-2
Fuel Air**	5-5-5-5	4-4-4-4	4-4-4-4	5-5-5-5
NO_x, lb/MBtu	0.69	0.42	0.42	0.45
Heat rate, Btu/kWh	8959	9093	9162	9100

* Top mill loading is displayed as a % of total coal flow. The remaining coal flow is distributed evenly among the other three mills.

** Damper position listed from top of boiler. 5 indicates fully open and 1 indicates fully closed.

rate. For example, these constraints might include high unburned carbon, low steam temperatures, elevated opacity, and furnace to furnace imbalances. By correcting these deficiencies, the engineer will be able to broaden the operating envelope and achieve lower NO_x and heat rate. This enhanced understanding will also be useful in diagnosing problems as they arise in the future.

Even more important, using systematic parametric tests increases the safety margin by allowing the engineer to identify and avoid unsafe settings. Thus, the possibility of undesirable conditions, such as flame attachment to the burner or an unstable flame, is greatly reduced.

Parametric testing also generates a database covering a wide range of operating conditions. This database can then be used to perform economic trade-off studies on unit operating strategies. Even more important, it provides the knowledge needed to re-optimize the boiler with different conditions and constraints, all without further testing.

Table 2

Control Settings and NO_x
Emissions from Unit 4 Test Results

	Baseline	Min. NO_x	NO_x = 0.45 lb/MBtu with Minimum Heat Rate
Economizer O_2, %	3.1	1.9	2.0
Burner tilt angle, °	19	+7	+14
Top mill load, % of total coal flow*	25	13.8	25
Auxiliary Air**	4-4-3-3-2	5-4-3-2-2	5-4-3-2-2
Fuel Air**	5-5-5-5	3-4-4-4	4-4-4-4
NO_x, lb/MBtu	0.62	0.39	0.45

* Top mill loading is displayed as a % of total coal flow. The remaining coal flow is distributed evenly among the other three mills.

** Damper position listed from top of boiler. 5 indicates fully open and 1 indicates full closed.

Finally, the database allows the NO_x Advisor to be used as a training tool. General correlations giving NO_x and heat rate as functions of control settings can be developed and then used by the Advisor in a training mode. The training mode allows simulation of plant testing for educating engineers and plant operators on NO_x and heat rate reduction techniques.

Summary

The NO_x Advisor, an intelligent software system for combustion optimization has been successfully tested at full load at PEPCO's Potomac River Unit 5. These tests show that the NO_x Advisor is capable of conducting a series of parametric tests, collecting and analyzing the data, and determining realistic optimal points. At full load conditions, 56 tests were run over the course of even days of testing. The minimum NO_x level found by the Advisor was 0.42 lb/MBtu, which is 39 percent below the baseline. These settings increased heat rate by 2.3 percent from the baseline and reduced load by 5 percent due to reductions in steam temperature. The Advisor can also determine the boiler settings required to minimize heat rate, subject to a target NO_x level. For example, at 0.45 lb/MBtu, these yield a 141 Btu/kWh increase in heat rate over the baseline.

The NO_x Advisor is currently being generalized to include a wide variety of tangentially-fired boiler configurations, including boilers with conventional or low NO_x firing systems with various numbers of mills. In addition, planning is underway for a version which can handle wall-fired boilers.

The NO_x Advisor utilizes a parametric testing approach to gather the data needed to perform the optimization. The use of this method yields some important additional benefits to the utility. By having the plant engineer involved in the testing, he/she develops a physical understanding of how NO_x and heat rate depend on the control settings. This can help the engineer overcome constraints on equipment that limit the ability to reduce NO_x or heat rate. It provides the physical insight needed for diagnosing problems which may arise in the future and it increases the safety margin by allowing the engineer to identify and avoid unsafe settings. This approach also generates the database needed for performing economic trade-off studies on unit operating strategies and for re-optimizing the boiler with different conditions and constraints, all without further testing.

Acknowledgments

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Session 6B
Selective Catalytic Reduction

Demonstration of Selective Catalytic Reduction (SCR) Technology for the Control of Nitrogen Oxides (NOx) Emissions from High Sulfur Coal-Fired Utility Boilers at Plant Crist SCR Test Facility

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ABSTRACT

This paper describes the status of the Innovative Clean Coal Technology project to demonstrate SCR technology for reduction of NOx emissions from flue gas of utility boilers burning U.S. high-sulfur coal. The project is sponsored by the U.S. Department of Energy, managed and co-funded by Southern Company Services, Inc. on behalf of the Southern Company, and also co-funded by the Electric Power Research Institute and Ontario Hydro; and is located at Gulf Power Company's Plant Crist Unit 5 (75 MW tangentially-fired boiler burning U.S. coals that have a sulfur content near 3.0%), near Pensacola, Florida. The test program is being conducted for approximately two years to evaluate catalyst deactivation and other SCR operational effects. The SCR test facility has nine reactors: three 2.5 MW (5000 scfm), and six 0.2 MW (400 scfm). Eight reactors operate on high-dust flue gas, while the ninth reactor operates on low-dust flue gas using a slip stream at the exit of the host unit's hot side precipitator. The reactors operate in parallel with commercially available SCR catalysts obtained from vendors throughout the world. Long-term performance testing began in July 1993. A general test facility description and the results from three parametric test sequences and long term test data through December 1994 are presented in this paper.

INTRODUCTION

Selective catalytic reduction (SCR) is a process in which ammonia is added to the flue gas to reduce NOx to nitrogen and water over a catalyst. The need within the utility industry for detailed information on SCR technology has never been greater. The 1990 U.S. Clean Air Act Amendments (CAAA) create two new nitrogen oxide (NOx) control requirements on fossil fuel-fired utility boilers. First, Title IV of the CAAA regarding acid rain addresses NOx emission limits on all coal-fired utility boilers. Second, Title I of the CAAA (attainment of the ambient air

quality standards) calls for certain areas presently not in attainment on ozone to consider NOx controls to achieve attainment. As a result, renewed focus has been placed on advanced NOx control technologies such as SCR, which may be required to meet compliance requirements associated with the CAAA.

SCR technology is in commercial use in Japan and Western Europe on gas-, oil-, and low-sulfur coal-fired power plants. There are now over 36,000 MW of fossil-fuel-fired SCR capacity in Japan, including 6,200 MW on coal. There are over 33,000 MW of fossil-fuel-fired SCR capacity in Western Europe, including 30,500 MW of coal-fired capacity.[1]

SCR DEMONSTRATION GOALS

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

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

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

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

These uncertainties are being explored by constructing and operating a series of small-scale SCR reactors and simultaneously exposing different commercially available SCR catalysts to flue gas derived from the combustion of high-sulfur U.S. coal. First, SCR catalyst performance is being evaluated for two years under realistic operating conditions found in U.S. pulverized-coal-fired utility boilers. Deactivation rates for the catalysts exposed to flue gas of high-sulfur U.S. coal are being documented to determine catalyst life and associated process economics. Second, parametric tests are being performed during which SCR operating conditions are being adjusted above and below design values to observe deNOx performance and ammonia slip. The performance of air preheaters installed downstream of the larger SCR reactors will also be observed to evaluate the effects of SCR operating conditions on heat transfer and boiler efficiency. Third, Honeycomb- and plate-type SCR catalysts of various commercial compositions from the U.S., Japan, and Europe are being evaluated. Tests with these catalysts will expand knowledge of the performance of SCR catalysts under U.S. utility operating conditions with high-sulfur coal.

The intent of this project is to demonstrate commercial catalyst performance and to determine optimum operating conditions and catalyst life for the SCR process. This project will also

demonstrate the technical and economic viability of SCR while reducing NOx emissions by at least 80% and maintaining acceptable ammonia slip (5 ppm).

SCR TEST FACILITY DESCRIPTION

The SCR demonstration facility is located at Gulf Power Company's Plant Crist in Pensacola, Florida. The facility treats a flue gas slip-stream from Unit 5, a commercially operating 75-MW unit, firing U.S. coals with a sulfur content near 3.0%. Unit 5 is a tangentially-fired, dry bottom boiler with hot- and cold-side electrostatic precipitators (ESPs) for particulate control. The SCR test facility consists of nine reactors operating in parallel for side-by-side comparisons of commercially available SCR catalysts obtained from vendors throughout the world. With all reactors in operation, the amount of combustion flue gas that can be treated is 17,400 scfm or 12% of Unit 5's capacity (about 8.7 MWe). Table 1 shows typical pilot facility flue gas constituent concentrations and particulate loadings.

The process flow diagram for the SCR test facility is shown in Figure 1. There are three large SCR reactors (2.5 MW, 5000 scfm) and six smaller SCR reactors (0.2 MW, 400 scfm). Eight of the nine reactors operate with flue gas containing full particulate loading (high dust) extracted from the inlet duct of the hot-side ESP, while one small reactor uses flue gas fed from the ESP outlet (low dust). Only eight reactors are now being operated, one high-dust small reactor is idle.

Each reactor train has electric duct heaters to control the temperature of the flue gas entering the reactor and a venturi flow meter to measure the flue gas flow. An economizer bypass line to the SCR test facility maintains a minimum temperature of 620 °F for flue gas supplied to the test facility. Anhydrous ammonia is independently metered to a stream of heated dilution air that injects the ammonia via nozzles into the flue gas stream upstream of each SCR reactor. The flue gas and ammonia pass through the SCR reactors, which have the capacity to contain up to four catalyst layers.

For the large reactor trains, the flue gas exits the reactor and enters a pilot-scale air preheater (APH). The APHs are incorporated in the project to evaluate the effects of SCR reaction chemistry on APH deposit formation and the effects of the deposits on APH performance and operations. All reactor trains, except the low-dust train, have a cyclone downstream of the SCR reactor to protect the induced draft (ID) fans from particulates. The exhaust for all the SCR reactors is combined into a single manifold and reinjected into the host boiler's flue gas stream ahead of the cold-side ESP. The preheated air from the APH on the large reactors is also combined into a single manifold and returned to the host boiler draft system at the air outlet of the existing APH. All of the particulate that is removed from the flue gas with the cyclones is combined and sent to an ash disposal area.

AIR PREHEATER TESTING PLANS

Air preheater tests include a number of performance measurements aimed at determining the effects of SCR installations on downstream air preheaters. These tests include pressure drop tracking, sootblowing efficiency tests, general heat transfer analysis, washing efficiency and requirements, corrosion analyses, and basket weight loss. By nature, these air preheater tests are long-term. Included in the project final report will be details of expected washing requirements, sootblowing requirements and basket replacement requirements for the operation of air preheaters in conjunction with SCR installations.

CATALYST TESTING PLANS

Six catalyst suppliers are participating in this project, providing eight different catalysts. The suppliers, corresponding reactor size, and catalyst configuration are listed in Table 2. The two suppliers from Europe and two from Japan provide one catalyst each. The two U.S. firms are supplying four of the catalysts. The catalysts being evaluated represent the wide variety of SCR catalysts being offered commercially and possess different chemical compositions and physical shapes. Of these eight catalysts, five have a honeycomb geometry while the remaining three are plate-type catalysts.

After start-up, the baseline performance of each catalyst was determined at design conditions which are being maintained for the two year test period. Once baseline performance was established, each reactor was sequenced through a series of parametric tests that varied the following variables around the SCR process baseline point: NH_3/NOx ratio, temperature, and space velocity. Space velocity is the ratio of flue gas volumetric flow rate to catalyst volume. With a fixed catalyst volume, variations in flue gas flow rates alter the space velocity around the design point.

DeNOx efficiency, pressure drop, SO_2 oxidation, and ammonia slip are determined at specific parametric test conditions. After each parametric test matrix has been completed, each reactor is returned to baseline conditions. This allows for steady-state operation to age the catalyst between parametric tests. The parametric test matrix is repeated every four months for each reactor train. The operating parameter ranges examined during the parametric tests and the long-term design conditions (baseline) are as follows:

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

PROJECT SCHEDULE AND STATUS

The demonstration project was organized into three phases. Phase I consisted of permitting, preparation of the Environmental Monitoring Plan, and preliminary engineering. Phase II included detailed design engineering, construction, and start-up/shakedown. Detailed design engineering began in early 1991 and concluded in December, 1992. Construction began at the end of March 1992 and was completed by the end of February 1993. Start-up/shakedown concluded in June 1993. Baseline commissioning tests without catalysts were conducted through June. The loading of all catalysts was completed at the end of June.

The operational phase for process evaluation, Phase III, commenced in July 1993. The process evaluation will last for approximately two years and will be followed by preparation of a final report, which will include catalyst technical performance and process economic projections. The major milestones on the schedule are shown in Table 3.

PARAMETRIC AND LONG TERM TEST RESULTS

The parametric and long term catalyst testing can be divided into four main categories: 1) test facility ammonia measurements, 2) test facility sulfur trioxide measurements, 3) test facility general reactor performance measurements, and 4) catalyst supplier laboratory tests. The ammonia measurements consist of measurements of intermediate ammonia (usually downstream of the first catalyst bed), and slip ammonia measurements (at the reactor exit). These ammonia measurements are used to assess the deNOx performance of the catalyst. Sulfur trioxide measurements are performed at the inlet and outlet of the reactor to determine the SO_2 oxidation characteristics of the catalyst. Other parameters are also measured to determine reactor performance. These parameters include NH_3/NOx distributions at the reactor inlet, particulate and velocity distributions within the reactor, and pressure drops across the catalyst beds. Also

included in the parametric testing are measures of HCl and N_2O concentrations at the inlets and outlets of the reactors.

Ammonia Measurements

Of particular concern in SCR operation is the relative decline of deNOx activity over time. This decline is often indicated in the field by increasing ammonia slip exiting the reactor over time. Figure 2 shows the measured ammonia slip for each catalyst as a function of time exposed to flue gas. The data show a slight upward trend in ammonia slip for the 7,000 hour operational period. In most cases however, this upward trend represents less than 1 ppm increase in ammonia slip. Some uncertainty exists in the exact evaluation of ammonia slip, since most measurements are near the ammonia detection limit and measurement accuracy is relatively poor ($\pm$ 0.3 ppm) at these very low ammonia concentrations. In addition, the plot assumes that the NH_3/NOx ratio is constant. In reality, slight variation exist in the NH_3/NOx ratio which contribute to the scatter of the data. This is especially true of early slip measurements where the NH_3/NOx ratio was consistently high due to calibration errors. The most important conclusion to be drawn from the figure is that ammonia slip has not significantly increased over the 7,000 hour operational period.

In addition to the evaluation of baseline slip over time, other important characteristics such as the catalyst response to changes in flow rate (space velocity) and temperature are also examined using intermediate and ammonia slip measurements. These responses can be measured in a variety of ways but are typically measured in the test facility as changes in first bed NOx reduction or changes in ammonia slip versus variations in reactor flow or temperature. The first parametric sequence (preliminary sequence) was an abbreviated parametric test sequence. This test sequence contained a large number of intermediate ammonia measurements with a relatively small number of corresponding ammonia slip measurements. For this reason, the first parametric test results showing flow or temperature dependency on deNOx capability are demonstrated as a function of first bed NOx reduction (calculated from ammonia measurements).

Figure 3 shows the relative first bed NOx reduction as a function of increased reactor flow rate for all catalysts that were operational at that time. The data show a general decline in first bed NOx reduction, as expected. The drop in NOx reduction is mitigated by improvements in mass transfer with increased flow. If no mass transfer limitations were present under the operational conditions, a more severe drop in NOx reduction would probably be noted.

Subsequent parametric test sequences focused more on ammonia slip measurements. Consequently, the effects of flow rate changes on deNOx characteristics are shown as a function of ammonia slip versus flow rate for the second and third parametric sequences. Figure 4 shows the measured ammonia slip versus flow rate using data acquired at a 0.8 NH_3/NOx ratio and 700 °F during the second parametric test sequence. As expected, a general increase in ammonia slip is noted with increasing flow rate. Relatively high slip values for catalyst #1 were measured due to an error in the setting of the NH_3/NOx ratio. For this catalyst, the data represent an NH_3/NOx ratio close to 1.0 rather than the 0.8 NH_3/NOx ratio used for the other catalysts.

Figures 5 through 7 show ammonia slip versus flow rate for three NH_3/NOx ratios (1.0, 0.8, 0.6 NH_3/NOx, respectively) measured during the third parametric test sequence. Figure 5, showing data for the 1.0 NH_3/NOx ratio, is presented to demonstrate a well known effect. The extremely high ammonia slip values at the 1.0 NH_3/NOx ratio indicate that in some cases more ammonia was being injected than there was available NOx to react, thus forcing the reactor to slip the excess ammonia. This high NH_3/NOx ratio tends to control the ammonia slip and masks potential changes in slip due to flow rate variations. Figure 6 and 7, using data at a 0.8 and 0.6 NH_3/NOx ratios, show trends more in keeping with the previously noted results from figure 5 (second parametric test). Again, slight increases in ammonia slip are noted with increasing flow rate.

Figures 8 through 12 show the effects of deNOx capabilities with changes in temperature. As with the flow rate effect data, the first parametric sequence data shows first bed NOx reduction, while subsequent measurements focus on changes in ammonia slip. Figure 8 depicts the relative first bed NOx reduction versus temperature for all catalysts operational at that time. In some cases, fairly significant increases in first bed NOx reduction are noted. Mass transfer limitations tend to inhibit improvements in apparent deNOx activity with increasing temperature. The temperature effect would be nearer to exponential in the absence of mass transfer limitations.

Figure 9 shows ammonia slip versus temperature data for the second parametric data set. Relatively little improvement (decrease) is noted in ammonia slip between the 700 °F and 750 °F conditions. At these temperatures, the kinetic rate has likely increased to a point where mass transfer limitations have become controlling and no significant improvements are noted with increased temperature. This is not the case however, for the 620 °F versus 700 °F case conditions. Over this temperature range, relatively significant improvements in ammonia slip are realized with increasing temperature. This can be interpreted to mean that kinetic rate is a fairly significant portion of the overall reaction rate and improvements in this parameter with increasing temperature result in the realization of ammonia slip improvements.

Figures 10 through 12 show data at 1.0, 0.8 and 0.6 NH_3/NOx ratios, respectively for the third parametric test sequence. Figure 10 shows ammonia slip versus temperature for the 1.0 NH_3/NOx ratio and design flow condition. The plot shows fairly significant improvements in ammonia slip for some catalysts while others show nearly constant ammonia slip across the temperature range. As with the slip versus flow rate data, high values of NH_3/NOx ratio tend to control the ammonia slip rather than the temperature. No definite conclusions should be made as to temperature effect using the plot. Figure 11a shows the effect on ammonia slip with variations in temperature at the 0.8 NH_3/NOx ratio. Relatively high ammonia slip values are noted for catalyst #1 due to erosion problems with the catalyst resulting in some flue gas channeling. Subsequent investigation of the catalyst revealed that a manufacturing problem contributed to weak areas in the catalyst which eroded quickly. The catalyst was replaced at the conclusion of the third parametric test sequence. Figure 11b is identical to Figure 11a, but with the catalytst #1 data excluded and the scale changed. This more clearly shows the temperature effect for the remaining catalysts.

The data in figure 11b show that there is relatively little improvement in ammonia slip with increasing temperature across the entire temperature range. Data for catalyst #7 and catalyst #2 show zero slip at the low and mid temperature conditions. These points are plotted as zero since the actual slip measurement was below the detection limit of roughly 0.5 ppm. Thus, the points for these two catalysts at the high temperature condition should not be construed as a significant increase in ammonia slip with increasing temperature. Figure 12 shows data acquired at a 0.6 NH_3/NOx ratio and design flow. This plot is very similar to Figure 11b, again showing little or no improvement in ammonia slip with increasing temperature.

Another important catalyst response to be considered is the ammonia slip versus NH_3/NOx ratio (i.e. NOx reduction). Information of this type is important because it defines the maximum NOx reduction that can be achieved within a particular maximum ammonia slip limit. Figure 13 shows the ammonia slip from each reactor versus the NH_3/NOx ratio. This plot is typical for this type of application showing very sharp increases in ammonia slip as NH_3/NOx ratio nears 1.0. As previously mentioned, NH_3/NOx ratios greater than 1.0 force the catalyst to slip ammonia, since no NOx is available to react with the excess ammonia. The plot clearly shows this effect at NH_3/NOx ratios above 1.0.

Sulfur Dioxide Oxidation Measurements

Another important reactivity characteristic of SCR catalysts in addition to NOx reduction activity, is their propensity to oxidize sulfur dioxide. This is an important aspect of SCR catalyst since increased SO_3 can exacerbate problems with equipment downstream of the SCR due to increased formation of acidic deposits. SO_2 oxidation is normally considered to be a first order reaction. Thus, in absence of mass transfer limitations, SO_2 oxidation should have a nearly linear relationship to reactor flow rate and an exponential relationship to temperature. SO_2 oxidation is normally considered to be constant with exposure time based on catalyst supplier historical experience. Figure 14 shows the baseline SO_2 oxidation rate as percent of inlet SO_2 oxidized to SO_3 as a function of catalyst exposure time. A fair amount of scatter is present in data, but in general, no significant trends are noted as a function of catalyst exposure time. Nearly all of the data show oxidation rates of 0.75% or less which was the original design specification for maximum allowable SO_2 oxidation.

Flow rate and temperature effects on SO_2 oxidation are shown in Figure 15 and 16 respectively. This data is from the second parametric test sequence. Figure 15 shows the flow rate effect on SO_2 oxidation for 0.8 NH_3/NOx and 750 °F conditions. The catalyst #8 data shows relatively high SO_2 oxidation characteristics (in keeping with suppliers quoted rates) and does show a nearly direct linear effect of flow rate on SO_2 oxidation. However, the other catalysts show little or no effect of flow rate on SO_2 oxidation with a relatively constant SO_2 oxidation rate across the flow rate range.

Figure 16 shows the effect of temperature on SO_2 oxidation rate. This plot shows that SO_2 oxidation rate is a much stronger function of temperature than of flow rate, as expected. The

effect across the temperature range is relatively linear for most catalyst with the exception of catalyst #8 which exhibits a more exponential SO_2 oxidation relationship to temperature.

Measurements of SO_2 oxidation subsequent to the second parametric test sequence were limited to only two or three conditions. The SO_2 oxidation rate for 0.8 NH_3/NOx ratio and standard flow at medium and high temperature was examined during the third parametric test sequence. These data are shown in Figure 17. As with Figure 16, the data show a consistent increase in SO_2 oxidation rate with respect to temperature.

General Reactor Performance Measurements

In addition to catalyst performance, reactor performance in terms of NH_3/NOx distributions, particulate distributions, and velocity distributions are critical to the efficient operation of an SCR unit. These parameters are controlled primarily by reactor design rather than catalyst design. Other reactor performance measurements such as reactor pressure drop and N_2O formation are primarily catalyst specific parameters.

Typically, each of these general reactor performance parameters are measured as part of each parametric test sequence. The following discussions outline the general results of these performance measurements.

Ammonia to NOx Distribution

A balanced NH_3/NOx distribution within an SCR reactor is critical to its efficient operation. Maldistribution of either component can create areas within the reactor where the local NH_3/NOx ratio is very high. This leads to an increase in overall ammonia slip from the reactor and a reduction in deNOx efficiency. Reactor design criteria are set to minimize this problem. In the pilot facility, design criteria are primarily set to maintain a smooth distribution of both ammonia and NOx resulting in an even NH_3/NOx ratio across the reactor. In full scale installations, the ammonia injection grid is tuneable, thus allowing ammonia to be injected in a profile which matches the NOx profile, thereby creating a smooth NH_3/NOx ratio distribution across the reactor. Although the test facility has this tuning capability, in practice most adjustments that have been made were aimed at smoothing the ammonia distribution.

The test facility design criteria requires an ammonia distribution within ± 10% of the average concentration. The NOx distribution within the reactors is normally smooth; therefore this ammonia distribution criteria controls NH_3/NOx ratios within the reactor cross section to within roughly 10% of the average value.

Ammonia distributions are normally measured in the test facility reactors just upstream of the first catalyst beds. Results to date have demonstrated that the ammonia distribution within each reactor is within the ± 10% design criteria originally set.

Velocity Distribution

The flue gas velocity distribution within an SCR unit is important for several reasons. First, severe maldistributions in velocity make tuning of ammonia injection difficult because the flue gas velocity maldistributions have the same effect as NOx maldistributions. Further exacerbating the problem is that velocity maldistributions are often strong functions of reactor flow rate. Thus, unit load changes would require constant adjustment to the ammonia tuning grid when velocity maldistributions are present. Another concern with velocity maldistribution is the possible erosion impact on the SCR catalyst. Localized high velocities can cause premature catalyst erosion in some areas of the catalyst, thereby reducing the overall life of the catalyst installation.

Measurements of test facility flue gas velocity distributions within the reactors have consistently shown very smooth distributions throughout the reactor (roughly less than ± 5%). The distribution measurements have proven to be very valuable in identifying problems with the catalyst such as areas of fouling or channeling which are indicated by a severe localized change in velocity near the problem area.

Particulate Distributions

Particulate distributions are an important parameter in SCR units primarily for the potential erosion problems that can occur with particulate maldistributions. Localized high particulate concentrations can accelerate erosion in particular areas, similar to localized high velocities. As previously mentioned, this increased localized erosion can reduce the overall life of the catalyst installation.

Particulate distributions measured within the test facility, normally at the reactor exits, have consistently been very smooth, within ± 5%.

Reactor Pressure Drops

Reactor pressure drops are controlled primarily by pressure drops across the individual catalyst beds. These catalyst bed pressure drops are dictated, of course, by specific catalyst design. Pressure drop is a major design concern for a utility boiler mainly due to the energy cost involved in overcoming the reactor pressure drop. In SCR retrofit situations, reactor pressure drop becomes critical and often dictates whether or not a fan retrofit or upgrade is required in conjunction with the SCR retrofit.

Long-term catalyst pressure drops are controlled by the original catalyst design and the effectiveness of sootblowing and catalyst cleaning over time. Consistent increases in reactor pressure drop over time indicate that adequate sootblowing/cleaning is not being performed.

Figure 18 shows the pressure drops for each of the SCR reactors as a function of time. These pressure drops represent the summation of pressure drops across all catalyst beds and dummy bed.

N_2O Formation

The possible formation of N_2O across SCR reactors was a concern with early SCR experience in the U.S. Subsequent investigations of the problem revealed that this was a sampling/analysis anomaly rather than the true formation of N_2O across SCR catalysts. Measurements at the test facility have included evaluation of N_2O concentrations to further confirm this finding. All measurements to date indicate that the test facility catalysts do not form N_2O during the course of normal operation. Table 4 shows N_2O concentrations acquired during the second parametric tests. The table shows reactor exit N_2O concentrations for each of the catalysts and gives an average reactor inlet N_2O value as measured in the test facility main inlet plenum (prior to splitting of the stream to the individual reactors).

Laboratory Tests

Several catalyst characteristics are measured in the laboratory to trace catalyst changes versus flue gas exposure time. Samples are normally extracted from the test facility catalysts on a quarterly basis and sent directly to the individual catalyst suppliers for evaluation. These tests normally examine parameters such as deNOx activity, SO_2 oxidation activity, BET surface area, porosity, and chemical composition.

Of particular interest are laboratory performed deNOx activity measurements. These measurements are very important because they typically demonstrate activity declines before they become apparent in test facility tests. The laboratory tests can often identify the source of catalyst deactivation such as chemical poisoning, loss of surface area, and sintering etc. This type of deactivation data is critical to estimating parameters such as total catalyst life. Figure 19 shows relative deNOx activity versus exposure time for three catalysts. This plot is representative of all catalysts in the test program. The deactivation rates shown are in keeping with catalyst supplier historical experience. No severe declines in deNOx activity are noted which is an encouraging finding, since there has been little previous knowledge of catalyst deactivation rates for U.S. coal applications. This deactivation will of course be tracked throughout the test program and will be used to estimate ultimate catalyst life.

ACKNOWLEDGMENT

We would like to thank the Southern Research Institute on-site team for their significant contribution in obtaining the data presented here; Mr. Robert Heaphy, Mr. James Garrett, Mr. Randy Hinton, and Mr. Richard Jacaruso. Sincere thanks also go to Mr. Charles Powell for his management of test facility operations, making data acquisition possible. We would also like to thank Mr. James Gibson of Spectrum Systems, Inc. for his diligent efforts in maintaining the SCR gas analyzer system and for his efforts in assuring data quality. Special thanks go to Mr. Gerald Bandura (SCS Engineering) and Mr. Danny Mabire (ICS) for their constant efforts in maintaining the instrumentation and controls for the test facility as well as maintaining and programming the digital control system. Our sincere thanks also go to Mr. Ken Pathak for his unfailing efforts in maintaining the huge quantities of data acquired from the test facility.

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1. A. L. Baldwin, J. D. Maxwell, U.S. Department of Energy's and Southern Company Services's August 24 -September 1, 1991, Visit to European SCR Catalyst Suppliers, U.S. DOE, Pittsburgh, PA, 1991, p 41-3.

Table 1

Test Facility Inlet Flue Gas Composition

Constituent	ESP Inlet		ESP Outlet	
	84 MW	43 MW	84 MW	43 MW
NOx	325	401	332	Not Available
SO_2 (ppm)	2340	1780	2030	1510
SO_3 (ppm)	32	42	14	20
HCl (ppm)	104	89	115	101
NH_3 (ppm)	<0.4	<0.4	<0.4	<0.4
Particulate (gr/dscf)	3.76	2.43	0.0018	BDL*

* Below detection limits

Table 2

SCR Project Catalyst Suppliers

Catalyst Supplier	Reactor Size	Catalyst Configuration
Nippon Shokubai	Large	Honeycomb
Siemens AG	Large	Plate
W R. Grace	Large	Honeycomb
W R. Grace	Small	Honeycomb
Haldor Topsoe	Small	Plate
Hitachi Zosen	Small	Plate
Cormetech	Small	Honeycomb
Cormetech	Small	Honeycomb (low dust)

Table 3

Project Schedule

Detailed Engineering	1/92 - 12/92
Construction	3/92 - 2/93
Start-up/Shakedown	1/93 - 6/93
Process Evaluation	7/93 - 6/95
Disposition/Final Report	7/95 - 10/95

Table 4

N_2O Concentrations - First Parametric Test

Location (Catalyst, reactor exit)	1	2	3	4	6	7	8
N_2O, ppmv* (dry @ 3% O_2)	1.8	1.8	1.6	1.6	1.2	1.0	1.7

* Average inlet conc. = 2.1 @ main plenum

Figure 1

Prototype SCR Demonstration Facility
Process Flow Diagram

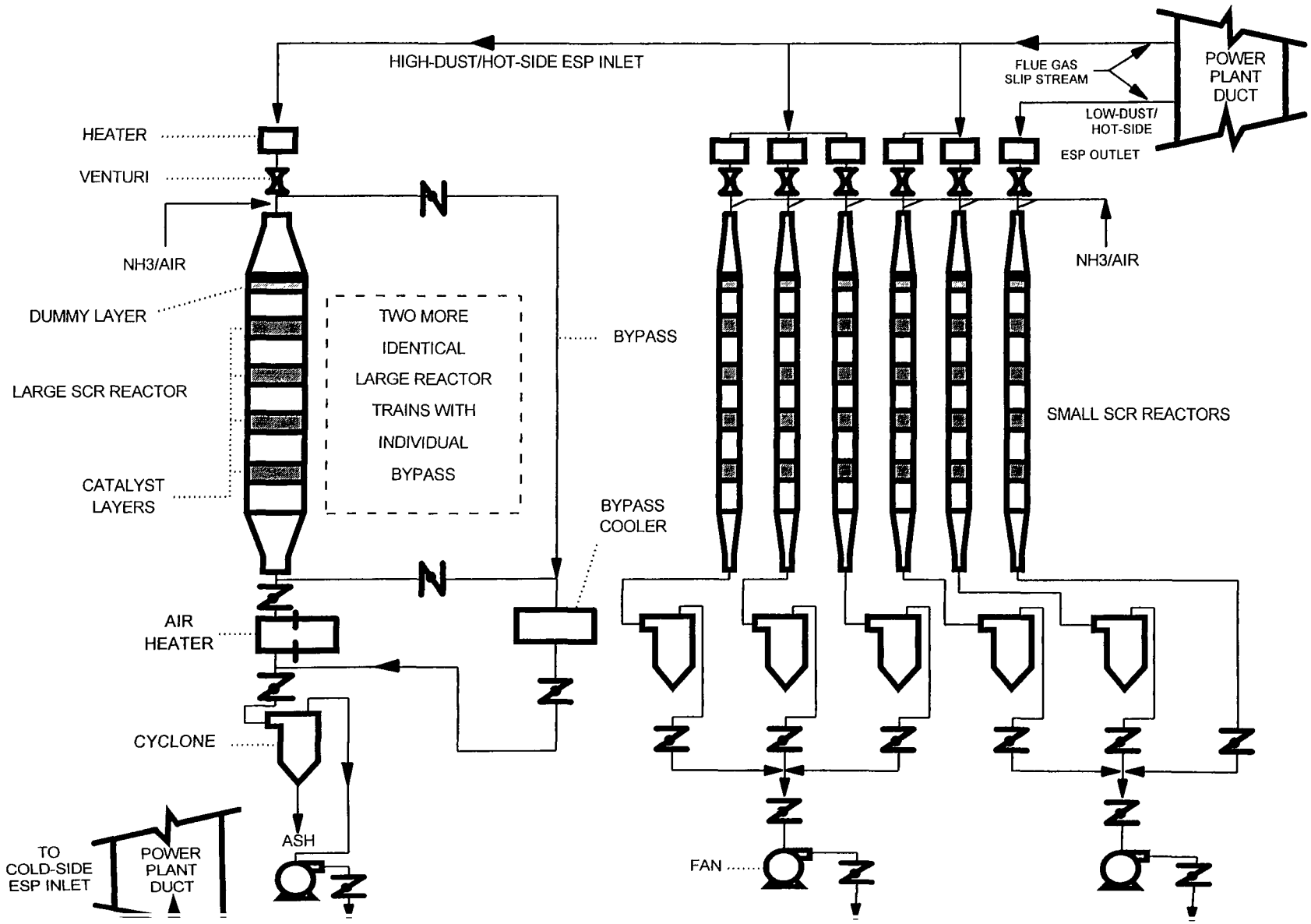
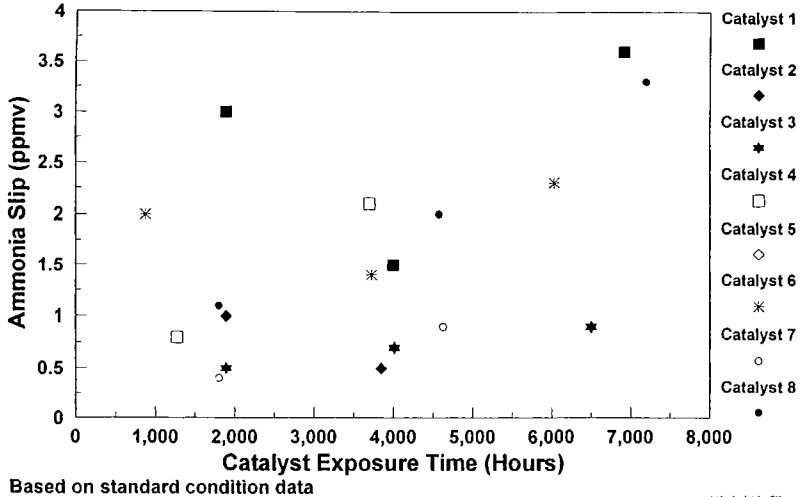


Figure 2

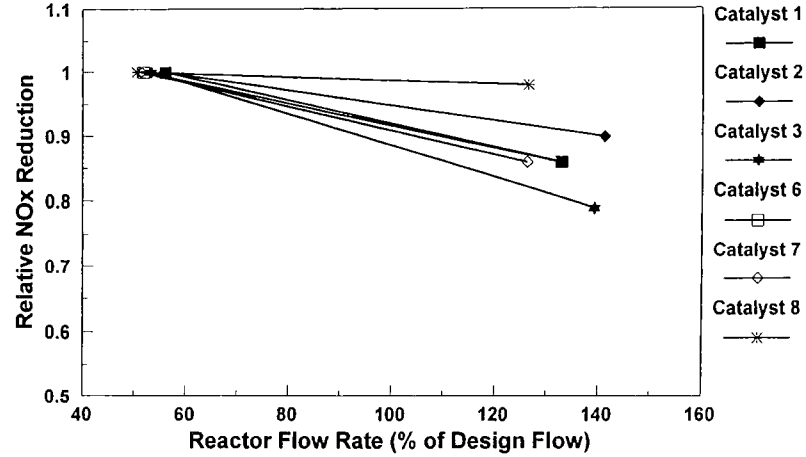
Ammonia Slip vs. Exposure Time



Based on standard condition data

Figure 3

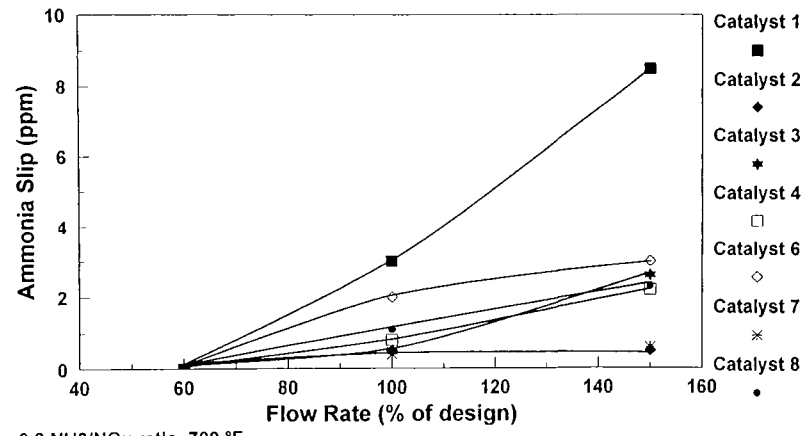
First Bed NOx Reduction vs. Flow Rate



0.6 NH3/NOx ratio, 620 °F
First Parametric Data Set

Figure 4

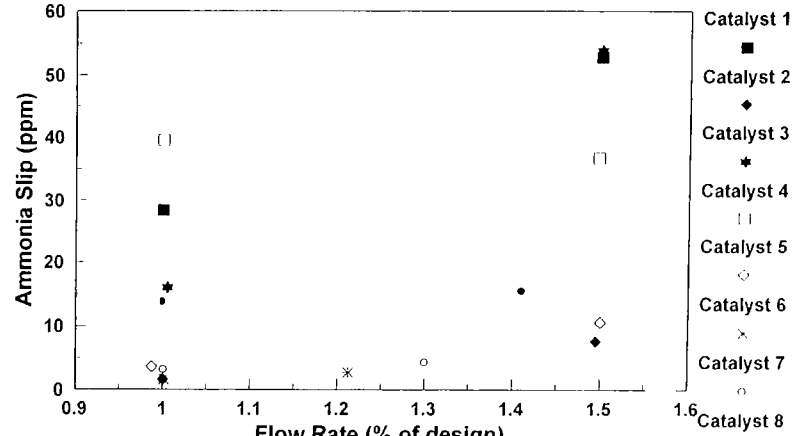
Ammonia Slip vs. Flow Rate



0.8 NH3/NOx ratio, 700 °F
Second parametric data set
Data for Catalyst 1 is for 1.0 NH3/NOx

Figure 5

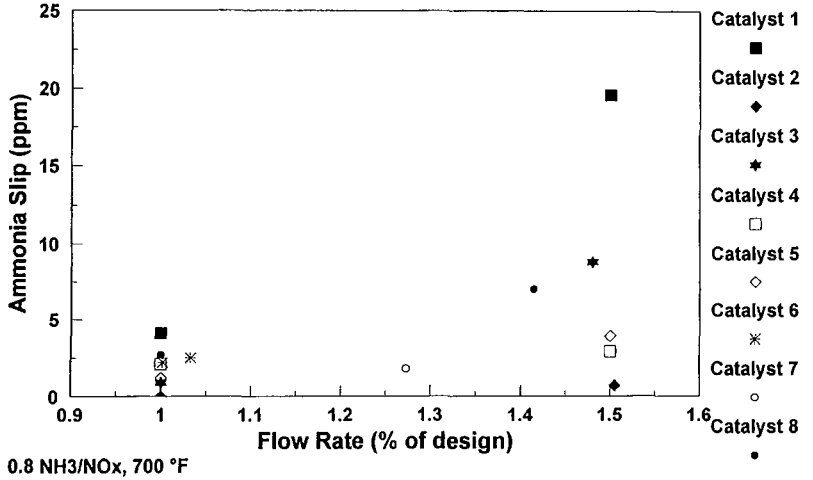
Ammonia Slip vs. Flow Rate



1.0 NH3/NOx, 700 °F
Third parametric data set

Figure 6

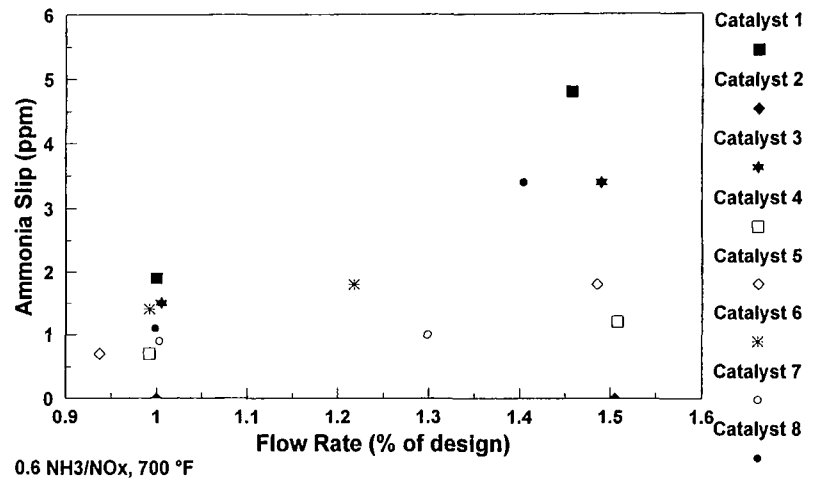
Ammonia Slip vs. Flow Rate



0.8 NH3/NOx, 700 °F
Third parametric data set

Figure 7

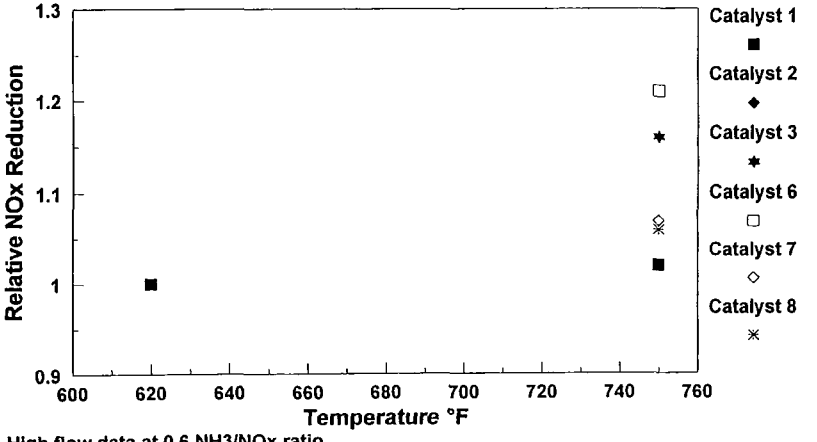
Ammonia Slip vs. Flow Rate



0.6 NH3/NOx, 700 °F
Third parametric data set

Figure 8

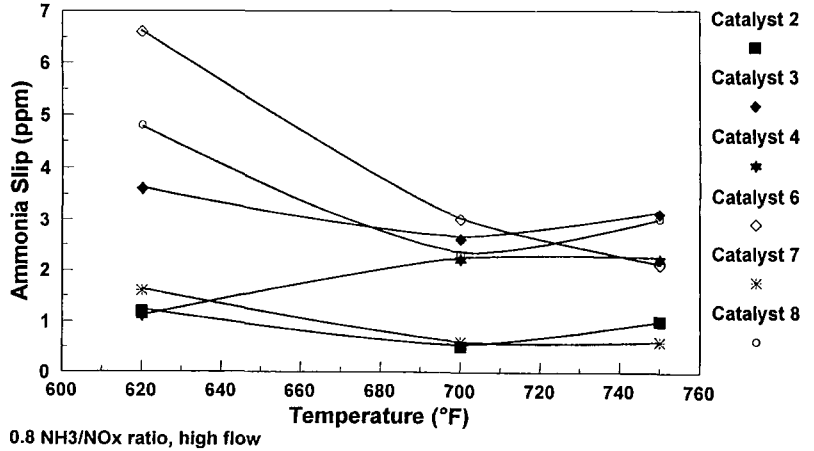
First Bed NOx Reduction vs. Temperature



High flow data at 0.6 NH3/NOx ratio
First parametric data set
Data for Catalyst 3 is for 1.0 NH3/NOx ratio

Figure 9

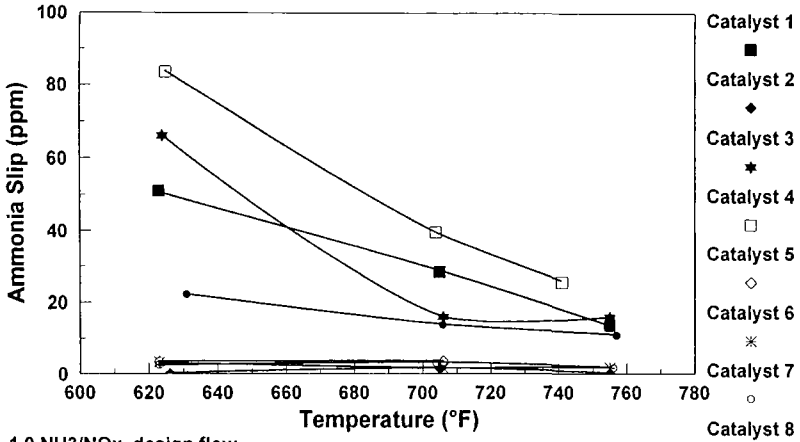
Ammonia Slip vs. Temperature



0.8 NH3/NOx ratio, high flow
Second parametric data set
Catalyst 1 omitted

Figure 10

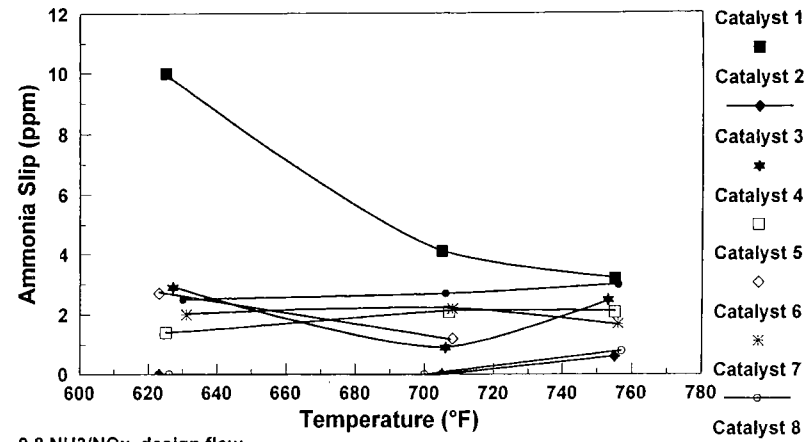
Ammonia Slip vs. Temperature



1.0 NH3/NOx, design flow
Third parametric data set
No high temp Catalyst 5 data available

Figure 11a

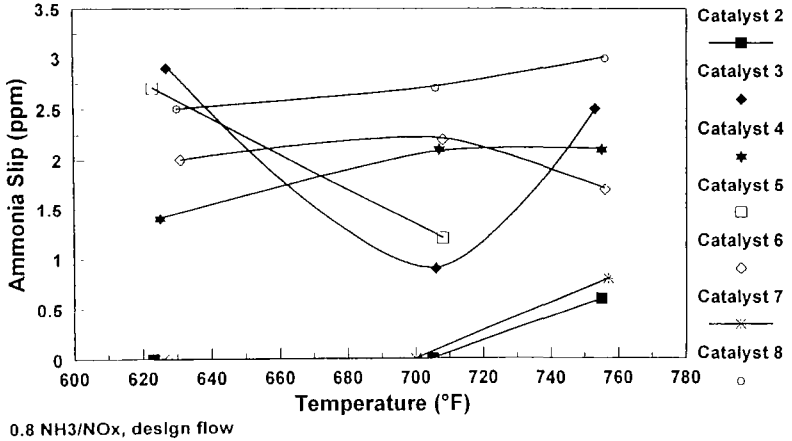
Ammonia Slip vs. Temperature



0.8 NH3/NOx, design flow
Third parametric data set
No high temp Catalyst 5 data available

Figure 11b

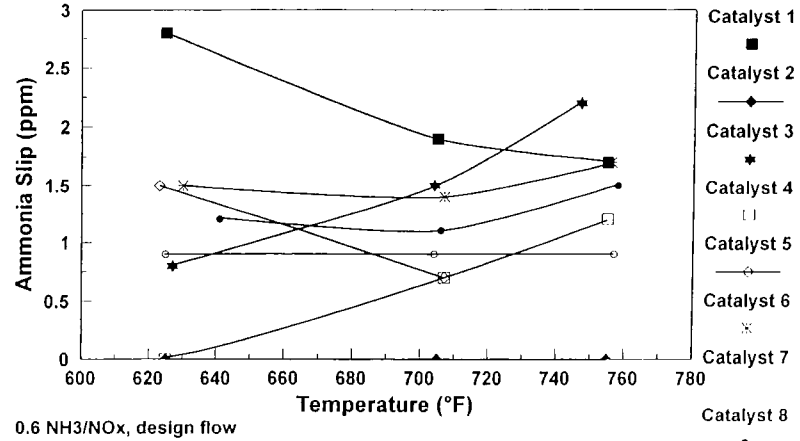
Ammonia Slip vs. Temperature



0.8 NH3/NOx, design flow
Third parametric data set
Data for Catalyst 1 omitted

Figure 12

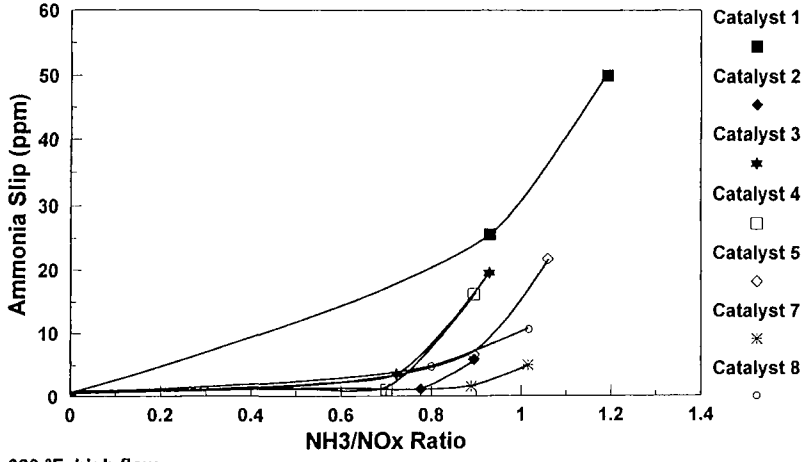
Ammonia Slip vs. Temperature



0.6 NH3/NOx, design flow
Third parametric data set
No high temp Catalyst 5 data available

Figure 13

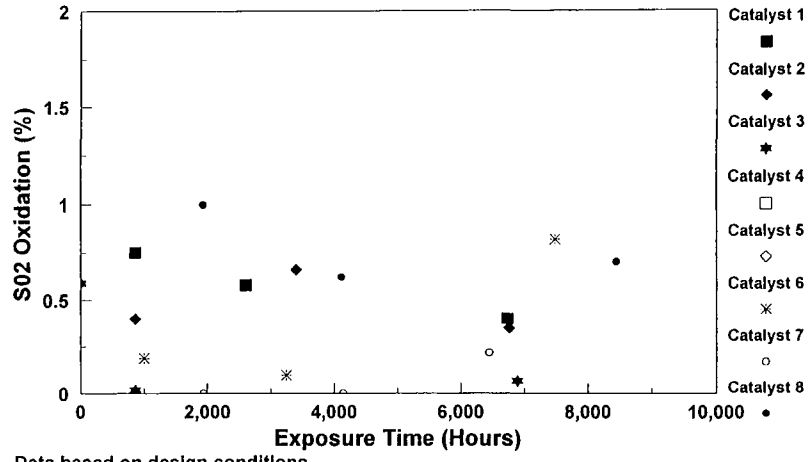
Ammonia Slip vs. NH3/NOx Ratio



620 °F, high flow
Second parametric data set

Figure 14

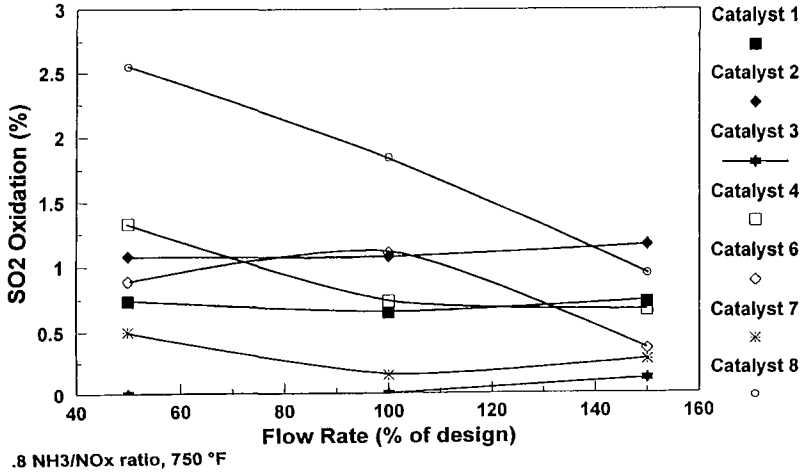
SO2 Oxidation vs. Exposure Time



Data based on design conditions
Linear corr. applied for small flow variations

Figure 15

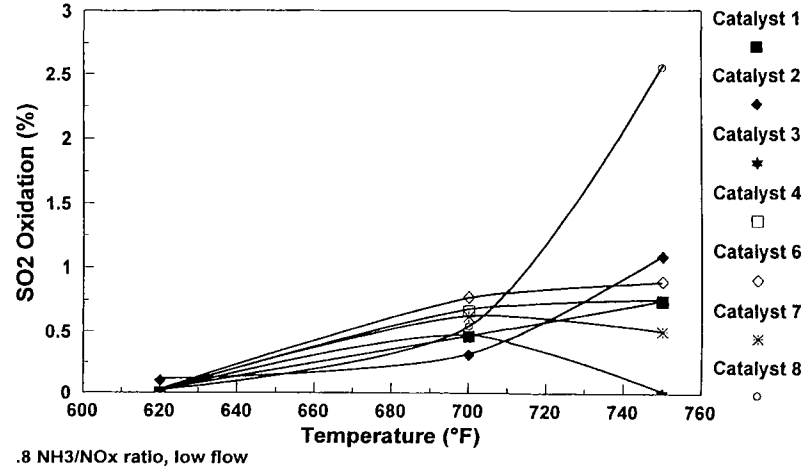
SO2 Oxidation vs. Flow Rate



.8 NH3/NOx ratio, 750 °F
Second parametric data set

Figure 16

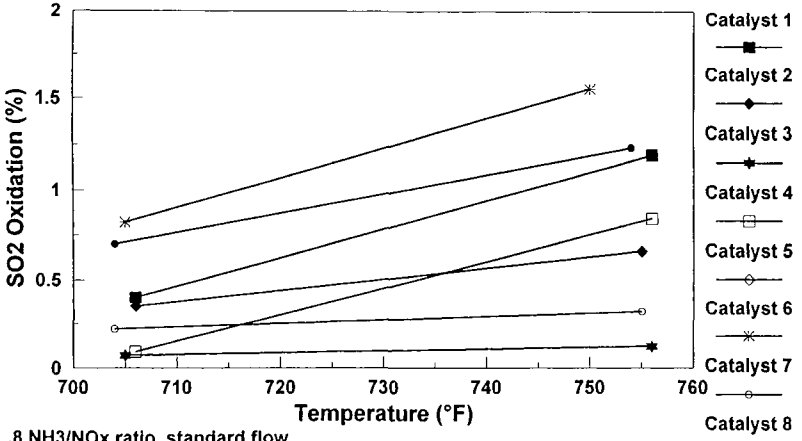
SO2 Oxidation vs. Temperature



.8 NH3/NOx ratio, low flow
Second parametric data set

Figure 17

SO2 Oxidation vs. Temperature



.8 NH3/NOx ratio, standard flow
Third parametric data set
No high temp Catalyst 5 data available

Figure 18

Reactor Pressure Drop vs. Time

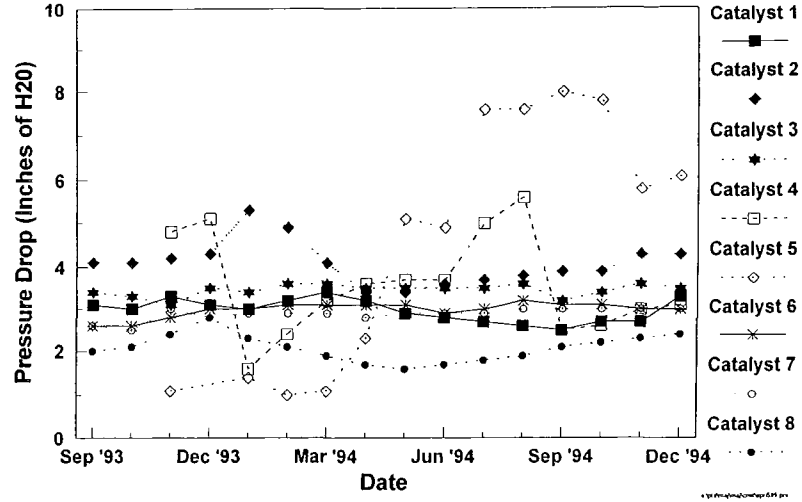
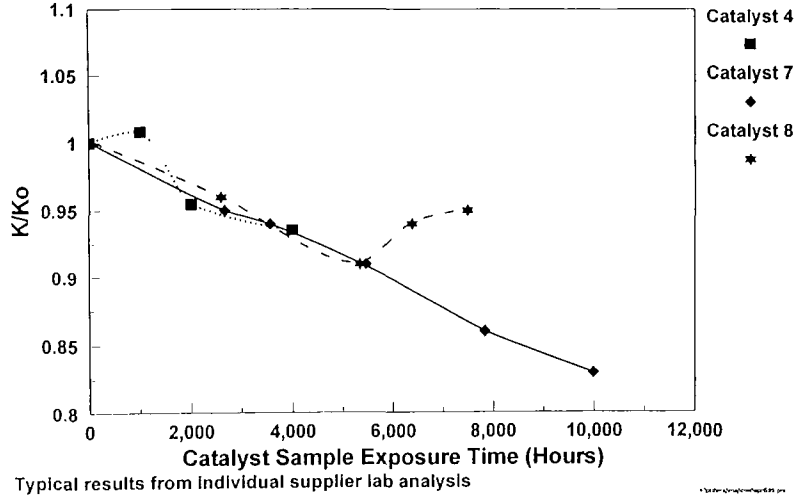


Figure 19

Laboratory K/Ko



Typical results from individual supplier lab analysis

STATUS OF TESTING OF SCR PILOTS: A REVIEW OF CURRENT EPRI-SPONSORED RESULTS

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Abstract

This paper provides an update of the ongoing testing at the New York State Electric & Gas (NYSEG) Kintigh Station SCR Pilot and the final results from the Tennessee Valley Authority (TVA) Shawnee Station and Niagara Mohawk (NiMo) Oswego Station SCR Pilots. The paper also previews the report, *SCR Design and Operational Recommendations: R&D Lessons Learned*, to be released later in 1995. The report will provide a comprehensive summary of key pilot results and their implications for full-scale SCR systems.

Recent pilot results have significant implications for full-scale SCR systems on U.S. coal- and oil-fired utility boilers. At TVA (high dust, high S coal), final results illustrate the effects of sulfur/ash on catalyst masking and deactivation, and the importance of reactor design features that counteract ash deposition. Results also show the potential for catalysts that are resistant to sulfur/ash masking through adjustment of physical catalyst properties. At NiMo (medium S oil), solids deposition from MgO injection at the host boiler caused significant catalyst plugging and deactivation of the uppermost catalyst layer. At NYSEG (post-FGD), exploratory catalyst testing shows the potential for reducing the operating temperature of post-FGD SCR, thereby reducing operating costs associated with flue gas reheat. In addition, heat pipe heat exchanger corrosion tests have significant cost implications for SCR in this configuration. In the future, this work will form the basis for the economic evaluation of post-FGD SCR.

The SCR design and operational recommendations report will be organized to allow the user to access information by topic with tables of SCR issues common to the pilots. The tables will incorporate short descriptions of the issues, associated results, and recommendations, with references to the specific report section that contains a full description of key pilot results.

Introduction

Selective Catalytic Reduction (SCR) has been widely demonstrated in Europe and Japan as a postcombustion NO_x control technology. However, most of this experience has been gained using relatively low-sulfur fuels, typically less than 1.5 percent. By comparison, the application of SCR in the United States has been much more limited, and to date the experience base is virtually non-existent for coal- and oil-fired boilers.

In addition to the foreign precedent, regulatory forces stemming from the 1990 Clean Air Act Amendments will inevitably lead to the more wide-spread use of SCR in the U.S. for both new and existing units. Although international experience with SCR is significant, it can not be directly transferred to the U.S. utility industry due to differences in fuel properties, furnace design, and the design and operation of U.S. powerplants.

Two of the most important differences between foreign and domestic SCR applications are the sulfur content and trace element concentration of fuels.[1] Higher fuel sulfur content corresponds to higher concentrations of SO_2 and SO_3 in the flue gas of typical U.S. SCR systems. This can lead to potential poisoning and more rapid deactivation of the catalyst. In addition, SCR catalysts have the potential to oxidize SO_2 and further elevate the concentration of SO_3 in the flue gas within and downstream of the SCR reactor. This can lead to serious problems with ammonium sulfate and/or bisulfate deposition in the air preheater, changes in the marketability of fly ash, and potential increases in plume opacity.

A number of compounds present in fly ash of coal- and oil-fired boilers may poison the active sites of an SCR catalyst. In Germany, for example, arsenic poisoning of SCR catalysts has been observed. Alkaline metals such as sodium and potassium have been demonstrated to deactivate SCR catalysts, are were believed to be the cause of reduced catalytic activity in SCR testing at EPRI's 2.5 MW Arapahoe test facility.[2] The relative lack of SCR experience for U.S. applications, coupled with significant differences in fuel composition represents a potentially serious situation for any U.S. utility faced with installation and operation of an SCR system on a high sulfur fuel source.

In response to uncertainties in the cost and feasibility of SCR for the U.S. utility industry, EPRI has sponsored a multi-pilot plant test program to evaluate the feasibility and cost of SCR as a function of fuel type and SCR/host boiler configuration. At the present time, pilot units have been operated at four utility sites. This paper discusses three of those pilots:

1. A high sulfur/high dust SCR unit was operated at the National Center for Emissions Research located at the Tennessee Valley Authority (TVA) Shawnee Power Station. This unit treated flue gas from the economizer outlet of a high sulfur coal-fired boiler.

2. A post-FGD unit is operating at EPRI's Environmental Control Technology Center located at the New York State Electric & Gas Corporation (NYSEG) Kintigh Station. This unit treats gas from downstream of an FGD system.

3. An SCR unit was operated on a residual oil-fired boiler located at the Niagara Mohawk (NiMo) Power Corporation Oswego Steam Station. This unit treated flue gas from the economizer outlet in a configuration similar to the TVA pilot.

Test program objectives for each pilot unit depend on the specific configuration of the pilot and fuel burned in the host boiler. Testing focused on the impacts of U.S. fuels on the catalyst, the reactor, and key process subsystems. Additional objectives include the development of testing and analytical procedures suitable for use on high sulfur SCR units, and identification and resolution of process control and continuous emissions monitoring issues.

The remainder of this paper contains a description of the distinguishing features of each pilot unit and their operating characteristics. Testing status and recent results are highlighted for each pilot, including updates to system performance and key operational issues. Finally, the forth-coming EPRI report, *SCR Design and Operational Recommendations: R&D Lessons Learned* is introduced. This report will contain all of the key design implications from testing at EPRI-sponsored pilots documented in a user-friendly format.

Pilot Unit Operating Characteristics

Schematic diagrams of the TVA, NYSEG, and NiMo SCR pilot systems are shown in Figures 1, 2, and 3, respectively. Each pilot represents a 1 MW(e) equivalent SCR reactor divided into two parallel sections to allow for simultaneous testing of two catalyst types. Each pilot features slip stream isolation dampers, flue gas heaters, air startup dampers, and ammonia injection control systems that calculate the required ammonia injection rate from flue gas flow and inlet NO_x signals. Operating conditions and catalyst characteristics are listed for each pilot in Tables 1 and 2, respectively.

The pilot SCR catalysts were designed to maintain certain performance criteria over guaranteed and design lifetimes of 2 and 4 years, respectively. Performance goals call for 80% NO_x conversion with residual ammonia (slip) levels of less than 5 and 2 ppm at the exits of the second-to-bottom and bottom catalyst layers, respectively. At TVA (5 catalyst layers), the 5 and 2 ppm slip limits apply to the outlets of the fourth and fifth catalyst beds, and at NYSEG and NiMo (3 layers), the limits applied to the outlets of the second and third catalyst beds.

TVA Hot Side/High Dust SCR Pilot

The TVA pilot was equipped with features to counter the effects of fly ash on the SCR catalysts. The catalyst was designed with relatively large cell openings (i.e., a large pitch) typical of a high dust application. A non-catalytic "dummy" catalyst layer was provided at the top of the reactor to straighten the gas flow prior to catalyst exposure, and to minimize catalyst erosion. Screens were installed above each catalyst layer and sootblowers were installed above the first and fourth layers to help keep the top surface of the catalysts free of solids. Reactor walls and catalyst baskets were also streamlined to eliminate many of the horizontal surfaces where ash could accumulate and subsequently fall onto the catalysts.

The original V/Ti catalyst was tested for the entire pilot test duration. Near the end of the test program, 6 of the 9 test elements in the center of catalyst beds 1 and 3 were replaced with fresh samples. The vendor theorized that a softer catalyst formulation would be more resistant to masking that had contributed to deactivation of the original catalyst over time.[3] The activity of the test elements was examined after 7,000 operating hours at the end of the test program. Results of the activity measurements are highlighted later in this paper.

The original TVA zeolite catalyst experienced substantial deactivation was replaced with a reformulated design about mid-way through the test program. Prior to replacement, the original catalyst was removed and water-washed to remove deposits that had plugged the catalyst channels. The purpose of the experiment was to determine whether catalytic activity could be restored, and to determine the effects of washing on the catalyst structure. These results are also discussed later in this paper.

NYSEG Post-FGD SCR Pilot

The distinguishing feature of the NYSEG pilot unit is the equipment used to raise the flue gas temperature from 120°F at the outlet of the wet FGD system to SCR operating temperatures (600 - 700 °F). Reheat requirements are minimized through the use of a recuperative heat-pipe heat exchanger. While the heat exchanger recovers heat from the gas exiting the reactor, an additional 185 °F of reheat input is required to maintain a reactor temperature of 650 °F.

Because the cold end of the NYSEG heat exchanger operates below the maximum condensation temperatures of ammonium sulfate/bisulfate and sulfuric acid, the test program was focused on evaluating exchanger fouling effects (i.e., heat transfer loss, increase in gas pressure drop, and corrosion).[4] These operating impacts are countered by the smaller reactor size and favorable catalyst environment (lower flue gas sulfur and ash content) in the post-FGD configuration.

The NYSEG reactor is typical of a post-FGD configuration in that less catalyst is required, and problems associated with high flue gas sulfur and fly ash content are avoided. No screens, "dummy" catalyst layers, or reactor sootblowers are required. The NYSEG reactor contains 3 layers, as opposed to 5 in the TVA pilot. The overall catalyst volume is lower than its high dust

counterpart because of the higher surface area-to-volume ratio inherent to a smaller pitch catalyst. The favorable reactor environment can also lessen the rate of catalyst deactivation, and further reduce catalyst volume requirements to achieve a given catalyst life. Lower flue gas SO_x and particulate content also reduces the potential for sulfur-related catalyst poisoning.[5]

NiMo Residual Oil SCR Pilot

The NiMo Oswego Station pilot is a hot side configuration similar to the TVA pilot. The host boiler is used for load following with frequently changing conditions. Typically, the unit cycles down to 20% MCR overnight, and hourly load changes of up to 60% MCR were routinely experienced during pilot testing. This characteristic presented challenges to the pilot process control system that were addressed during the test program.

Although the particulate content of test flue gas from the NiMo host boiler is considerably less than that at TVA, problems with catalyst deposition and pluggage were encountered throughout the test program. Reactor deposits were found to consist of oil ash and magnesium oxide and magnesium sulfate; the magnesium source being fuel oil additives. Catalyst pluggage was particularly troublesome for the composite V/Ti catalyst which had a very aggressive catalyst pitch (3.6 mm) for this service. Plugging countermeasures for the pilot were limited to soot-blowers above the first catalyst layer and routine catalyst cleaning during system shutdowns.[6]

A unique ammonia injection system design was evaluated at the NiMo SCR pilot. The TVA and NYSEG units employed high pressure, vaporized anhydrous ammonia injection systems typical of past designs in Europe and Japan and in early combustion turbine SCR systems in the U.S. At the NiMo pilot, direct liquid phase injection of aqueous ammonia was employed, using a twin fluid atomizing nozzle to facilitate droplet evaporation within the duct upstream of the catalyst. The benefits of such a system include the avoidance of aqueous ammonia vaporization costs, and the use of droplet injection to enhance cross-duct penetration and mixing of ammonia with flue gas upstream of the catalyst.

Pilot Status and Highlights of Testing Results

The following sections detail the status of each EPRI-sponsored pilot, and include brief descriptions of the key pilot performance issues and results. Pilot measurements include flue gas pressure loss and observations of catalyst pluggage, NO_x CEM and manual ammonia measurements to determine on-line system performance, and periodic vendor analysis of catalyst samples to determine activity changes and changes in chemical and physical catalyst properties. Manual SO_2 and SO_3 measurements were also made to determine the potential for SO_2 oxidation across each catalyst over time. In addition, special tests were conducted to examine the effects of SCR on key pilot unit subsystems (e.g., the effects of ammonia slip and SO_2 oxidation on the NYSEG recuperative heat exchanger).

TVA Hot Side/High Dust SCR Pilot

The TVA pilot was operated between May 1990 and May 1994. During this period, the unit operated with flue gas flowing for a total of approximately 22,000 hours (about 2.5 years). The relatively low availability of the pilot unit was the result of both short-term outages caused by events such as boiler tube leaks or reactor inspections, and longer-term outages caused by extended boiler outages or failures of certain pilot unit equipment (e.g., the pilot ID fan). Key pilot results center around the effects of high sulfur and fly ash on catalyst deactivation, and the countermeasures to ash deposition and catalyst plugging that were evaluated during the program.

Catalyst Activity. Both TVA test catalysts exhibited significant deactivation which was exacerbated by frequent pilot shutdowns early in the test program. Analysis of ash samples from the reactor identified a mechanism in which the ash deposits become enriched with sulfur via interaction with ambient moisture during shutdowns. As the moist acidic deposits reacted with alkaline ash constituents, hard deposits were formed that permanently plugged a number of catalyst channels. This mechanism may have also occurred on a smaller scale on the catalyst surface and within catalyst pores, and contributed to formation of a masking layer and consequential loss of catalyst activity.

The original V/Ti catalyst was tested for the entire pilot operating duration (about 22,000 hours) Results of catalyst sample activity measurements by the manufacturer are shown in Figure 4. The measurements were made on small sections of the catalyst sample that did not contain plugged channels; therefore, results were normalized with data shown in the figure from selected European experience.[7,8,9] In all cases, the samples from bed 1 exhibited a higher activity than those from bed 3, which may indicate the positive influence of sootblowers located above the first catalyst layer, but not above the third layer.

Figure 4 also shows the activity curve for replacement V/Ti test elements installed in the center of catalyst beds 1 and 3 near the end of the test program. The replacement elements featured different hardness values than the original catalyst charge. Although the replacement elements exhibited a lesser rate of deactivation, the positive effects of altering catalyst hardness are not entirely conclusive because a lower sulfur coal was being fired while the replacement elements were in place.

The original zeolite catalyst was replaced after approximately 12,000 hours with a reformulated zeolite design from the same vendor. The original catalyst failed to meets its performance criteria after only 5,000 hours of operation. Laboratory activity measurement results for the original and reformulated catalyst samples are shown in Figure 5. When examining the activity curves, it should be noted that the range of European experience represents traditional V/Ti catalysts, not zeolites. The reformulated catalyst showed improvement in its baseline activity and in the rate of deactivation compared to the original catalyst.

Figure 5 also shows the positive effects seen from water-washing of the original zeolite catalyst. However, subsequent inspections revealed damage to the physical structure of the catalyst, and swelling of the catalyst contributed to a significant rise in reactor pressure drop. Long-term effects of these physical changes were not evaluated at the pilot.

Bulk and surface chemical measurements were also made by both catalyst vendors to monitor changes in the composition of the catalyst and the accumulation of potential catalyst poisons. Bulk analysis results for the V/Ti catalyst are shown in Figure 6. These results indicate increases in the concentrations of arsenic, sodium, and potassium with increasing exposure time. Vendor analysis of catalyst cross sections indicated that these elements had penetrated the catalyst and were evenly distributed within the interior of the catalyst matrix. Other elements, including silicon, iron, aluminum, and sulfur were present in high concentrations on a masking layer on the outer catalyst surface.

The V/Ti catalyst vendor conducted a test in which they essentially sandblasted the masking layer from the catalyst surface. Subsequent activity tests conducted on the abraded catalyst sample revealed that the sample activity had increased from a K/K_o value of 0.63 to 0.99 as a result of the procedure. Because arsenic was found throughout the depth of the catalyst, this result indicates that the V/Ti catalyst activity loss was the result of masking of the catalyst surface by sulfated fly ash, and not by arsenic poisoning.

Catalyst Plugging Countermeasures. The TVA pilot represented a severe environment with respect to potential catalyst plugging with fly ash. Periodic reactor inspections revealed considerable buildup of solids on the outer catalyst blocks, which resulted from "wall effects" in the relatively small reactor. Over the course of the test program, the flue gas pressure loss across the V/Ti catalyst increased from below 4 inches to over 10 inches of water. Manual counting of the plugged channels showed that nearly 55 percent of all V/Ti catalyst channels had become permanently plugged by the conclusion of testing.

A number of strategies were implemented or considered during the test program to limit increases in catalyst channel plugging. These include screens, sootblowers, vacuuming, and moisture avoidance. All strategies employed at the pilot were successful to a certain degree, but their practicality for full-scale SCR application will vary.

Screens and sootblower above the catalyst beds were installed to help keep the top surface of the catalyst free of solids. Screens were installed above all 5 catalyst layers, and the optimum mesh size appeared to be slightly smaller than the catalyst pitch so that any ash that passed through the screen would also pass through the catalyst. Air sootblowers were installed above the first and fourth catalyst layers, and their effectiveness was evident through observations of pluggage in these beds compared to the others. Sootblowers were operated once each 8-hour shift while the pilot was operating.

After the initial experiences with outage-related catalyst plugging, vacuuming was employed to remove ash deposits during pilot shutdowns. Early in the test program, step changes in reactor pressure drop were observed following outages. This is consistent with the aforementioned sulfur enrichment mechanism that lead to hard, permanent deposits within the catalyst channels. Pilot operating procedures were later modified to incorporate vacuuming of the top surfaces of the catalyst at the start of all extended outages. Thus, ash deposits were removed before they could absorb significant ambient moisture.

In theory, sulfur enrichment of ash deposits could be minimized by preventing moisture from entering the reactor during outages. This could be accomplished by isolating or bypassing the reactor, or using a dry air purge or heaters to maintain a margin above the ambient dew point within the reactor. The design of the pilot, however, did not permit complete isolation of the reactor to allow for demonstration of this approach.

Other Operating Issues. Several operational issues were encountered during the TVA test program that provided pilot experience with full scale SCR design implications. These include:

- Ammonia injection nozzle pluggage in the high sulfur/high dust environment.
- Artifact reactions over sampling probe materials during NO_x and ammonia sampling.
- Process control issues associated with zeolite catalyst ammonia adsorption/desorption times.
- CEM system maintenance and sample preconditioning issues specific to high dust/high sulfur SCR systems.

These pilot experiences and resulting implications are beyond the scope of this paper. Detailed descriptions will be included, along with catalyst performance data and catalyst protection recommendations in EPRI's forthcoming SCR design and operational recommendations report.

NYSEG Post-FGD SCR Pilot

The NYSEG SCR Pilot began testing in December 1991 and is currently operating. At present, test catalysts have been exposed to flue gas for approximately 21,000 hours. Key pilot results include catalyst performance in the relatively clean post-FGD environment and cost issues associated with flue gas reheat. The reheat system includes a heat recovery exchanger, part of which operates below the condensation temperatures of ammonium-sulfur compounds and sulfuric acid. Adverse heat exchanger performance and corrosion effects from fouling were examined in terms of their potential cost implications for the post-FGD SCR configuration.

Catalyst Activity. Both of the catalyst designs tested at NYSEG have shown very good resistance to deactivation in the low dust, lower sulfur environment. The original extruded V/Ti catalyst and the replacement composite V/Ti catalyst showed no measurable activity change in

pilot tests over 13,100 and 12,700 hours, respectively. Therefore, given the same amount of catalyst and the same performance goals, a post-FGD catalyst would be expected to exhibit a substantially longer catalyst life than its high sulfur, high dust counterpart.

After being proven over 1.5 operating years, the original extruded catalyst was exchanged with a new extruded V/Ti formulation designed for higher absolute catalyst activity. In the reformulated design, the physical structure was further optimized to reduce reactant diffusion rate limitations and increase overall catalyst activity by up to 50%. The top layer of the original catalyst was moved to the fourth bed in the current arrangement to allow for continued flue gas exposure beyond 1.5 years. The reformulated catalyst currently occupies reactor levels 1 - 3. Baseline testing of the new catalyst has not been completed as of this time.

The original composite V/Ti catalyst exhibited severe deactivation and was replaced after only 5,300 operating hours. Activity changes were found to occur exclusively during pilot shutdowns, which indicated that ambient moisture had aided in the mobilization and penetration of catalyst poisons throughout the active catalyst surface layer. Contaminants that penetrated the catalyst include silicon, sodium, potassium, phosphorus, and sulfur, while calcium and iron were concentrated at the surface. The source of the contaminants is the fine ash and FGD carryover solids that are lightly deposited on the catalyst surfaces during operation.

The reformulated composite catalyst has shown no measurable performance decline over approximately 1.5 years of operation and numerous pilot shutdowns. According to the vendor, the original catalyst was manufactured using titanium with an incorrect size distribution. Pilot catalyst performance measurements confirmed that the absolute activity of the replacement catalyst exceeded that of the fresh, original catalyst. This data appears to confirm that manufacturing error caused, or at least contributed to the failure of the original composite catalyst.

Prior to the exchange of the extruded V/Ti catalyst, a short testing period was dedicated to catalyst evaluation at temperatures below the typical lower limit of 600 °F. The objective was to determine the potential for reducing the operating temperature in the post-FGD configuration, thereby reducing operating costs associated with flue gas reheat. To accomplish this, SCR catalysts would need to overcome performance effects from kinetic limitations at low temperature, and from possible fouling due to condensation of ammonium-sulfur compounds on the catalyst surface and within the catalyst pores. Exploratory testing provided insight to these effects that could lead to development of specific catalysts for this application.

At a reactor temperature of 550 °F, the extruded V/Ti catalyst exhibited marginally lower, but steady performance over 1,400 operating hours even though catalyst fouling was detected. The composite catalyst exhibited a more severe performance decline that varied with changes in the inlet NO_x concentration from the host boiler. This indicated that the degree of catalyst fouling was varying with the reactor ammonia concentration, which varies in proportion to host boiler NO_x levels at a fixed ammonia-to-NO_x ratio. At lower temperatures (i.e, 500 °F) both catalysts

exhibited substantial performance deterioration. After each test period, both catalyst's performance was restored to original levels by raising the temperature to baseline (650 °F) conditions.

Heat Exchanger Fouling Effects. Fouling within the recuperative heat pipe heat exchanger occurs as the gas exiting the SCR reactor is cooled from the normal reactor operating temperatures of 600 - 700 °F to the heat exchanger discharge temperature of 200 - 250 °F, depending on the reactor temperature and gas flow rate. The heat pipes downstream of the reactor operate at temperatures approximately 60 °F lower than the surrounding gas temperature. Internal surface temperatures of approximately one-third of the heat pipes in the cold end are below the sulfuric acid dewpoint (270 - 280 °F, based on SO_3 levels of 7 10 ppm). About half of the heat pipes operate below the maximum temperature threshold for condensation of ammonium-sulfur compounds (approximately 330 °F, based on 7 ppm SO_3 and 1 ppm NH_3). Figure 7 shows the range of heat exchanger fouling observed after operating periods with different levels of ammonia slip from the SCR reactor.

Figure 8 shows the relative decline in heat transfer and increase in flue gas pressure drop across the return side of the heat exchanger during operating periods with distinct ammonia slip levels. In the figure, heat transfer is expressed as a fraction of the design rate to normalize exchanger efficiency for changes in gas flow and reactor temperature. The figure also shows the effects of internal water-washing between operating periods. Water-washing was very effective in dissolving and removing deposits, and consequently in restoring heat exchanger performance and pressure drop to original conditions.

Figure 8 also illustrates the importance of minimizing ammonia slip from SCR catalysts in the post-FGD configuration. At the pilot unit, severe heat exchanger performance degradation was avoided when the average ammonia slip was held to below 2 - 3 ppm. During these periods, the total flue gas reheat loss equates to 10 - 12 °F at pilot design conditions over an approximate 3 month operating period. Longer term testing has shown that much of heat transfer loss occurs over the first 1 - 2 months after cleaning, and declines more slowly beyond that time. Over an approximate 8 month operating period at low (less than 2 ppm) average ammonia slip, the total flue gas reheat loss was only 12 - 14 °F. Therefore, a full-scale unit may be expected to sustain operation between annual outages without forced shutdowns for heat exchanger cleaning, provided a strict limitation on SCR ammonia slip is maintained.

Analysis of deposits proved that the cold end of the exchanger operates in a highly corrosive environment. Deposits contain ammonia, SO_3, heat exchanger metals, and constituents of fine solids within the pilot flue gas. The relative amounts of ammonia and SO_3 vary with the average catalyst ammonia slip and with the operating temperature in the deposit location. Overall, deposits are highly acidic, primarily due to the presence of ammonium bisulfate and free sulfuric acid. Heat exchanger wash-water pH values are typically 1.5 2.0 regardless of the deposit location or pilot operating conditions.

One of the pilot heat pipes from the cold end of the heat exchanger was removed and destructively examined after approximately 9,100 hours of pilot operation. The heat pipe was constucted of duplex 2205 steel, with 409 stainless steel fins. The cold end of the heat exchanger is considered the most severe environment in terms of corrosion effects because it operates below the sulfuric acid dewpoint on both the supply (upstream of the SCR system) and return (downstream of the SCR system) sides of the exchanger. The 409 stainless steel fins exhibited severe pitting and a maximum 17.5 mils/year corrosion rate. The maximum 2205 duplex steel heat pipe corrosion rate was 9.7 mils/year, and corrosion was found to be relatively uniform. A higher corrosion rate (18.1 mils/year) was found in a highly localized area on the supply end of the heat pipe.

Based on pilot corrosion measurements, the 2205 duplex steel heat pipes would not sustain a full 15 year equipment life in cold-end heat exchanger service. The high localized corrosion rate occurs where a small pocket of non-condensibles exists within the pipes at the upper end of the heat pipe incline. Assuming this area could be protected via enameling or by other means, the next highest corrosion rate (9.7 mils/year) equates to a projected tube life of 6 - 7 years. With the pilot heat pipe thickness (65 mils), the material would not be acceptable for full-scale service unless a heat pipe replacement strategy was incorporated into the full-scale design.

Test heat pipes constructed of carbon steel, CORTEN B, and AL6XN were also installed in the exchanger and destructively examined after 3,300 operating hours. In the most severe test location (at the cold end) various corrosion phenomena were observed, and none of the materials were found to be acceptable. Upstream of the cold-end heat pipe module (between modules 1 and 2 as shown in Figure 7), CORTEN B exhibited acceptable corrosion while AL6XN and carbon steel were marginal. Overall, deposits in the higher temperature areas of the exchanger (i.e, modules 2 and 3) are much less corrosive because of the lack of free sulfuric acid in the higher operating temperature range. For full-scale applications, carbon steel would likely be acceptable for these areas.

Additional implications from heat exchanger testing were developed during the test program, the details of which are beyond the scope of this paper. Some of these issues include heat exchanger water-wash techniques and disposal issues, potential stack particulate emissions impacts, and other recommendations from corrosion testing. These issues will be detailed in the forthcoming SCR design and operational recommendations report.

NiMo Residual Oil SCR Pilot

The NiMo pilot unit was operated between October 1991 and October 1993. During this period, the unit operated with flue gas flowing for approximately 4,800 hours (about 6.5 months). Low pilot unit availability was caused by low Oswego Station demand, resulting in a low host boiler capacity factor throughout the test program. Key pilot results include the effects of vanadium within the fuel on catalyst activity, and catalyst pluggage and deactivation from flue gas solids from the host boiler.

Catalyst Activity. Activity changes were measured by both catalyst vendors on samples taken after 2,400 and 4,100 operating hours. The relative activity of the corrugated plate catalyst increased during both test intervals, and exceeded the original activity by 24% by the end of the test program. Similar activity increases were measured across all three catalyst layers. The effect was attributed to deposition of vanadium from the host boiler on the catalyst surface. SO_2 oxidation rates increased with time over both catalysts, which is a further indication of vanadium deposition. The measured fuel oil vanadium content varied between 55 and 170 ppm during the test program.

The activity of the top layer of composite V/Ti catalyst decreased somewhat during the test program, but no overall performance change was detected via pilot NO_x conversion and ammonia slip measurements at the reactor exit (after 3 layers). After 4,100 hours, the activity of the top layer declined by roughly 20% based on the average of values for samples taken from the tops of the first and second layers. Essentially no activity change was seen in the second and third beds during the course of the test program. Overall, catalyst performance was well within its design limitations at the conclusion of testing despite significant catalyst pluggage.

Deactivation at the inlet of the composite catalyst was attributed to masking by a thin layer of solids found on the catalyst surface. Although a significant fraction of catalyst channels were completely blocked, laboratory activity measurements and calculations were based only on open catalyst channels within the samples. The catalyst vendor concluded that solids deposition and consequential activity loss at the top of the reactor was exacerbated by the aggressive catalyst pitch (3.6 mm). A more conservative catalyst design and additional measures to prevent solids deposition (i.e., sootblowers above every catalyst level) are advisable for full-scale SCR systems in similar heavy oil service.

Catalyst Plugging and Deposition. Severe reactor solids deposition and catalyst plugging was experienced over the course of the test program. Flue gas pressure loss across the plate catalyst (4.6 mm hydraulic diameter) increased from 3.0 to 3.8 inches water during operation, which corresponds to overall pluggage of 15-20% of the catalyst channels. The composite catalyst (3.6 mm pitch), exhibited nearly 50% overall pluggage, resulting in a pressure drop increase from 5.5 to over 10.5 inches water.

Reactor solids consisted of primarily oil ash, magnesium oxide (MgO), and magnesium sulfate. Deposits were also found to contain insulation from the host boiler, which may have contributed to catalyst pluggage early in the test program. MgO is added to the fuel oil and is also injected at the economizer inlet to control SO_3 via reaction to form magnesium sulfate. One mole of MgO is required to bind every mole of SO_3. Analysis of reactor deposits indicates that the overall additive/injection rate employed by the station equates to nearly two moles of MgO per mole SO_3. Although not proven at the pilot scale, more strict control of MgO usage may reduce solids deposition and catalyst pluggage effects in full-scale SCR systems for residual oil boilers. In addition, sootblowers were found to be highly effective in preventing catalyst pluggage in this service in a detailed evaluation at another EPRI-sponsored pilot.[10]

Other Operating Issues. Operational lessons from the NiMo pilot study include the demonstration of direct liquid ammonia injection, and process control issues associated with inconsistent aqueous ammonia concentrations and deep cycling of the host boiler. These issues will be summarized in EPRI's SCR design and operational recommendations report, a preview of which follows.

SCR Design and Operational Recommendations Report

Results from all EPRI-sponsored pilots are currently being incorporated into a guidance document entitled *SCR Design and Operational Recommendations: R&D Lessons Learned* (EPRI Report TR-105103). The report will be released later in 1995, and will include results and design implications from the three pilot studies described in this paper. In addition, the report will include the results of testing at the advanced SCR pilot system at the Pacific Gas & Electric Company's Morro Bay Station, and at the multi-pilot SCR system at Southern Company Service's Plant Crist sponsored under the DOE's Clean Coal Technology Program. Other EPRI-sponsored SCR studies to be detailed in the report include:

- Results of SCR catalyst research which focused on the examination of SCR poisoning and deactivation mechanisms.
- Recent updates to SCR cost studies for utility power station applications.
- An evaluation of Hybrid SCR designs.
- SCR procurement guidelines.

The report is being developed to help those responsible for designing, specifying, and trouble-shooting SCR systems for utility boiler applications. The report will be organized by topic to combine the related lessons from the various pilot studies. Overall SCR design and operational issues will be described with pilot implications included where applicable. Tables of important pilot results will highlight each specific issue and full-scale recommendation, with reference to the report section or appendix that fully details each key pilot result.

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Table 1

Typical SCR Pilot Operating Characteristics

	TVA Shawnee Station	NYSEG Kintigh Station	NiMo Oswego Station
Host Boiler Fuel Type	High (2.5-5.0%) S Coal	Medium (1.5-2.5%) S Coal	1.5% S Residual Oil
Pilot Configuration	Hot Side/High Dust	Post-FGD	Hot Side
Total Flue Gas Flow Rate, scfm	2100	2000	2000
Reactor Temperature, °F	700	650	700
Inlet NO_x, ppm	450	300	200-1000
Inlet SO_2, ppm	2000	150	800
Inlet SO_3, ppm	20	5	23
Particulate, gr/dscf	3.0	0.0012	0.091

Table 2

SCR Pilot Catalyst Characteristics

	TVA Shawnee Station			NYSEG Kintigh Station				NiMo Oswego Station	
Catalyst Type	Extruded V/Ti	Zeolite	Reformulated Zeolite	Original Extruded V/Ti	Reformulated[a] Extruded V/Ti	Original Composite V/Ti	Replacement[a] Composite V/Ti	Composite V/Ti	Plate V/Ti
Total Flue Gas Exposure, hr	21800	11900	7900	14700	1600	5300	12700	4600	4600
Cell Size (Pitch), mm	5.95	7.60	5.59	4.2	4.0	3.6	3.2	3.6	4.6[b]
Wall Thickness, mm	0.80	1.68	0.86	0.78	0.69	0.66	0.30	0.66	1.50
Cell Density, cpsi	18	11	21	37	39	50	64	50	18
Void Fraction, %	73.7	61.0	73.0	64.8	66.0	67.0	80.0	67.0	66.0
Geometric Surface Area, ft^2/ft^3	174	125	170	229	244	277	338	277	177
Space Velocity, hr[c]	2470	1677	2204	4660	4590	9180	6750	10260	4360
Area Velocity, ft/hr[c]	14.1	13.4	13.0	20.4	18.8	33.1	20.4	25.0	25.0
Linear Velocity, ft/sec[c]	10.4	12.5	10.5	11.8	11.6	11.4	9.6	11.4	5.4

[a]Exposure hours current as of March 31, 1995 in continued test program
[b]Hydraulic Diameter
[c]At 1000 scfm, 60°F, 5 catalyst layers at TVA and 3 layers at both NYSEG and NiMo

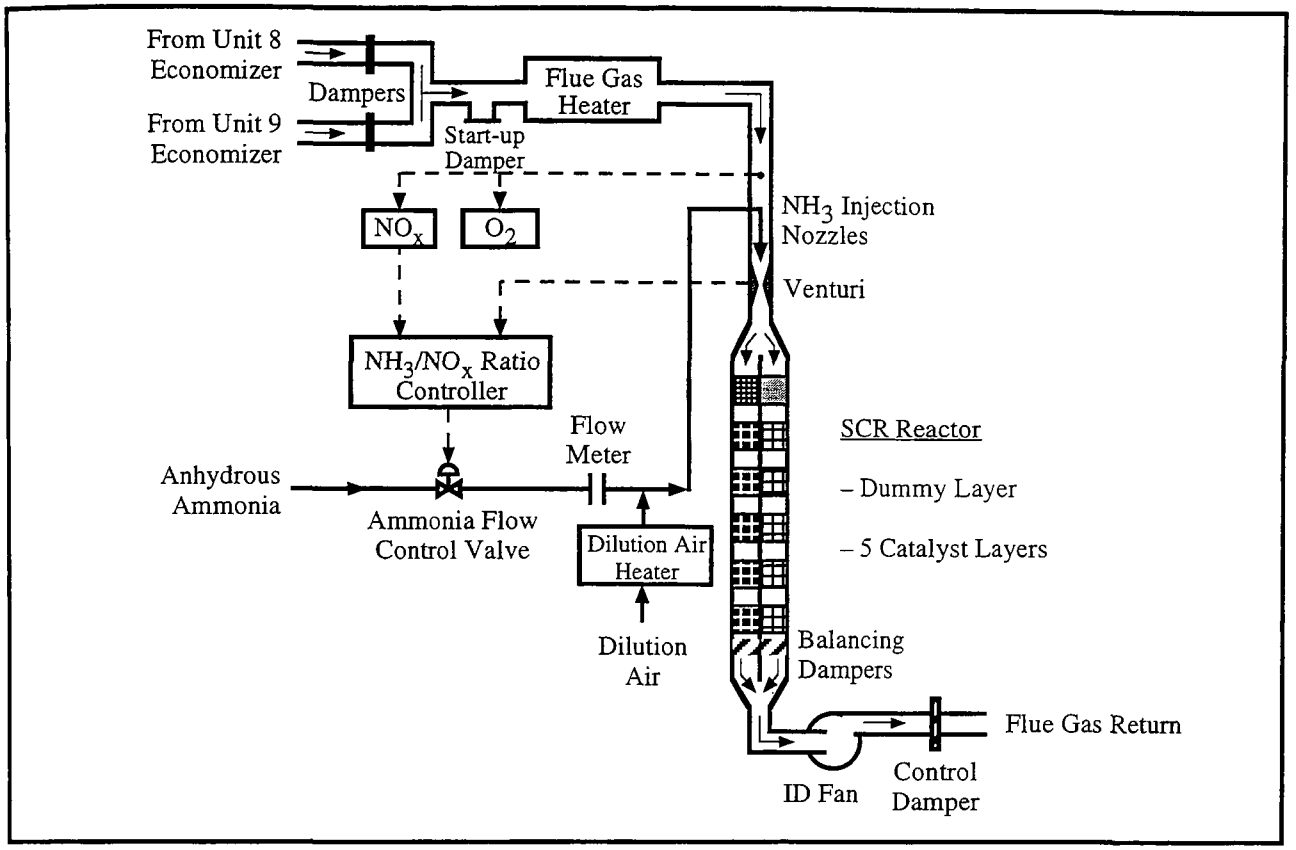

Figure 1
TVA Hot Side/High Dust SCR Pilot Unit Process Flowsheet

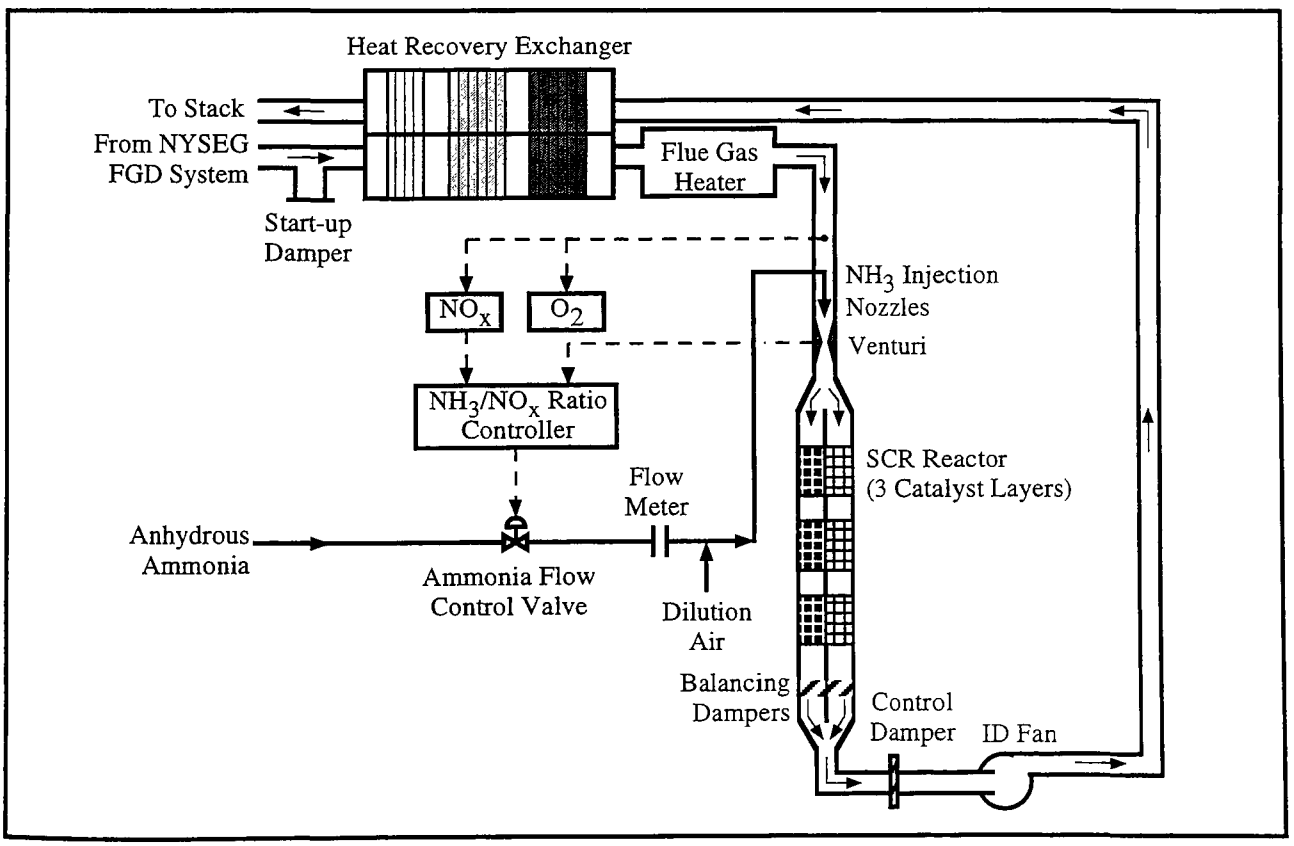

Figure 2
NYSEG Post-FGD SCR Pilot Unit Process Flowsheet

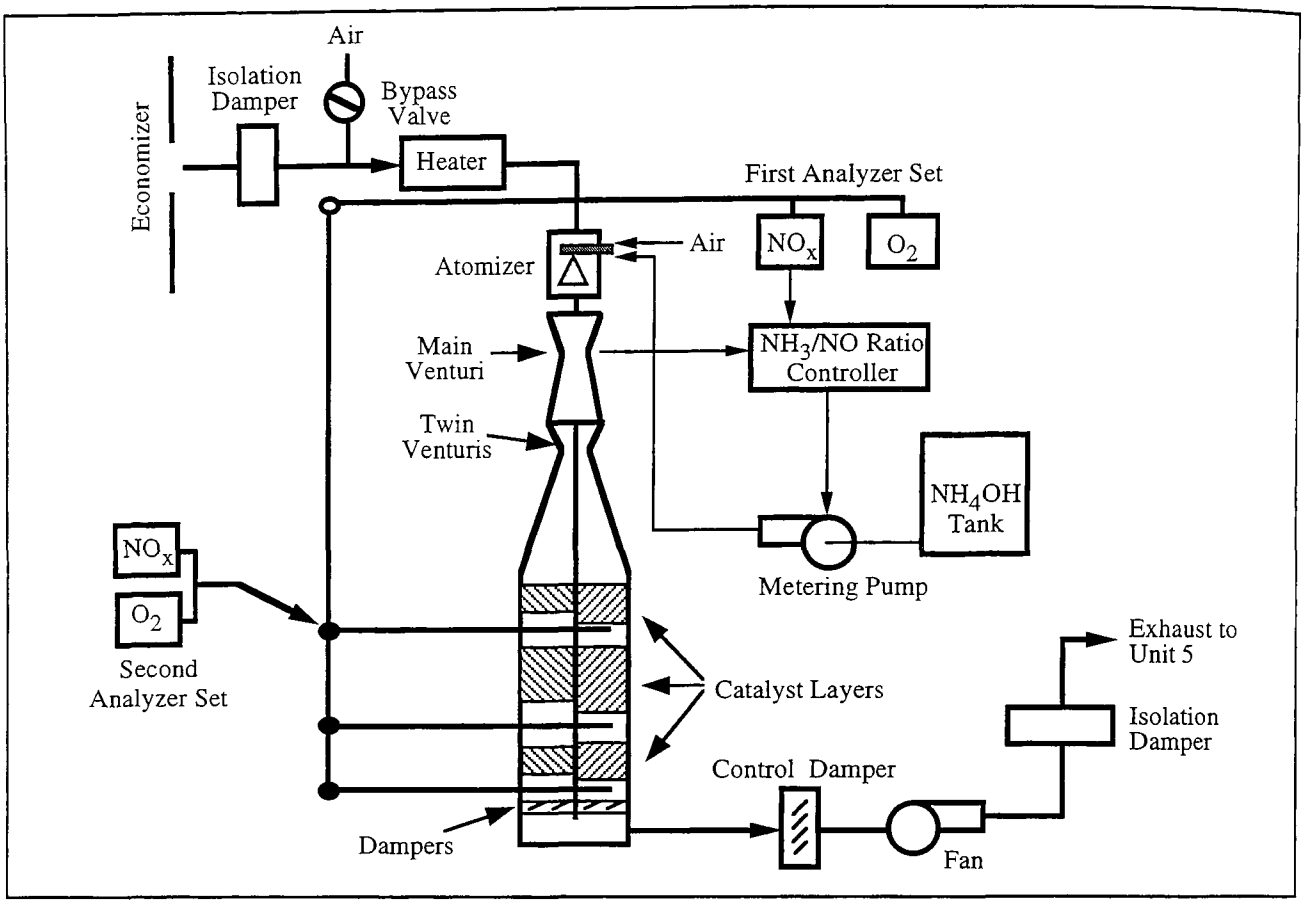

Figure 3
NiMo Hot Side SCR Pilot Unit Process Flowsheet

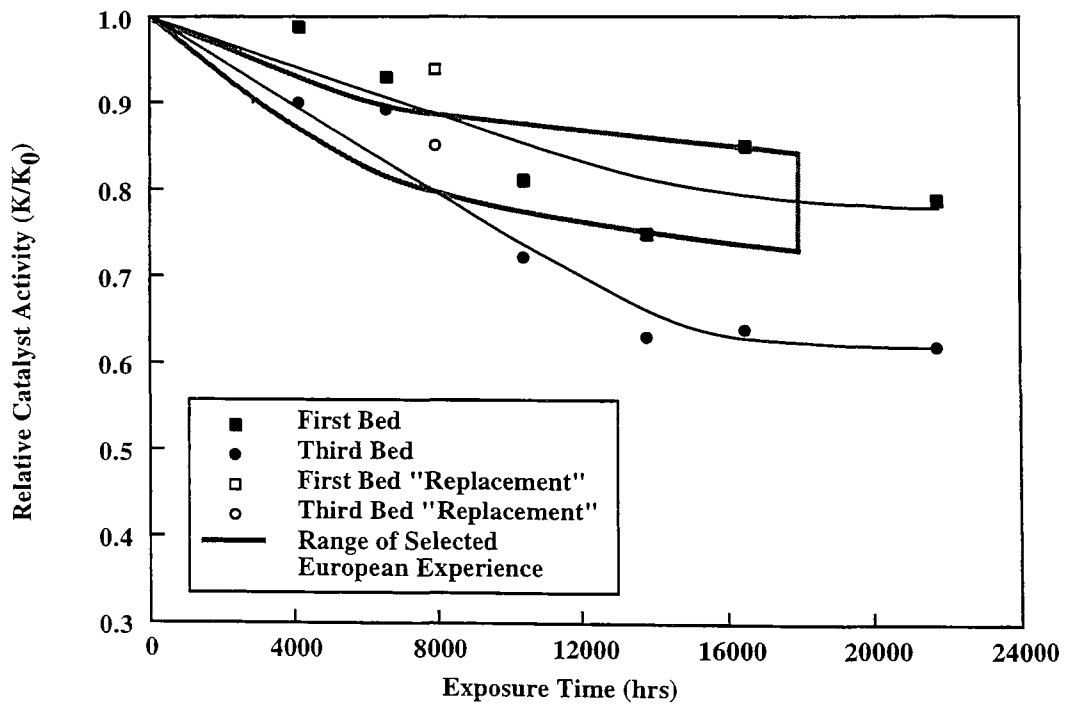

Figure 4
Relative Changes in TVA V/Ti Catalyst Activity vs. Exposure Time

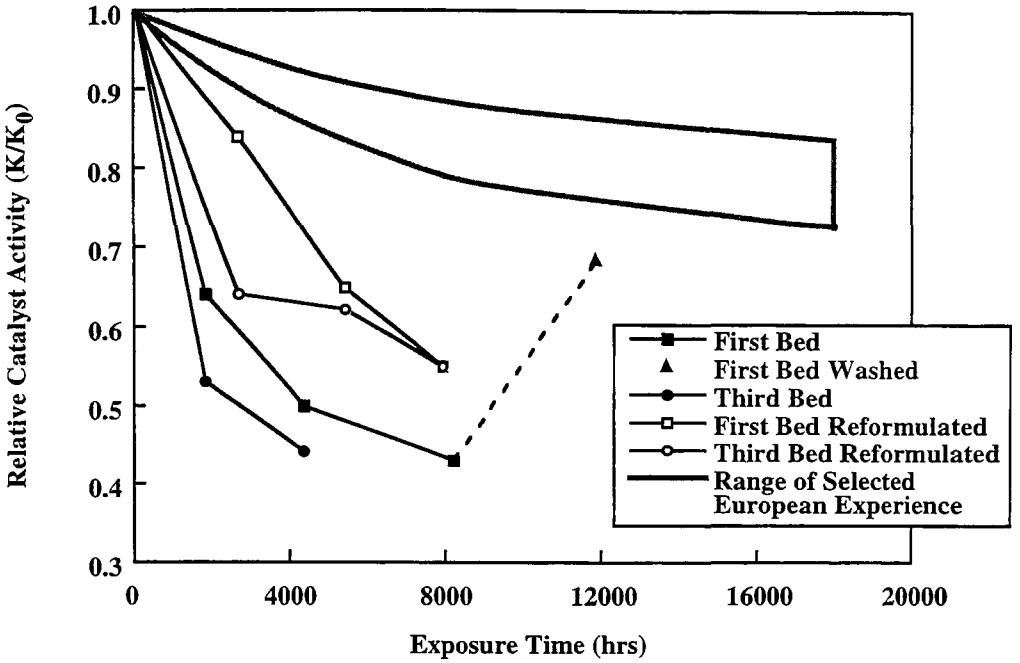

Figure 5
Relative Changes in TVA Zeolite Catalyst vs. Exposure Time

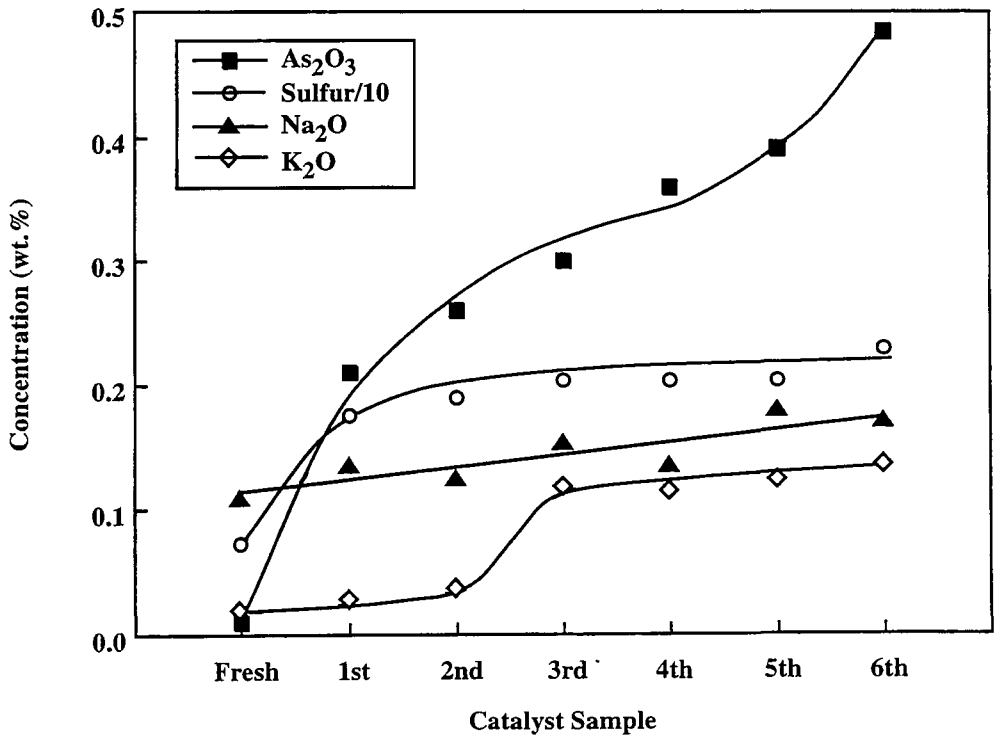

Figure 6
Bulk TVA V/Ti Catalyst Contaminants

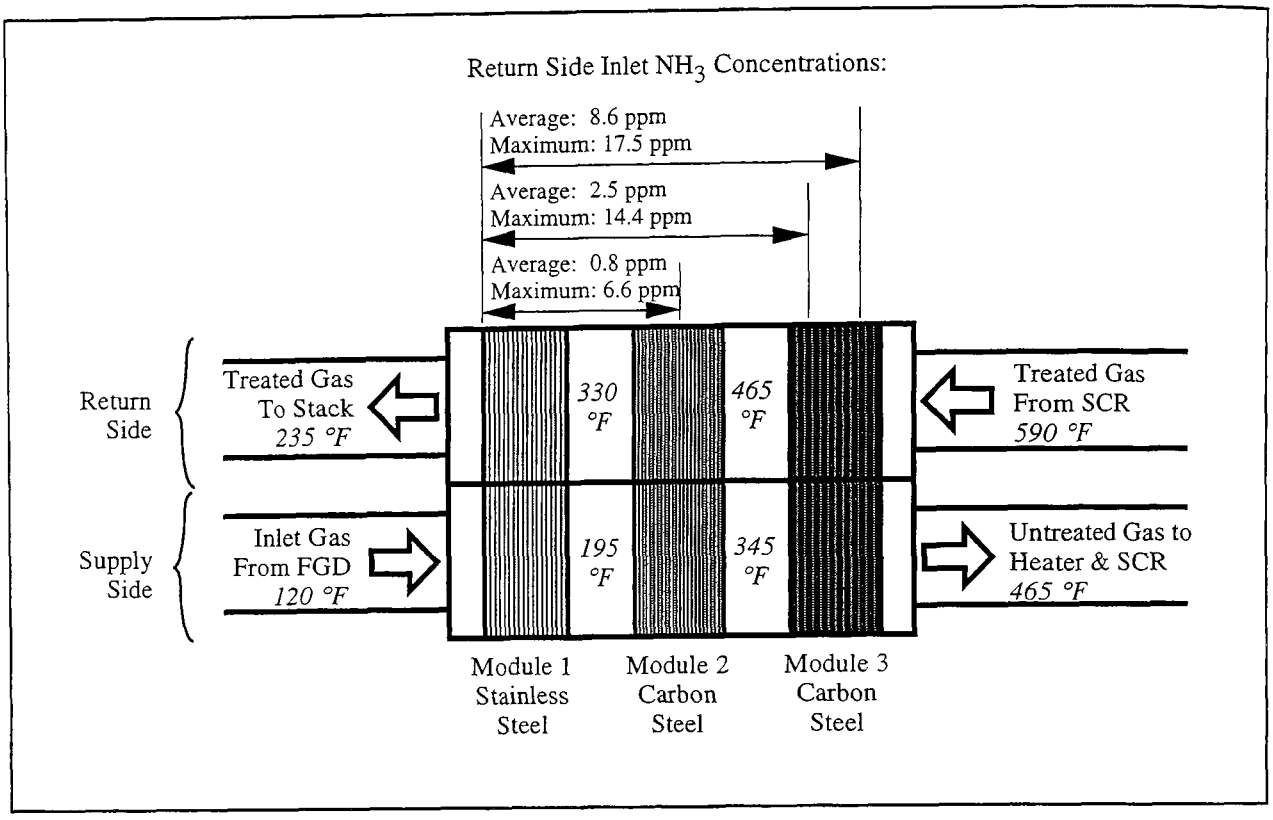

Figure 7
Range of Fouling Within Return Side of NYSEG Pilot Recuperative Heat Exchanger

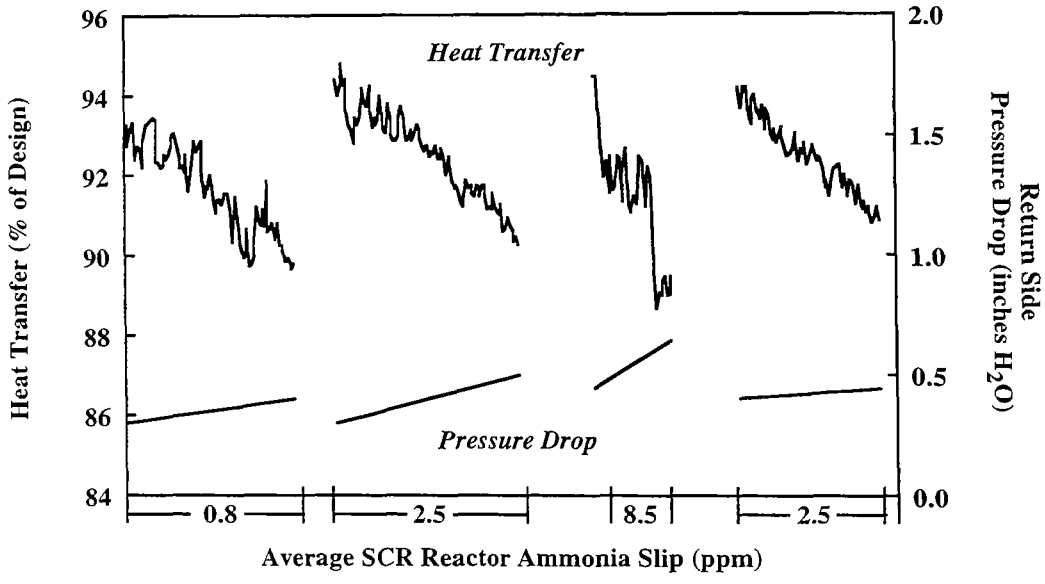

Figure 8
NYSEG Pilot Heat Exchanger Performance and Return Side Pressure Drop vs. Time (Heat Exchanger Was Water-Washed Between Operating Periods)

RESULTS OF CATALYST TESTS AT THE PG&E/EPRI ASCR PILOT PLANT

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Abstract

This paper presents results from the PG&E/EPRI advanced selective catalytic reduction (ASCR) 3-MWe slipstream pilot plant located at PG&E's gas/oil-fired Morro Bay Power Plant. Specifically, results of tests using an "in-duct" ASCR configuration and a catalytic air preheater (APH) are covered. Results from the in-duct ASCR tests include catalyst performance during oil-fired operation, APH deposition during oil-fired operation, and the effect of selective catalytic reduction on visible plume characteristics during oil-fired operation. Initial performance results for the catalytic APH tests are presented.

Introduction

Regulations requiring major reductions in NO_x emissions from fossil-fired power plants in Northern California have recently been adopted by a number of air quality regulatory agencies. Table 1 summarizes these regulations.

Table 1

NO_x Emission Regulations in the PG&E Service Territory

Species	Fuel	Allowable Emissions
NO_x	Natural gas	10 ppm at 3% O_2 dry
NO_x	No. 6 oil	25 ppm at 3% O_2 dry
NH_3	Gas/oil	10 ppm at 3% O_2 dry

These low NO_x levels will almost certainly require the use of some form of selective catalytic reduction (SCR) technology with ammonia. Although SCR technology has been widely applied commercially in a number of countries (notably, Japan and Germany) for many years, and more recently in Southern California, costs for these technologies are still excessive, especially given the rapidly evolving competitive environment for power generation in California. Furthermore, from a technical perspective, it is believed that additional, significant cost reduction opportunities remain to be exploited in both SCR technology and combustion control technologies. For this reason, PG&E has initiated a significant research effort to realize these opportunities. Major activities currently underway include a full-scale burner demonstration (Babcock and Wilcox S-burner) at the Morro Bay Power Plant and the ASCR pilot plant (the subject of this paper).

ASCR Technology Description

Since most costs associated with SCR are not related to the catalyst itself, but rather to associated equipment such as duct modifications, reactor vessels, structural steel, and earthwork/foundations, PG&E's research has been directed at minimizing, or even eliminating, these latter expensive components.

In-duct SCR arrangements deploy the catalyst within the fuel gas path between the economizer outlet and APH inlet. Current in-duct SCR configurations enlarge the ductwork to lower velocities, accommodate a larger volume of catalyst, and reduce pressure drop. A number of such designs have been recently used in Southern California utility boiler installations. An even lower cost SCR technology, "true in-duct" SCR, would utilize the existing ductwork with only minor modifications. In addition to the in-duct ASCR research activities, another interesting technology with the potential for eliminating the previously mentioned expensive components is the catalytic APH. All of these ASCR techniques are actively being investigated by PG&E's research programs at the ASCR pilot plant test facility.

ASCR Pilot Plant Description

As might be inferred from the above discussion, the principal purpose of the ASCR pilot plant is to permit a facility which allows investigation of primarily engineering/construction, i.e., cost, and operations-maintenance, related issues. A brief description of the pilot plant facility follows.

The ASCR pilot plant is located at PG&E's Morro Bay Power Plant. The pilot plant is capable of drawing a 5,000-scfm slip stream from either the Unit 3 or Unit 4 boilers at Morro Bay. The facility is designed to operate on a continuous, unattended, 24-hour per day basis, and duplicates full-scale utility boiler design and operation. The pilot plant simulates boiler components from the economizer inlet through the outlet of a Ljungstrom regenerative APH, including allowance for installation of catalytic elements in the APH. Initially, a Haldor Topsøe, Inc. (HTAS), catalyst was installed in the in-duct catalyst section of the reactor. A catalytic air heater element provided by a joint effort of ABB Air Preheater, Inc., and Engelhard Company has also been installed.

The test facility consists of the following major components:

- An in-duct SCR reactor.
- A Ljungstrom APH that duplicates the thermal history (metal, air, and gas temperatures vs. axial distance) of a full-scale PG&E APH.
- Flue gas and air fans, flow measurement venturis, flue gas heater, ductwork, insulation, and structural and electric components.
- A computer-driven PLC control system.
- State-of-the-art emissions measurement instrumentation for process control and catalyst performance evaluation.
- Multiple locations for aqueous or vaporized NH_4OH injection.
- SO_3 injection system

A photograph of the ASCR pilot plant is presented in Figure 1. As shown, the electric heater and gas fan are located upstream of the economizer and catalyst sections. Venturi flow meters on the gas and air circuits of the pilot plant provide flow measurement capability. Pressure transducers and thermal RTDs measure gas pressures and temperatures at several locations throughout the pilot plant.

A Haldor Topsøe catalyst has been initially installed in the pilot plant using an in-duct configuration. The reactor can accommodate up to three layers of catalyst; however, only two layers, or catalyst beds, are currently installed. Each layer is nominally 2 ft high by 4 ft wide, with a depth of 1.67 ft. The catalyst has a hydraulic diameter (i.e., characteristic dimension opening size) of 3.3 mm and a nominal wall thickness of 0.5 mm. At the design flow rate of 5,000 scfm, the overall space velocity is approximately 26,000 h^{-1} at the outlet of the fist catalyst bed and 13,000 h^{-1} at the outlet of the second.

After exiting the catalyst beds, the flue gas flows through a Ljungstrom APH. This device is a scaled-down version of the full-scale APHs installed on PG&E boilers. The design was selected to simulate the flue gas, air, and metal temperature time history of a full-scale APH, specifically in relation to heat transfer, element configuration, and construction materials. The ASCR APH is supplied with a sootblower in the air inlet duct at the cold end of the APH and blows from the cold air side towards the hot. The sootblower is an oscillating pipe with a single steam nozzle.

The ABB/Engelhard catalyst elements replaced the original hot-end elements with catalytically coated ones. A high heat and mass transfer geometry was selected. The catalytic elements have a hydraulic diameter of about 8 mm with a nominal thickness of 0.7 mm. Based on the total volume of the 16-inch-deep high-temperature elements, the overall space velocity is approximately 14,400 h^{-1} for a slipstream flue gas flow rate of 4500 scfm. However, since only 5.5 of the 12 sectors (46%) are exposed to the flue gas at any point in time, the actual operating space velocity of the catalytic APH is about 31,400 h^{-1}. A high activity vanadia/titania catalyst is bonded to the surface of a treated metal plate.

ASCR Project Goals

The goals of the ASCR pilot plant project are summarized below. While cost is a major factor in the research activities, other areas are also being pursued, for example minimization of by-product environmental impacts associated with SCR technology.

- Cost minimization.
- Technical risk management/risk minimization.
- Permit high reliability/availability at full capacity soon after startup.
- Minimize of operations and maintenance requirements.
- Minimize of adverse environmental impacts associated with SCR technology.

Specific research activities completed or currently underway are summarized below:

- Evaluate advanced catalysts (Haldor Topsøe and ABB/Engelhard).
- Examine APH deposition/fouling during oil-fired operation.
- Assess/avoid visible plumes during oil-fired operation.
- Evaluate the minimum acceptable catalyst size opening for the short periods of oil burning normally experienced at PG&E.
- Develop advanced flow modeling performance predictive techniques.
- Develop true in-duct ASCR configurations to minimize retrofit costs.

As a result of this research and other ongoing projects, a cost savings goal of $30-50 million has been targeted.

Results

Advanced flow modeling/predictive performance techniques are covered separately by another presentation at this conference ("A New Design Tool for SCR Systems" by L. J. Muzio et al.). True in-duct SCR work is in progress and will not be covered at this time.

Air Preheater Deposition/Removal

The goal of the APH fouling tests was to investigate the effects of an SCR catalyst on APH operation during oil firing. As is well known, ammonium sulfate and bisulfate salts can form due to SO_3 present in the flue gas, and NH_3 slip emitted by the SCR process at temperatures below nominally 400°F. These ammonium salts may then deposit in the APH baskets. Deposition of these salts can lead to increased pressure drops across the APH, and if unchecked, limit the full-load capability of the unit. Corrosion of the APH could also result.

During this testing, the formation and deposition of ammonium salts from SO_3/NH_3 reactions within the APH was evaluated by monitoring the pressure drop across the APH, measuring the concentrations of various NH_4^+ compounds entering and leaving the APH, and visual inspections. In addition, the effectiveness of current sootblowing procedures for removing any deposits was evaluated. SO_3 and NH_3 concentrations were varied in a controlled manner (up to 20 ppm each were investigated).

Test results can be summarized as shown in Figure 2. For reference, APH pressure drop is about 3-3½ inches H_2O during natural gas operation. As can be seen, the pressure drop began to increase rapidly during the two periods when the sootblowers were not operated. More important, operation of the sootblowers per the normal design/procedures permitted rapid recovery of the pressure drop to normal values. Other tests and inspections verified this result.

The tests conducted showed that the existing air heater/sootblowing design and operating procedures are readily able to control any deposition, and no modifications are required.

In-duct Catalyst Pluggage during Oil-fired Operation

The goal of this testing was to determine if the 3.3-mm catalyst opening size is acceptable during the normal, brief (2-4 weeks per year) oil burning periods that PG&E has experienced in recent years. This is important since a catalyst with a smaller opening results in lower costs. Tests were conducted without sootblowers or turning vanes because this would be the preferred lowest cost approach for the predominant natural gas-fired operation of PG&E plants.

Test results are shown in Figure 3. As can be seen, pressure drop in the catalyst increased rapidly due to the buildup of particulates from the oil-fired operation: pressure drop increased from about 3½ inches H_2O to 11 inches in less than 500 hours (about 3 weeks). It is estimated that this increase in pressure drop would result in a 40% derate in full-load capacity—an obviously unacceptable situation.

Fortunately, these deposits were concentrated at the inlet face to the catalyst and were quite friable. Subsequent tests with a conventional sootblowing system has shown these deposits to be easily removed; pressure drop was held at low levels during several recent months of oil-fired operation.

As a consequence of these tests, it was concluded that the 3.3-mm opening is acceptable for oil firing as long as adequate sootblowing capability is provided.

It should be emphasized that the deposition observed in not a unique feature of the Haldor Topsøe catalyst investigated; any catalyst of similar pitch would exhibit the same behavior.

Catalytic APH Technology

It is possible that catalytic APH technology could play an increasingly important role as part of an overall system to achieve the lowest cost strategy for complying with NO_x regulations. The catalytic APH system selected for evaluation at the ASCR pilot plant is under development by ABB/Engelhard. As noted previously, only the hot element layer was replaced for this installation. Greater NO_x reduction capability is possible—at least in some cases—by extending the catalytically active length of the hot element. However, such an approach requires a complete intermediate and cold element replacement at a significantly increased cost.

Results of the initial tests to characterize the performance of the fresh catalyst at full-load design and operating conditions are shown in Figure 4. For example, NO_x removal and ammonia slip were 54% and 24 ppm, respectively, at an NH_3/NO_x ratio of 1.0. This relatively modest performance is not unexpected given the high space and area velocities associated with the system. Despite these apparent limitations, such performance, when coupled with other NO_x control technologies, could play an important role in providing the necessary degree of control at the lowest cost.

Future tests will compare catalyst performance for various operating conditions and evaluate catalyst deactivation as a function of time.

Visible Plumes

One of the most important objectives of the pilot plant tests was to collect data regarding possible impacts of the SCR system on plume visibility during oil-fired operation. This issue is important because PG&E's fossil power plants are often located in densely populated areas; as a result, operation with visible plumes is unacceptable. Furthermore, the plants do not incorporate particulate control devices such as precipitators or baghouses.

Since particulate control devices are not available, prevention of visible plumes can only be effected by controlling the basic chemical constituents leading to the condensed sulfate/bisulfate species (i.e., SO_3 and NH_3 slip) and, realistically speaking, only the NH_3. NH_3 slip, in turn, is determined by the allowable levels that are covered by the amount of catalyst and, ultimately, by the bid specification. It is clear that specifying these allowable levels too high can lead to the formation of plumes; however, specifying the levels too low can result in significantly added cost for the SCR system.

There is little doubt that SO_3 and NH_3 condense to form sulfates/bisulfates at temperatures around 400 °F. However, depending on the physical phenomenon occurring, the resulting condensed compounds may or may not contribute to plume visibility. If, for example, the condensation occurs in an environment with many condensation nucleation sites available (e.g., particulates that occur in oil firing), it is possible that little or no plume visibility impacts would occur; in effect, the condensation products formed a thin shell around existing particles.

However, if the particles condense via a homogeneous nucleation process, it is possible that highly visible submicron particles would result. The resulting fumes, which occur in the relatively particle-free environments of natural gas combustion, are known to yield, at least in some situations, particles in the 0.5-micron range. These particles are known to scatter light in the visible portion of the spectrum that is most sensitive to the perception of the human eye.

Because of the low particle concentrations, small path lengths, the non-steady nature of the tests involved, and the desire to focus on submicron particle sizes, conventional measurement techniques, such as impactors, were judged not to be suitable. After evaluating various alternatives, a methodology was adopted based on optical measurements of the particle concentrations using the Insitec[1] optical particle counter system. This technique yields particle size distribution data down to about 0.2 micron, which covers the portion of the visible spectrum of interest. In the next step of the methodology, these particle size distribution measurements are then converted into a visibility parameter using previously developed software techniques (PLUVUEII).

The key to this approach is to establish baseline particle size distribution—visibility values based on the normal operating conditions of the boiler, which are known to have invisible plumes. This baseline value could then be compared to that resulting from various SCR operating conditions (e.g., various levels of SO_3 and NH_3, sootblowing events, etc.). Actually, as shown below, calculation of the visibility parameter was not necessary since no meaningful changes in particle emissions characteristics could be observed due to the presence of the SCR system.

Typical results from the plume tests at the ASCR pilot plant are illustrated in Figure 5. This figure presents the sub-2 micron (i.e., visible portion) particle concentrations as a function of time for two operating situations. The upper curve shows the baseline case without any NH_3 or additional SO_3 injection. The peaks in the particle counts reflect the various boiler sootblower events as indicated. The lower graph presents the same information, but for the ASCR system operating (in this example, NH_3 = 8 ppm and SO_3 = 11 ppm). These levels of SO_3 and NH_3 are known to have a potentially large impact on plume characteristics (for example, if the NH_3 and SO_3 were quantitatively converted to 0.5 micron NH_4HSO_4 particles, the resulting particle count would increase by approximately 287,000 particles/cc). Note that the ASCR sootblower system is inoperative in both cases. As can be seen, within the repeatability of the sootblowing procedures, no increase in particulates, and hence plume visibility, is observed. Additional tests were conducted to evaluate the impacts of catalyst, and APH, sootblowing. Similar results were obtained.

Based on these results, it can be concluded that for the sootblower designs and operating procedures used by PG&E, levels of SO_3 and NH_3 up to about 10 ppm are acceptable, and no visible plumes will be created.

Future Activities

As indicated earlier, future research activities are aimed at two major thrusts: developing/ evaluating true in-duct ASCR, and catalytic APH systems into commercially viable options. In addition, PG&E R&D is closely monitoring the progress of so-called "ultra-low NO_x burners." Although major technical barriers must be overcome, success in this area presents major additional cost savings opportunities.

Conclusions

Based on testing at the ASCR pilot plant test facility, the following conclusions have been reached:

- At up to 20 ppm of NH_3 and SO_3 entering the APH, results demonstrate that the current sootblower design and sootblowing operating procedures are effective in maintaining the pressure drop across the APH.

- A greater than two-fold increase in pressure drop was observed across the catalyst beds in the absence of sootblowers and turning vanes during the relatively short oil burn due to oil-fired particulate deposition. In a full-scale PG&E application, this increase in pressure drop would have resulted in a major capacity derating estimated to be on the order of 40%. Subsequent testing with a sootblowing system permitted operation for several months with no pressure drop increase.

- Initial results of catalytic APH performance are encouraging. Future tests will evaluate the long-term activity.

- For the APH/sootblower designs and operating procedures currently in operation, NH_3 and SO_3 levels up to 10 ppm will not impact plume visibility.

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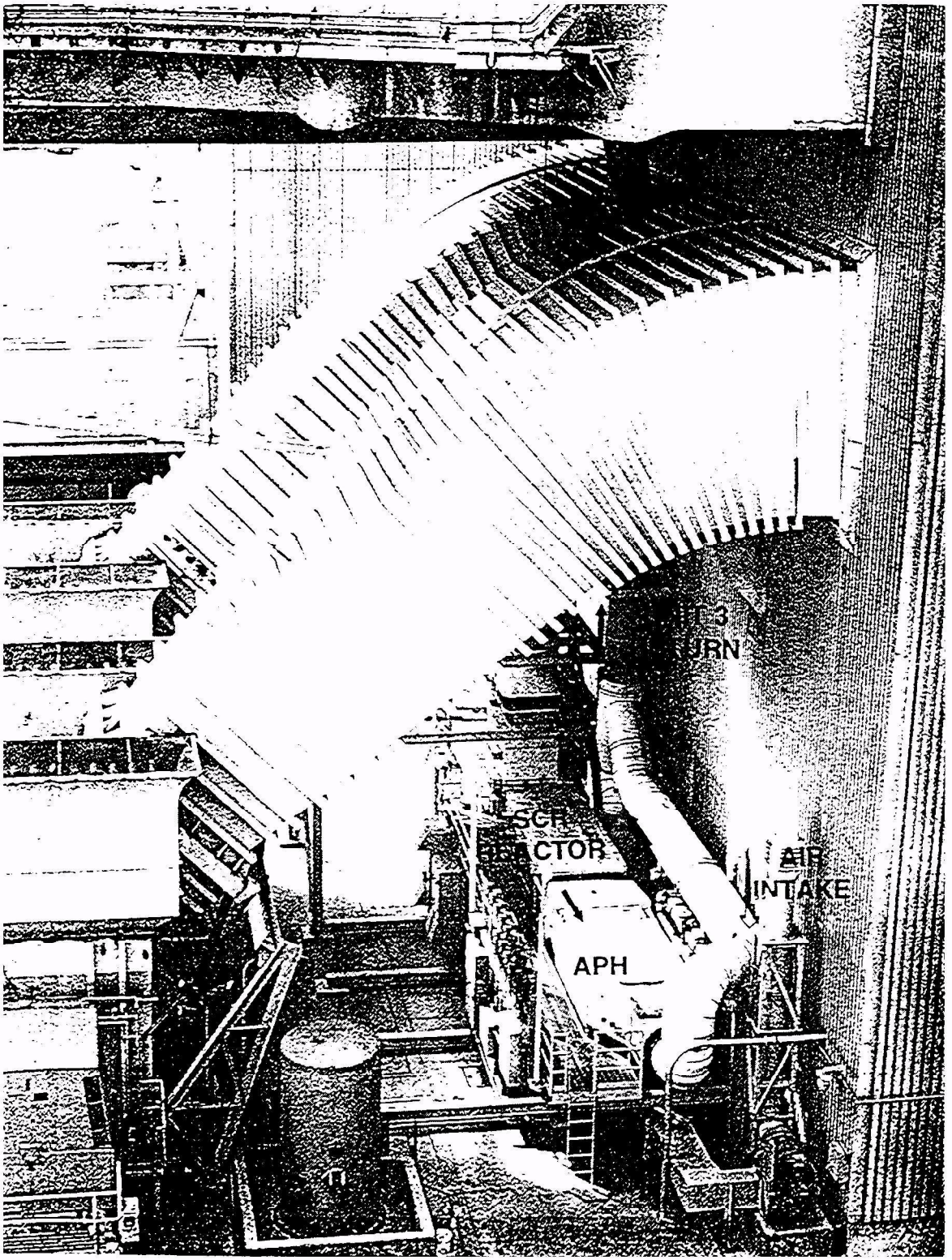


Figure 1
Morro Bay ASCR Pilot Plant

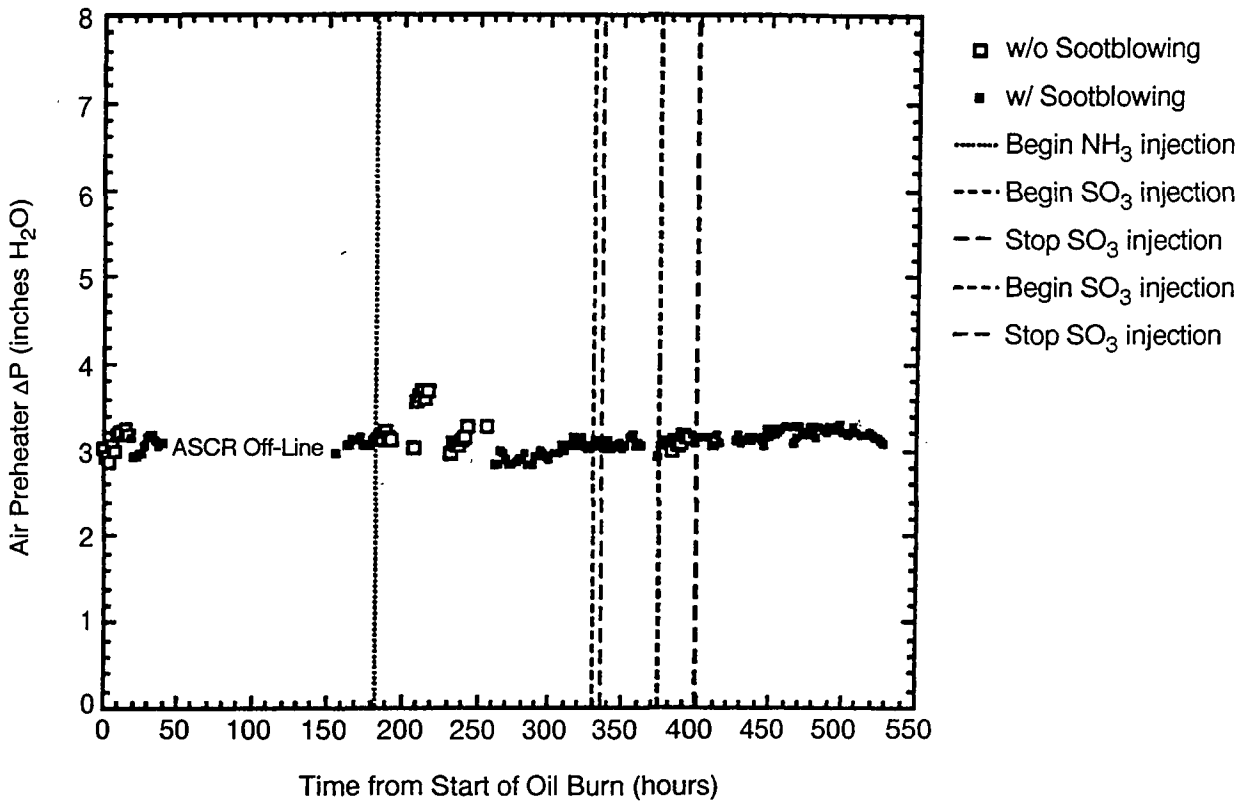


Figure 2
APH Pressure Drop vs. Time

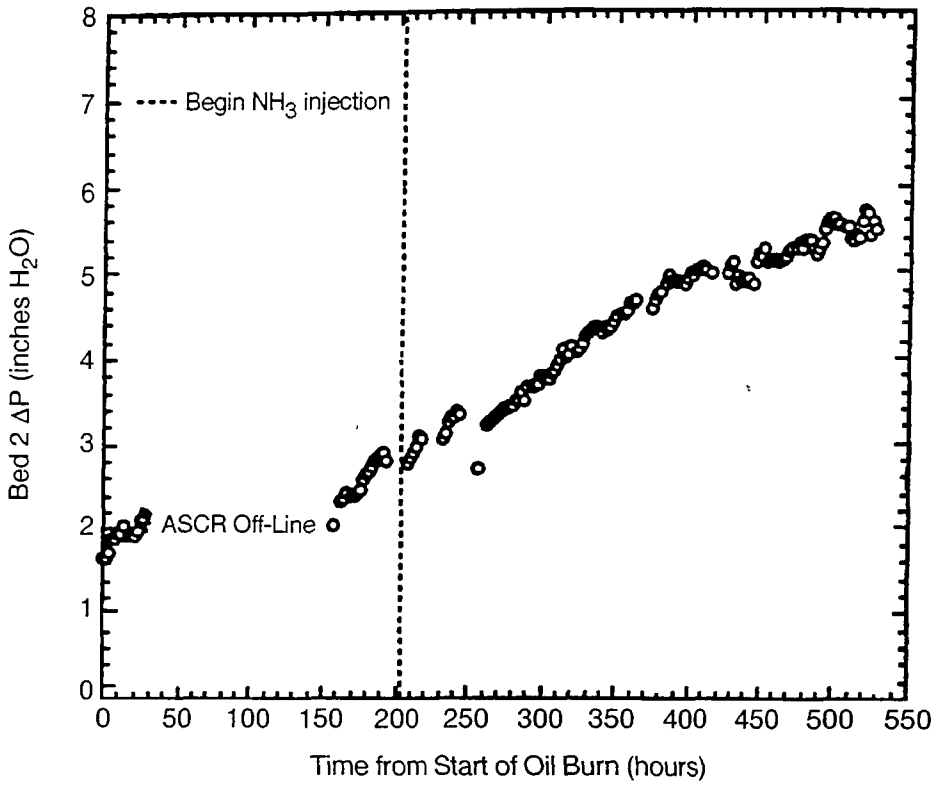
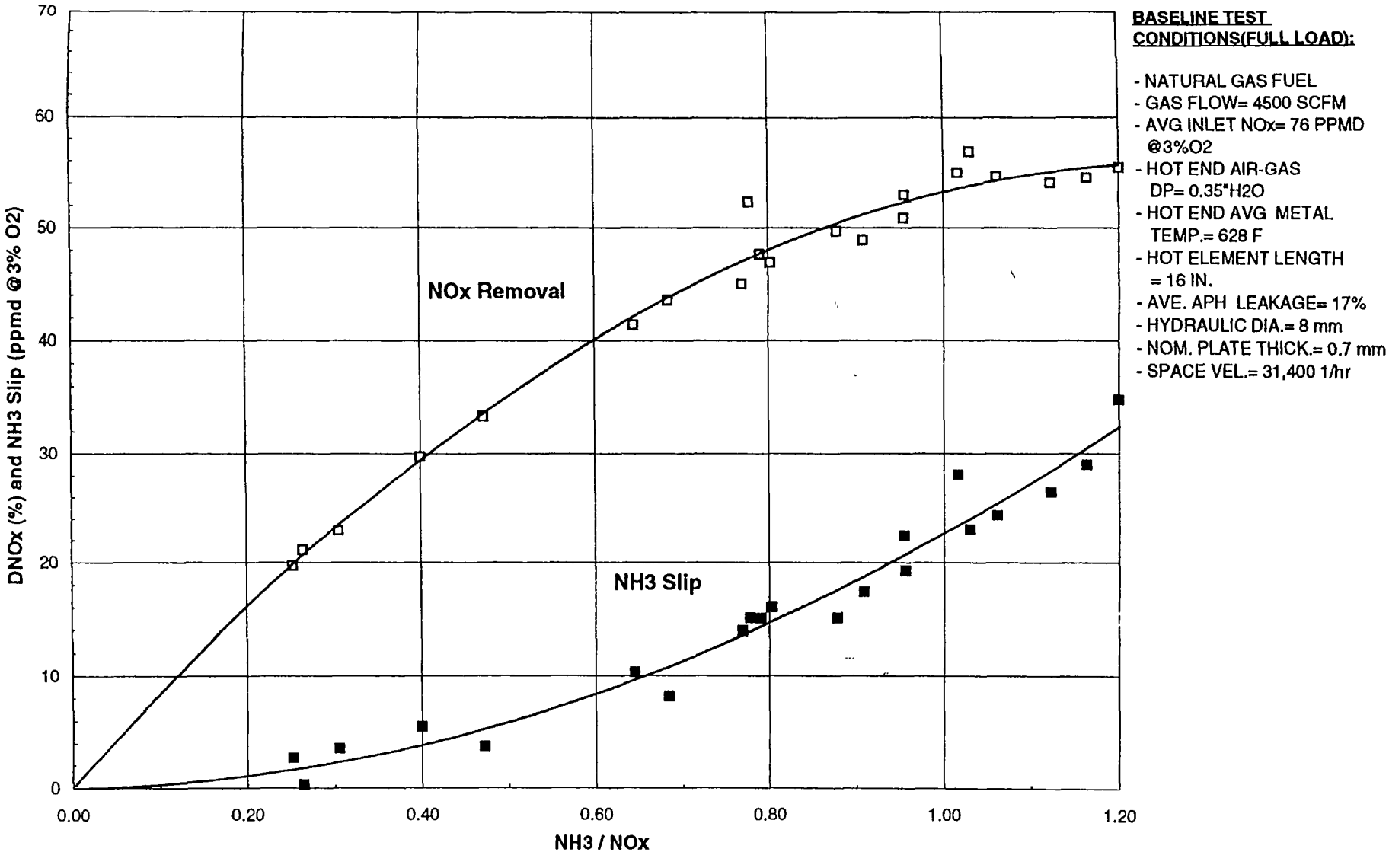


Figure 3
Bed 2 Pressure Drop during Oil-fired Operation



INITIAL PERFORMANCE OF ABB/ENGELHARD CATALYTIC AIR PREHEATER, LCAP™

Figure 4

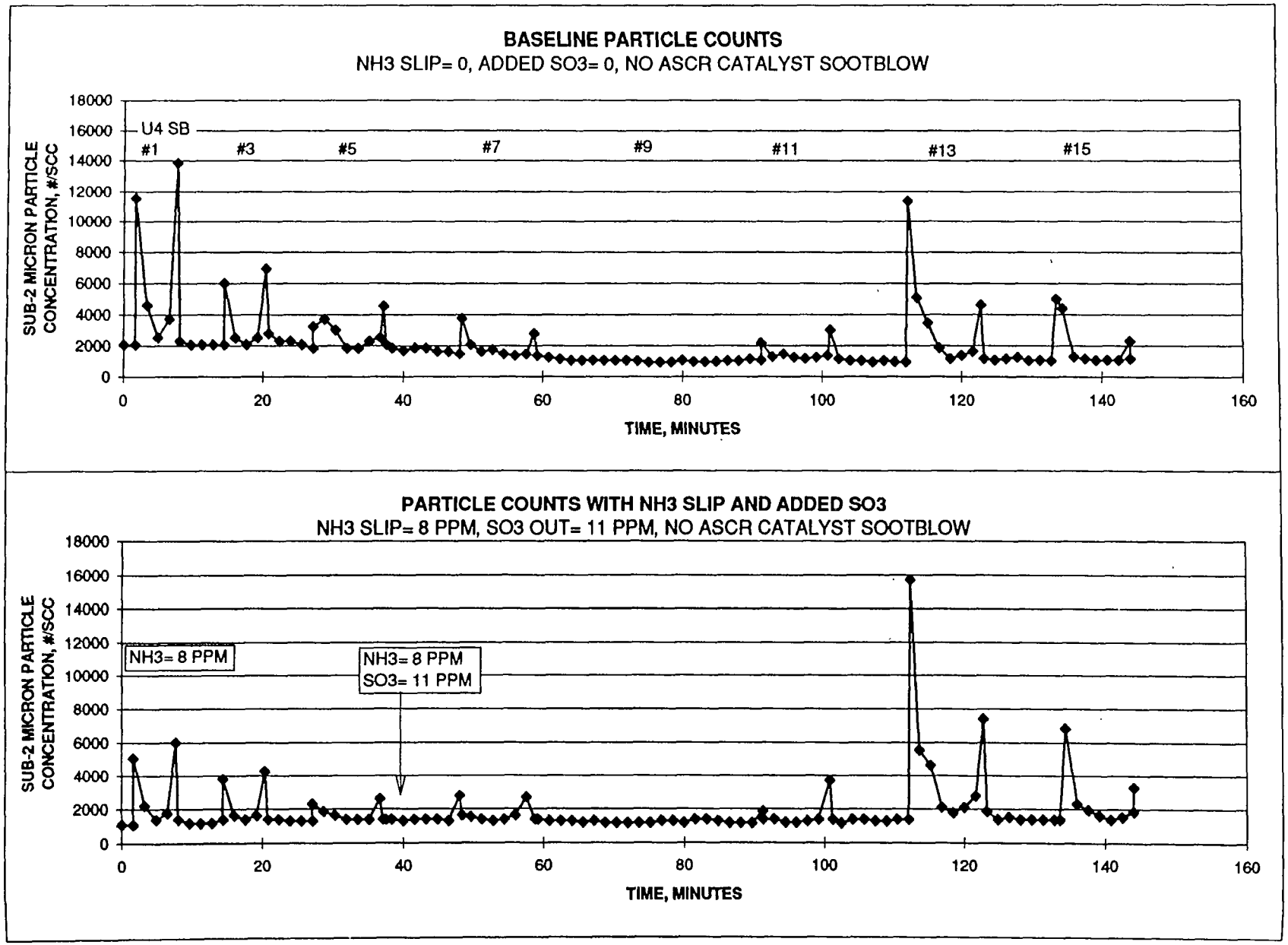


Figure 5
Effect of NH_3 Slip and SO_3 on Plume Visibility - No ASCR Catalyst Sootblow

A NEW DESIGN TOOL FOR SCR SYSTEMS

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Abstract

A new technique to support the design of SCR systems for natural gas, oil, or coal applications, and accurately predict the performance (NO_x removal and NH_3 slip) in commercial applications is described. The approach takes into account features such as actual duct layout and ammonia injection grids. The procedure is based on integrating two major efforts: laboratory cold flow modeling tests; and "ideal" (i.e., uniform ammonia and velocity distribution) catalyst performance characteristics. A key feature is the inclusion of the NH_3 injection process in the flow model by use of tracer gas techniques. Results of detailed test to validate the design procedure based on data from a laboratory cold flow model and the PG&E-EPRI 3MW ASCR pilot plant are presented. Use of the design technique in commercial applications is discussed.

Introduction

To meet more stringent NO_x emission regulations, the utility industry is moving in a direction of utilizing Selective Catalytic Reduction (SCR), in addition to combustion modifications, as a way of achieving compliance. At the same time, efforts are being made to reduce costs by exploring "in-duct" SCR arrangements. In-duct SCR arrangements incorporate the SCR reactor basically along the existing flue gas path from the economizer to the air preheater.

In the extreme, there would be no changes to the ductwork for an "in-duct" SCR and the catalyst inserted within the existing ductwork, so-called "true in-duct SCR". True induct systems are likely to be most applicable when used in conjunction with other NO_x controls (e.g., low NO_x burners, SNCR, air preheater catalyst) where the SCR NO_x reduction requirements is reduced to the range of 60-80%. In many cases, "in-duct" arrangements will involve enlarging the ductwork to lower velocities, accommodate a larger volume of catalyst, and reduce pressure drop. This "in-duct" concept is contrasted to a more traditional retrofit approach associated with Japanese or German SCR installations, where a large separate SCR reactor is built with fairly extensive ductwork changes to route the flue gas to and from the SCR reactor.

While the "in-duct" arrangement can potentially reduce retrofit costs, it also poses greater challenges to the designer. In addition to the type and amount of catalyst, the process designer must also address velocity uniformity, and NH_3/NO_x uniformity across the catalyst face. In the traditional separate SCR reactor approach, there is usually sufficient ductwork to allow a uniform velocity to be generated at the ammonia injection grid. A uniform velocity profile at the ammonia injection grid, greatly simplifies the ammonia/NO_x mixing process. Likewise, with the separate SCR reactor, there will be sufficient space to design ductwork expansions and turning vanes to meet velocity uniformity requirements at the catalyst face. As a consequence, previous design approaches specified criteria such as the standard deviation in the velocity profile, or NH_3/NO_x profile, at the catalyst face. If these criteria are met, then the SCR system will perform as designed.

With an "in duct" arrangement, it may not be possible to: (1) provide a uniform velocity at the ammonia injection grid, or (2) accommodate traditional design guidelines in terms of duct expansion angles, etc. As a consequence it will be more difficult, if not impossible in some instances, to meet traditional criteria in terms of a velocity uniformity and/or NH_3/NO_x uniformity at the catalyst face. However, this does not mean that an "in duct" arrangement should be discarded as a viable approach. Rather, it points to the need to develop better engineering design tools to assess the performance of these systems.

Before discussing the specific objectives of this work, it is of value to briefly discuss the consequences of non-uniformalities of both, velocity and NH_3/NO_x across the catalyst face. The impact of velocity uniformity depends on the type of fuel being fired. If the unit fires only natural gas, then the primary impact of velocity non-uniformities is varying local space velocities across the catalyst face. Regions where the velocity is higher than average, correspond to high space velocity regions; conversely, low velocity regions correspond to locally low space velocity regions. Since high and low space velocity regions tend to compensate for one another velocity non-uniformity tends to have only a small effect on SCR NO_x removal and NH_3 slip.

If the fuel being burned is oil, or coal, where the flue gas is laden with particles, the velocity impacts for the SCR are different. In these cases, velocity uniformity is more important. Regions of high velocity can lead to accelerated catalyst erosion. Also, for the case of a particulate laden flue gas, it is important to align the particles as parallel as possible with the catalyst channels, in order to minimize deposition. This can be accomplished using tuning vanes or other low cost techniques.

The NH_3/NO_x distribution across the catalyst face has a much greater impact on SCR performance. While high and low velocity regions tend to compensate one another, this is not the case with the NH_3/NO_x distribution. An imbalance in the NH_3/NO_x distribution can lead to increased NH_3 slip. In fact, if the NH_3/NO_x distribution has significant regions with NH_3/NO_x ratios greater than unity, overall performance may not be improved to satisfactory levels by increasing the amount of catalyst (i.e., decreasing space velocity).

This paper describes a new methodology to support the designed SCR system. The method integrates the results of cold flow modeling with an SCR process model to quantitatively account for site specific conditions. The paper will describe the methodology, outline the approach to

cold flow modeling, provide validation of the method, and briefly discuss the use of the methodology for a full-scale SCR design.

SCR Process Design Methodology

Physical cold flow modeling has been used previously during the design of SCR systems. In most cases, the cold flow modeling has focused on developing uniform velocities at the catalyst face and/or at the plane of the ammonia injection grid, and reducing pressure drop. In some instances, tracer gas techniques have been used to simulate the ammonia injection process. To provide more input to the SCR design process a methodology has been developed that integrates cold flow modeling results along with a model of the SCR process to predict performance for a given specific configuration.

The process model integrates: (1) the cold flow velocity distributions, which define local variations in space velocity, (2) tracer gas results, which define local variations in NH_3/NO_x ratio, and (3) ideal catalyst performance data (i.e., NO_x removal and NH_3 slip versus NH_3/NO_x and space velocity) to calculate overall NO_x removal and NH_3 slip for a given catalyst and NH_3 injection configuration. The model also estimates the pressure drop across the catalyst.

Figure 1 shows the type of ideal catalyst performance data used as an input to the calculations. It should be emphasized that the model calculates the performance for a specific SCR configuration relative to this ideal catalyst performance. The model currently incorporates a generic representation of a titania/vanadia SCR catalyst. For an actual full scale design, a higher degree of confidence in the absolute performance values would be achieved if data for the specific catalyst were available (either from the catalyst vendor, or independently-developed test data).

Cold Flow Modeling Principles

Cold flow modeling is widely used throughout the industry to address a number of flow problems in ducts, ESP's, furnaces, SNCR systems, etc., and the underlying foundation is fairly well established. The same modeling principles are used for SCR systems, although the presence of the catalyst and its pressure drop characteristics impose some additional requirements.

The following issues need to be addressed in cold flow modeling of SCR systems:

- Overall geometric similarity.
- Dynamic similarity including the catalyst pressure drop.
- Modeling the NH_3 injection process.

Each of these is addressed below:

Geometric Similarity

The cold flow SCR models are geometrically scaled (typically 1 to 12). This includes all ductwork, SCR reactor, NH_3 grid, etc. The model should also include all internal duct structures that would impact the overall flow; typically this includes structures larger than 6-8 inches.

Dynamic Similarity

Cold flow modeling is based on the assumption that the overall flow patterns will be similar between the model and full scale if the model flow is fully turbulent. As such, it is not necessary to operate the model at the same Reynolds number as full scale; rather, the cold flow models are operated at Reynold's numbers greater than 20,000 [1].

In most cold flow modeling applications, dynamic similarity is entirely met by maintaining the Reynold's number criteria (Re > 20,000). This is not the case with an SCR system, however, due to the pressure drop of the catalyst. The catalyst pressure drop results in a redistribution of the velocity at the catalyst face. The amount of velocity redistribution depends on amount of pressure drop relative to the dynamic pressure of the flow field (i.e., $1/2\rho V^2$). This is characterized by the pressure coefficient (PC).

$$PC = \frac{\Delta P_{cat}}{\frac{1}{2}\rho V^2} \tag{1}$$

To insure that this is correctly modeled, the cold flow model is designed such that the pressure drop across the simulated model catalyst is identical to the projected pressure drop across the full-scale catalyst.

$$\Delta P_{\text{model cat}} = \Delta P_{\text{full-scale cat}} \tag{2}$$

Additionally, the model is operated such that the dynamic pressure of the model is identical to the dynamic pressure of the full-scale system. This means that the model and full-scale pressure coefficients are equal.

NH_3 Injection Process

An important part of the SCR system is the NH_3 injection grid and the mixing between the NH_3 and the combustion products. The NH_3 mixing process depends on four basic parameters: location of the grid in the flue gas flow path, geometry of the grid (number of injection holes, size), amount of carrier gas used to transport the ammonia, and the resulting momentum of the jets leaving the injection holes. Because of density differences between the model and full scale, it is not possible to simulate all of the above at one time. To simulate the NH_3 injection process, two criteria are adhered to. First, the NH_3 injection grid is geometrically scaled. Second, to best

simulate the mixing process, the model ammonia injection grid is operated at the same jet momentum ratio as full scale:

$$J = \frac{\rho_{jet} V_{jet}^2}{\rho_{flue\ gas} V_{flue\ gas}^2} \quad (3)$$

Another issue that needs to be addressed is the flow conditions inside the injection grid. Ideally, each injection lance in the grid should function as a plenum with equal flow from each individual injection hole along the lance. The degree to which this occurs can be characterized by the ratio of the average velocity out of an injection orifice to the average velocity at the entrance to the lance.

$$VSR = \frac{V_J^2}{V_{lance}^2} \quad (4)$$

If this ratio is less than 10-15, unequal flow can result from each of the orifices. Alone, this may or may not be an issue for a particular SCR injection grid. However, it does raise an issue in terms of modeling the actual injection grid. At full scale, the air/ammonia or flue gas/ammonia mixture will heat up as it flows from the entrance of a lance to the last injection orifice at the exit of the lance. The resulting density gradient along the lance will affect the flow rates from the individual injection orifices. Techniques to account for this phenomenon in cold flow ammonia injection grid design have been developed and verified.

Test Facilities

The test work to validate the design methodology utilized PG&E's advanced SCR pilot facility located at the Morro Bay Power Plant, along with a physical cold flow model of the pilot plant. Only a brief description of the ASCR facility is provided in this paper as it is described in detail in another paper at this symposium[2].

ASCR Pilot Plant

The ASCR pilot plant draws a 5000 scfm slipstream from either Morro Bay Unit 3 or Unit 4 boilers. The design of the pilot plant is a 1 to 11 scale simulation of an in-duct SCR concept for a typical large PG&E fossil plant. As such, the pilot plant simulates boiler components from the economizer inlet through the outlet of a Ljungstrom regenerative air preheater.

Figure 2 shows a plan and side view of the reactor. The basic reactor starts just above a simulated economizer. The slip stream flows through a series of perforated plates that are used to simulate the full-scale economizer. The flue gas proceeds through a right angle turn then passes through a relatively high velocity "throat" region. Exiting the throat, the reactor undergoes an expansion in order to reduce the velocity and accommodate the catalyst. As can be seen in Figure 2, the reactor expands both vertically and horizontally. In the horizontal direction

the entire expansion occurs along one wall. This is done in order to accommodate the existing location of the air preheaters in the full scale system (the ASCR pilot plant simulates one duct of the full scale system). The inside dimensions of the reactor are 24.5 inches high by 45 inches wide. The actual ASCR reactor section that holds the catalyst is longer than would be for the full scale system. This was done to accommodate up to three layers of catalyst and to provide for experimental flexibility in the test program.

The SCR reactor currently incorporates two layers of a Haldor Topsøe SCR catalyst. Each layer is nominally two feet high by four feet wide with a depth of 1.67 feet (0.5 m). The catalyst cell opening has a hydraulic diameter of 3.3 mm and a nominal wall thickness of 0.5 mm. At the design flow rate of 5000 scfm, the two layers of catalyst provide a space velocity of about 13,000 hr^{-1}.

Exiting the reactor, the slip stream passes through a Ljungstrom® regenerative air preheater, similar to full scale and is then returned to either Unit 3 or Unit 4.

For this study aqueous ammonia (NH_4OH) is used to supply NH_3 to the ASCR pilot plant at the following locations:

- upstream of the flue gas boost fan,
- downstream of the simulated economizer,
- at the "throat" region,

At the first location upstream of the gas fan, a twin fluid atomizer is used to spray the NH_4OH directly into the slipstream. This injection location provides the highest degree of mixing as the NH_4OH vaporizes and flows through the fan and serves as the "baseline" injection configuration.

For NH_3 injection at the other locations, the NH_4OH is prevaporized in a heated air stream, and injected into the reactor, along with the carrier air through a simulated ammonia injection grid.

Cold Flow Model

To support the original design of the ASCR pilot plant and the ongoing test work, a one half physical scale model of the ASCR pilot plant was designed and fabricated.

The model was constructed mostly from clear acrylic. It encompassed the entire ASCR pilot plant from the fan outlet to the air heater inlet. This included the circular inlet duct, the economizer and economizer hopper, catalyst bed, and the inlet transition duct to the air heater. A schematic of the model is shown in Figure 3. A 5000 scfm blower operating in an induced draft mode was used to provide air flow to the model.

Validation Test Results

Validation of the design methodology was done through a series of tests at the ASCR pilot plant and on the cold flow model. This section of the paper will present the validation results in the following order:

- First, detailed velocity profiles at the exit of the first bed of catalyst (Bed 1) will be presented and compared to comparable measurements made in the physical cold flow model.

- Next, a number of ammonia injection scenarios will be discussed in terms of the resulting uniformity of the NH_3/NO_x ratio across Bed 1. Again, a direct comparison will be made to comparable tracer gas tests performed in the physical cold flow model.

- Finally, the performance predictions will be presented which integrate the cold flow modeling data (or NH_3/NO_x distributions at the pilot plant) with a SCR process model.

One of the primary objectives of this test effort was to validate the cold flow modeling by direct comparisons to the pilot. To make this comparison as direct as possible, the measurements at the ASCR pilot plant and cold flow model used the same number, and layout, of sample points to develop the profiles; 27 points total, 3 vertical locations with 9 points horizontally. While the number and location of sample points was comparable, basic differences in the two facilities dictated that some of the measurement techniques differ. These differences will be discussed, where appropriate.

Detailed Velocity Measurements

During this test series, detailed velocity measurements were made at the exit of catalyst Bed 1 at the design flow rate of the ASCR pilot plant. At the design flow rate of 5000 scfm the pressure coefficient in the pilot plant is 30. The physical cold flow model was operated at a flow rate to yield, as close as practical, a similar pressure coefficient.

In the cold flow model, the velocity measurements were made using a thermal anemometer which has good resolution and linear response over the velocity range of interest. However, the thermal anemometer cannot operate at the near 700°F temperatures of the pilot plant. This necessitated the use of a 1/4"-S-type pitot probe to obtain the detailed velocity profiles under "hot" flow conditions at the pilot plant. An electronic pressure transducer with a full scale range of 0.5 inches of H_2O was used with the S-type pitot probe.

The detailed velocity profiles from the two facilities are shown in Figure 4, both contour plots and wireframe diagrams are included. Good agreement is seen between the two profiles. A high velocity region is seen near the center, with a low velocity region on the right hand side, which represents the region where there is a horizontal expansion as the gas exits the throat region and transitions to the catalyst. Quantitatively, the standard deviation in the pilot plant velocity measurements was 39 percent at full load (PC-30). The comparable cold flow model test (PC=35) yielded a standard deviation in velocity of 27 percent. As will be discussed later, the slightly higher pressure coefficient, 35 versus 30 should decrease the velocity standard deviation by about five percentage points.

Figure 5 summarizes the effect of the pressure coefficient on velocity uniformity at the exit of Bed 1 from a number of tests. Included are data from the physical cold flow model, and pilot plant (both with hot flue gas and ambient air). As expected, the velocity uniformity increases as the pressure coefficient increases (i.e., higher pressure drop per unit of velocity head in the

flow). The effect of pressure coefficient on velocity uniformity is quite steep below PC=60. Since most full scale system will likely operate with pressure coefficient greater than 50-60, the inherent velocity uniformity should be below 20%. Note, the design pressure coefficient for the two beds of catalyst at the ASCR pilot plant is 60.

Ammonia Injection

The next portion of the test program addressed ammonia injection and the impact that various ammonia injection parameters have on the NH_3/NO_x uniformity at the catalyst face. It should also be emphasized that the purpose of this test series was not to develop an optimized injection system for the ASCR configuration, but, rather, to investigate the ammonia injection process and validate the cold flow modeling techniques. With this goal, the ammonia injection parameters were chosen to primarily show changes in NH_3/NO_x distribution, in order to validate the cold flow modeling approach. The parameters that will be addressed include:

- Injection location (ahead of the gas fan, economizer exit, throat)
- Number of injection holes
- Amount of transport air and injection momentum

Before presenting the results, it is worthwhile to briefly review the measurements that were made and the manner in which the data will be presented. For each test condition, the primary goal was to determine the NH_3/NO_x distribution across the catalyst face; all measurements were made downstream of Bed 1, which represents a space velocity of 26,000 hr^{-1}.

At the pilot plant, the inlet NO_x distribution is uniform across the catalyst face. The local NH_3/NO_x ratio was determined by measuring the local NO_x removal across the catalyst along with the local NH_3 slip. Assuming that the injected ammonia either reacts with NO_x (1 mole of NH_3 per mole of NO, and 1.33 moles of NH_3 per mole of NO_2), or exits the catalyst as slip, the local NH_3/NO_x ratio is calculated as

$$\left(\frac{NH_3}{NO_x}\right)_{local} = \frac{\left[\Delta NO + 1.33\ \Delta NO_2 + NH_{3\ slip}\right]_{local}}{NO_{x\ inlet}} \qquad (5)$$

The cold flow model tests were done by injecting a tracer gas into the simulated injection flow and measuring local tracer gas concentrations at the exit of Bed 1 in the cold flow model. The normalized tracer gas concentration represents the local NH_3/NO_x ratio for an overall NH_3/NO_x ratio of unity.

Baseline Ammonia Injection. To demonstrate a uniform NH_3/NO_x profile across the catalyst face, a test was conducted injecting the ammonia upstream of the gas fan. The mixing through the fan provides good mixing between the ammonia and the slipstream. This data set also serves as a baseline data set where the NH_3/NO_x ratio is nearly uniform, and the only impacts on SCR performance are due to velocity non-uniformities. The results of this baseline

test are not shown but the NH_3/NO_x distribution was quite uniform exhibiting a very small standard deviation in the NH_3/NO_x profile of 2 percent.

Economizer Exit Injection: Variation in the Number of Holes. Three injection grids were investigated at the economizer exit location, which varied the total number of injection holes (15, 50 and 100 holes). Each injection grid consisted of five lances at the economizer exit which entered the ASCR from the left hand side. For these three tests, the amount of transport gas and the momentum ratio, J, was held constant. This necessitated decreasing the hole diameter as the number of holes increased. The results of these tests are shown in Figure 6. Overall, both the pilot scale and cold flow results show an increase in uniformity as the number of injection holes was increased from 15 to 50 (i.e., 3 injection holes per lance increased to 10 per lance). A further increase to 100 holes showed no further improvement.

For the 15 hole injector tests the cold flow model results showed a higher overall non-uniformity in the NH_3/NO_x ratio than the pilot scale tests. This is evident looking at the detailed profiles in Figure 6, where the cold flow model exhibits a region of lower NH_3/NO_x ratio on the left hand side. This is the side where the injectors penetrate the reactor. This general characteristic is attributed to flow dynamics inside each lance. In the cold flow model, the flow inside the lance is isothermal, while in the pilot plant the ammonia-air mixture inside the lance heats up as it flows through the lance. Since these particular lances were designed with a VSR=6 (see Equation 4) the non-isothermal flow in the pilot plant lance behaves somewhat differently than the cold flow lances. This effect is particularly apparent for the fifteen hole case shown at the top of Figure 6. With only three injection holes per lance, any flow maldistribution effects would be expected to have a larger effect on the tracer gas profiles. For this case, the vertical contour lines (Figure 6(a)) indicate that a smaller amount of a tracer gas is being injected on the left hand side near the lance inlet. This local stratification leads to the higher standard deviation (i.e., 21% versus 14%) in the cold flow profiles, compared to the pilot plant NH_3/NO_x profiles.

Throat Injection: Effect of Momentum Ratio. Next, a 15 hole injection grid was set up at the "throat" region, just ahead of the duct expansion out to the catalyst dimensions. While this is a relatively high velocity region, just downstream of a 90° turn, there are some practical reasons why this could be an attractive location for the ammonia injection grid. In the PG&E units which utilize FGR for NO_x control, the FGR takeoff is generally in the vicinity of the economizer exit. To avoid injecting NH_3 into the FGR stream, it may be desirable to locate the ammonia injection grid downstream of the FGR takeoff.

The results of these tests are shown in Figure 7 where the momentum ratio was varied from 70 to 150 (corresponding to ratios of injection air to flue gas of 0.5 - 0.75 percent). The contour plots show that the pilot and cold flow profiles are generally similar. The peaks and valleys are in the same locations, and there is excellent quantitative agreement in terms of the overall standard deviation of the NH_3/NO_x profiles.

SCR Performance Predictions

The velocity profiles and NH_3/NO_x profiles obtained from the cold flow model and/or pilot plant can now be used along with the process model to predict SCR performance. By utilizing the process model, a direct assessment can be made of the impact of velocity and NH_3/NO_x non-uniformities on performance (NO_x removal and NH_3 slip). To accomplish this comparison, the NH_3/NO_x profiles (cold flow tracer gas, and ASCR pilot plant) were put into the process model and the performance calculated. All of the calculations presented in this section were performed for an initial NO_x concentration of 75 ppm (dry @ 3% O_2) and a space velocity of 26,000 hr^{-1}. The ideal catalyst performance shown in Figure 1 was used. To reiterate, the model basically calculates the effect of both velocity and NH_3/NO_x uniformities on performance, relative to an ideal performance curve for the catalyst.

Performance predictions based on the pilot scale velocity and NH_3/NO_x profiles for the baseline test (NH_3 injection upstream of the gas fan) are shown in Figure 8. Recall that the standard deviation in velocity is 39% (Figure 5), and that the NH_3/NO_x uniformity is quite good for this case (2% standard deviation). In Figure 8, three curves are shown for both the predicted NO_x removal and NH_3 slip. The solid line represents the ideal performance at a space velocity of 26,000 hr^{-1} (i.e., perfectly uniform velocity and NH_3/NO_x). The line through the triangle points is the baseline prediction corresponding to the 39% standard deviation in velocity and 2% standard deviation in NH_3/NO_x ratio. For comparison, a calculation was performed using the velocity non- uniformity (39% standard deviation) but assuming perfectly uniform NH_3/NO_x (the curves through the square symbols in Figure 8). Comparing this latter calculation to the ideal performance illustrates the effect of velocity non-uniformities alone. As can be seen, even with a large velocity non-uniformity, there is a small impact on performance. At an NH_3/NO_x ratio of 0.9, the NO_x reduction is only decreased by about 2 percentage points and NH_3 slip increased by nominally 1 ppm due to the velocity non-uniformity. A small non-uniformity in the NH_3/NO_x ratio (2% standard deviation) has almost the same impact on performance as a 39% standard deviation in velocity. This clearly illustrates that in the SCR design process, NH_3/NO_x uniformity has a far more important impact on performance than velocity non-uniformities.

The performance predictions for the economizer injection configurations (varying the number of injection holes, from 15, 50 and 100) are shown in Figure 9. Figure 9 shows three separate figures for the 15, 50 and 100 hole case, respectively. Each of the figures contains curves for the ideal performance, performance based on the cold flow tracer gas results, and performance based on the velocity and NH_3/NO_x profiles from the pilot plant. These results show very good agreement between the pilot and cold flow model predictions for all three cases. The other observation that is noteworthy, is the effect of the number of injection holes on the performance. An improvement in performance can be seen when the number of injection holes was increased from 15 to 50. No further increase in performance is predicted when the number of holes is increased to 100.

A further example of the performance prediction methodology is shown in Figure 10 for the case of injection at the throat location, varying the injection momentum (i.e., amount of transport air). For these injection scenarios, the NH_3/NO_x non-uniformities were higher than for the economizer exit configurations. As a result, particularly for the J=70 case, there is a marked

departure from ideal performance (both in NO_x reduction and NH_3 slip). As the injection momentum is increased, the performance is seen to increase. Again, excellent agreement is seen between the cold flow model and pilot scale.

These results clearly show the utility of using this methodology to help guide the design of SCR system. With the current methodology, design approaches can be tested, and the impact on performance quantified. Thus the designers can revisit previous rule-of-thumb design criteria such as requiring velocity or NH_3/NO_x uniformity to be below a certain standard deviation, or requiring a set number of injection holes per unit duct area. This should allow more aggressive SCR concepts to be confidently pursued, whereas previously they may have been discarded for violating design criteria which evolved from prior empirical experience.

Full Scale Application

The design methodology described in this paper has been applied in a number of ways. It has been used to aid in the detailed design of full scale SCR systems and it has also been used as a screening tool to evaluate various retrofit concepts. For example the procedure was used to assess a retrofit concept of an "in duct" SCR for a utility boiler. The initial concept utilized a catalyst space velocity of 11,800 hr^{-1} and had a target performance goal of 90% NO_x removal with 10 ppm slip.

The cold flow modeling showed a standard deviation in the velocity profile across the catalyst of 10 percent. With a particular ammonia injection grid design the performance shown in Figure 11 was calculated. As can be seen in Figure 11 the predicted performance is under the target of 90% NO_x removal with 10 ppm slip. For this particular case, pressure drop was not an issue and the target performance goals could be met by increasing the depth of the catalyst. This is illustrated in Table 1 where the model was used to assess the effect of decreasing the space velocity; reducing the same velocity to 10,000 hr^{-1} achieves the required performance goals.

Table 1

Full Scale Retrofit: Effect of Increased Catalyst Depth on Performance

Space Velocity hr^{-1}	ΔNO_x % @ 10 ppm NH_3 slip	ΔP in. H_2O
11,800 (design)	87	0.93
11,000	89	0.99
10,500	90	1.04
10,000	91	1.09

Discussions and Conclusions

The paper addressed a new SCR design methodology. The new methodology integrates the results from physical cold flow modeling with an SCR process model. The procedure easily allows the effects of velocity and NH_3/NO_x nonuniformities on SCR performance to be quantified. With this new methodology, the designer does not need to rely on previous "rules-of-thumb" engineering criteria. For instance, for a natural gas-fired unit is it really necessary to engineer the system to have a velocity profile with a standard deviation of 10%? Likewise, specification on the performance of the ammonia injection grid may be relaxed in some cases in favor of an overall specification on performance. For instance, injection of ammonia at the throat location may be dismissed based on prior design criteria. Now its feasibility can be investigated and quantified. By offering a more quantitative approach for SCR design it is expected that this will also translate into more aggressive, and therefore lower cost, designs.

References

1. a. Patterson, R.C., Abrahamsen, R.F., *Flow Modeling of Furnaces and Ducts,* Combustion, March 1962.
   b. Industrial Gas Cleaning Institute Inc., *Gas Flow Model Studies,* Publication No. EP-7.

2. Teixeira, D.P., et al, *Results of Catalyst Tests at the PG&E - EPRI ASCR Pilot Plant,* presented at the 1995 Joint EPRI/EPA Stationary NO_x Symposium, Kansas City, Missouri, May 16-19, 1995. [conference paper]

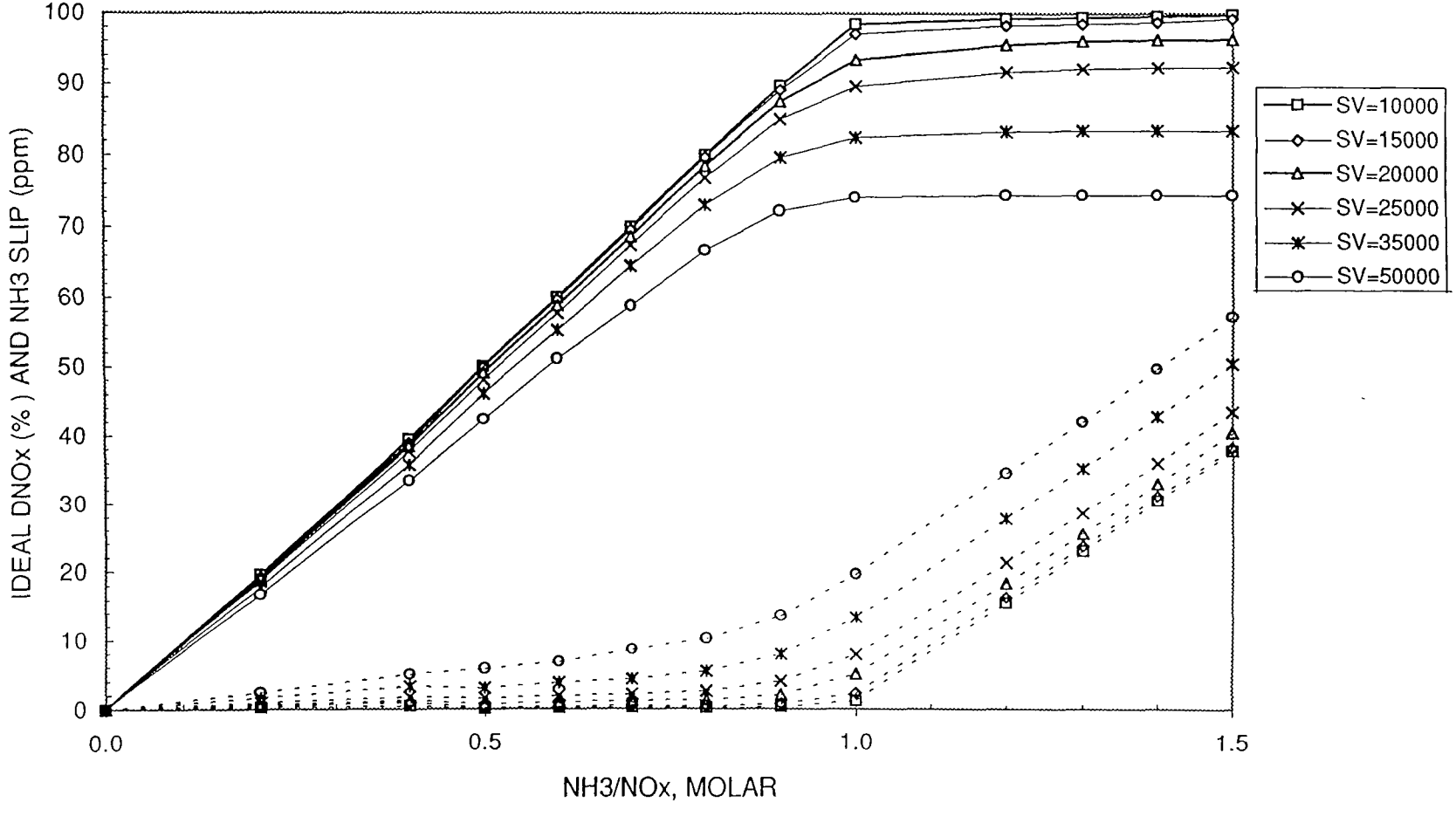
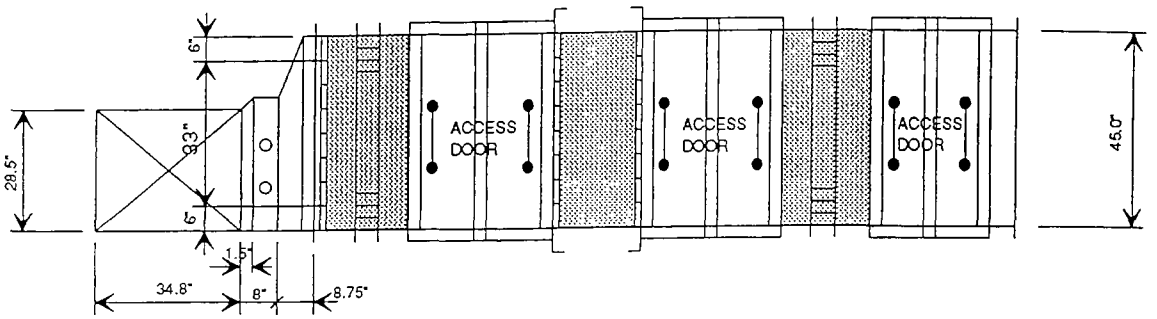
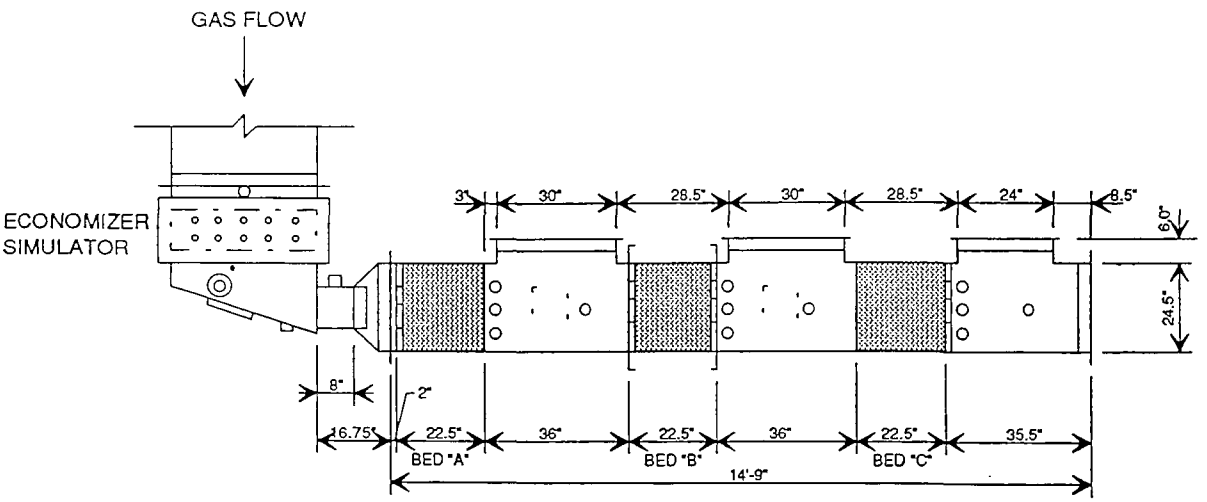


Figure 1
Ideal Catalyst Performance (Solid: DNO_x, Dotted: NH_3; Initial NO_x = 75ppm)



(a)
Plan View



(b)
Side View

Figure 2
Plan and Side View of the ASCR Reactor

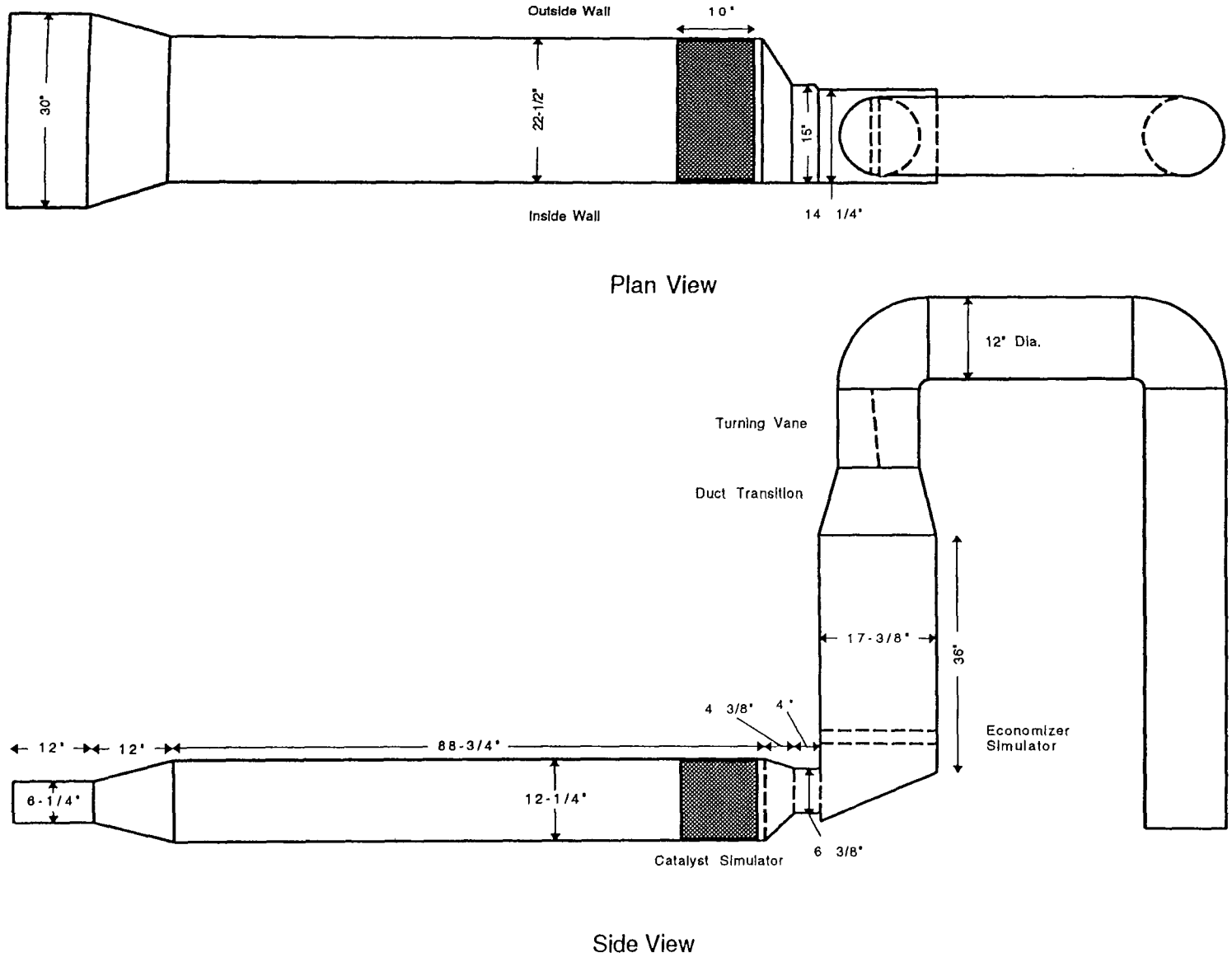
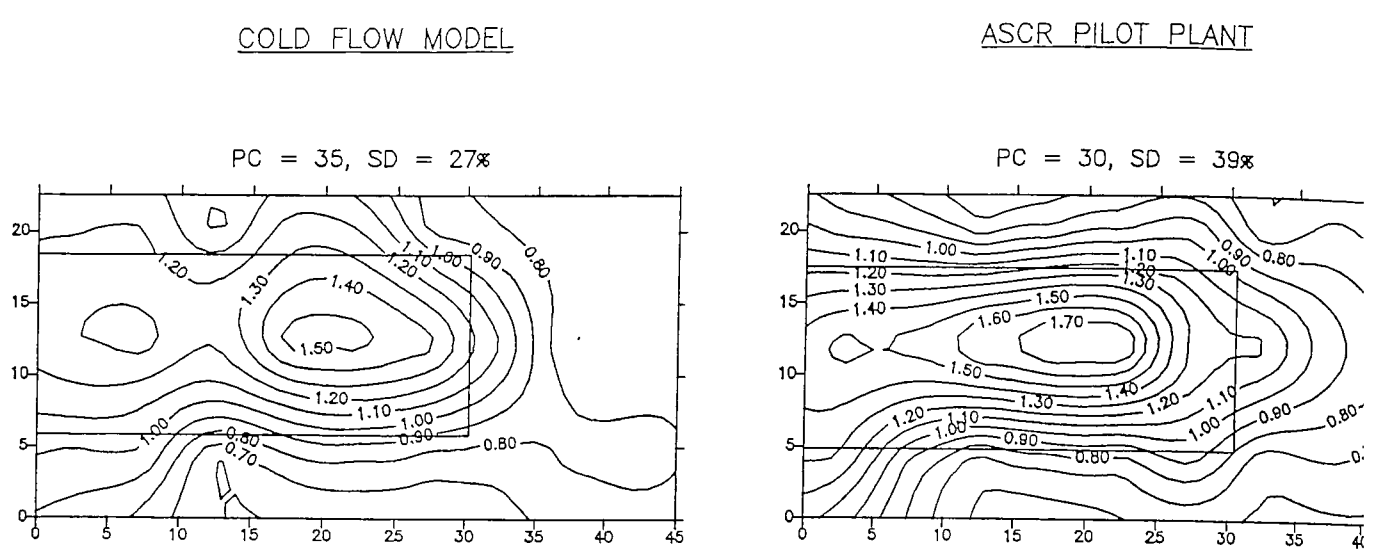
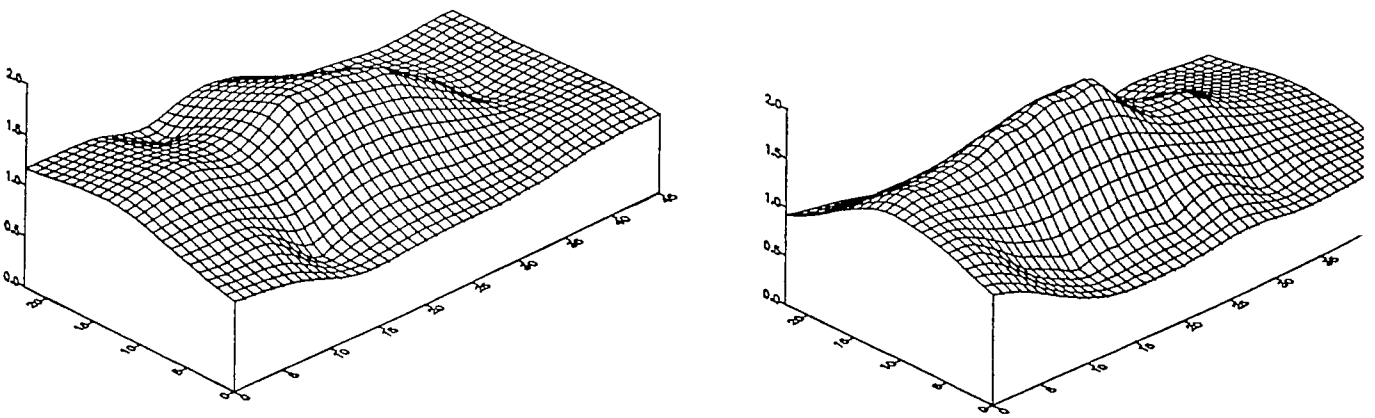


Figure 3
ASCR Cold Flow Model Schematic



(a)
Contour Plots



(b)
Wireframe Diagrams

Figure 4
Velocity Profiles at the Exit of Bed 1: Pilot Plant and Cold Flow Model
(Pilot Plant Conditions: T-680°F, Flow-5000 scfm, PC=30)
(Note: The Rectangle Shown on the Pilot Contour Plot Indicates the Location of the "throat" region)

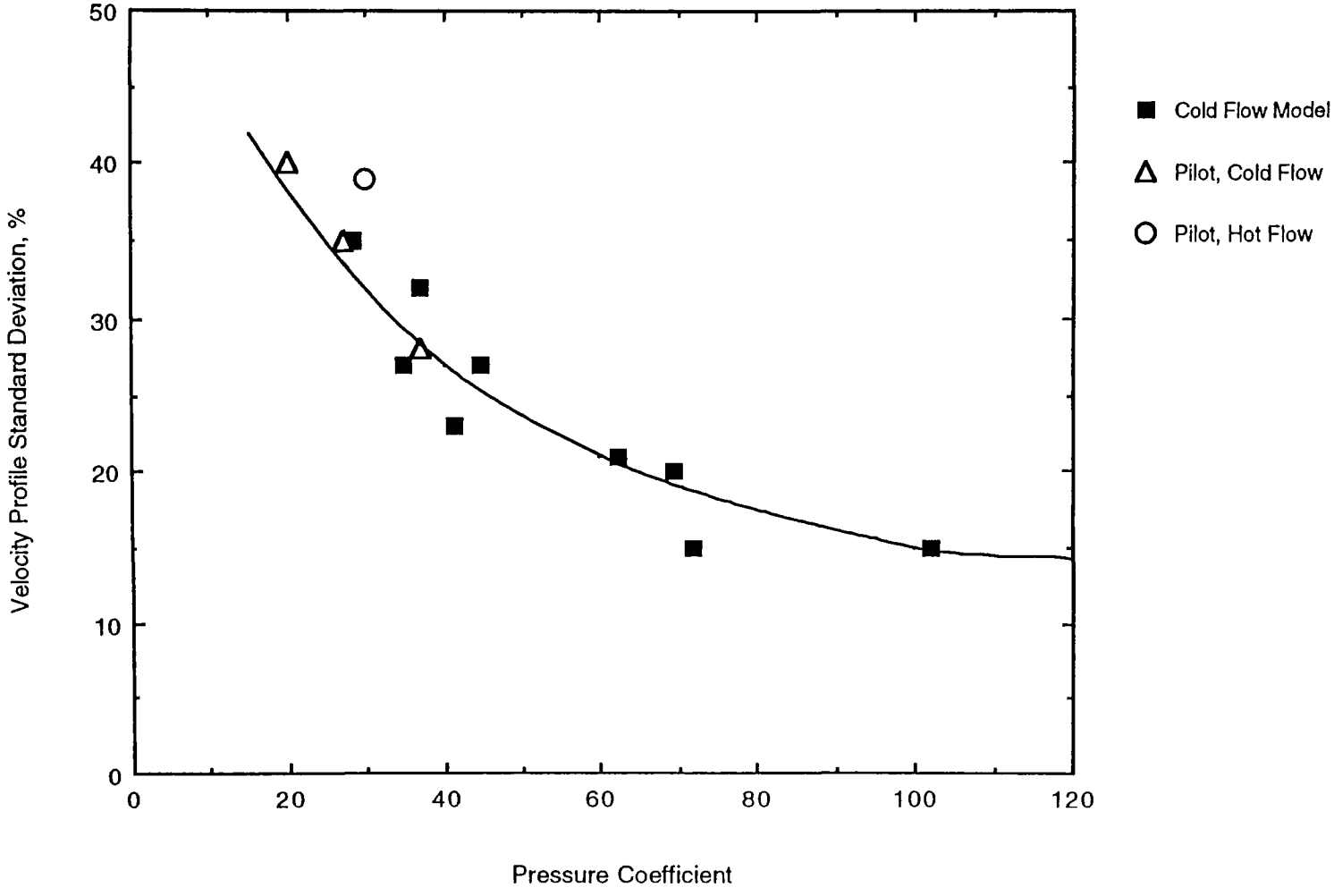


Figure 5
Effect of Pressure Coefficient on Velocity Uniformity

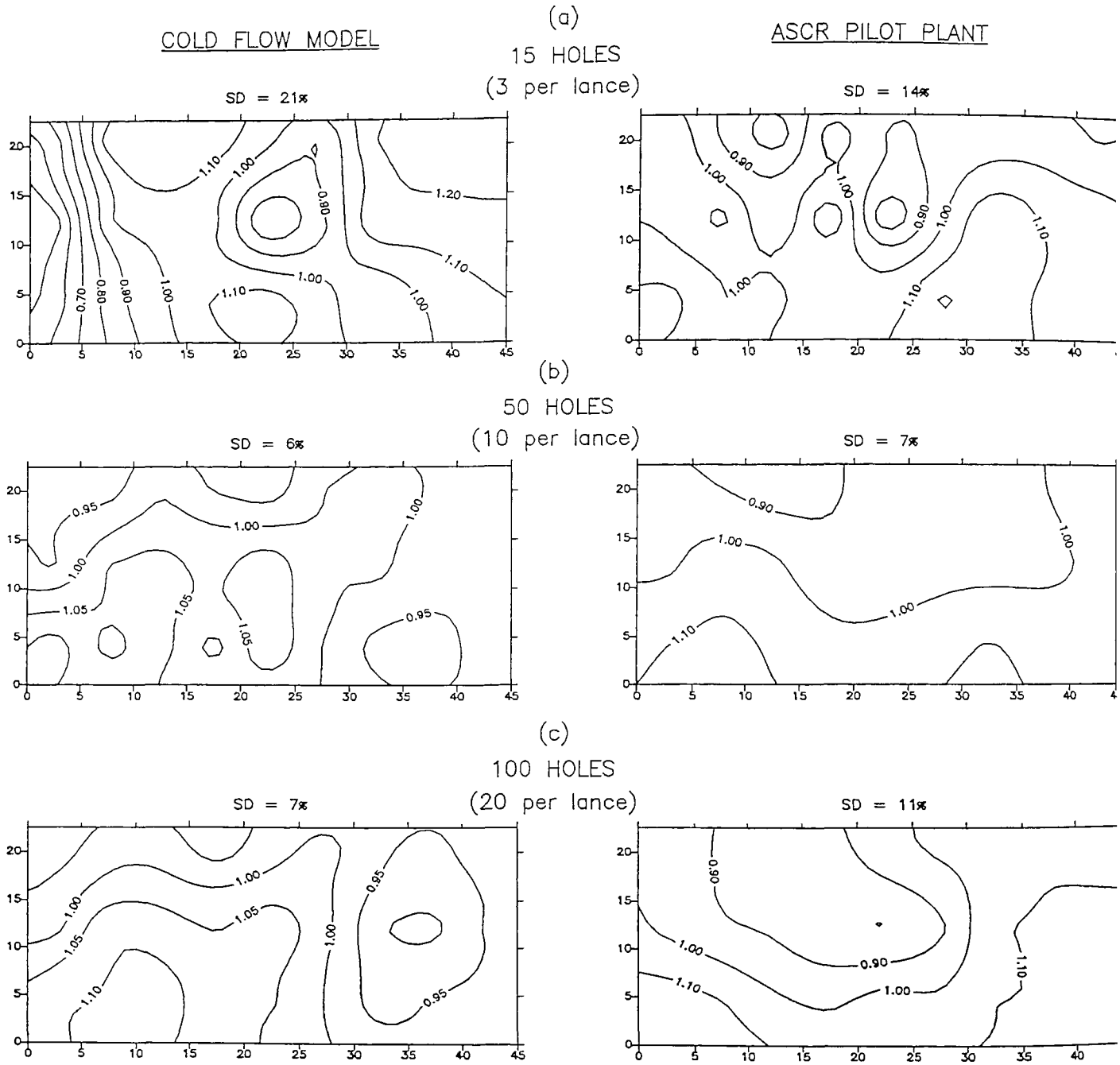


Figure 6
NH_3/NO_x Profiles: Economizer Exit Injection:
(Effect of the Number of Injection Holes)

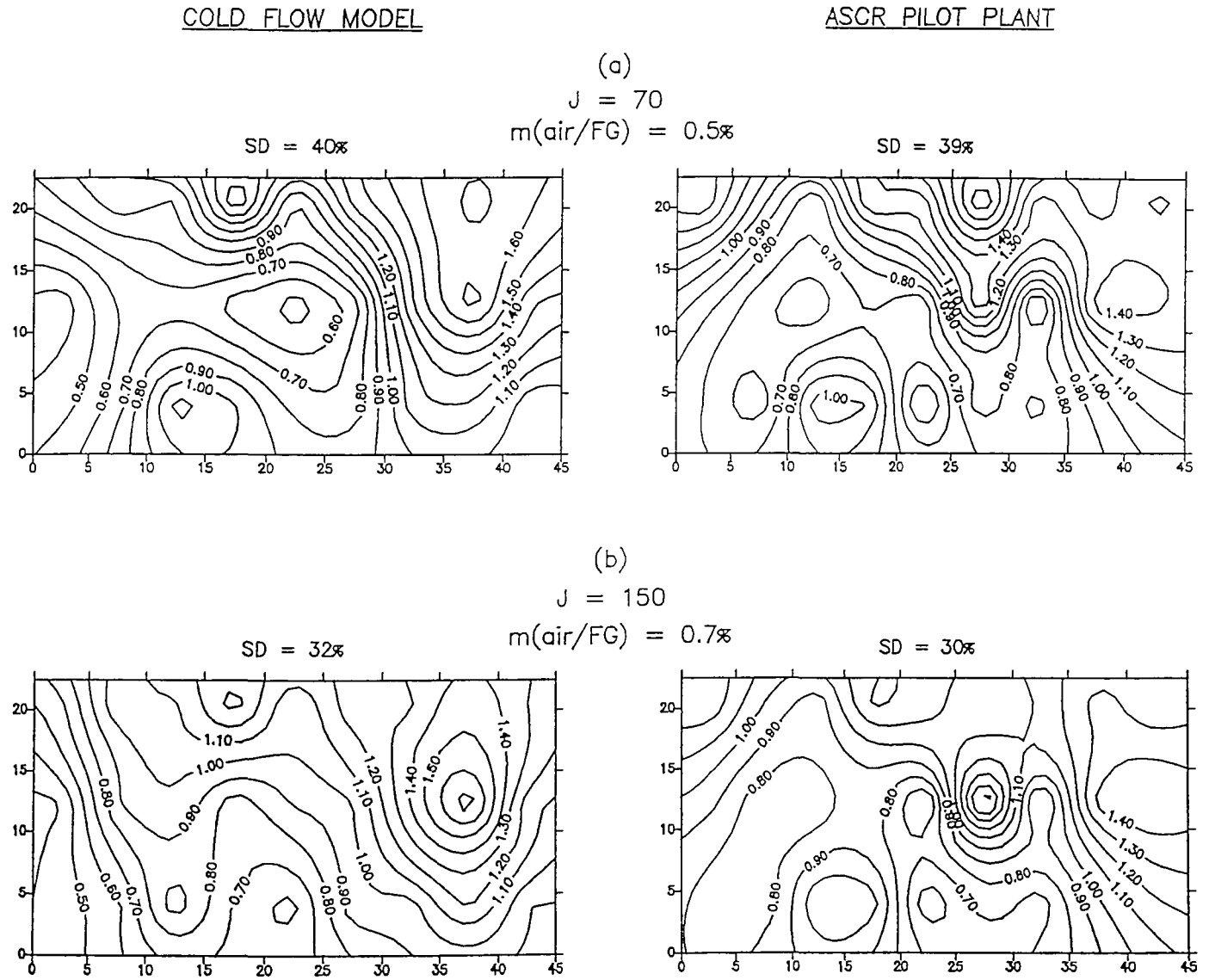


Figure 7
NH_3/NO_x Profiles: Throat Injection, 15 Injection Holes
(Effect of Injection Momentum)

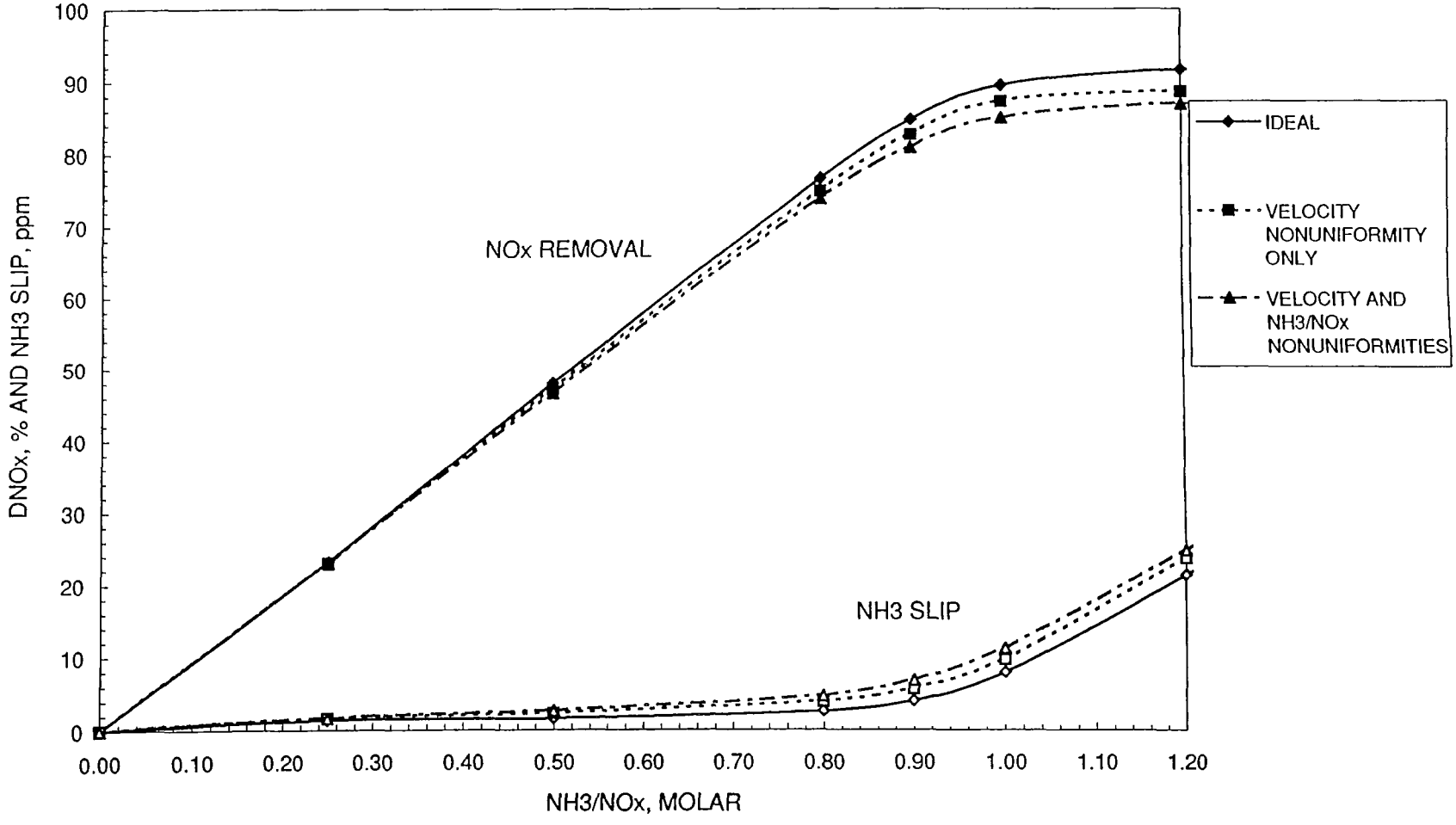


Figure 8
Baseline SCR Performance Predictions: NH_3 Injection Upstream of Gas Fan
(Velocity Standard Deviation: 39%, NH_3/NO_x Standard Deviation: 2%)

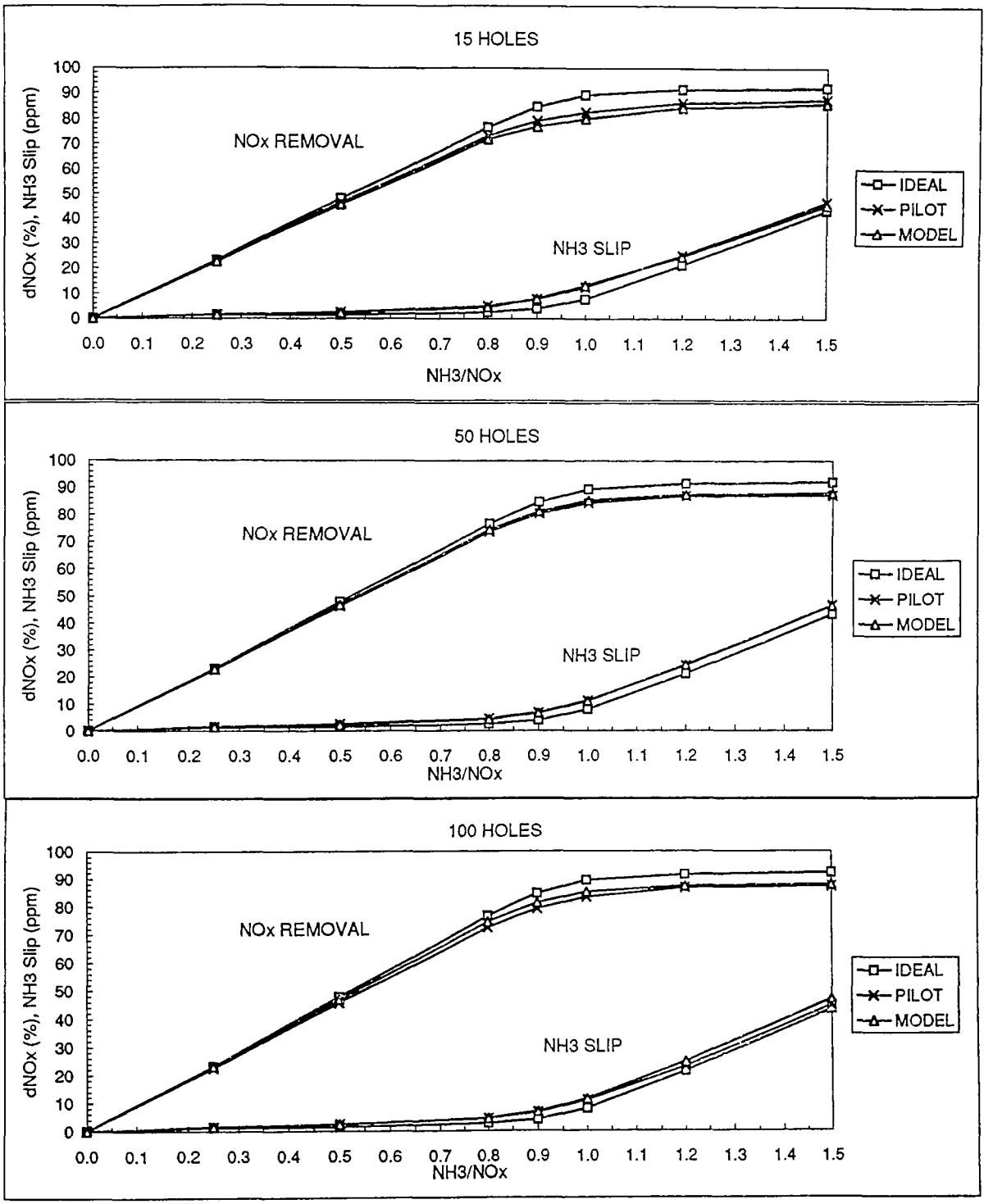
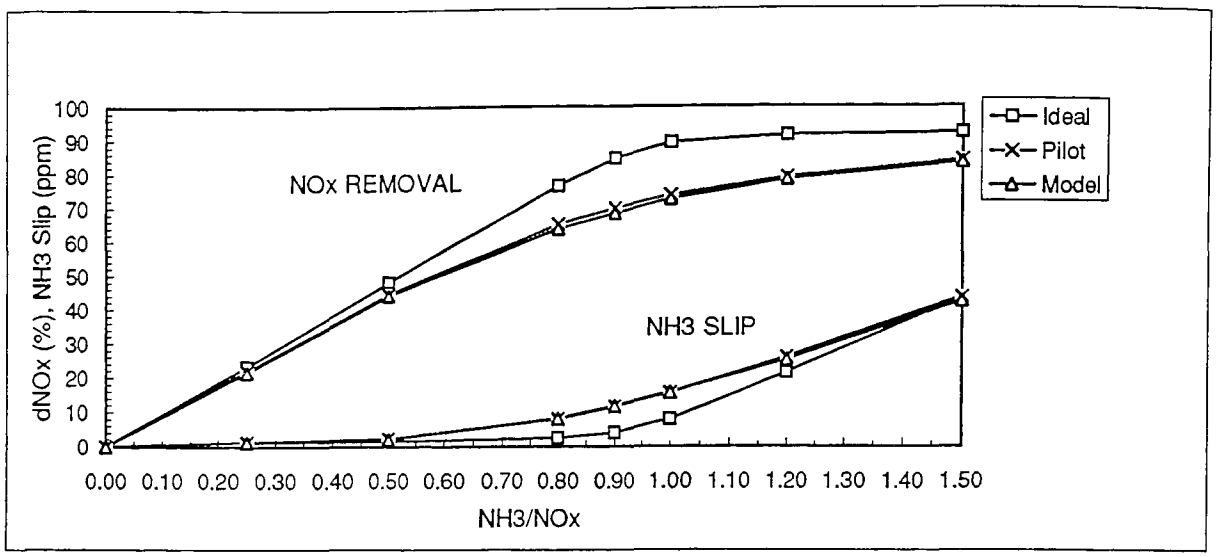
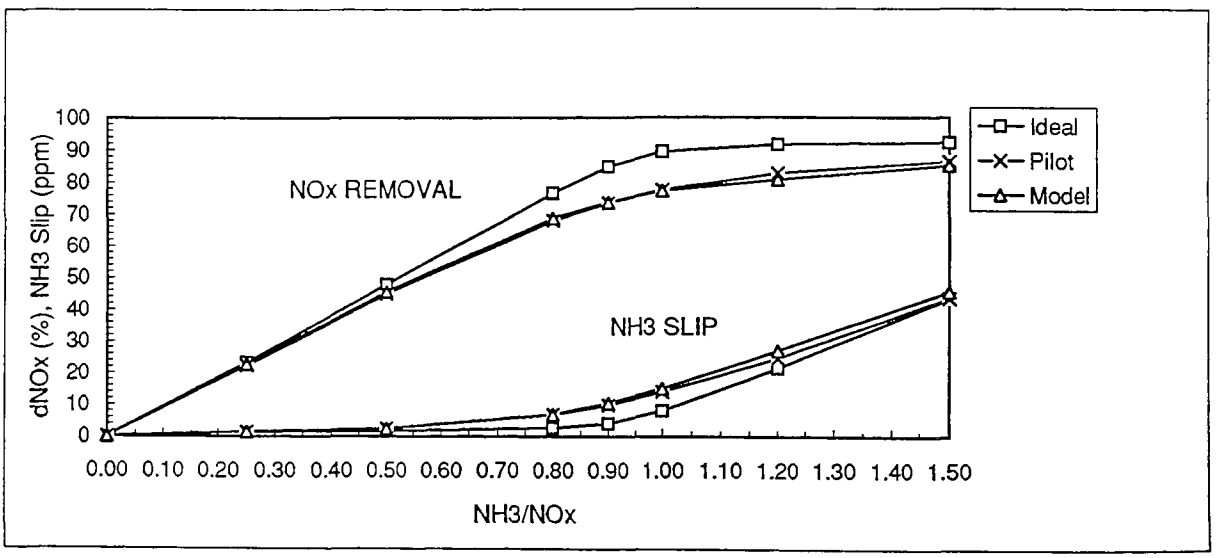


Figure 9
SCR Performance Predictions: Economizer Exit Injection
Varying the Number of Injection Holes (SV=26,000 hr^{-1}, J=2200)



(a)
J = 70



(b)
J = 150

Figure 10
SCR Performance Predictions: Throat Injection
Varying the Injection Momentum (SV=26,000 hr^{-1}, NO_{xi} = 75ppm)

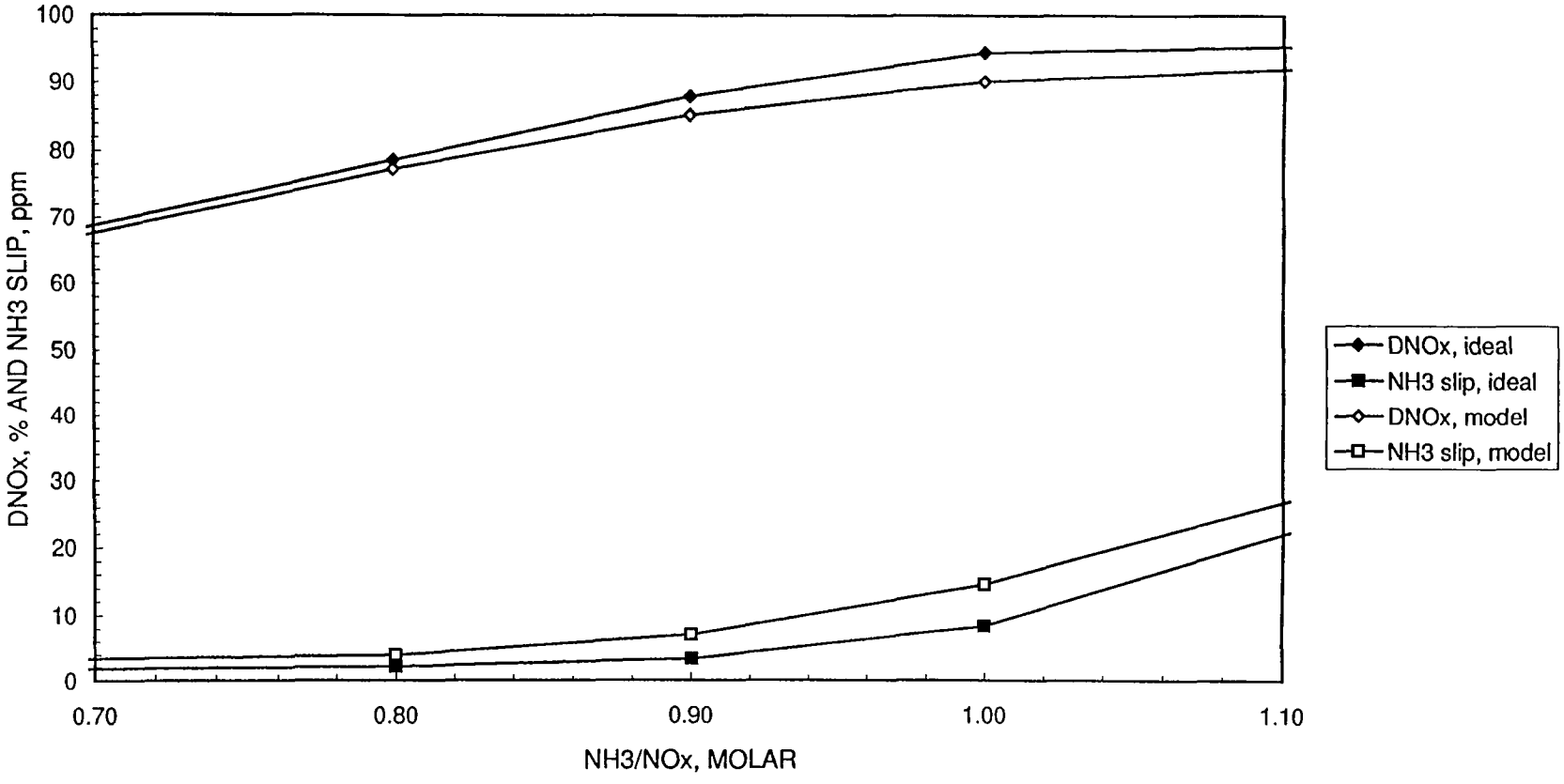


Figure 11
Full Scale Retrofit Performance Predictions

Various Types of SCR Plants
Full-Scale SCR
Compact SCR
In-duct SCR
Air Preheater SCR
Under Consideration of Achievable NOx Removal Rate and Cost-Effectiveness of Catalyst Use

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Abstract

SCR technology has proven itself for nitrogen oxide reduction in fossil-fired power plants. Depending on spatial conditions, different types of SCR plants can be realized in a power plant. In some individual cases, detailed investigation is required to determine how the requirements of a specific NOx reduction rate can be fulfilled.

This paper describes the primary differences in these various SCR plant types based on catalyst designs, such as achievable NOx removal rates and cost-effectiveness of catalyst use.

Introduction

The technique of selective catalytic reduction (SCR) has proven itself as a secondary measure for the reduction of nitrogen oxide emissions from fossil-fired power plants due to its capacity and operating reliability both in the backfitting of existing combustion facilities as well as in new plants. NOx removal rates of up to 95% with simultaneous adherence to an NH_3 slip of 5 ppmv or less can be achieved with this process.

In planning for a new plant or in backfitting with favorable space conditions, the SCR plant is generally placed between the steam generator and air preheater, using the optimum temperature range for the catalytic reaction, in a "high-dust" configuration (Fig. 1). The location of the SCR plant in the flue gas path as well as structural constraints for the amount of catalyst material to be installed and hence the performance data of the system represent a further type of classification. In addition to conventional SCR technology - installation of a "full-scale" SCR - newer systems such as compact SCR, in-duct and air preheater SCR have recently been discussed as concepts or have already been implemented, particularly for backfit measures (1), (2).

Types of SCR Plants

Common to all types of SCR plants are the components reducing agent storage and evaporation, reducing agent injection, catalyst, enclosure and instrumentation.
The significant characteristics of the various types of high-dust SCR plants are the intake velocity of the reactors and the number of catalyst levels (Figure 2).

Full-scale SCR plants for coal-fired power plants are designed for a velocity of approx. 14-17 ft/s based on the catalytic reactor cross section and on 3 to 5 reactor levels of installed catalyst material. The pressure drop on complete filling with catalyst is approx. 4 in. water gauge.

The reactor of a **compact SCR** plant is characterized by an increased gas velocity of 22-25 ft/s. To limit the pressure drop on the flue-gas side to roughly 4 in. water gauge, the number of the catalyst levels must be reduced to 2 or max. 3 levels.

In **in-duct SCR** plants the catalysts in the existing flue gas duct upstream of the air preheater are operated at typical gas velocities of roughly 32-38 ft/s. The volume is limited to 1 or max. 2 levels to ensure that a pressure drop of roughly 4 in. water gauge is not exceeded.

In the **air preheater SCR** process, catalytically active heat transfer elements are used on the hot side of the air preheater in 1 or max. 2 layers. A complete replacement of all layers of heat transfer elements by catalysts is not possible due to the decreasing gas temperature in the flow direction. On the one hand, the catalyst activity decreases with decreasing temperature, and on the other there is a danger of ammonium sulfate formation in the low temperature range (2). The pressure drop across the air preheater is not influenced by the installation of the catalyst. Heat transfer properties are only insignificantly impaired.

The installation of plate-type catalysts is possible in all SCR plant types. These converters are characterized by a high erosion resistance, even at high inlet velocities and ash contents. Temperature cycle resistance is outstanding, especially when the catalysts are implemented in the air preheater.

SCR Catalyst Design

In order to represent the principal differences of the aforementioned SCR plant types, the flue gas data for a fictitious power plant, equipped with a coal-dust fired generator with dry ash removal were applied as design conditions. For comparison purposes, all SCR plant types which are described below were designed with this power plant data.

Power plant and flue gas data

Table 1 shows a summary of the relevant data for dimensioning the catalyst.

Table 1:

Fuel fired	lb/hr	150,000
Flue gas flow rate	lb/hr	2,000,000
HHV	BTU/lb	13,000
Gas temperature	°F	700
Inlet NOx	lb/MMBTU	0.4
H_20	Vol. %, wet	8
O_2	Vol. %, wet	4
SO_2	ppmvd	1000
Particulate loading	gr/cf	3.5

Catalyst design conditions

NOx Reduction Efficiency. Inlet NOx concentration and the permissible NOx emissions limit stipulated by law are the factors on which the required NOx removal rate is based. This removal rate can be used as a selection criterion for determining which of the various SCR plant types is best suited. Achievable removal rates of the respective SCR plant types are discussed in the below.

NH_3 slip. In addition to the NOx removal rate, the maximum allowable NH_3 slip at given flue gas conditions is of particular importance in determining the required catalyst volume.

In coal-fired plants, the NH_3 slip downstream of the SCR plant, which is gaseous at first, is increasingly adsorbed on the fly ash on cooling of the flue gases in the air preheater. In dependence on the type and preparation of the coal fired, a fraction of approx. 70-100% of the

NH_3 slip is retained in the fly ash. Recycling of hard coal fly ash in such areas as the flooring or concrete industry can be limited by bad smells in processing if the NH_3 content of the ash exceeds specific limits (80-150 ppmv).

Based on experience from the operation of SCR plants in Germany, a maximum NH_3 slip of 2 ppmv is applied for the design to ensure the recyclability of ash from plants firing high-grad coals. As the following analysis shows, this value can be realized in full-scale and compact SCR plants at NOx removal rates > 60% given appropriate dimensioning.

Operating time. The operating time is that time during which the catalyst achieves the required removal rate. Catalyst activity decreases over time due to aging and catalyst poisons, until it is finally insufficient for the required NOx removal. A specific deactivation curve based on type of fuel fired and continuous feedback of information from existing plants is used for dimensioning of the catalyst. The volume is then dimensioned such that the required removal rate is still achieved at the end of the guaranteed operating time.

Depending on the type of SCR plant certain operating periods can reasonably be achieved until a first catalyst exchange or an additional catalyst loading is necessary. These are typically as follows:

Full-scale SCR	24,000 h
Compact SCR	16,000 h
In-duct SCR	14,000 h
Air preheater SCR	24,000 h (in case of NH_3 slip control only)

Catalyst selection. The selection of a suitable catalyst composition and geometry are of decisive significance for reliable operation of SCR plants. Siemens is the only catalyst manufacturer offering both honeycomb and plate-type catalysts, and is therefore also experienced in selecting the optimum catalyst type.
For implementation in ash-containing flue gases and especially at very high flue gas velocities, plate-type catalysts have proven themselves due to their high erosion resistance and mechanical stability as well as their low pressure drop in comparison to honeycomb catalysts (3).

The catalyst designs for all of the SCR plant types considered here are based without exception on a highly active, selective Siemens plate-type catalyst. The specific surface area is 110 ft^2/ft^3, corresponding to a pitch of 0.2 inches. The individual catalyst elements have a cross section of 18.2 inches * 18.2 inches with a plate height of 22.4 inches in the flow direction, and can be installed in all types of SCR plants. However, depending on the requirements, other deviating catalyst geometries can be implemented, such as those with lower or higher specific surface areas. This is particularly important if a specific pressure drop is determining for the design.

Inlet flow conditions. For optimum utilization of the catalyst, minimum requirements must be met for the intake flow to the SCR reactor as regards temperature, flue gas velocity and reactant concentration. It is advantageous to optimize the flow using model experiments (4).
An inlet flow quality corresponding to the state of the art is assumed for the comparison of the SCR systems.

Representation of Catalytic Reactor Design Data for Individual SCR Systems

In addition to the linear velocity limits mentioned above, the total number of levels available for loading with catalyst was limited to obtain similar pressure losses (app. 4 in. water gauge) for each system with a full catalyst load. This enabled a reasonable comparison of the full-scale, compact and in-duct SCR systems. Table 2 shows an overview of the SCR plant types examined.

Full-scale SCR

The result of the catalyst design is shown in Table 3. The design yielded a reactor with a total of 4 levels, each of which can be charged with 2 layers of catalyst. With an initial loading of 9310 ft^3 of catalyst on 5 layers and an initial operating time of 24,000 operating hours (3 years), an NOx removal rate of 80% is achievable. This configuration retains 3 layers free for subsequent recharging.

Compact SCR

The result of the catalyst design is shown in Table 4. A reduction of the reactor cross section by approx. 35% over that of a full-scale plant and limiting the number of available levels to 3 yielded a compact SCR plant configuration.

This design type is characterized in the selected example by a reactor with a total of three available levels. For a catalyst initial operating time reduced to 16,000 hours an initial catalyst charge volume of 6100 ft^3 in 5 layers can achieve an NOx removal of 60%. In this case only a single catalyst layer is available for future recharging with catalyst.

In-duct SCR

The result of the catalyst design is shown in Table 5. The design assumed the existence of two 50% flue gas ducts upstream of the air preheaters. Each of the two flue gas ducts is charged with 1625 ft^3 of catalyst on 2 levels of 2 layers each. Due to space limitations and to limit pressure drop, no free space is planned for recharge catalyst.

Under the very high space velocities (SV > 7000 1/h, flue gas volume flow rate divided by catalyst volume) typical for this application, it is not possible to adhere to an NH_3 slip of 2 ppmv

at relevant NOx removal rate over operating times which are reasonable for the utility. An NOx removal rate of 25% can be achieved with the aforementioned volume, even if the NH_3 slip is increased to 5 ppmv for an initial operating time of 14000 hours.

APH SCR

The result of the catalyst design is shown in Table 6. The design assumed two 50% flue gas ducts and air preheaters.
Charging the hot end area of each air preheater with one catalyst layer, based on the volume through which the flue gas flows (only roughly half of the installed volume), yields space velocities of 31,000 1/h. These conditions result in a low NOx removal rate of approx. 5%.

However, an APH SCR system can be used to reduce the NH_3 slip from upstream systems such as compact SCR or in-duct SCR. This enables the achievement of higher NOx removal rates in upstream systems. For example, if the in-duct SCR plant discussed here is operated in combination with an APH SCR system and the overall NOx removal rate remains at 25% the ammonia slip will then be reduced to 2 ppmv.
Similar results are obtained with the compact SCR plant in combination with an APH SCR. The NOx removal rate increases from 60% to 73% while the ammonia slip is maintained at a level of 2 ppmv.

Catalyst Management Plan

All of the above catalytic reactor design performance data are based on the initial period of operation. At the end of this period NH_3 slip exceeds the permissible limit due to declining catalyst activity. It is therefore necessary to either install additional catalyst material in the reactor or to replace a portion of the deactivated material with fresh catalyst. A catalyst management plan should be established during the planning phase to determine catalyst material requirements.

To compare the SCR systems described with regard to catalyst use, a catalyst management plan is determined for an operating time of roughly 20 years. Especially in high-dust SCR plants, catalyst utilization has a significant influence on operating costs. The catalyst replacement strategy is highly dependent on the specific design of the SCR reactor (availability of reserve levels). Plate-type catalysts are constructed such that two catalyst layers can be installed on each level. In many cases the addition or replacement of only a single layer (1/2 level) is sufficient to ensure adherence to the emission limits until the next or next two inspections.

A replacement strategy which foresees the possibility of charging additional catalyst when the allowable NH_3 slip limit is first reached uses a higher fraction of the catalyst activity associated with the catalyst charged. In the event that adherence to the required NOx removal rate necessitates early replacement of portions of the installed catalyst, the relatively high residual activity (generally roughly 30% of the initial value) of catalyst remains unused.

Therefore, the design of the catalyst as an "environmental protection system" should not only account for the cost optimization, but should also be viewed with regard to the minimization of waste materials generated.

Full-scale SCR

The catalyst management plan is shown in Figure 3.
At the end of the third operating year, the half-level which is still empty is filled with additional catalyst elements, and at later points in time the two remaining half-levels are filled. The complete charge enables NOx removal to continue for more than 8 years without removal of catalyst. After this time, further operation of the plant is ensured by the successive replacement of one complete catalyst level at a time, starting with the most fully exhausted catalyst level in the charge. This ensures maintenance of the guaranteed removal rate over at least two further operating years after each recharging or replacement.

Compact SCR

The catalyst management plan is shown in Figure 4.
In this type, the SCR reactor is completely filled by the addition of a half-level of catalyst after 16,000 hours. This enables operation for a further two years. Subsequently, the replacement of at least one level at each interval is required for continued operation with acceptable service intervals (a minimum of 2 years).

In-duct SCR

The catalyst management plan is shown in Figure 5.
According to the design, an NH_3 slip of 5 ppmv is attained after 14,000 hours operation. It is not possible to charge additional catalyst, i.e. after approx. 2 years time 50% of the initial catalyst charge in each of the two ducts must be replaced by new material.
This replacement however enables continued operation for only slightly more than one operating year.

APH-SCR

It is assumed in this case that deactivation takes place here as it does in other catalyst systems. This does not, however, decrease the systems's ability to significantly reduce low concentrations of NH_3. Consequently APH catalyst materials for NH_3 reduction can remain in service for extended periods without the need for replacement. Cost-effectiveness assessment is based on an assumed service life of 24,000 h.

As a result, complete replacement of the catalyst material is necessary about once every three years. This design therefore does not require a catalyst management plan.

Discussion

Careful consideration of all economic aspects reveals that the installation of an SCR plant tailored to the individual power plant, i. e. smallest possible SCR plant, is not always the optimum solution.

As the aforementioned designs for catalyst management of the individual SCR plant types have already shown, the catalyst management plans differ greatly. This leads to corresponding differences in the expenditure required to maintain the removal rate, in the catalyst requirements and in the utilization of the catalyst.

Another factor to be considered is the number of unit outages for the addition of catalyst or catalyst replacement. As it can be seen in the figures for catalyst management plans, the number of unit outages is much higher for all plant types other than full-scale. The duration of the plant outage for the addition of catalyst can be taken as approximately equal for the individual SCR plant types, and is roughly 3 to 5 days .

Figure 6 shows the NOx reductions achievable with the individual SCR plant types, expressed in tons of NOx over a time frame of 20 years. Full-scale, compact SCR and the combination of compact SCR with an APH SCR achieve the highest values, corresponding to the higher NOx reduction rates. As an APH SCR plant is not especially effective as an NOx reduction measure, it should therefore only be implemented for NH_3 slip reduction in combination with other measures such as compact or in-duct SCR.

Figure 7 shows catalyst consumption including the initially installed volume over a 20 year period. Compact and full-scale plants have virtually the same material requirements. An in-duct plant has only slightly lower catalyst material requirements. A combination of compact or in-duct SCR plants with an APH SCR plant results in the highest catalyst material requirement. It is imperative that these generalizations be considered in the light of the NOx reduction requirements for individual plants in question.

Figure 8 shows specific catalyst consumption (ft^3 catalyst/1000 t NOx reduced). The specific requirements are considerably higher for all other than full-scale SCR plants.

Figure 9 shows the volume of catalyst material to be disposed of over a 20 year period. As is evident, differences are significant, and the full-scale plant is most advantageous in terms of disposal requirements.

The comparison of various SCR plant types shows that the catalyst is best utilized in full-scale SCR plants. High priority should be given to the possibility of later addition of catalyst in planning an SCR plant. This enables the most cost-effective use of the catalyst. A catalyst management

practice which foresees the possibility of adding catalyst when the ammonia slip begins to exceed the guaranteed ammonia slip limit utilizes more of the catalyst activity then does the practice of replacing the catalyst charge.

At first glance, the combination of a compact or in-duct SCR plant with an APH SCR plant is interesting for improved NOx removal. However, the high requirements over a longer time frame show, that the associated high specific catalyst requirements are not cost-effective.

Inclusion of the expenditure for plant construction in the comparison does not lie within the scope of this paper. However, roughly the same expenditures are required for all SCR plant types for the following components:

NH_3 storage
NH_3 injection and control
Plant instrumentation
I&C.

In summary, it can thus be stated that a full-scale SCR system represents the most favorable solution with regard to the parameters considered. The other SCR plant types discussed, such as compact and in-duct SCR or combinations with an APH SCR system, could serve reasonable niche functions under the consideration of special requirements such as

- Limited available space
- Use in a peaking plant with low yearly operating hours
- Relatively short remaining plant service life
- Acceptance of lower NOx reduction rates or higher NH_3 slip values.

Literature:

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2. K. Hüttenhofer, J.-K. Beer, H. Smeets, J. van der Kooij. "The DeNOx Air Preheater Downstream of a Coal-Fired Boiler." EPRI/EPA 1993 Joint Symposium on Stationary Combustion NOx Control, Miami Beach, Florida.
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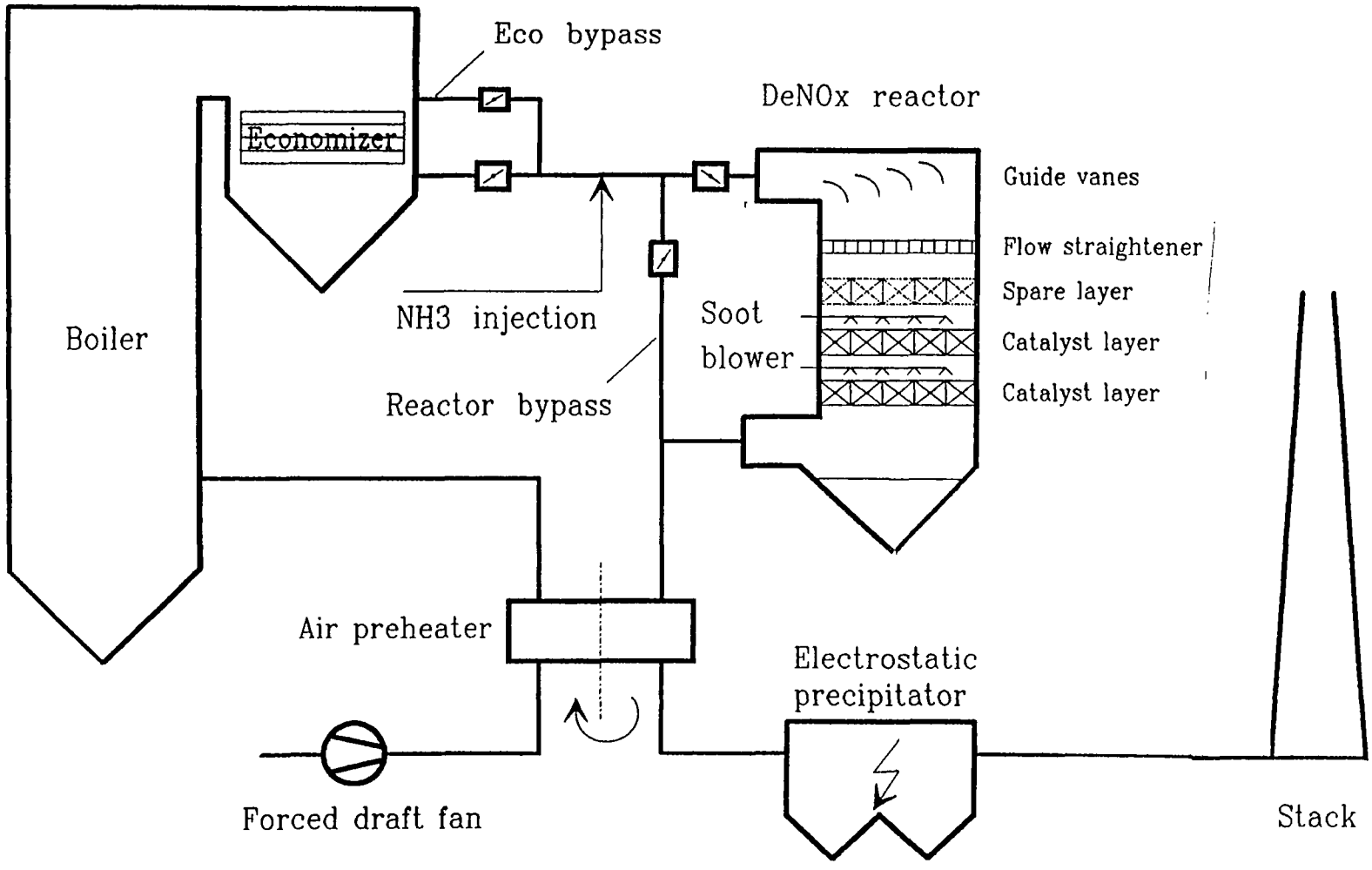


Figure 1: High-Dust SCR Arrangement

	Full-Scale SCR	Compact SCR	In-duct SCR	APH SCR
Flue gas velocity (ft/s)	14 - 17	22 - 25	32 - 38	40 - 44
Number of layers (initial load)	5 - 9	4 - 5	3 - 4	1 - 2
Number of catalyst spare layers	3	1	0	0

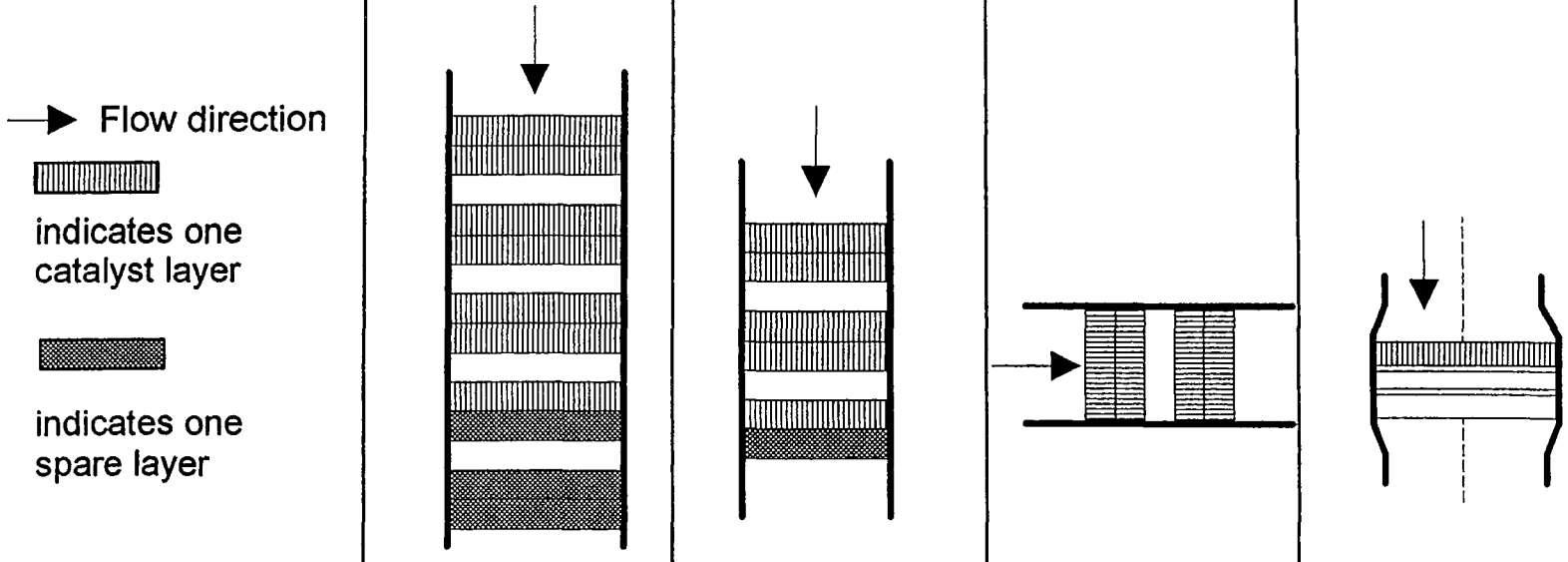


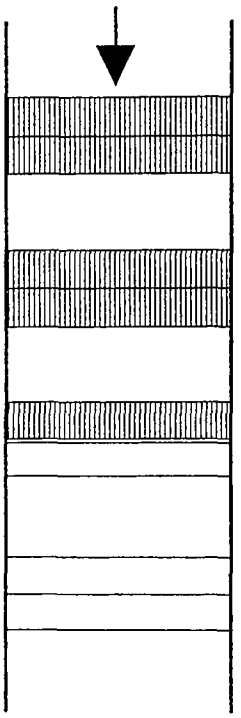
Figure 2: Characteristic of different SCR-Types

	Full-Scale SCR	Compact SCR	In-duct SCR	APH SCR
Catalyst Volume [ft³]	9,310	6,100	2 * 1625	2 * 777
Spec. surface area [ft²/ft³]	110	110	110	110
Height of one catalyst layer [inch]	22	22	22	22
Initial catalyst layers	5	5	4	1
Spare layers	3	1	0	0
SV value [1/h]	2,580	3,940	7,390	31,000
Flue gas velocity [ft/s]	16	24	31	39
Pressure drop [in. wg]	2	3	4	5
Guarantee period [h]	24,000	16	14	24,000
NO_x reduction efficiency [%]	80	60	25	-
NH_3 reduction efficiency [%]	-	-	-	50 - 70
NH_3 slip at the end of the guarantee period	2	2	5	-

Table 2: Overview of the SCR plant types examined

Catalyst Data:

Specific surface area	110	ft^2/ft^3
Pitch	0.2	inch
Height of one layer	22.4	inch
Catalyst volume	9310	ft^3



Reactor Data:

Level 1 with 2 catalyst layers	78.0 tons
Level 2 with 2 catalyst layers	78.0 tons
Level 3 with 1 catalyst layer	53.0 tons
Level 4 with 2 spare layers	

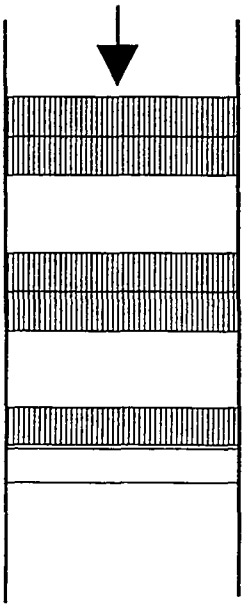
Reactor cross section area	33' * 34'	
Flue gas velocity	15.7	ft / s
Pressure drop (initial)	1.85	inch wg.
(final)	3.0	inch wg.
Initial guarantee period	24,000	hours
NO_x reduction efficiency	80	%
NH_3 slip	< 2	ppmvd
SO_2 conversion rate	< 1.5	%

Table 3: Catalyst Design Data Sheet for Full-Scale SCR

SIEMENS

Catalyst Data:

Specific surface area	110	ft^2/ft^3
Pitch	0.2	inch
Height of one layer	22.4	inch
Catalyst volume	6100	ft^3



Reactor Data:

Level 1 with 2 catalyst layers	51.0 tons
Level 2 with 2 catalyst layers	51.0 tons
Level 3 with 1 catalyst layer	35.0 tons

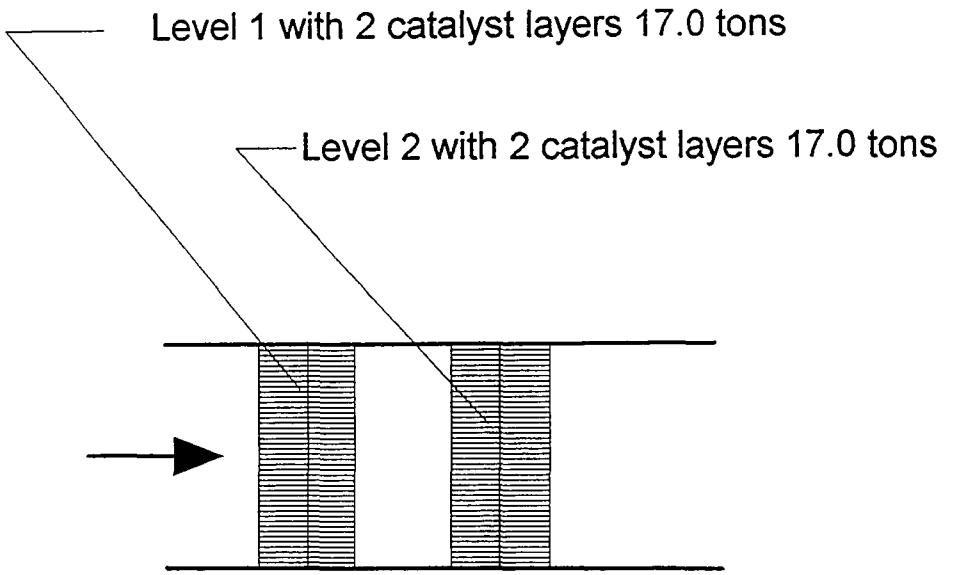
Reactor cross section area	25' * 30'	
Flue gas velocity	23.9	ft / s
Pressure drop (initial)	3.1	inch wg.
(final)	3.7	inch wg.
Initial guarantee period	16,000	hours
NO_x reduction efficiency	60	%
NH_3 slip	< 2	ppmvd
SO_2 conversion rate	< 1.0	%

Table 4: Catalyst Design Data Sheet for Compact SCR

Catalyst Data:

Specific surface area	110	ft^2/ft^3
Pitch	0.2	inch
Height of one layer	22.4	inch
Catalyst volume	2 * 1625	ft^3 (two ducts are equipped with catalysts)

Reactor Data:

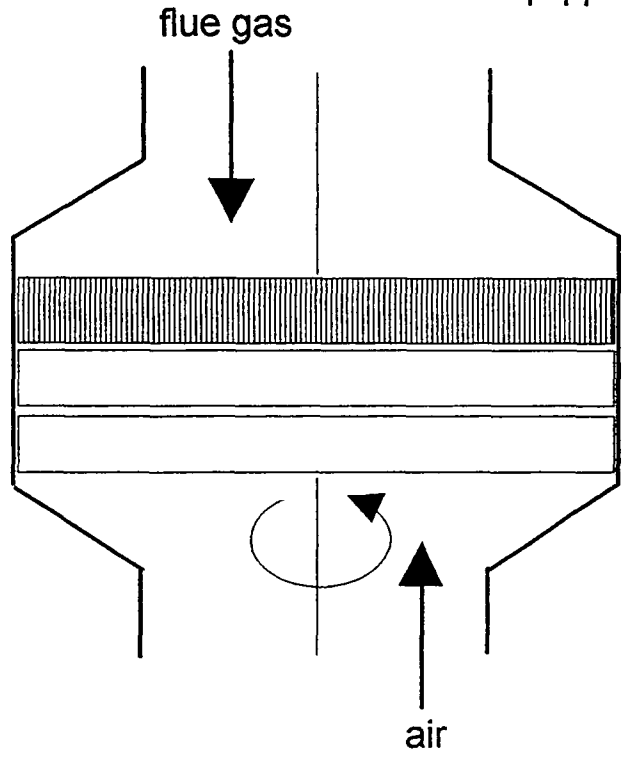


Dust cross section area	20' * 13'	
Flue gas velocity	30.8	ft / s
Pressure drop (initial)	3.8	inch wg.
Initial guarantee period	14,000	hours
NO_x reduction efficiency	25	%
NH_3 slip	< 5	ppmvd
SO_2 conversion rate	< 0.5	%

Table 5: Catalyst Design Data Sheet for In-duct SCR

Catalyst Data:

Specific surface area	110	ft^2/ft^3
Pitch	0.2	inch
Height of one layer	22.4	inch
Catalyst volume	2 * 805	ft^3 (two APH are equipped with catalysts)



Cross section area	200	ft^2
Flue gas velocity	39.4	ft / s
Pressure drop (initial)	4.7	inch wg.
Initial guarantee period	24,000	h (for NH_3 control)
NH_3 reduction efficiency	50 - 70	%
SO_2 conversion rate	< 0.2	%

Table 6: Catalyst Design Data Sheet for APH SCR

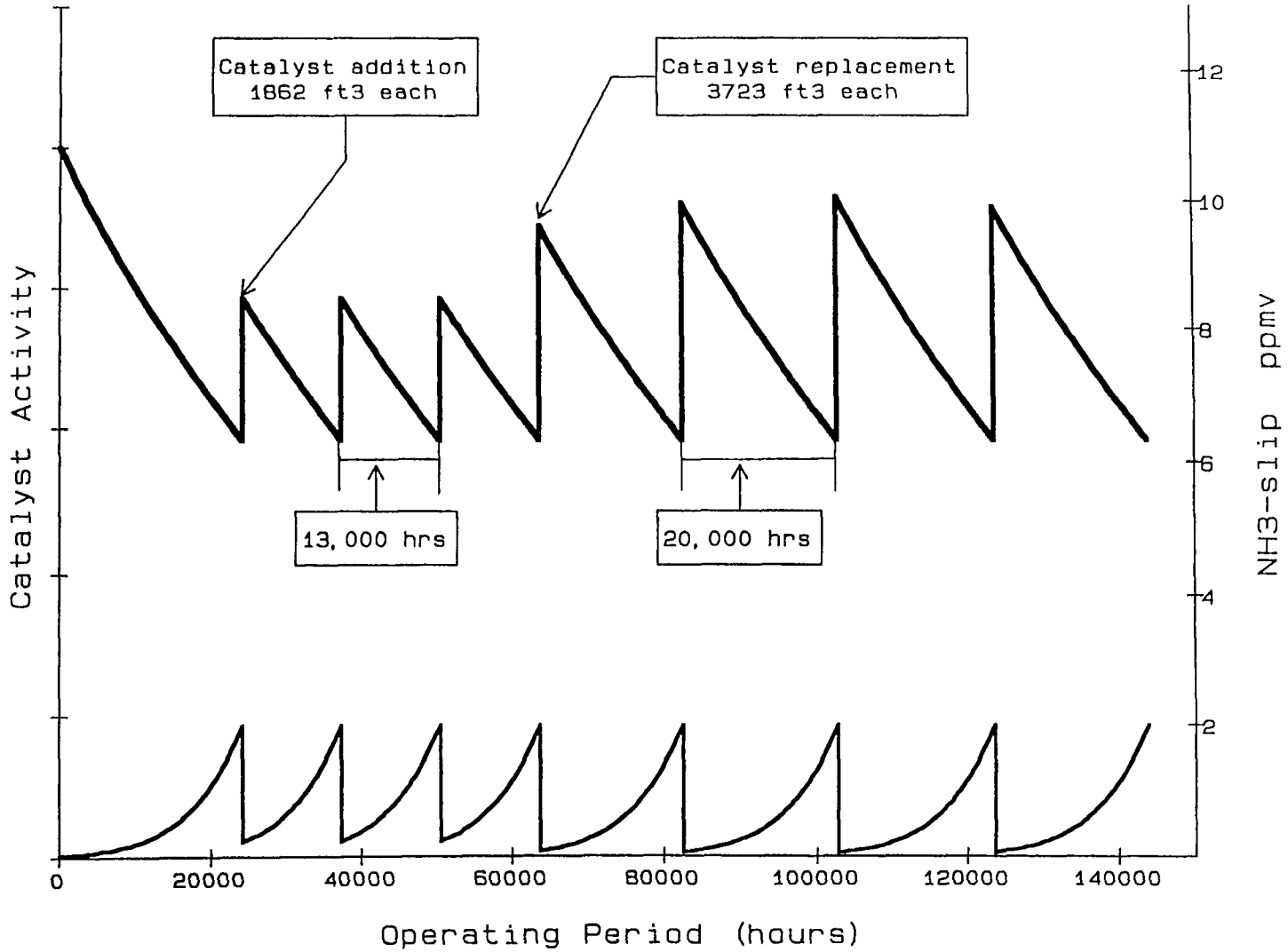


Figure 3: Catalyst Management Plan for a Full-Scale SCR

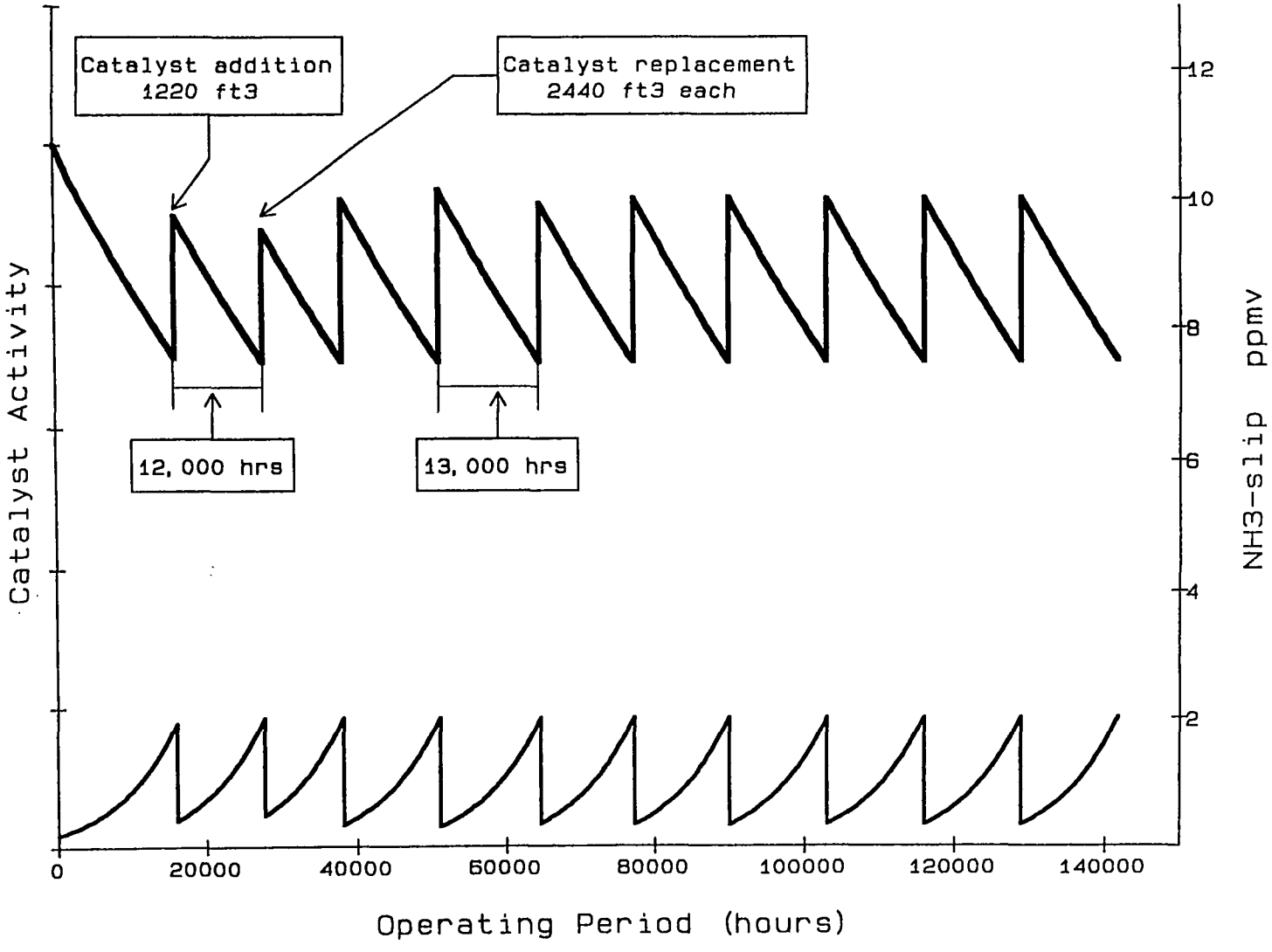


Figure 4: Catalyst Management Plan for a Compact SCR

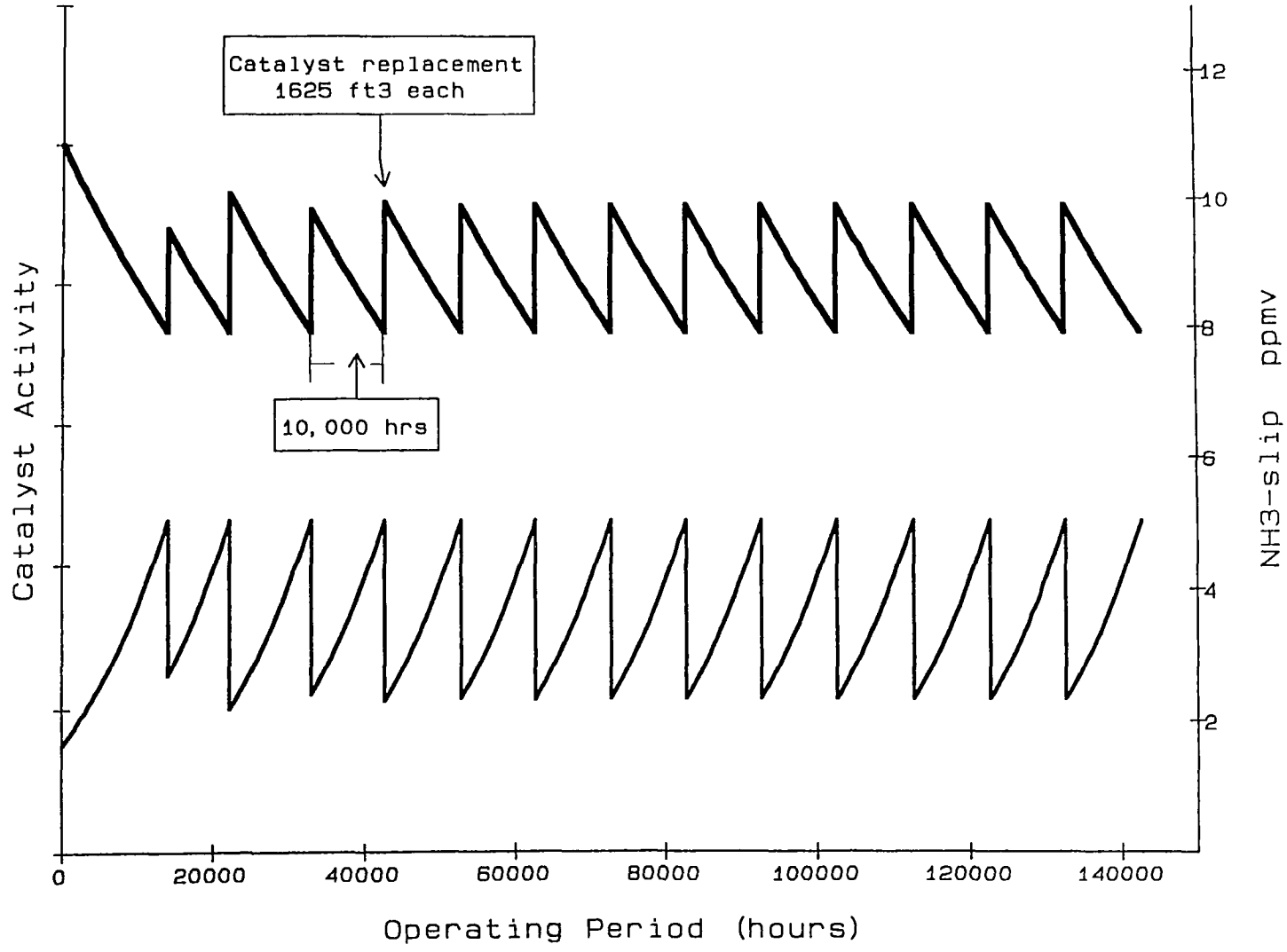


Figure 5: Catalyst Management Plan for an In-duct SCR

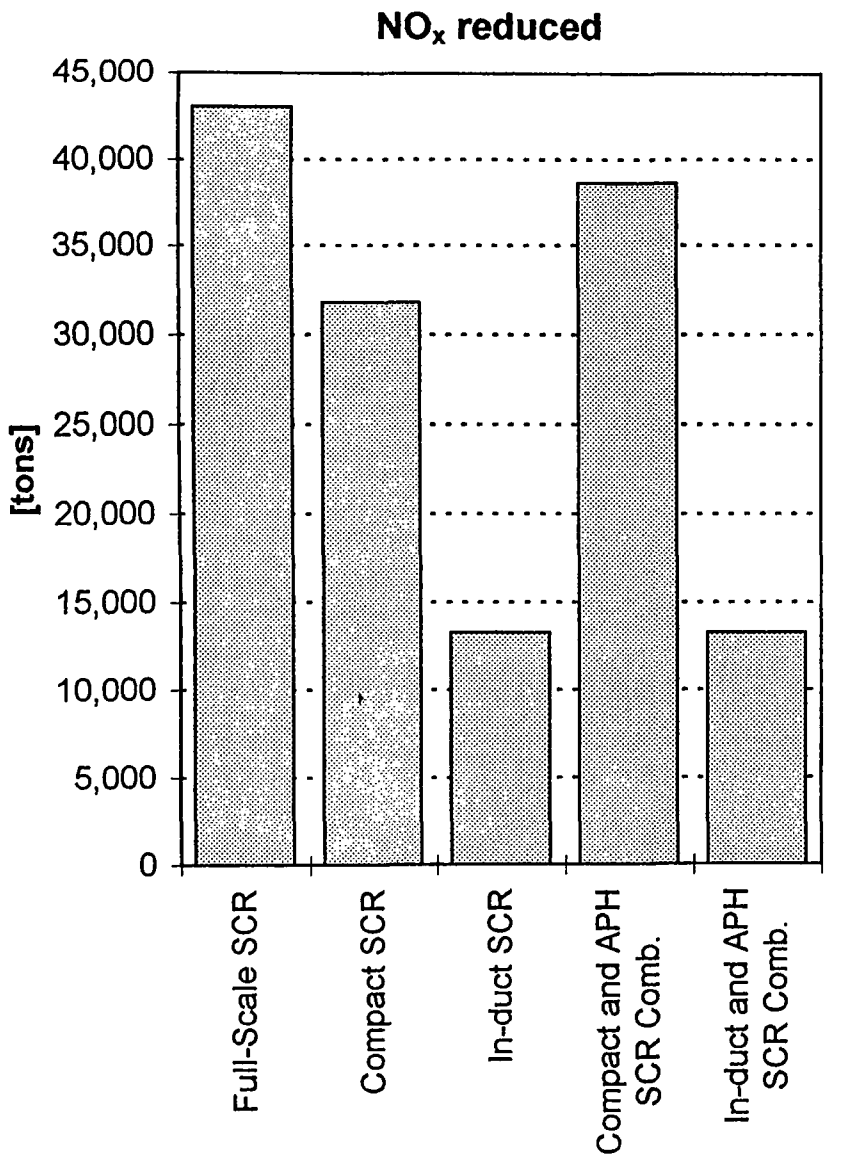


Figure 6: NO_x Reduced [tons] over an Operating Period of 20 Years

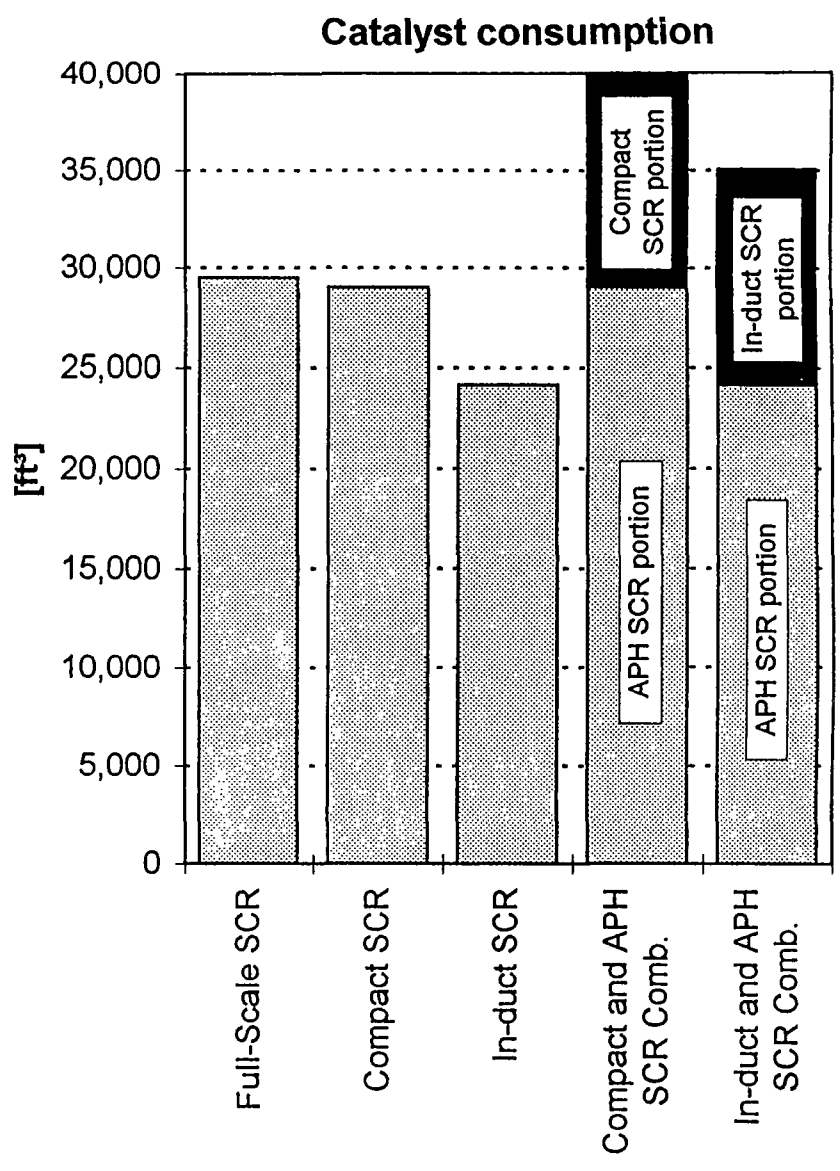


Figure 7: Catalyst Consumption [ft³] for an Operating Period of 20 Years

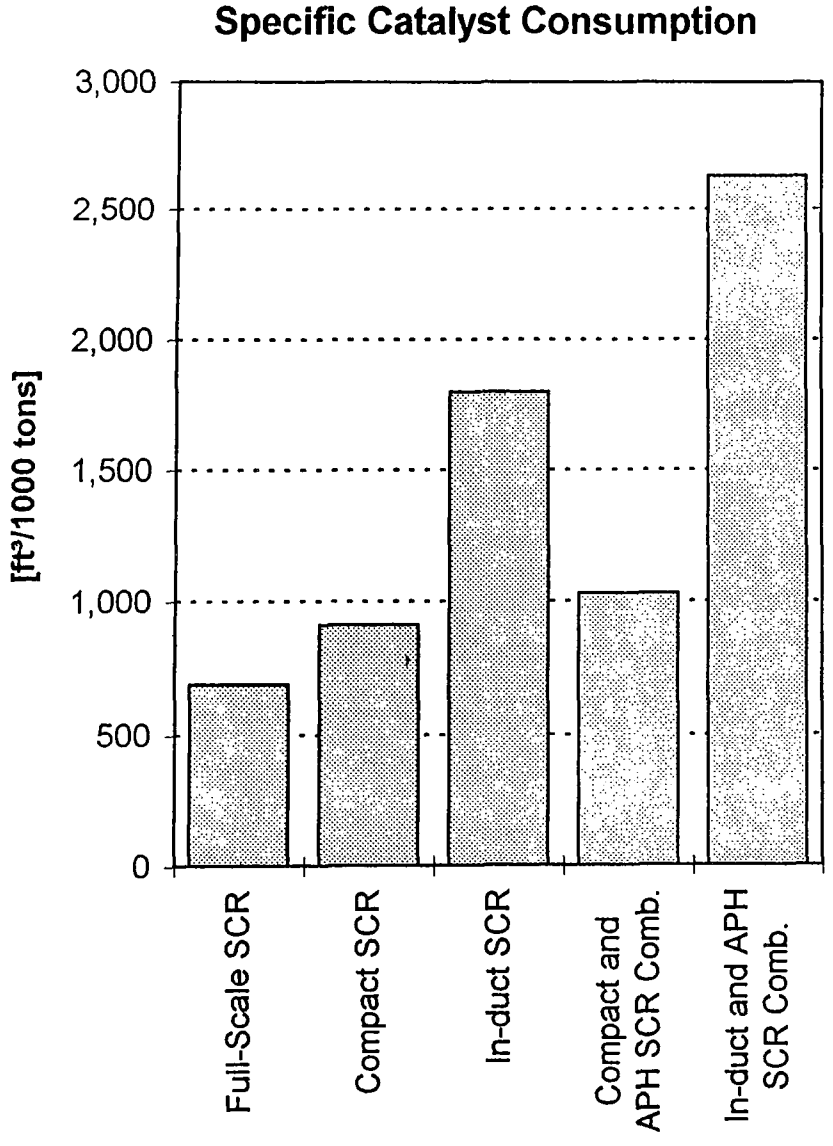


Figure 8: Specific Catalyst Consumption for an Operating Period of 20 Years [ft³ / 1000 t NOx]

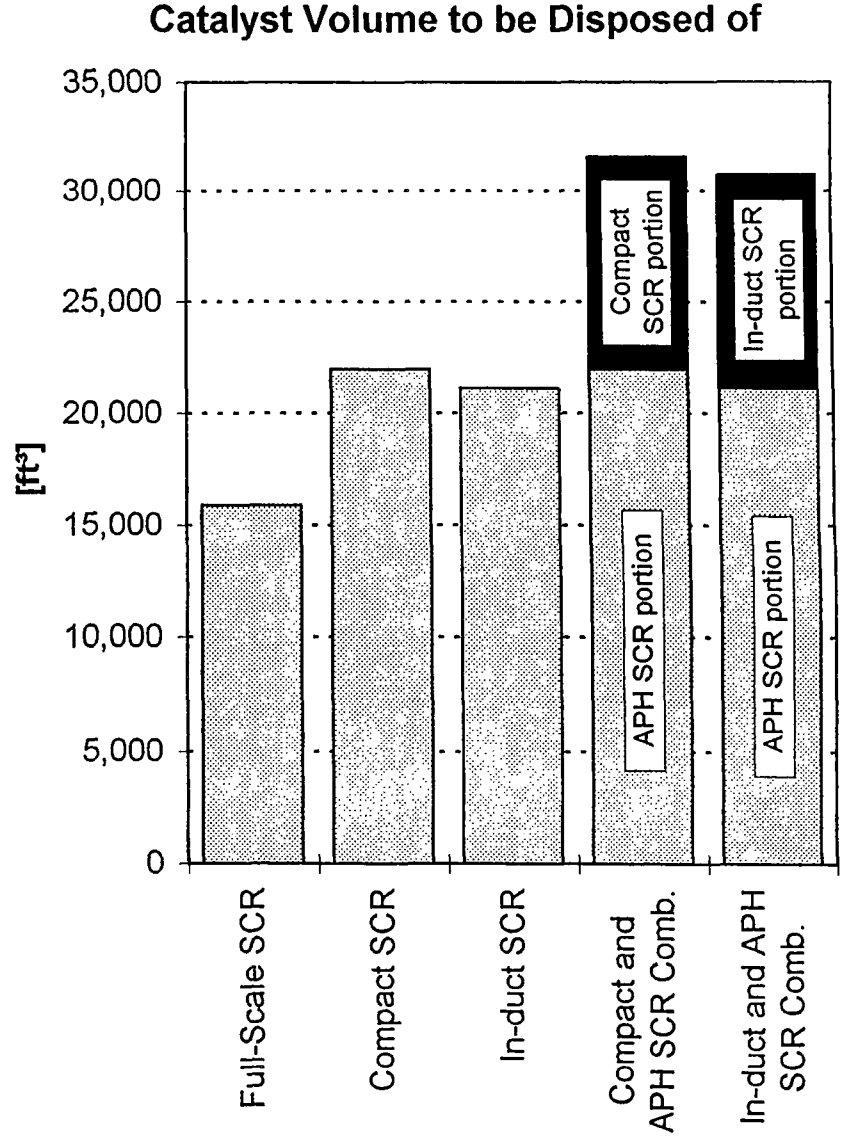


Figure 9: Catalyst volume to be disposed of over an Operating Period of 20 Years [ft³]

OPTIMIZING SCR CATALYST DESIGN AND PERFORMANCE FOR COAL-FIRED BOILERS

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Abstract

The design of an SCR system for a coal-fired application is a challenge due to particulate, catalyst poisons, and SO_2 in the flue gas. Our experience shows that coal-fired SCRs are successful when the system impacts and catalyst deterioration factors are understood and specific countermeasures are implemented in system and catalyst design.

There are several factors that the system and catalyst designers must consider in order to assure success. They include: type of boiler, required performance, fuel and ash analysis, ash loading, type of SCR, inlet conditions, catalyst deactivation mechanisms, and impact on downstream equipment.

This paper explains each system impact and catalyst deterioration factor that we have experienced. For each factor, it presents the specific countermeasures that have resulted in successful SCR systems. Also, it discusses the design of improved catalysts derived from our experience. Finally, two case studies illustrate the impact of optimized design.

Introduction

Selective Catalytic Reduction (SCR) is recognized worldwide as the most effective NOx Control technology for utility boilers and combustion turbines when substantial NOx reduction of 50% to 95% is required. In addition to its proven high performance, it has also become an economically viable solution, with current installed costs, in the United States, estimated at between $20/kw for natural gas and $50/kw-$70/kw for coal units. The technology has even given some utilities the capability to achieve lower heat rates by allowing optimization of burner operation and reduction or omission of flue gas recirculation (FGR).

Mitsubishi Heavy Industries (MHI) pioneered the development and application of SCR technology starting in the late 70's in Japan, and installed the world's first High-Dust SCR application on a 175 MW coal fired boiler in 1980. MHI transferred their technology through licensing agreements to several companies in Europe and to Cormetech in the United States. Over 319 units including 56 coal fired boilers worldwide operate successfully using this technology.

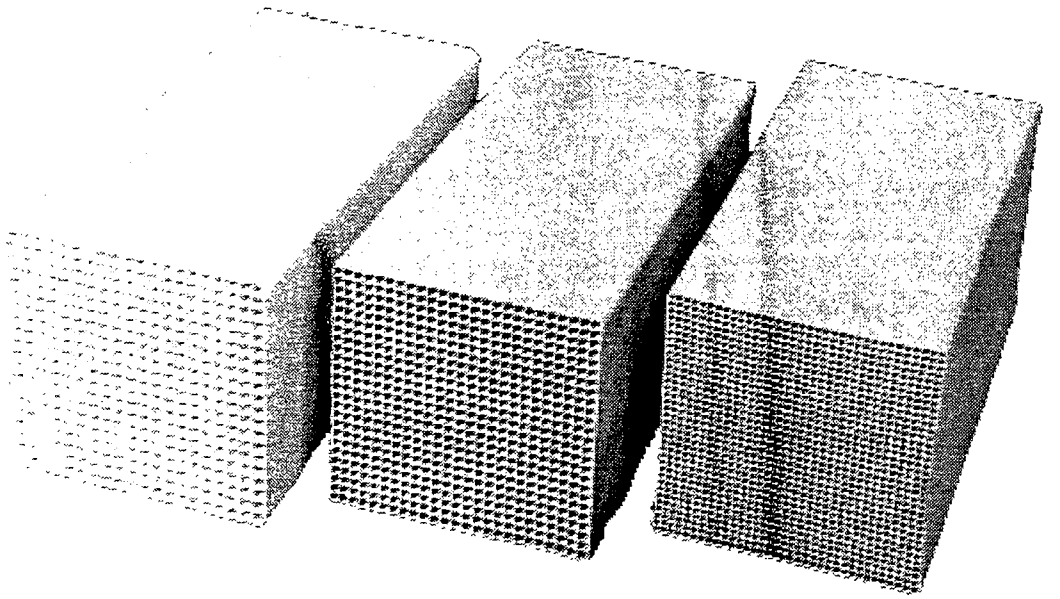

Figure 1
Catalysts for Coal, Oil, and Gas Applications

Cormetech is a joint equity company of Corning Incorporated, MHI, and Mitsubishi Chemical Company (MCC). Cormetech designs and manufactures homogeneous Titanium-Tungsten-Vanadium (Ti-W-V), extruded honeycomb catalysts for fossil-fuel-fired applications. *Figure 1* shows a photograph of catalysts for coal, oil, and gas applications. Cormetech draws upon the vast experience database of MHI, the extrusion and catalyst know-how of Corning, and the catalyst technology of MCC to provide an optimized product to the market.

This paper addresses the methods successfully implemented to design and optimize catalyst performance in coal fired boiler applications. The basics of deNOx catalysis are reviewed in preparation for the discussion on system impact, catalyst deterioration mechanisms, and countermeasures that follows. Additionally, catalyst design optimization is presented by a discussion of improved catalysts for coal applications. Two case studies (pulverized coal and cyclone boilers) comparing conventional and improved catalyst are presented.

Background of SCR Reaction

In coal fired boilers in which SOx is present in the flue gas, two chemical reactions that occur in the presence of the SCR catalyst are of most importance to the following discussion.

One reaction, of course, is the deNOx reaction:

$$4NO + 4NH_3 + O_2 \xrightarrow{\text{catalyst}} 4N_2 + 6H_2O \qquad (1)$$

Figure 2 illustrates this primary reaction mechanism. Starting in the upper left hand corner of the figure and proceeding clockwise is:

1. Active catalytic site (Me = metal)
2. Adsorption of the ammonia molecule
3. Reaction of NO with NH_3 forming N_2 and H_2O
4. Regeneration of active site by oxidation

The second reaction is oxidation of sulfur dioxide:

$$SO_2 + \frac{1}{2}O_2 \xrightarrow{\text{catalyst}} SO_3 \qquad (2)$$

The NOx conversion rate and SO_2 oxidation rate will partly depend upon the rate in which the reactants diffuse through the walls of the porous catalyst to reach active sites where reaction takes place. Through our experience we have determined the rate of diffusion of reactants for our catalyst. For deNOx conversion the rate of chemical reaction is fast relative to the rate of diffusion. Therefore, the catalyst is effective primarily at the surface of the wall. On the other hand, the SO_2 oxidation chemical reaction rate is slow relative to the rate of diffusion so that the reactants diffuse into the entire wall. This conclusion is relevant to the subsequent discussion on system impacts, catalyst deterioration mechanisms, and countermeasures.

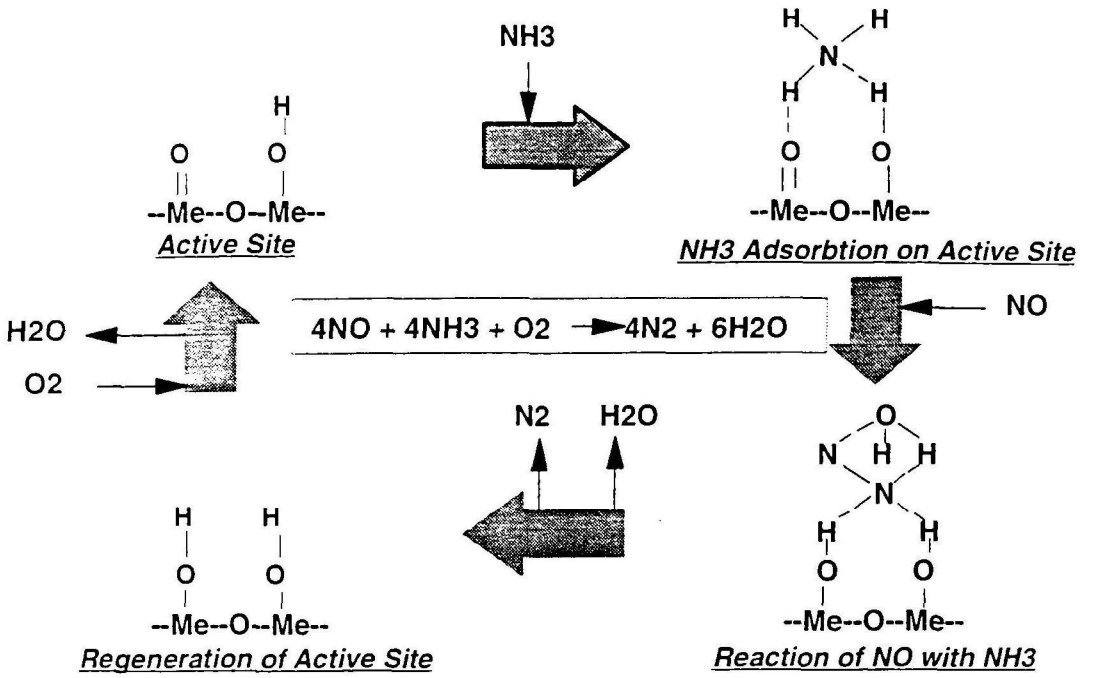

Figure 2
SCR Surface Reaction Mechanism

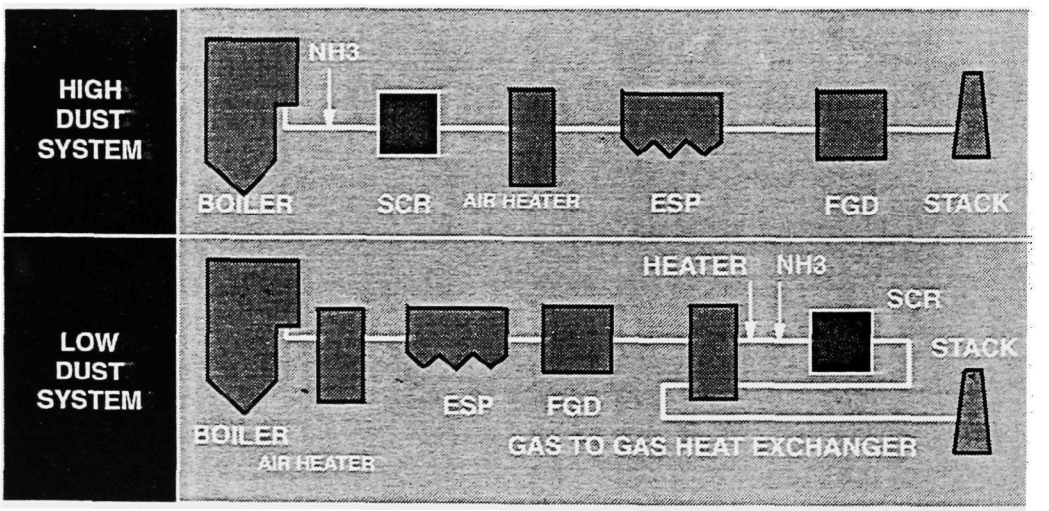

Figure 3
SCR Configurations

System Design Impacts & Countermeasures

Figure 3 shows two typical layout options of SCR systems applied to coal fired boilers. The Low-Dust or Tail-End option shows the installation of the SCR reactor after the air preheater (APH), Electrostatic Precipitator (ESP), and Flue Gas Desulfurization (FGD) systems. This option requires the flue gas to be reheated to acceptable SCR temperatures, typically 550°F - 750°F. Low-Dust designs have primarily been used on wet bottom boilers which have ash recirculation, due to concern over catalyst degradation caused by arsenic poisoning. The High-Dust option locates the SCR reactor between the economizer exit and the APH inlet. This is more traditional for dry bottom boilers and less costly, and is now being applied to wet bottom boilers based on economics, advancements in catalyst resistance to arsenic poisoning, and limestone addition to the fuel which will be discussed later.

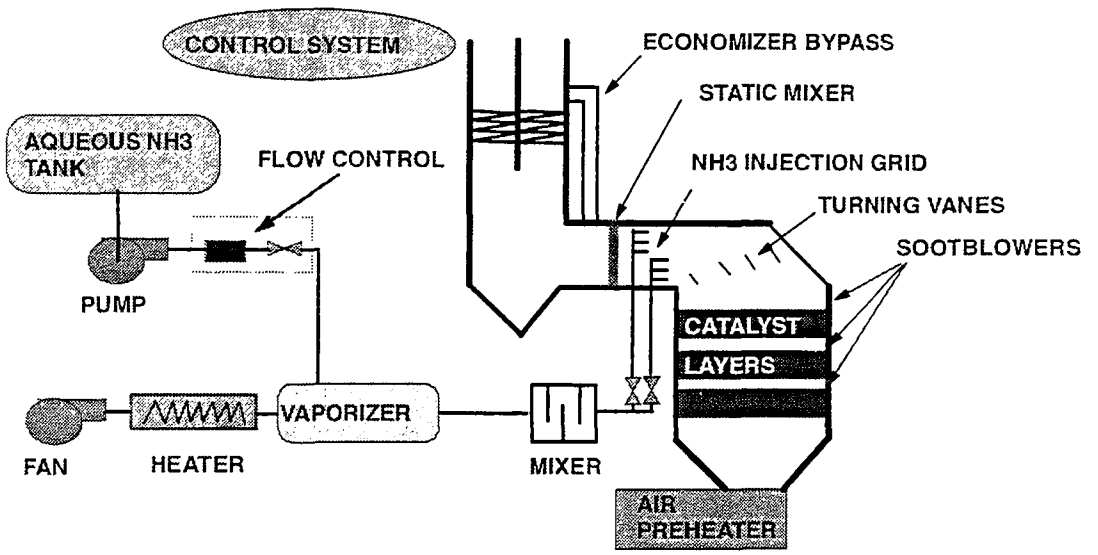

Figure 4
SCR Components

The SCR system consists of many components in addition to the catalyst. *Figure 4* depicts a high-dust SCR system which includes the catalyst reactor, economizer bypass used for part-load temperature control, static mixer used for temperature and/or ammonia mixing, turning vanes, ammonia vaporization system, ammonia injection grid (AIG), sootblowers, and the control system. Since these components directly influence overall system effectiveness, Cormetech works closely with system equipment suppliers to ensure proper design.

Operational impacts of an SCR system on coal fired boilers include draft loss, SO_3 formation, and ammonia emission. Pressure drop is caused by the installation of the catalyst material in the flue gas stream and is typically minimized by expanding the flue gas path. Typical pressure drop values for conventional high dust applications range from 0.5 to 1 inch of water per catalyst layer with 2 - 3 layers installed, with an additional 1 - 2 inches for the other components of the system.

As shown in equation (2) the SCR catalyst will convert a small amount of SO_2 to SO_3. This conversion rate is typically less than 0.5% per layer depending upon the catalyst formulation, deNOx requirements, and operating temperature. If the SO_2 oxidation rate is too high, corrosion and plugging problems will occur in the APH due to the formation of H_2SO_4 and NH_4HSO_4.

As a countermeasure, our catalyst is designed to achieve high deNOx activity while keeping SO_2 oxidation activity low. When considering the catalyst formulation, a high concentration of vanadium will result in a high deNOx activity, but will also result in a high SO_2 oxidation activity. One countermeasure to this is a vanadium concentration is selected such that there is sufficient deNOx activity with an acceptable level of SO_2 oxidation.

As a system countermeasure, the cold end surface of a regenerative-type APH may be enameled and/or sootblowing capabilities may be enhanced.

Ammonia slip may contaminate flyash, and in combination with SO_3, cause APH plugging. Designers can combat the effects by considering:

- 2 ppm to 5 ppm NH_3 slip at end-of-life.
- ammonia injection grid with adequate coverage and distribution capability.
- multi-point outlet NOx sampling control grid to assure representative NOx input value to control system.
- permanent measuring grid for AIG tuning to assure proper distribution of ammonia.
- flyash ammonia concentration monitoring.

Catalyst Deterioration Mechanisms & Countermeasures

When designing for a given application, designers must carefully review fuel and ash constituents (see *Table A*). Under ideal conditions the catalyst will reduce NOx for an unlimited period of time. However, at actual full scale operating conditions, catalyst deactivation will occur over time. *Table B* lists the primary deterioration mechanisms versus the type of coal fired unit, wet or dry bottom.

Ash Moisture, %	6 - 33
Total Sulfur in Coal, %	0.6 - 1.6 *
Trace Material, ppm	
Ni	3 - 40
Cr	7 - 46
As	1 - 25
Cl	41 - 1,900
Ash Analysis, %	
Sl O2	41 - 71
As2 O3	2 - 33
Fe2 O3	2.5 - 10
Ca O	2.4 - 26
Mg O	0.7 - 49
Ti O2	0.1 - 1.8
Mn O	0.02 - 0.2
V2 O5	0.01 - 0.1
Na2 O	0.05 - 1.6
K2 O	0.1 - 4.0
P2 O6	0.06 - 1.3
S O3	1.6 - 16.5

*** Application Experience on Oil up to 5.4% Sulfur**

Table A
Typical Fuel/Ash Composition

Boiler Type	Wet Bottom	Dry Bottom
Sintering	Negligible	Negligible
Accumulation of Alkaline Metal	Small	Small
Alkaline Earth Metal Masking	Large	Large
Accumlation of Arsenic Oxide	Large with Ash Recirculation	Moderate
Flyash Deposition*	Small	Small
Erosion*	Small	Small

***Highly dependent on proper design of SCR system**

Table B
Main Causes of Catalyst Deterioration

Based on a sound understanding of the deterioration mechanisms, specific countermeasures are implemented, such as:

- system design evaluation.
- catalyst formulation.
- catalyst volume selection.
- catalyst geometry.
- catalyst macro and micro pore design.
- catalyst characterization testing and predictive tools.

Thermal Sintering

Thermal sintering is the growth of primary catalyst particles resulting in a reduction of catalyst surface area which reduces catalyst performance. *Figure 5* illustrates this mechanism. Thermal stability is maximized with the incorporation of Tungsten in the catalyst formulation. As a result, sintering is negligible at normal SCR operating temperatures.

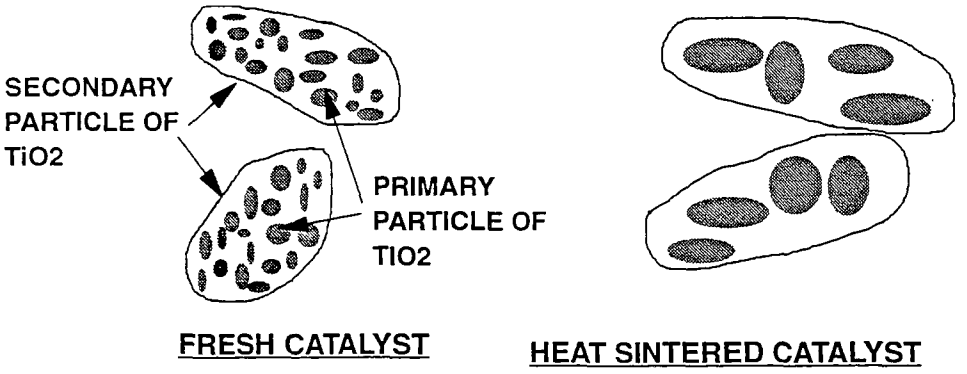

Figure 5
Thermal Sintering

Alkaline Metal (Na, K)

Alkaline metals may directly react with active sites and render them inert as shown in *Figure 6*. Since the deNOx reaction takes place primarily on the surface, the degree of deactivation depends on the surface concentration of the alkaline metals. In a water soluble form, these alkaline metals are highly mobile and will migrate throughout the catalyst material. Since the walls of our catalyst consist entirely of catalyst material, the surface concentration of alkaline metals is diluted by this migration, minimizing the deactivation rate.

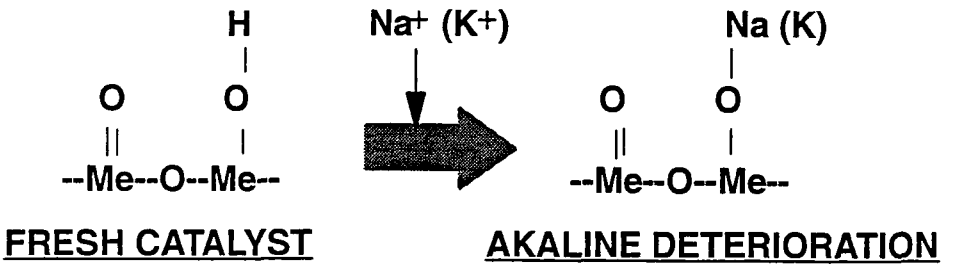

Figure 6
Alkaline Deterioration Mechanism

For the subject catalyst in coal-fired applications, it is our experience that this type of poisoning has a small impact, since most of the alkaline metals in the coal ash are not water soluble. The impact is larger in oil fired applications where the majority of the alkaline metals in the ash are water soluble or when the thickness of catalyst material is low. For example, given the same concentration of alkaline metals in the flue gas, this catalyst will have a lower surface poison concentration than a coated product and, thus have a lower deactivation rate and longer useful life as shown in *Figure 7*.

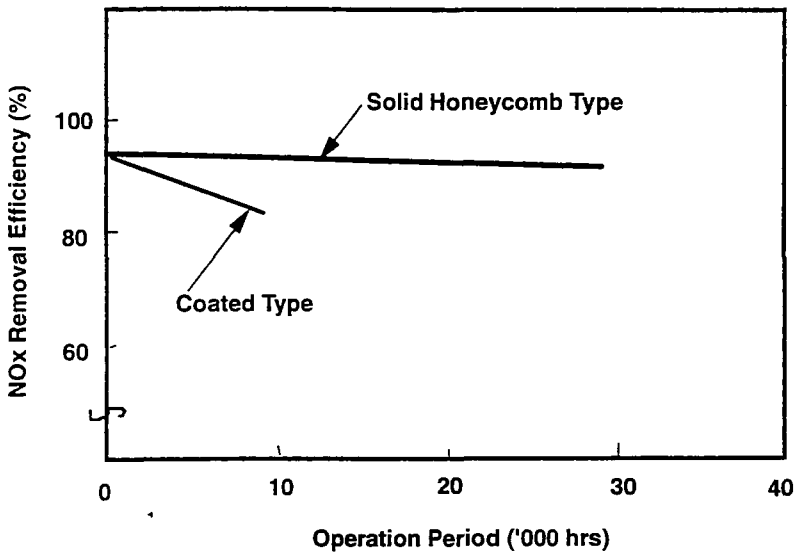

Figure 7
Comparison of Durability (Oil Firing)

Arsenic (As)

Arsenic poisoning is caused by gaseous arsenic As_2O_3 in the flue gas. The As_2O_3 diffuses into the catalyst and solidifies on both active and non-active sites, as shown in *Figure 8*.

As with alkaline metals, arsenic poisoning is best abated by use of homogeneous catalyst compositions, which effectively reduce the surface poison concentration. Since the mechanism of arsenic poisoning is more diffusion-limited and subject to capillary condensation, optimized catalyst pore structure also lessens its effects. Further, engineers use accumulation rates determined by both lab and field tests to ensure a proper catalyst volume is provided for each specific application.

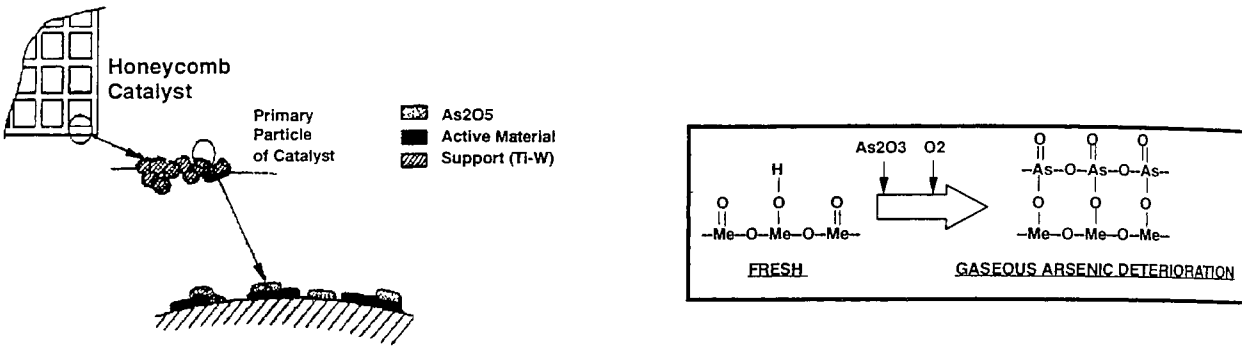

Figure 8
Model of Arsenic Compound Accumulation

A system countermeasure incorporates the use of a fuel additive. As mentioned previously, wet bottom boilers with 100% ash recirculation present a worst case scenario for catalyst deterioration caused by arsenic poisoning. In order to combat the high levels of gaseous arsenic (As_2O_3) in cyclone boilers, limestone may be added to the fuel. A typical limestone to fuel ratio is approximately 1:50. *Figures 9a, 9b and 10* show how the addition of limestone effectively reduces gaseous arsenic at the entrance to the catalyst. The free CaO in the limestone reacts with arsenic, forming a solid, $Ca_3(AsO_4)_2$, which will not poison the catalyst. The impact of adding limestone will be addressed further in the wet bottom boiler case study.

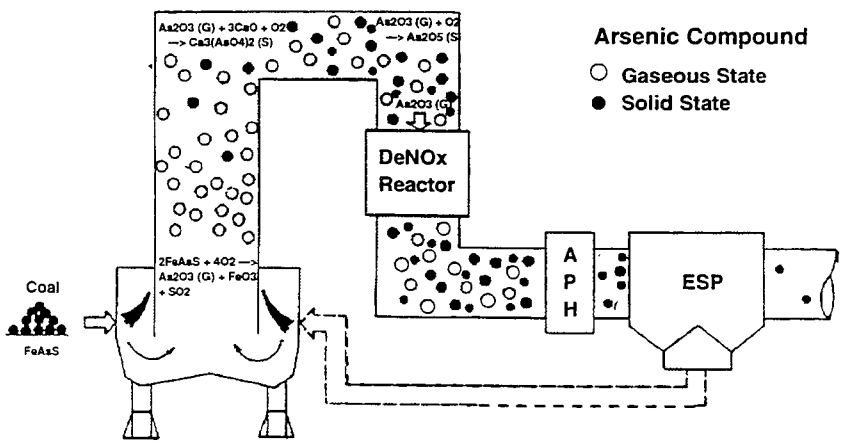

Figure 9a
Arsenic Compounds in Flue Gas

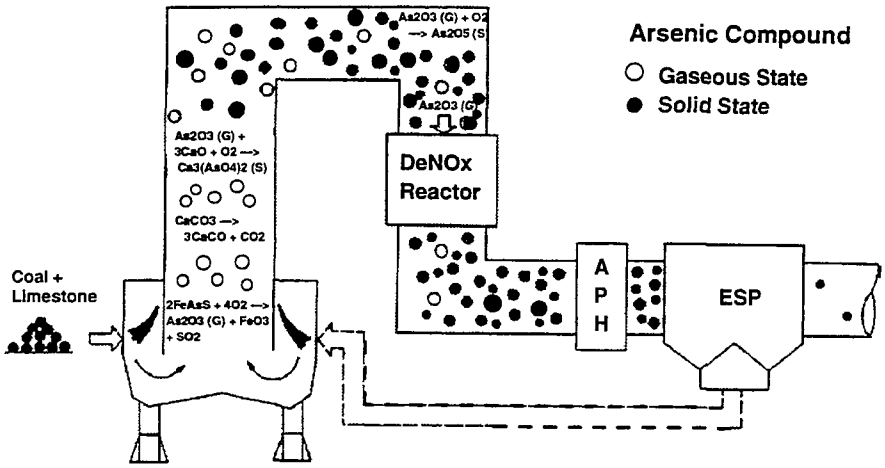

Figure 9b
Effect of Limestone Injection

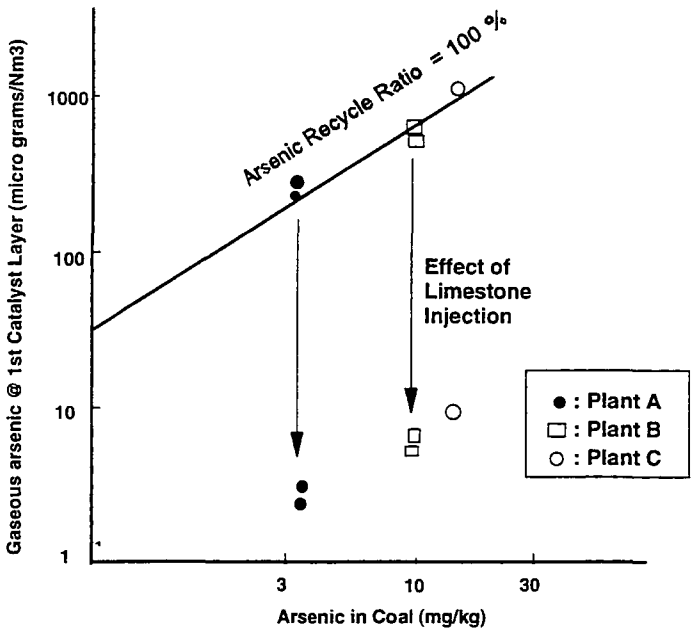

Figure 10
Relationship Between Arsenic in Coal and Gaseous Arsenic

Alkaline Earth Metal (Ca)

Alkaline earth metal poisoning primarily occurs when free CaO in the flyash

reacts with SO_3 adsorbed on the catalyst surface forming $CaSO_4$. The $CaSO_4$ may cause catalyst surface masking, preventing the reactants from diffusing into the catalyst. *Figures 11 and 12* illustrate this mechanism.

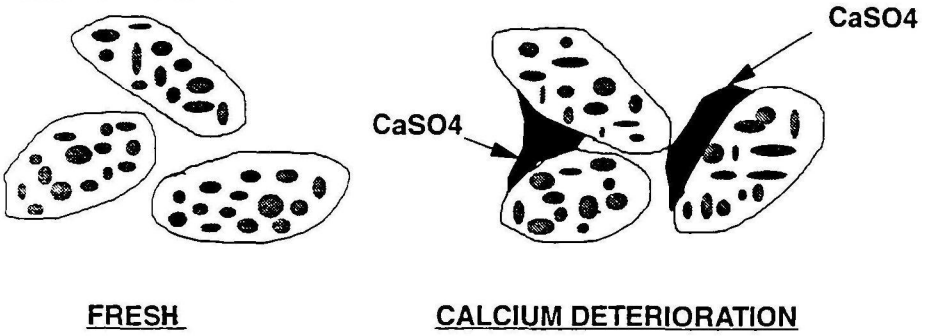

Figure 11
Mechanism of Calcium Deterioration

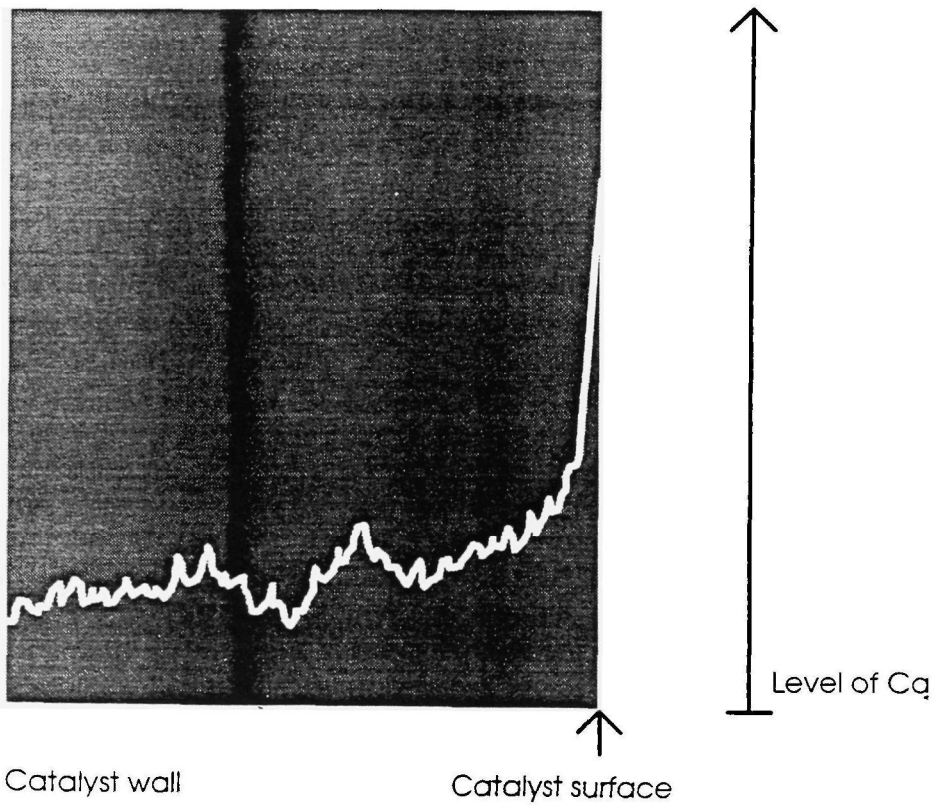

Figure 12
Accumulation of Calcium Composition in Catalyst

In order to address masking, particularly in dry bottom boilers where the free CaO content is nearly double that of a wet bottom boiler (see *Figure 13*), characterization of the mechanism and accurate prediction of deactivation rate is

employed. Experience shows that the quantity of free CaO in the fly ash governs the amount of deactivation. Extensive modeling of deactivation from full scale and laboratory data has been developed providing high confidence in design to counter this phenomenon.

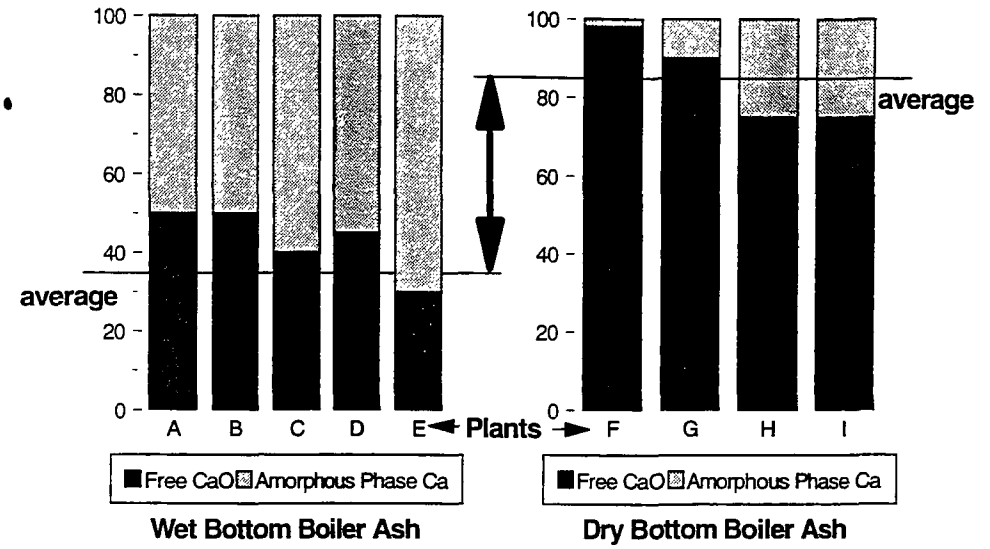

Figure 13
Free CaO Ratio in Fly Ash

Catalyst Plugging

Catalyst plugging has two primary causes: (a) ammonia salt deposition and (b) flyash deposition. Proper selection of catalyst pitch and cell opening size minimizes plugging. Good system design is also a primary prevention tool.

Ammonia salt formation/deposition will not occur provided the SCR inlet temperature is maintained above the salt formation level. Typically temperature maintenance is only a problem during part load operation and is elevated by installing a water- or gas-side economizer bypass in the system. If a gas side bypass is used, careful attention must be paid to provide adequate temperature mixing upstream of the AIG. This avoids low temperature streamlines which can cause localized salt formation at the AIG and catalyst.

Flyash deposition is minimized through proper flow distribution across the catalyst face. This is accomplished through engineering of SCR system flue layout, Computational Fluid Dynamics (CFD) codes, and/or cold flow modeling.

The design must limit the number of areas where dust accumulation may occur, i.e.,

- avoiding flat surfaces such as long leading edges to turning vanes;
- installing ash deflection devices on support beam flanges and sootblower pipes;
- avoiding flue dead legs between dampers where ash accumulation may occur in closed position and exhaust onto catalyst when opened, etc.

Erosion

Catalyst erosion is caused by the impingement of flyash on the catalyst face. Catalyst erosion is a function of gas velocity, ash character, angle of impingement, and catalyst properties. Cormetech has experience with flyash loadings as high as 30 g/Nm^3 and flue gas velocities up to 6.2 m/s.

Experience has proven that no significant erosion will occur with proper system design, catalyst material durability, and catalyst edge hardening. Poor flow and ash distribution at the inlet to the catalyst has been the source of most problems, and has improved significantly over the entire SCR experience history. Careful attention must be paid to utilize flow modeling (CFD and cold flow modeling) to properly simulate flow profiles through the SCR system. A flow rectifier grid should be installed to straighten flow into the catalyst bed. Since most catalyst erosion occurs at the catalyst leading edge which is exposed to the direct impingement of dust particles the catalyst may also be hardened at the entrance region to provide further protection as shown in *Figure 14*.

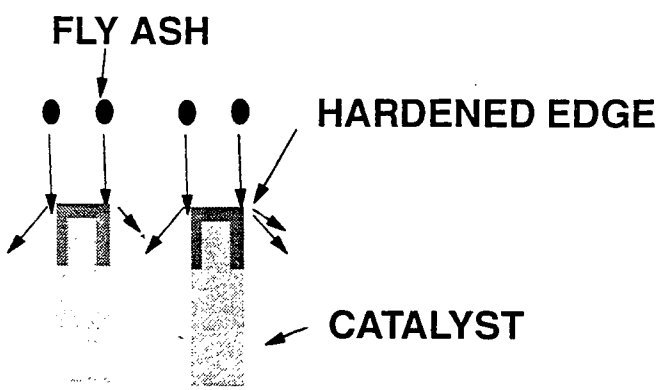

Figure 14
Honeycomb Catalyst With Hardened Edge

Although in most cases In-Duct SCR is not attractive due to extremely high system draft loss and difficulty in control of gas flow and ammonia distribution, some utility sites are considering the concept due to site plan limitations.

Therefore, we will pilot test our hardened edge catalyst at velocities > 12 m/s in 1995.

Catastrophic Failure of SCR Catalyst

Catastrophic failure is defined as sudden and permanent loss of catalytic performance. Based on our experience, catastrophic failures are extremely rare. The primary cause is associated with the ignition of ash buildup. The intense heat of a fire can irreversibly damage any SCR catalyst. Cormetech's ceramic honeycomb catalyst, unlike plate catalyst with stainless steel mesh substrates, will not promote oxidation, therefore fires are less likely to spread and are more easily contained.

Coal Catalyst Developments

Based on our understanding of reaction and deactivation mechanisms we are improving our coal catalyst. The objective is to increase the deNOx reaction rate without increasing the SO_2 oxidation rate. One method is to modify the pore structure of the catalyst wall in order to reduce diffusion resistance. The other is a more novel method and requires further explanation.

Since the deNOx reaction is only effective at the surface, we have developed a improved catalyst of the same composition as our conventional catalyst with the exception that the vanadium is preferentially distributed to the surface of the catalyst wall. In this manner, we can maximize the concentration of vanadium in the effective fraction of the wall, maximizing deNOx activity without incurring high SO_2 oxidation. *Figure 15* shows the performance of this improved catalyst compared to that of conventional catalyst. Since the product is still made entirely of catalytic material, resistance to poisons is maintained.

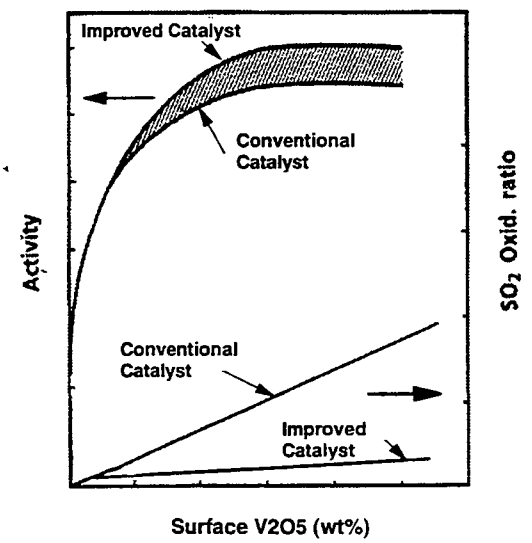

Figure 15
Comparison of Catalyst Performance

Development and Qualification of Improved Coal Catalysts

We are developing the improved coal catalysts described above according to the schedule shown in *Figure 16*.

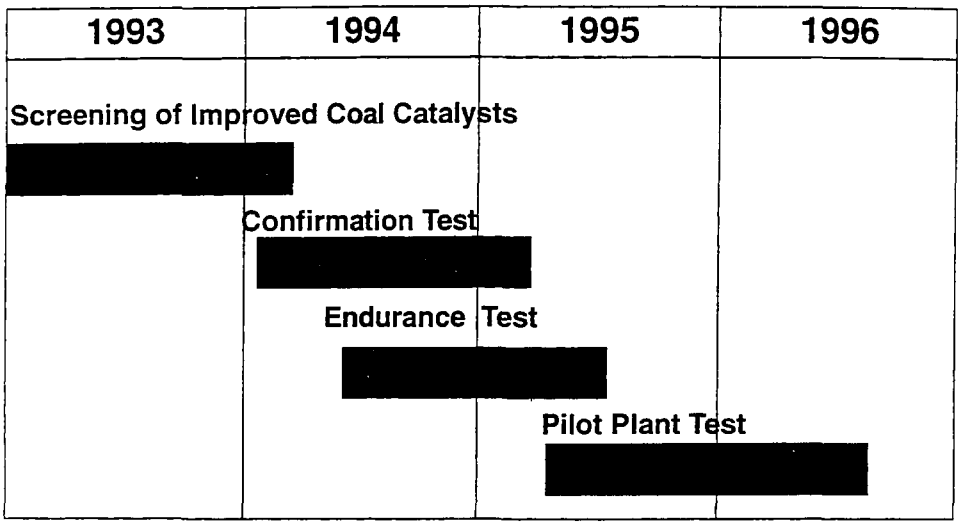

Figure 16
Improved Coal Catalyst R&D Schedule

We have already completed laboratory performance confirmation and durability tests. The performance of the improved catalyst is approximately 15% higher than our conventional catalyst, without impact on SO_2 oxidation.

We have tested durability to gaseous arsenic poisoning by an accelerated method using the apparatus in *Figure 17*. The test results are shown in *Figure 18*. The improved catalyst maintains its performance advantage over conventional catalyst at any arsenic poisoning level.

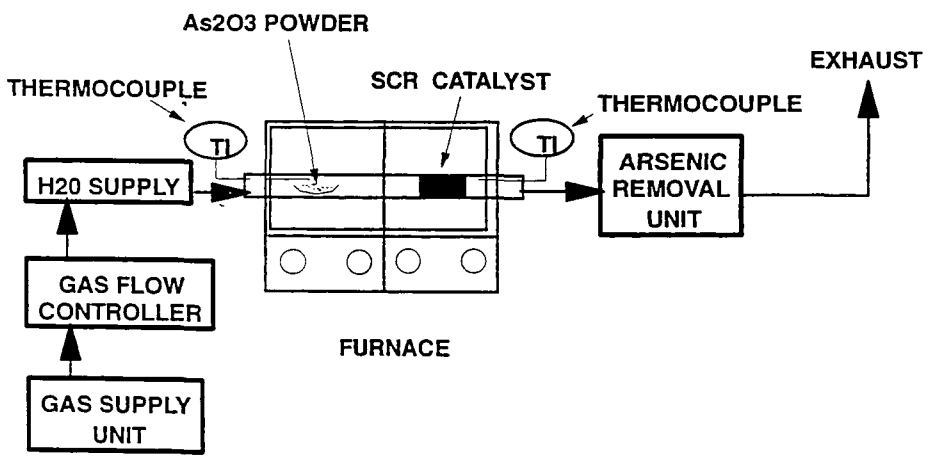

Figure 17
Schematic of Arsenic Testing Apparatus

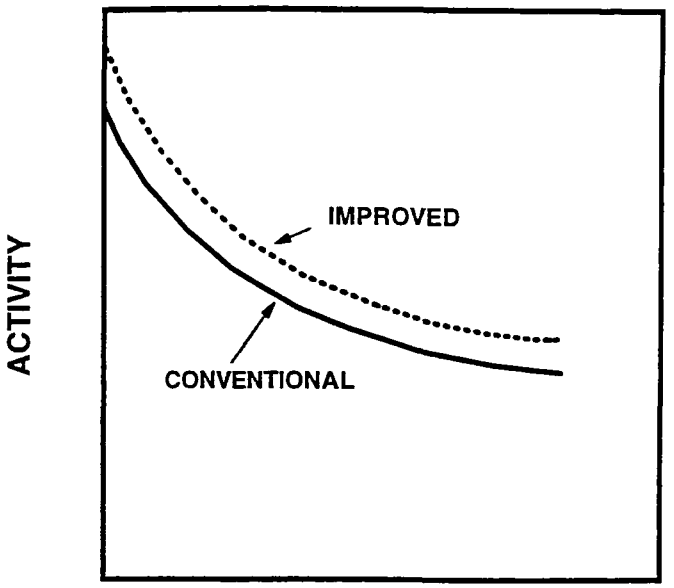

Figure 18
Arsenic Durability

Since the degree of deactivation due to surface masking by $CaSO_4$ is dependent only on the fraction of surface masked we expect that the rate of deactivation is the same for both conventional and improved catalyst. Therefore, given the same installed catalyst volume, the improved catalyst will maintain field performance much longer due to the higher initial activity, as shown in *Figure 19*.

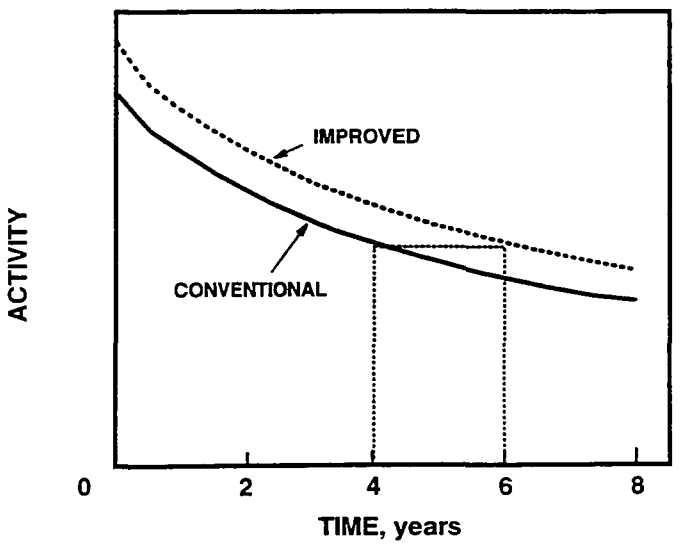

Figure 19
CaO Durability

Before commercialization, the endurance of the catalyst will be confirmed by operation in an actual plant. Initial endurance tests will start by charging catalyst samples into commercial SCR plants in Japan. During 1995 catalyst samples will be charged into a coal unit in the United States and three coal units in Germany. The performance of the catalyst samples after aging in the commercial plants will be compared to fresh performance in order to evaluate endurance.

Impact of Improved Coal Catalysts on SCR Design for Coal Fired Boilers

Whether achieving an increase in performance by optimizing the pore structure of the catalyst or by the method described above, the impact on SCR design is significant. In order to illustrate the impact of improved catalysts on the SCR design, two SCR case studies are described below comparing improved with conventional catalyst.

Case Study 1: Dry Bottom Boiler

A 250 MW unit is selected as an example unit. *Table C* shows the design data for this unit. *Figure 20* shows the 10 year catalyst management plan. Since the activity of the improved catalyst is higher by 15%, the catalyst volume of the improved catalyst is decreased by approximately 15% from the conventional catalyst volume.

Design Condition:	
Fuel	Coal
Plant Output, MW	250
Flue Gas Flow Rate, Nm3/Hr	784,000
NOx Inlet, ppmvd @ 3% O2	154
NOx Outlet, ppmvd @ 3 % O2	31
NOx Removal Efficiency, %	80
Ammonia Slip, ppmvd @ 3% O2	5
Specification:	
Type of System	SCR
Type of Reactor	Vertical Flow-Fixed Bed
Type of Catalyst	Honeycomb
Number of Reactors	1

Table C
Dry Bottom Boiler

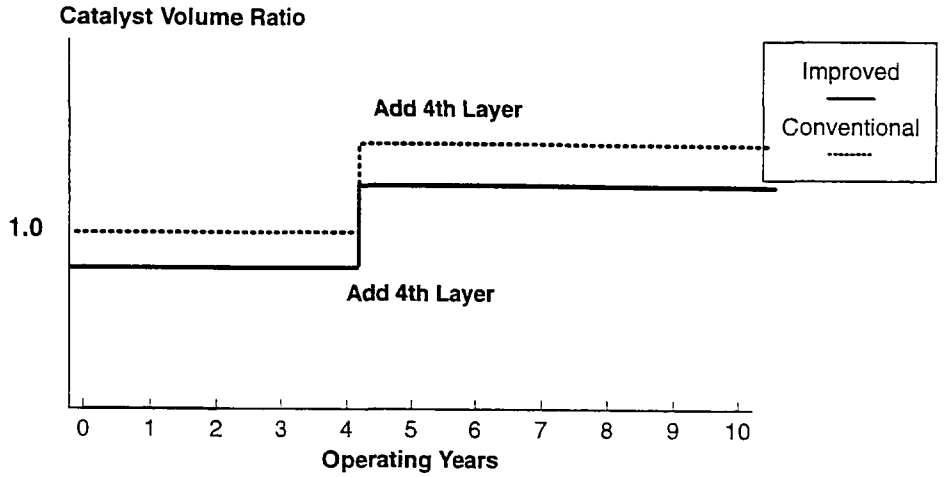

Figure 20
Impact of Improved Catalyst
Dry Bottom Boiler

Case Study 2: Wet Bottom Boiler with Ash Recirculation

A 320 MW unit is selected as an example unit. *Table D* shows the design data for this unit. *Figure 21* shows the 10 year catalyst management plan. The total catalyst volume of the improved catalyst for ten years of operation is two-thirds that of the conventional catalyst. As previously mentioned, limestone injection will extend the catalyst life. The impact on the catalyst management plan is shown in *Figure 22*. With limestone injection, the need for additions and replacements of catalyst are significantly delayed.

Design Condition:	
Fuel	Coal
Plant Output, MW	320
Flue Gas Flow Rate, Nm3/Hr	1,125,000
NOx Inlet, ppmvd @ 3% O2	1,652
NOx Outlet, ppmvd @ 3 % O2	530
NOx Removal Efficiency, %	68
Ammonia Slip, ppmvd @ 3% O2	5
Flyash Recirculation Rate, %	100
Specification:	
Type of System	SCR
Type of Reactor	Vertical Flow-Fixed Bed
Type of Catalyst	Honeycomb
Number of Reactors	1

Table D
Wet Bottom Boiler

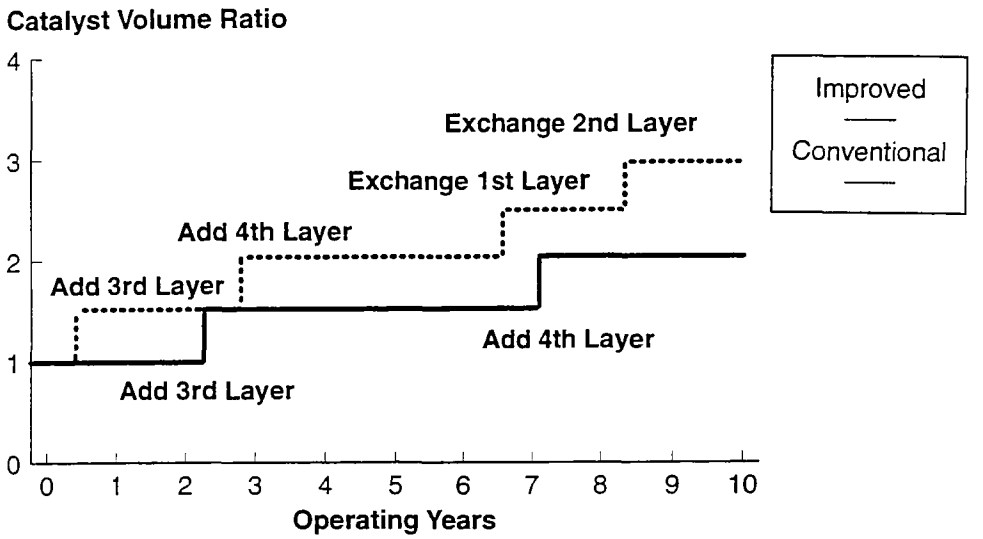

Figure 21
Impact of Improved Catalyst
Wet Bottom Boiler

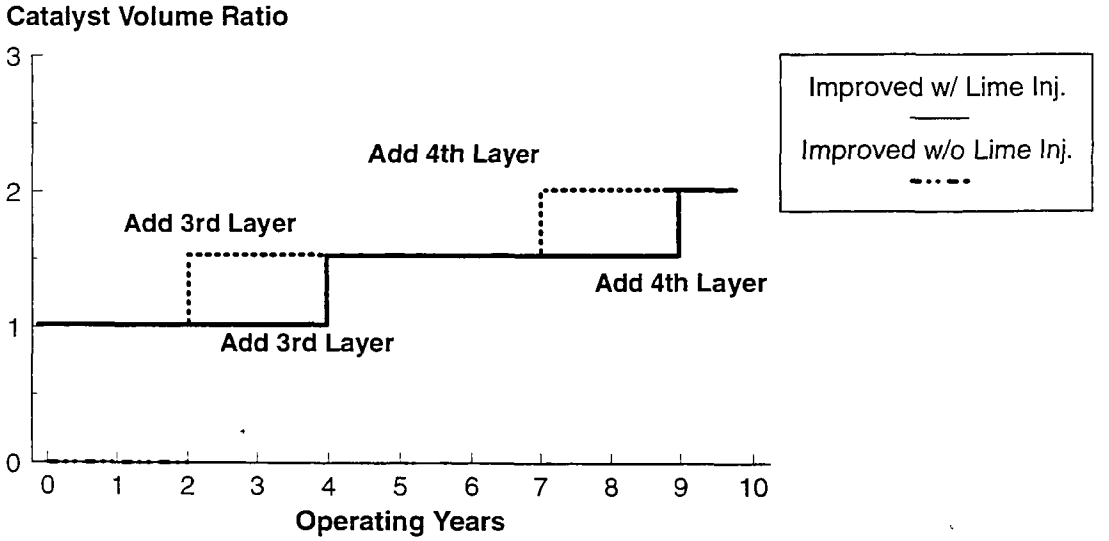

Figure 22
Impact of Limestone Injection
Wet Bottom Boiler

Conclusions

Our experience shows that coal-fired SCRs are successful when proper attention is paid to system and catalyst design.

Proper system design is crucial to maximizing catalyst performance potential and achieving the lowest annualized cost.

By exploiting the experience and know-how gained on existing facilities it is possible to further optimize catalyst design.

Fluid Dynamic Optimization of SCR Plants by Modeling Demonstrated by the SCR Plant Logan Generating Plant

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Abstract

The SCR process for the reduction of nitrogen oxides has proven itself in practical application in many fossil-fired power stations.
The concept of integrating a high-dust SCR plant between the boiler and air preheater has often been hampered by structural constraints. As a result of the limited space available, the plant's ducting will have in most cases a high degree of complexity. In order to achieve efficient SCR plant functioning, however, the flow conditions have to be analyzed in detail and to be optimized. The goal is to obtain uniform flow distribution upstream of the catalysts and a homogeneous distribution of the reducing agent ammonia, while incurring as small a pressure drop as possible. To this end, fluid dynamic measures are implemented such as the installation of guide vanes and static mixing systems.
This paper describes the procedure used to meet the requirements mentioned above. It enables the changes and modifications to be developed efficiently in terms of time and cost. Experience shows that the results obtained can indeed be transferred to and confirmed in a full-scale plant.
The fluid dynamic solutions derived by modeling and the adaptation of a static mixer to the geometry of the flue gas ducting are demonstrated using the example of the SCR plant at the Logan Generating Plant operated by US Generating Company. The results of the measurements are presented to illustrate the mechanisms of the modifications.

Introduction

The department responsible for heat transfer and fluid dynamics at Siemens/KWU has been optimizing SCR plants for approx. ten years. The stringent requirements with regard to homogeneity of the flue gas flow led to the development of special static mixers for the needs of SCR plants.

The experimental procedure most widely used currently has the advantage over the numerical method that it is faster and less expensive. The experimental procedure and the design of a suitable static mixer is explained using the example of the Keystone coal-fired power plant, and examples are shown of the measures taken and the results achieved.

General Requirements for Flue Gas Flow

It is evident that an optimized design of the SCR system requires proper distribution of flue gas and ammonia at the catalytic reactor inlet. To this end, a scale model of the SCR system was designed and constructed from the economizer outlet to the air heater inlet. The effects of different measures such as guide vanes, mixers and the dust behavior was then studied under conditions reflecting the most significant boiler load cases. The final SCR duct system, including positioning of guide vanes, mixers and ammonia injection grid reflects the results of the flow model test.
Variations over the reactor cross section area as they are considered for catalytic reactor design and as they are state of the art in many SCR systems are (Figure 1):

- Temperature distribution
  The catalyst activity is a function of operating temperature. In reactor areas in which the temperature deviates from the design temperature, the degree of NOx conversion will deviate from the design value. Ordinarily, the catalytic reactor design is based on an uneven temperature distribution of +/- 15 K deviation from the mean value.

- Inlet flow
  The inlet flow devices should be designed to obtain the most even possible inlet flow in all areas of the reactor. Only then is the residence time of the flue gases and hence of the NOx conversion equal in all parts of the catalytic reactor. Catalytic reactor design ordinarily accounts for an uneven velocity distribution of +/- 15% (deviation from the mean value) over 90% of the cross-sectional area and +/- 20% over the remaining 10% of the cross-sectional area.

- NH_3/NOx Concentration Ratio
  The NH_3/NOx concentration ratio required for a specific NOx conversion should not deviate more than +/- 5 % from the theoretical value. Larger deviations can result in excessive NH_3 slip concentrations in parts of the SCR reactor.

A further general criterion is that the flow in the flue gas ducting and in the reactor involve the smallest possible pressure drop. To achieve this, model flow experiments are used to optimize the location of guide vanes.

Guide vanes for flow direction and other reactor internals must be designed such that no ash streams arise and that ash deposits are prevented in the downstream zone.

Static Mixer for Homogeneous Flue Gas Flow in SCR Plants

Fig. 2 shows a typical static mixing system with NH_3 injection for an SCR plant with economizer bypass. The flue gas mixer (premixer) heats the main gas flow by mixing with the bypass flow. In addition, the flue gas mixer equalizes any uneven flue gas velocity distribution.

The downstream NH_3 mixer (main mixer) homogeneously mixes the NH_3 which is injected through lances into the flue gas flow.

The mixers extract energy from the flue gas flow and convert this energy into secondary flow and turbulence. Optimum adaptation of the mixer to duct geometry of the SCR-plant results in the extraction only of as much energy as is absolutely necessary, thus minimizing the pressure drop.

Special Conditions for Flow Direction of Flue Gas in the SCR Plant

The Logan coal-fired cogeneration facility, formerly referred to as the Keystone Cogeneration facility, was constructed in Logan township, New Jersey by Bechtel for the U.S. Generating Company. Both the steam generator and the SCR unit were supplied by Foster Wheeler Energy Corporation.
As is shown in Figure 3, the SCR unit is placed between the economizer and the air heater. This location is chosen because it gives the optimum gas temperature range for commercially available catalysts. The flue gas flows vertically downward through the catalyst layers.

Significant characteristics of the flue gas flow in the Logan plant are:

- Economizer bypass duct
  During part load operation (e.g. 30%), the temperature of the flue gas leaving the economizer is low and consequently there is a potential for formation of ammonium sulfate/bisulfate in the catalyst which would have harmful effects on the catalyst. Therefore, an economizer bypass duct is installed to maintain the flue gas temperature above approximately 320 °C (610 °F).

- Short mixing paths from NH_3 injection point to catalytic reactors
  The distance from the ammonia injection grid (AIG) to the first catalyst level is 50 feet. This corresponds to roughly 1.5 times the hydraulic diameter and is considerably less than in most of the SCR plants constructed to date.

- 180° flow deflection from flue gas duct to reactor
  The flue gas flow is redirected by 180° just upstream of the catalytic reactors, with a simultaneous increase in the flow cross-sectional area from that of the duct to that of the reactor.

Fundamentals of Model Flow Experiments

Flow Model

A 1:20 scale Plexiglas model was built to investigate flow direction, uneven distributions and other flow aspects. The following components of the SCR plant in the Logan power plant were modeled:

- Boiler economizer area
- Bypass ducts
- Flue gas duct from the boiler to the SCR reactor with flue gas mixer and NH_3 injection system
- SCR reactor
- Flow straighteners in the SCR reactor

The pressure drop across the catalytic reactors was simulated using sheets of perforated metal. Over the course of the investigations, various internal components such as guide vanes and mixers were tested in this model and their arrangement as well as size was optimized.

The model experiments were performed with air as the flow medium. Significant criteria which are to be considered and which therefore can be transferred from the model experiments to the full scale plant, are described in the following.

Similarity Theory

In the case of an isothermal, single-phase, steady-state flow, similarity theory requires the following to ensure that the model results can be transferred:

- Adherence to the geometrical similarity of the model, i.e. the model must be reduced to a suitable scale and must properly represent the geometry of all parts of the flow space.

- Adherence to the Reynolds number Re:

$$Re = \frac{w * D_h}{\nu}$$

The following difficulties arise when these criteria are imposed:

The Reynolds number of the model cannot correspond to the Reynolds number of the full-scale plant, as d_h in the model is reduced by the model scale of 20, whereas ν is only reduced by a factor of 4. To compensate for this, the velocity would have to be increased, but this in turn would endanger the similarity due to compressibility influences. For industrial investigations, the strict adherence to constant Re can be relaxed, as the flow conditions in the present highly turbulent flow ($Re > 2*10^5$) are hardly still dependent on the Reynolds number.

The adherence to geometrical similarity (1:20 scale) in general leads to technical difficulties in manufacturing the internals. For this reason, the model internals are replaced by geometrically simpler components, which however are similar to the real internals in relation to their fluid dynamic effect. The important effect of the internals is that they cause a change in the flow distributed over the duct cross section due to their pressure drop. The components used in the model must also cause the same effect. This is achieved if the flow resistance of the component in the original system and in the model are the same, i.e. the friction coefficient (ξ) in the model and original must be equal. In most applications, perforated metal sheets are used to simulate the friction coefficient.
For components which have a clear spatial structure and which offer different resistances in the three orthogonal directions, special constructions must be used with which the spatial flow processes can be simulated.

Similar considerations hold for the transfer of the tracer experiments used to verify the propagation of an NOx or temperature striation from the boiler until upstream of the first catalyst level or the distribution of the injected NH_3. The determining physical processes are here on the one hand molecular mixing and on the other turbulent mixing of the gas striations in the surrounding gas flow. For the gases considered in the full-scale plant, as well as in the model, turbulent mixing is the determining factor, and not molecular diffusion. Mixing is thus purely a flow phenomenon.

Individual Experiments and Results

Temperature Distribution across the Reactor Cross Section

These experiments are intended to show whether nonhomogeneous temperature distributions arise upstream of the first catalyst level following the combination of the bypass flows with the main flow. Figure 4 shows the experimental equipment and instrumentation. A fine mesh

injection grating was used to input the tracer gas homogeneously in the two bypass flows. The tracer gas was a 4% vol. propane/air mixture which can be detected using a flame ionization detector. The concentrations determined in this way were recalculated to temperatures, since, as was already explained, mixing of temperature and concentration are caused exclusively by turbulence in the highly turbulent model flow.
Following admixture of the bypass streams, the maximum temperature deviations upstream of the first catalyst level were +50°F/-70°F without the use of the flue gas mixer, and +25°F/-20°F with flue gas mixer (see Figures 5 and 6).

Velocity Distribution Upstream of First Catalyst Level

A hot wire anemometer was used for the velocity measurements. A thread probe was used to detect crossflow and backflow. As the Reynolds number was in the highly turbulent range in the model, the velocity measurements can be directly transferred to the original.

The velocity profile upstream of the first catalyst level was influenced not only by the arrangement of the guide vanes, as is shown in Figure 7, but also by the flue gas mixer, Figure 8.

In addition to the guide vanes, the mixing system also smoothed the uneven velocity distribution, which arises from the two flow redirections downstream of the boiler outlet. The flue gas mixer thus also takes on the function of the guide vanes, as was already be seen from the effect on the temperature distribution.

Simulation of Concentration Distribution of Injected Ammonia

One of the prerequisites for these experiments was scale modeling of the NH_3 injection system. For the measurements, tracer gas was evenly injected from the nozzles of the injection system into the main flow. To enable direct transfer of the results of the NH_3 concentration measurements to the original system, injection in the model was performed with the same momentum ratio as in the full-scale system. The momentum ratio was selected as the ratio of the momentum of the nozzle exit flow to that of the main flow.
Injection in the model was performed such that approximately the same volume flow rate of tracer gas was injected by all nozzles.
The concentration distribution upstream of the first catalyst level was determined using a flame ionization detector. The result is shown in Figure 9. Based on the deviation of +8%/-16% in the concentration distribution determined in the experiment it can be concluded that suitable regulation of ammonia injection in the full-scale system will present no problems in achieving the required allowable deviations of +/- 5%.

Following this, an experiment was performed using "selective injection", i.e. injection through only one nozzle. This should verify that the NH_3 concentration can be controlled in selected parts of the reactor by trimming the individual NH_3 nozzles. As is shown in Figure 10, this is possible, despite the presence of the flue gas mixer upstream of the NH_3 injection.

Minimizing Pressure Drop

All of the individual steps for optimization of the arrangement of the guide vanes were also evaluated with regard to pressure drop.
The pressure drop measurements confirm that an even flow also results in the minimum pressure drop. The arrangement of the guide vanes in the 180° redirection results in a reduced pressure drop over that for redirection without fluid dynamic internals due to the evenly distributed redirection of momentum.

The pressure drop due to the flue gas mixer is roughly 1.3 mbar at full load.

Minimizing the Possibility of Fly Ash Deposits

Fly ash deposits can be minimized by optimization of the flue gas ducting and by the installation of corresponding guide elements. Therefore, even in the velocity measurements in the model, emphasis was placed on the fluid dynamic internals generating no ash deposits which could impair operation. This was achieved by such measures as avoiding long horizontal sections in designing the guide vanes. The guide vanes furthest out in the reactor enclosure have a 60° redirection angle instead of a 90° angle. This shortens the horizontal section of the upstream edge, thus preventing ash deposits on the guide vanes.
Two processes were investigated in the fly ash experiments. First, the location and amount of deposits were determined, and then the behavior of ash deposits at increased flow velocity was determined.

Fly ash from the power plant was used in order to enable simulation of the deposition mechanism. The experiments were performed using an accelerated procedure in which the ash concentration was three times higher than that in the full-scale system. This enabled drawing of conclusions after relatively short experiment times, concerning where fly ash sedimentation can be anticipated at lower load conditions and whether a limiting condition is attained. The experiments also yielded indications as to whether deposits which would impair operation could be anticipated in the full-scale system.

In the part-load case, all significant plant components remained nearly free of deposits, even after a six-hour experiment with an ash-laden medium. The ash deposits on the flow straightener were minimal, and the attached deposits were oriented vertically, which indicates vertical impingement. Increasing the flow velocity did not result in abrupt removal of the ash deposits, which were gradually eroded away. This provides additional confirmation of the optimum design of the guide vane configuration as well as of the duct geometry.

Summary

The results obtained in the model flow experiments were implemented in the full-scale system. These are as follows:

Static flue gas mixer,
Location of the NH_3 injection,
Locations of the guide vanes and their geometry.

Without model experiments, the need for a flue gas mixer would not necessarily be recognized immediately. The model experiments made it clear that problems would have arisen in the plant without this component. The mixer developed by Siemens/KWU is simple and effective, as has since been verified in other implementations, including Mercer Station.
In addition to the flue gas mixer designed for this specific application, there is also a mixing system for the injected NH3. However, this system was not required in the case of the Logan plant.
The model experiments also served to verify that the planned NH_3 injection design fulfills the requirements for NH_3 distribution and control.
Adherence to the specified pressure drop for the overall plant was both verified and made possible by the model experiments.
Overall, it can be stated that the performance of model experiments is recommended, as the results are available before construction of the full-scale system. These results can then be used to make any necessary component modifications without the necessity of first suffering from negative operating experience. The expenditure for model experiments is thus slight in comparison with possible later modification work in the full-scale system.

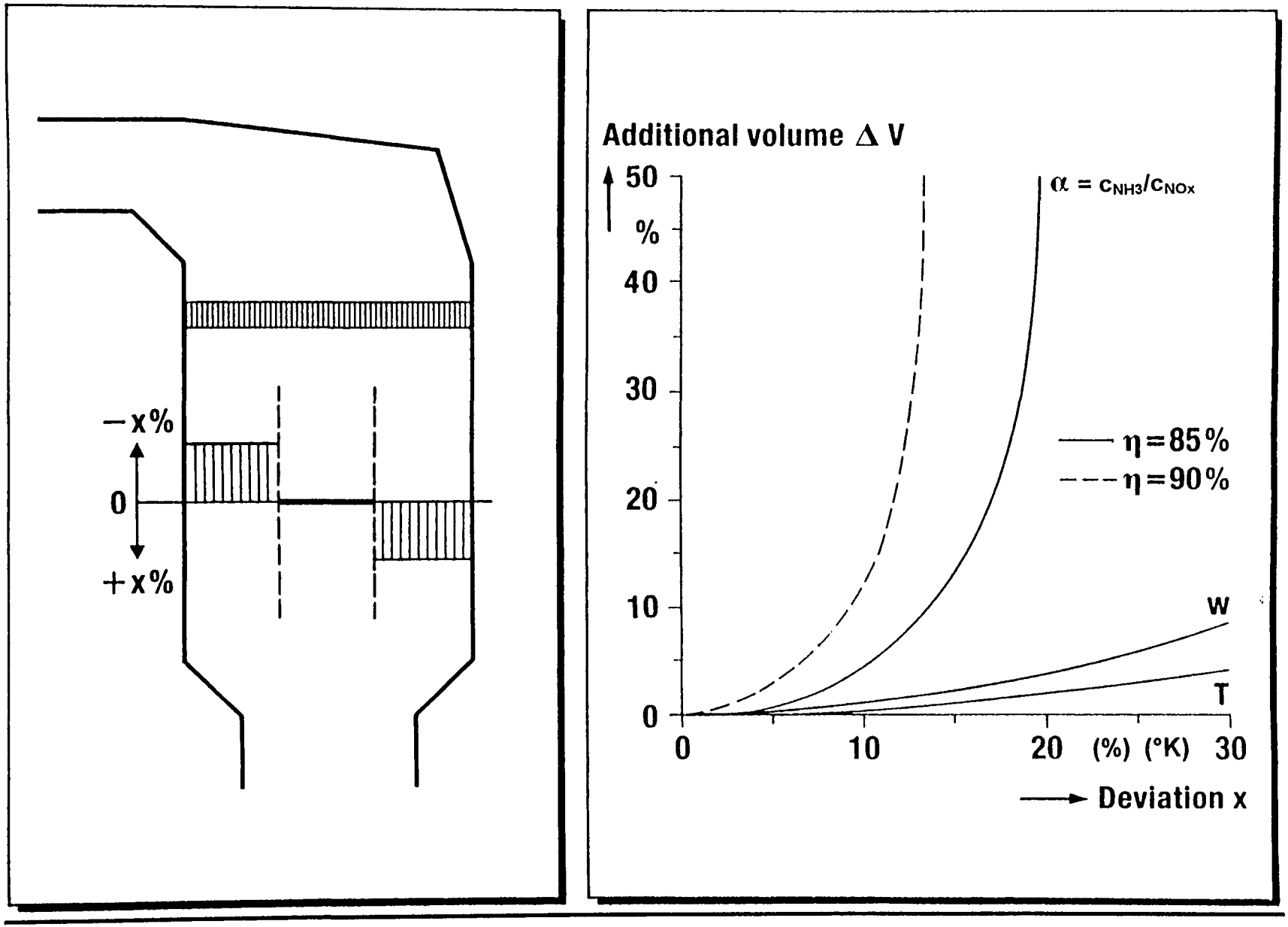

Figure 1: Effects of an Unequality on the Catalyst Volume

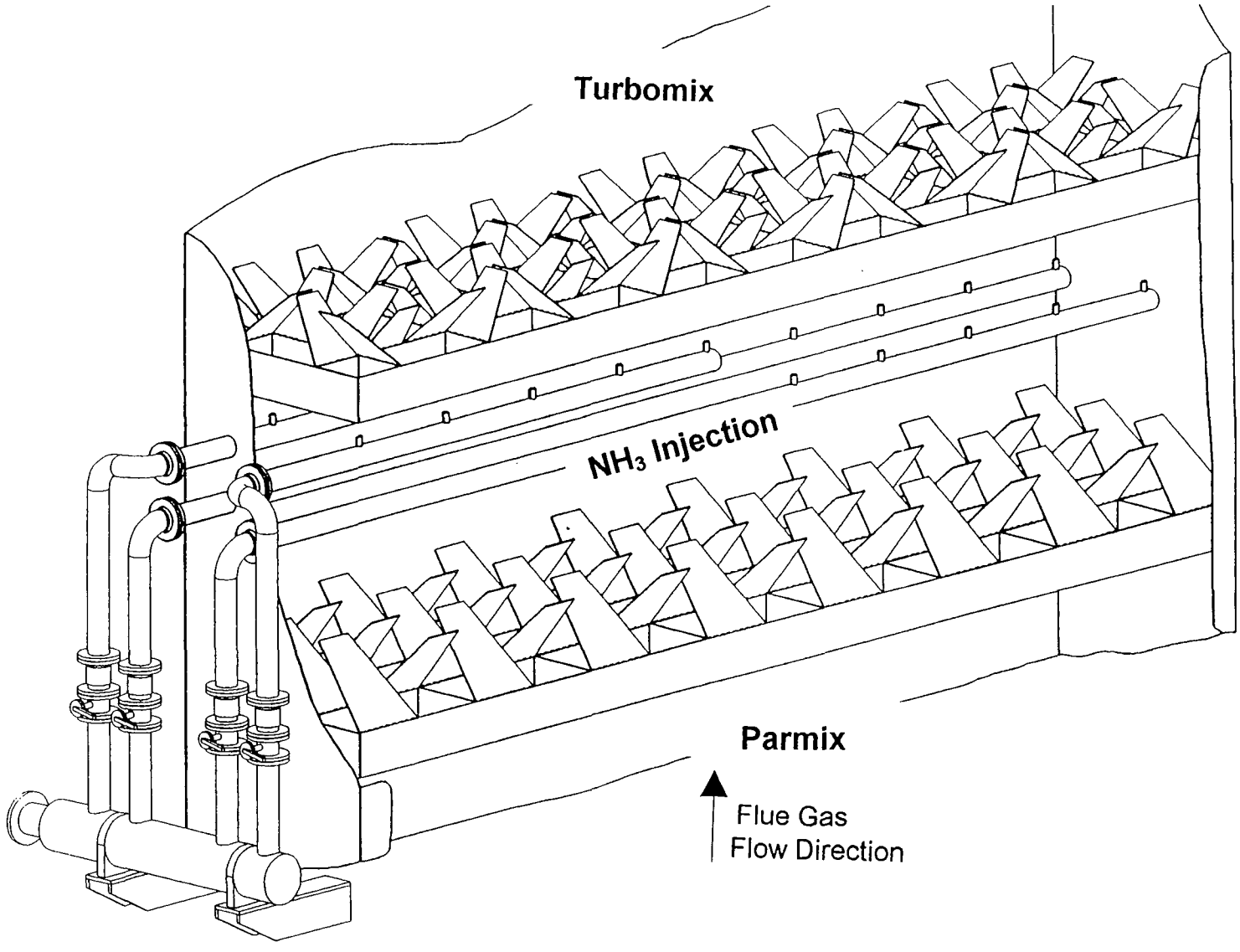

Figure 2: Siemens Static Mixing System Parmix and Turbomix for improvement of homgeneous Flue Gas distribution and Ammonia Injection

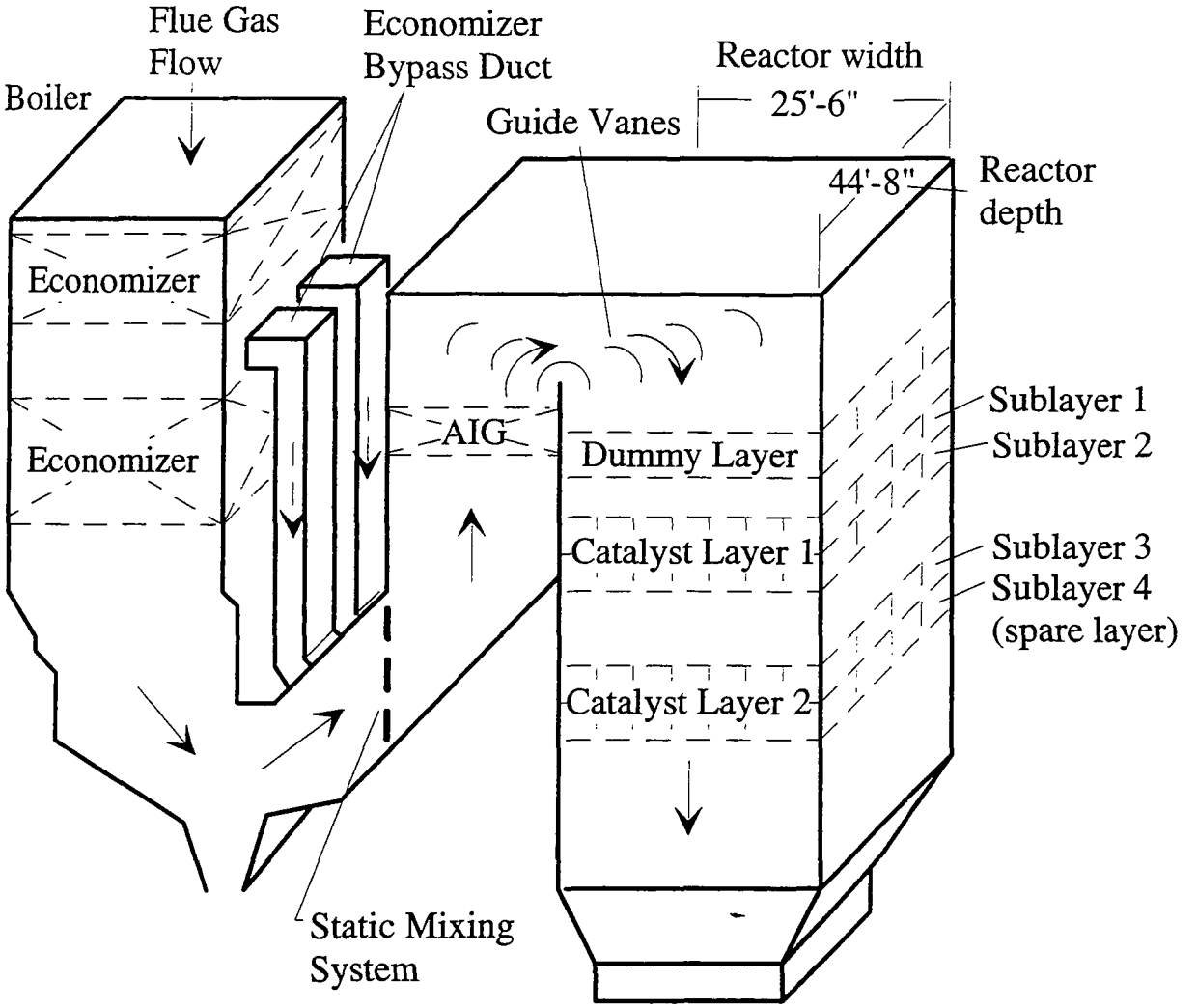

Figure 3: Arrangement of Boiler - Duct - Bypass - Static Mixer - AIG - Guide Vanes and SCR Reactor with Dummy Layer and Catalyst Layers

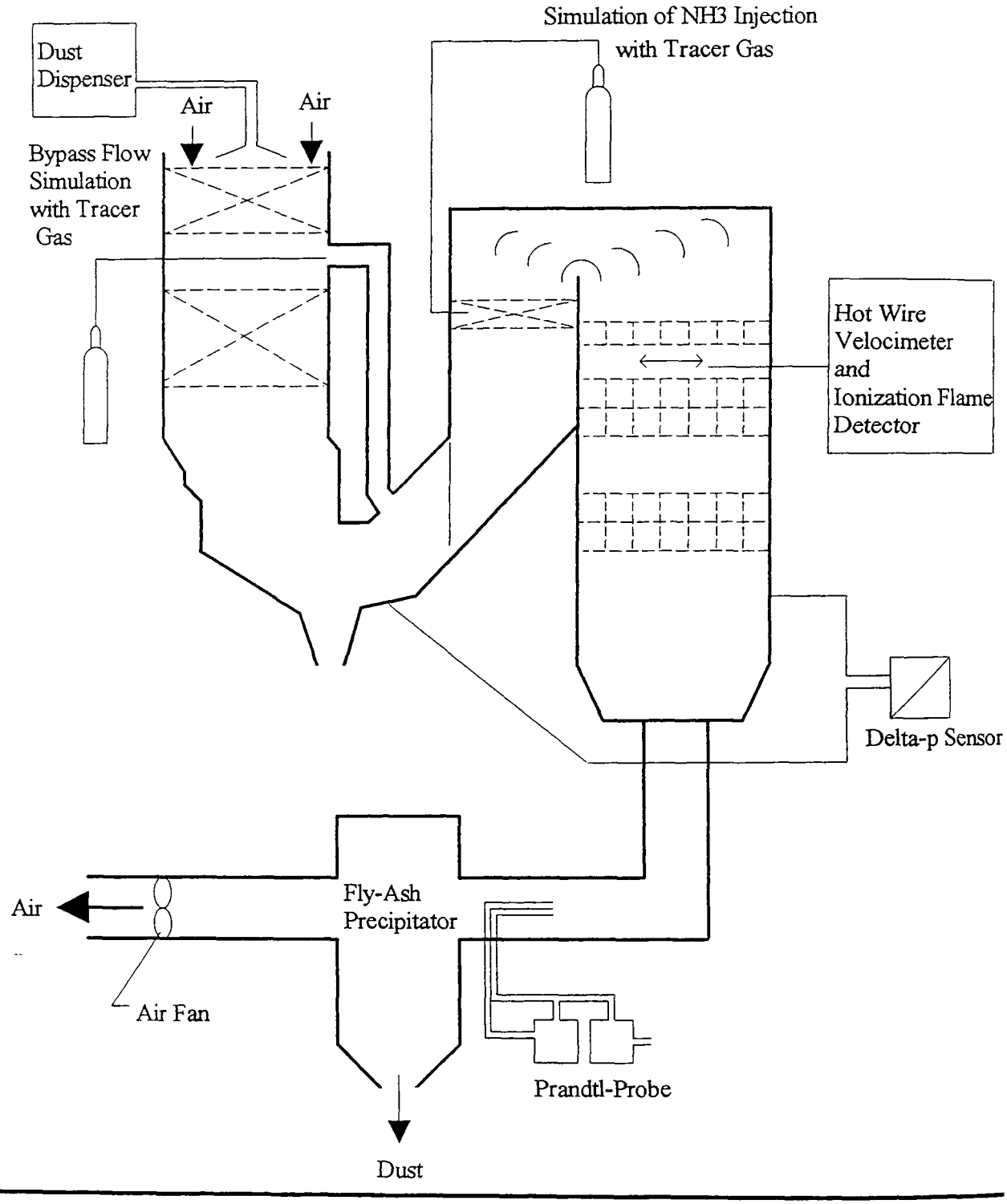

Figure 4: Schematic of Test Model with Instrumentation Equipment

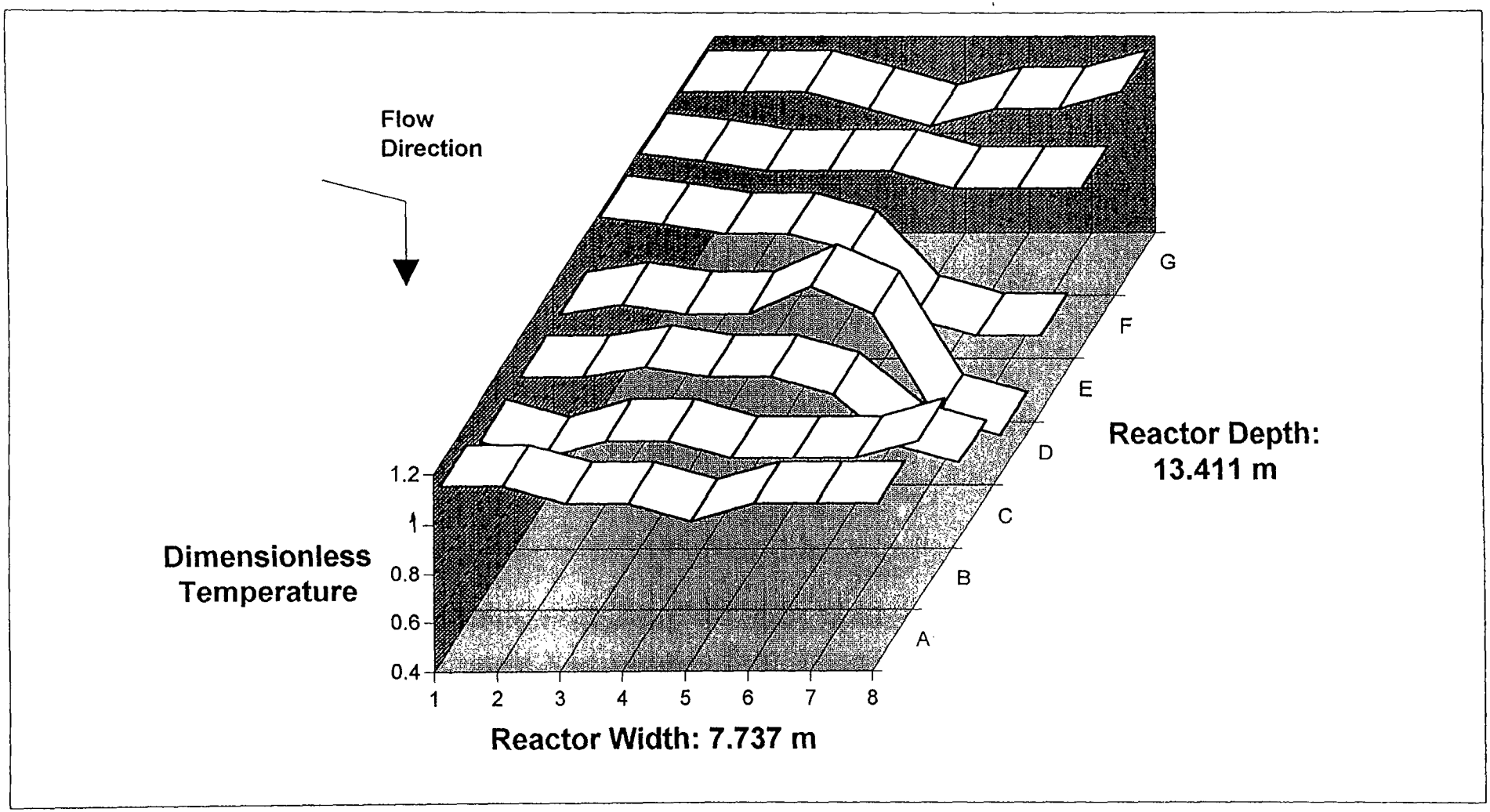

Figure 5: Temperature Distribution upstream of First Catalyst Layer without Internals such as Guide Vanes or Static Mixing System (RMS = 17.7%)

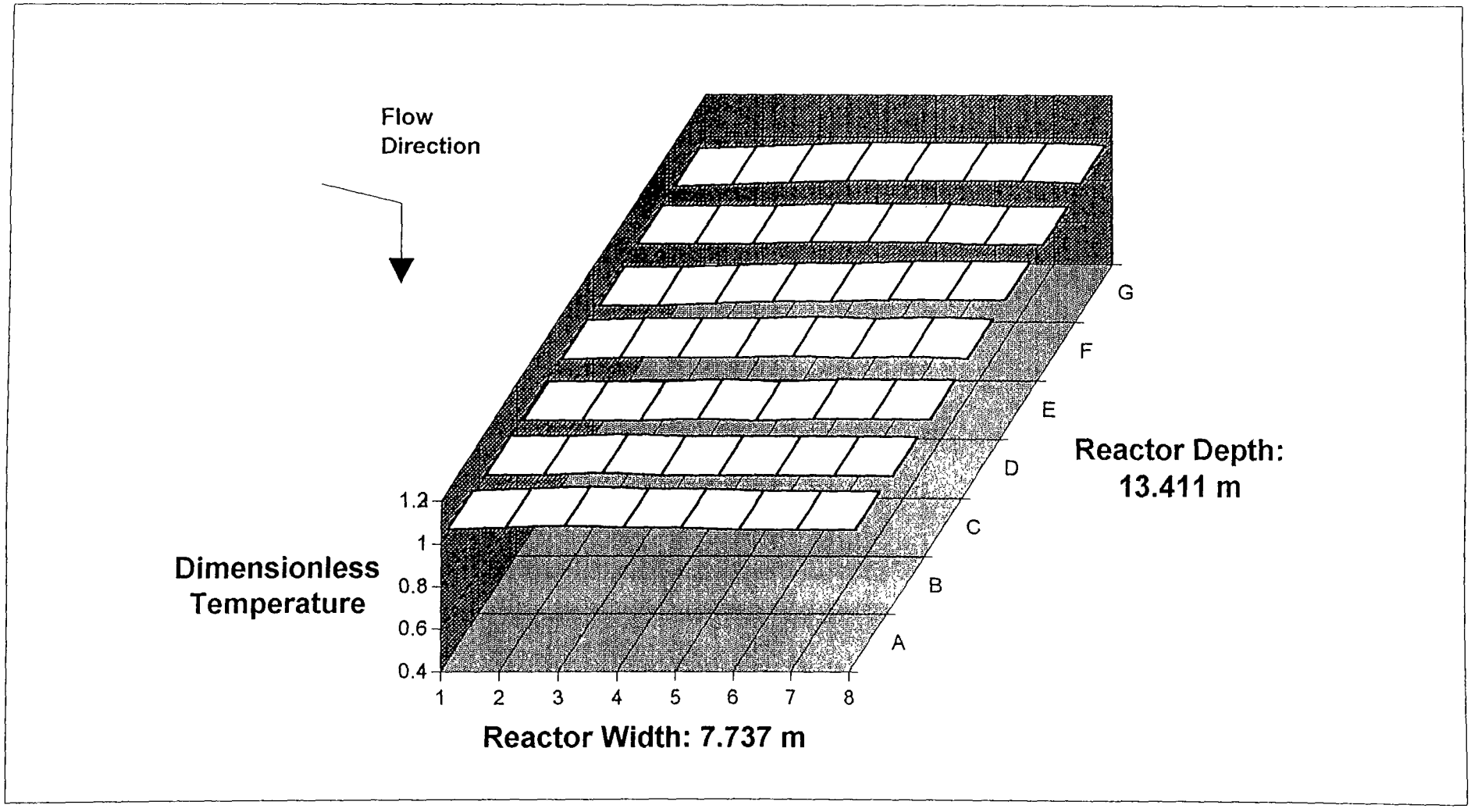

Figure 6: Temperature Distribution upstream of First Catalyst Layer after Installation of Guide Vanes and Static Mixing System (RMS = 1.6%)

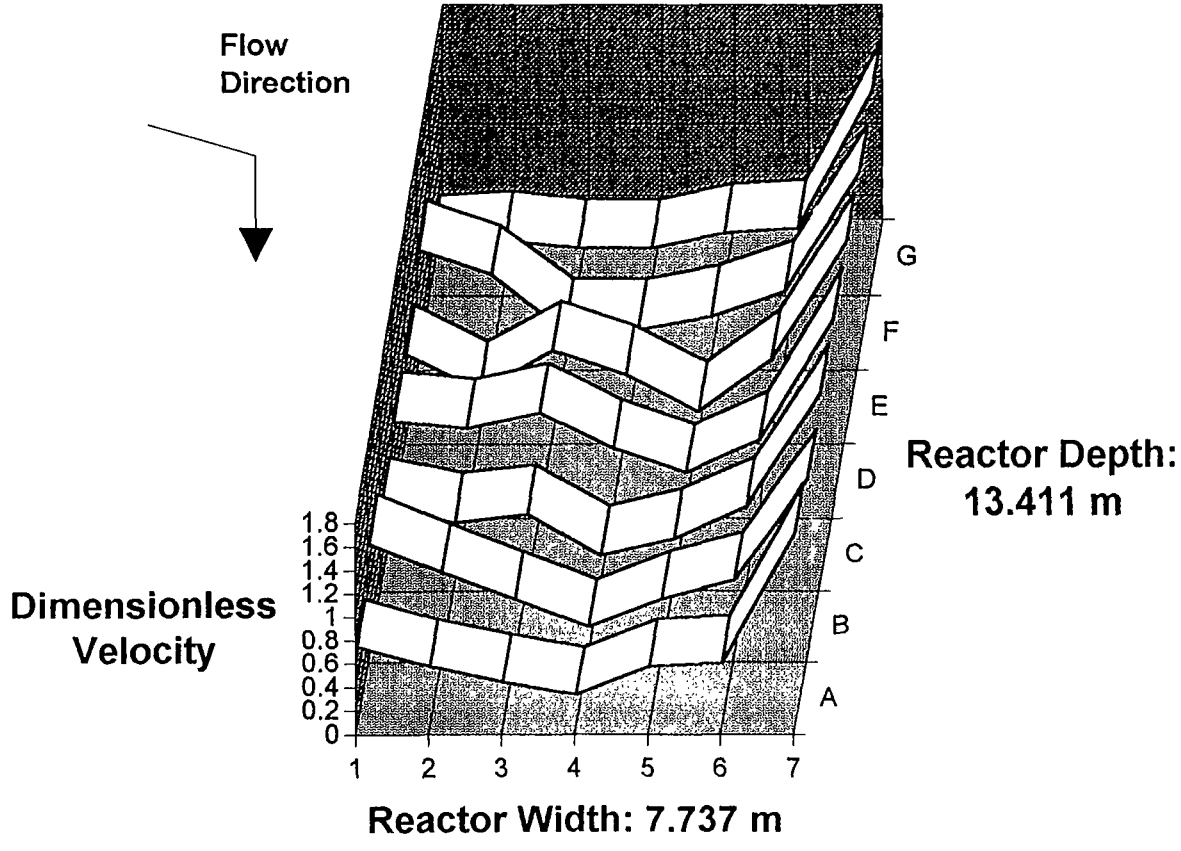

Figure 7: Velocity Distribution upstream of First Catalyst Layer without Internals such as Guide Vanes or Static Mixer (RMS = 65.3%)

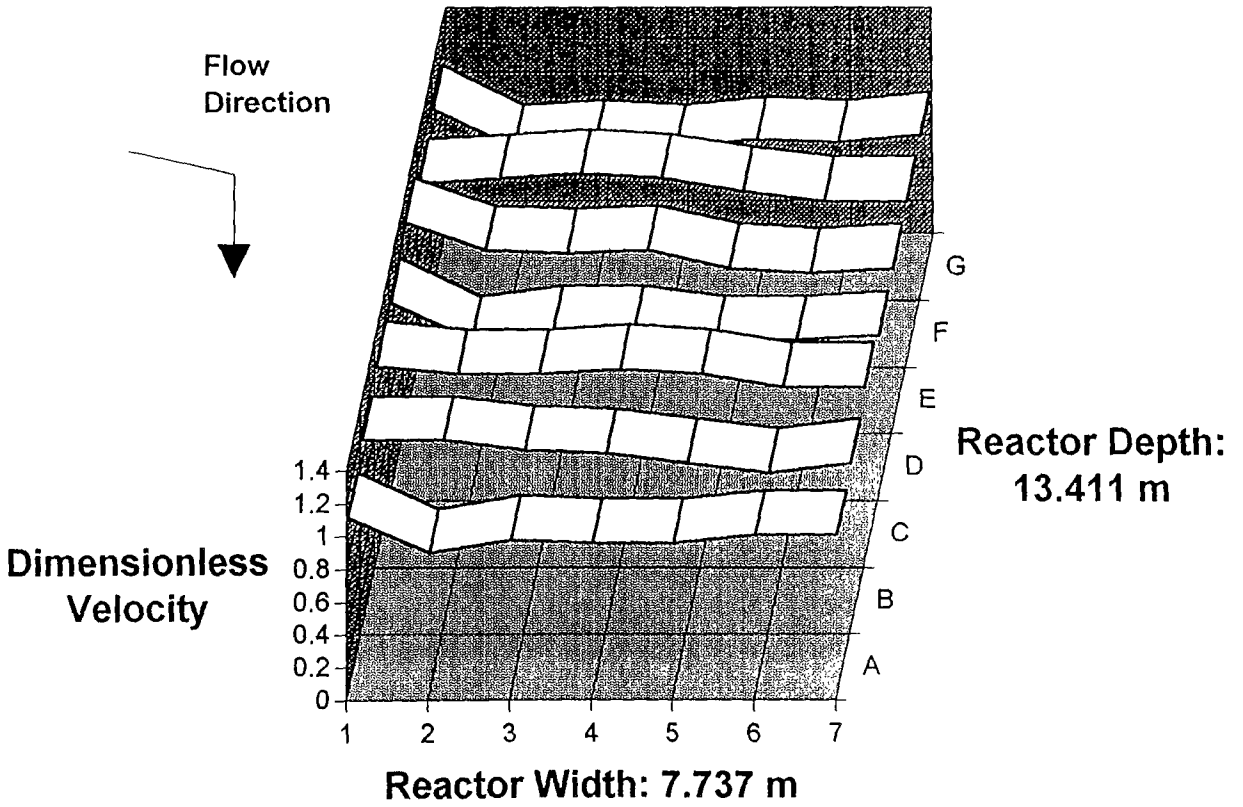

Figure 8: Velocity Distribution upstream of First Catalyst Layer after Installation of Guide Vanes and Mixing System (RMS = 8.8%)

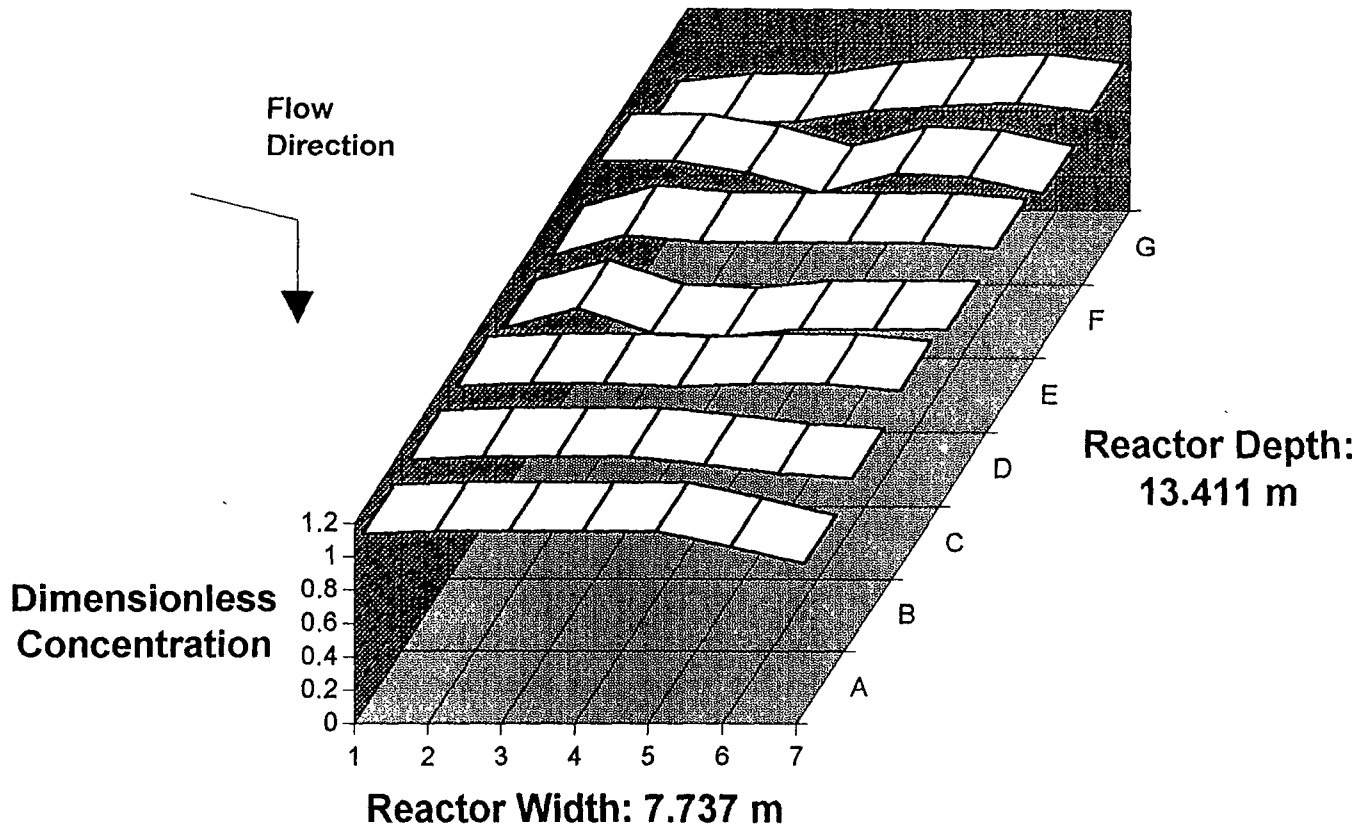

Figure 9: Concentration Distribution upstream of First Catalyst Layer after Installation of Guide Vanes and Static Mixing System (RMS = 7.6%)

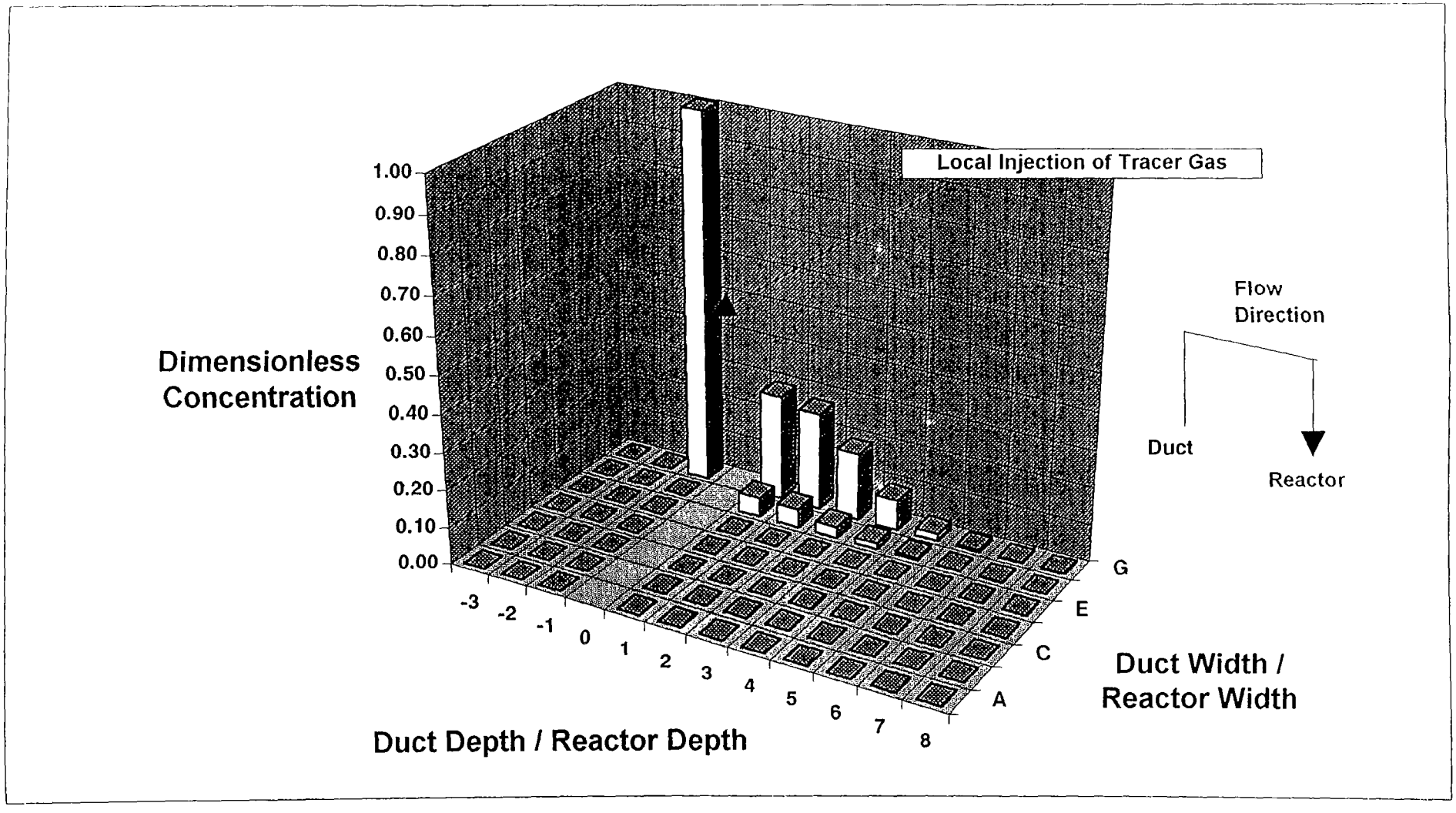

Figure 10: Simulation of the Effect of Local Ammonia Injection

Session 7A
Fundamentals and Modeling

PARTICLE SEGREGATION BY SIZE TOWARDS LOW NOx BURNER DESIGN.

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Abstract

Among the combustion modifications developed in order to reduce the levels of NOx emissions in Pulverised Coal combustion for power generation, the low NO_x burner is one of the most developed and popular. However, it is still considered that the maximum potential for NOx reduction via burner systems has not yet been fully realised. The continuation of the development of low NOx burners continues to be recommended by the power generation industry.

The flow conditions required for low fuel-NO_x formation during the combustion of pulverised coal particles, must allow for maximum volatile yield in oxygen depleted atmospheres. This decrease in O_2 availability (rich stoichiometry) affects the rates of combustion and ignition, leading to an increase of carbon in the fly ash.

Previous studies show that small particles release their volatile more rapidly than larger ones. Particle segregation by size, followed by the placement of particles of differing sizes in the appropriate regions of the inner recirculation zone of a power station pulverised coal burner would lead to a more balanced compromise between low NO_x and unburned carbon content in the fly ash.

The present paper explores the effect on NO performance of appropriately manipulating the distributions of particle sizes and velocities in the burner internal recirculation zone. It is concluded that such manipulation has the potential to achieve a significant reduction in NO emissions.

Introduction

Within the context of the environmental constrains imposed on coal utilisation, the need to reduce NO_x emissions from existing pulverised coal fired furnaces in the power generation industry can be satisfied by number of options. The commercially available options and their comparative costs are summarised in Table 1.

The current German limit on NOx emissions for coal fired power stations is 120 ppm. As a comparison, boilers in the UK with low-NOx burners installed operate at NOx

level generally below 350 to 400 ppm, depending on whether they are corner fired or wall fired. Of the possible choices shown in Table 1, only selective catalytic reduction (SCR) could be considered to reach the required performance. However this technique is expensive and it has not been fully proven on flue gases from high sulphur or high chlorine coals[2]. Nonetheless, the other and cheaper options can not match individually the effectiveness of SCR.

Of these options, it is considered that the maximum potential for NOx reduction via burner systems has not yet been realised. In addition, the capital costs of installing this type of equipment is generally lower than other NOx reduction measures, hence, the continuation of the development of low NOx burner techniques is recommended by the power generation industry[2].

Volatile NOx formation can be minimised by promoting devolatilisation in zones of high temperature and low stoichiometry. Low NOx burners are generally designed to achieve delayed combustion through the way air and fuel are introduced. The level of available oxygen is decreased in zones that are critical for NOx formation and the amount of fuel burned at peak temperature is also decreased. By fuel and air staging, the coal is devolatilised under conditions of rich stoichiometry promoting the conversion of fuel nitrogen to molecular nitrogen rather than oxidation to NO. As a consequence, the reduction in NOx achieved by low NOx burners is typically accompanied with a small reduction in combustion efficiency reflected on higher residual carbon content, due to the less intense nature of the combustion.

A number of fundamental studies on nitrogen release and particle devolatilisation rate, have reached to the conclusion that the particle size plays an important role in the mechanisms of devolatilisation[3,4,5]. Additionally, the amount of NO formed from the nitrogen released as volatile is also a function of O_2 availability, particles residence time and trajectories, and other parameters that can be addressed as aerodynamic effects[6].

The present paper studies the possibility to further reduce the NOx emissions in low NOx burners that are already in commercial use by segregating the pulverised fuel particles by size and by optimising the use of the aerodynamic conditions in the near burner region.

Background

The flow conditions required for low fuel-NO_x formation during the combustion of pulverised coal particles must allow for maximum volatile yield under oxygen depleted conditions. The complete nitrogen chemistry involves well over 100 elementary reactions. The main pathways of gas phase N-chemistry related to combustion may be supposed to be those illustrated by Figure 1, taken from De Soete, 1993[7].

One of the objectives of low NOx techniques is to promote the reduction via hydrocarbons given the appropriate O_2 concentration conditions so that the following reaction path is emphasized.

$$NO + CHi \rightarrow HCN + Products \quad (1)$$

$$HCN \xrightarrow{O} NCO \xrightarrow{H} NHi \xrightarrow{NO} N2 \quad (2)$$

This process emphasises the role played by hydrocarbon radicals (CHi), the existence of which depends on a number of factors such as devolatilisation mechanism, identity of volatiles, oxygen availability and the aerodynamic effects.

Based on previous studies[3,8] wich showed that small particles release their volatile more rapidly than larger ones, and that lower NOx emissions are produced during the combustion of pulverised coal of small size grind, when the particles can partially penetrate the inner recirculation zone (IRZ), it was concluded that particle segregation by size, followed by placement of different size particles in optimal regions of the IRZ would lead to a more balanced compromise between low NO_x and unburned carbon content in the fly ash.

as simulated In calculations performed by Romo-Millares[9], where two burner geometries were compared as shown in Figure 2, it was shown that because of their greater inertia, larger particles penetrate the IRZ further than the smaller particles. Smaller particles are flung rapidly outwards, by the effect of the secondary air swirl, towards the zones of greater oxygen availability. These are hot, lean stoichiometric conditions that will promote the formation of NO. Because staged air burners reduce the availability of O_2 in the inner zone, we want the smaller particles, with their greater reactivity, to concentrate there, near the centre-line where temperature should remain at a level that will enhance devolatilisation. The larger ones, with their lower devolatilisation rates, should be caused to populate the regions where temperatures are higher. Thus, the NO formed from the volatiles released by the larger particles in the outer regions of the IRZ will be reduced to N_2 when it passed into the centre of the recirculation zone due to the presence of CHi generated by the devolatilisation under oxygen depleted conditions of the smaller particles.

In studies performed by Abbas et al.[8] comparing the performance of two burner types, it was found that for the case of a single central orifice burner (SCO), with a flow pattern that gives longer residence times to the larger particles and allows the small ones to penetrate the IRZ, also increasing their residence times in this lower temperature region, the conditions for low NO_x are inherent. However, for a single annular orifice burner (SAO), which represents the geometry of many industrial burners, this was not the case. The small particles do not penetrate the recirculation bubble, rather they are reversed towards the outer regions of the recirculation zone, as shown in Figure 2, while the larger ones spend shorter residence times in this region. With this geometry, if the smaller particles could be directed into the inner region, the stoichiometry will be richer, contributing to a decrease in NO_x formation.

In order to cause the smaller particles to penetrate the recirculation zone further, greater initial velocities should be provided. Because we want to maintain the rich conditions, the amount of primary air should not be increased, the increase in velocity should be achieved either by reduction of the cross-sectional area, or by the increase in flow rate by the addition of an inert gas (e.g. EGR). Further penetration of the particles could also be achieved by the slight penetration of the burner pipe into the

quarl region. This has been achieved by Visser et al[10] where the penetration of the burner gun into the quarl by 1/3 of its depth results in a flow field where the primary flow penetrates through the recirculation bubble. This is not a desirable situation as far as low NOx production is concerned, as the residence time of the particles in this zone is reduced. The longer the residence time in the IRZ, the more efficient devolatilisation in the O_2 depleted zone and the greater the burnout.

The introduction of a flow through the pilot gun in the SAO type burner, should have a similar effect, i.e. a forward flow that will weaken the reverse flow (coupled with a desired reduction of temperature) generated by the recirculation of the secondary air. This effect should ideally be confined to the near burner region inside, or just outside the quarl, avoiding full penetration of the recirculation bubble (to minimize the carbon content in the fly ash). This option, with full penetration of the recirculation bubble, has been studied by Smart and Morgan[11] while testing the use of different re-burning fuels fed through the inner pilot tube of the burner.

Burner Design

Based on the research work performed by IFRF[10,11], proven low NO_x burner industrial practices[12] and burner manufacturer recommendations[13], and the experience gained in our laboratories on NOx emission studies[14] together with the preceding considerations, it was concluded that, as a first step for the introduction of an innovation into the existing low NOx burner designs, the possibility for a selective location of particles into the IRZ should be pursued.

As a starting point the basic design of typical present day low NOx commercial burners is adopted. These burners make use of fuel staging measures that separate the coal flow inside the primary tube in 4 jet streams. The ability to seaparate the flow in such a way, with a low pressure drop and minor erosion, is determined by the pulverised coal being introduced tangentially into the primary pipe, giving a toroidal trajectory to the coal particles within the tube wich is subsequentially eliminated by 4 pulverised coal collectors, named FFRs, located inmediately before the burner mouth. However, the swirling movement, which in the absence of the FFRs would be imparted to the particles, should inherently classify them by size due to their centrifugal force, with the larger particles populating the outer regions of the primary tube.

As the first step in the present design procedure, a geometrical burner modification has been introduced so that the smaller particles will be encouraged to penetrate the IRZ, without passing through the recirculating bubble; i.e. to follow a similar trajectories to those of the larger particles shown in Figure 2.

The following design of a low NO_x burner has been undergoing isothermal testing and computational simulation of its performance under firing conditions. A flow separator located in the primary line, as shown schematically in Figure 3, leads to a selection of the smaller particles and an increase of their initial velocity. With the smaller particles located in the annulus of smaller radius at the burner mouth and with an increased

initial momentum, the aim of further penetration of these particles into the O_2 depleted region of the IRZ is achieved.

Scaling Criteria

Due to inherent difficulties in conducting full-scale tests and the associated costs, experiments are often carried out on rigs of significant by smaller scale. The immediate question is of course: how representative of the full industrial scale burner is the experimental burner performance, measured at the reduced scale?. In an ideal scaling exercise, the strict similarity of all the fluid dynamic and thermo-chemical related processes should be maintained. In practice, compromises are necessary since of the various physical and chemical processes scale in different ways.

In order to adequately simulate the two phase interaction between the coal particles and the aerodynamic flow field, the mean particle size has to be reduced by the square root of the linear scale factor. However, experimentally, this requirement is normally not practicable, due to the high cost and inconvenience involved in grinding the coal to the required scaled size and, in particular, since the mean size is altered and consistency of particle properties across the size distribution cannot be maintained.

In this study, a 1/6 scale isothermal model has been built, making use of the constant velocity scaling criterion. This maintains the full scale burners velovities at the reduced scale and ensures that the minimum velocity required for the transport of the pulverised coal will be reached. To simulate the cyclone separation effect given by the geometry of the primary tube at the inlet, the main scaling criteria used were the air inlet velocity into the cyclone and the degree of swirl. In this way, it may be aargued that the particle separation is be properly simulated.

Results and Discussion

Making use of the primary tube design explained above, preliminary measurements of particle size and velocities at the exit of the tube have been performed in a reduced scale isothermal rig. The purpose of these measurements was to assess the effectiveness of the particle separation by size of the cyclone geometry of the primary tube coupled with the flow separator -1- of Figure 3.

Polydisperse coal of a size typical of industrial pulverised fuel was introduced into the primary tube. The experimental conditions are given in Table 2.

Figure 4 shows the particle size distributions of the feed coal and of the samples taken at the inner and outer tube outlets (i.e. numbers -4- and -5- in Figure 3). As expected from the design considerations, the coal feed has been segregated by the flow separator into coal of 75% under 20 μm size in the inner outlet and 75% under 90 μm in the outer outlet.

The air axial velocities at the exit of the two outlet annuli and the contour velocity profiles downstream were measured making use of a pitot tube 3mm in diameter. The data is shown in the shape of a velocity contour in Figure 5.

With the data on initial conditions thus obtained, computer simulations for their effect on NO emissions were performed. A two dimensional numerical code, 2-D CINAR, developed by this combustion group for simulation of the flow aerodynamics, combustion and heat transfer of two-phase flows was employed. The code is well documented having been used and validated against a variety of experimental data[15,16]. For the NO calculations, a simplified chemistry model is employed, which at this moment does not take into consideration theNO reduction mechanism via CHi.

A comparison was performed between two cases: the "test" case, wich uses the burner geometry of Figure 3 and the measured boundary conditions of the reduced scale isothermal experiments, and a hypothetical case, the "base" case, where no particle segregation occurs and the primary stream has a uniform initial velocity.

In the "test" case, small particles, of a typical 20 μm size, are injected at the inner outlet position with an initial velocity of 40 m/s. The larger particles, of a typical size of 70 μm, are injected at the outer outlet position at 27.0 m/s initial velocity. For the "base" case, both particle size groups are injected at an initial velocity of 32.0 m/s. Figure 6 depicts the downstream predictions of the particles trajectories as a function of initial velocities and size. In the "base" case, the small particles populate the regions closer to the quarl walls where they meet the secondary air flow at an early stage, while in the "test" case, these particles populate the central region of the flow, up to the exit of the quarl and beyond delaying their mixing with the secondary flow.

In Figure 7 the predicted NO profiles are shown. It can be observed that lower levels of NO are found for the "test" case where the smaller particles have been injected at greater velocities closer to the axis of the burner. These preliminary results are indicative of the effectiveness of the burner modifications proposed, towards the reduction of NO emissions. It must be emphasised here, that these predictions do not represent a commercial design, they are studies done with the sole purpose of observing trends, that the modifications sugested in the present paper, will provide. In addition, as the models employed in the present study do not simulate the reduction reactions via CHi, the levels of NO found in a practical applications should be much lower.

Further work is being undertaken on both: experiments, where laser techniques will be employed to study particles distribution, and computations, modifying the conditions to simulate better the practical case.

Conclusions

1. Existing burner designs do not match the particle size and velocity distributions in the inner recirculation zone in a manner conducing to optimizing NO reduction.
2. Improved distributions can be achieved through a simple modification of the upstream burner aerodynamics.

3. The experimental and computational work of this paper for one such burner modification demonstrate that NO emission reduction is indeed procured.
4. Further research should be undertaken to determine an optimised burner design based on the innovative concept of this paper.

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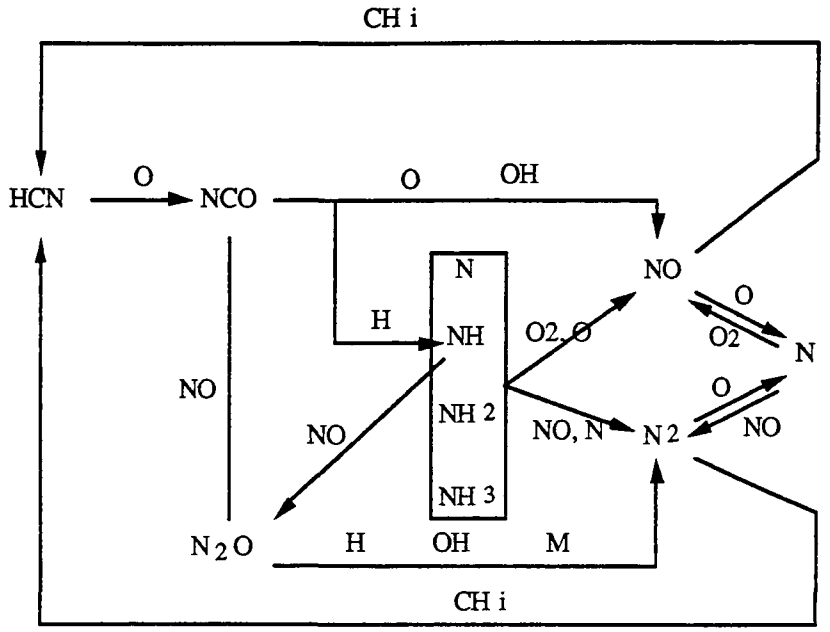

FIGURE 1 Main ways of gas phase N-Chemistry, reproduced from De Soete, 1993[7]

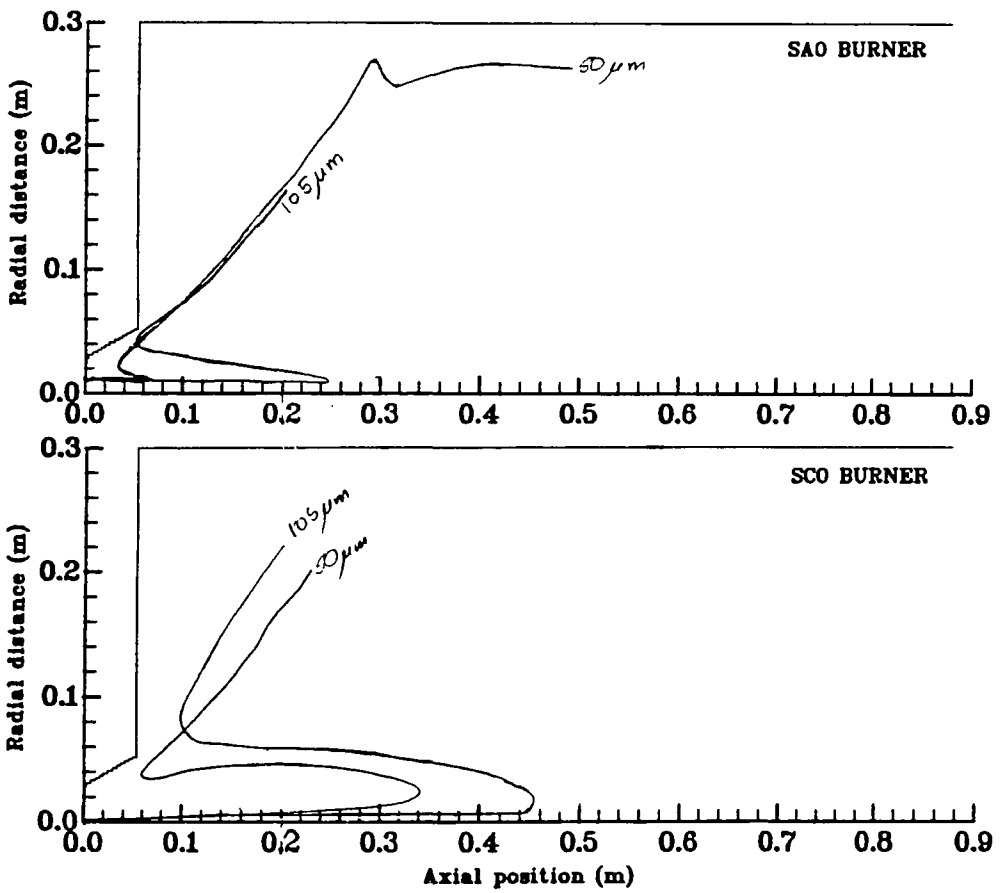

FIGURE 2 Calculated[9] particle trajectories for two burner geometries: (a) SAO and (b) SCO burners.

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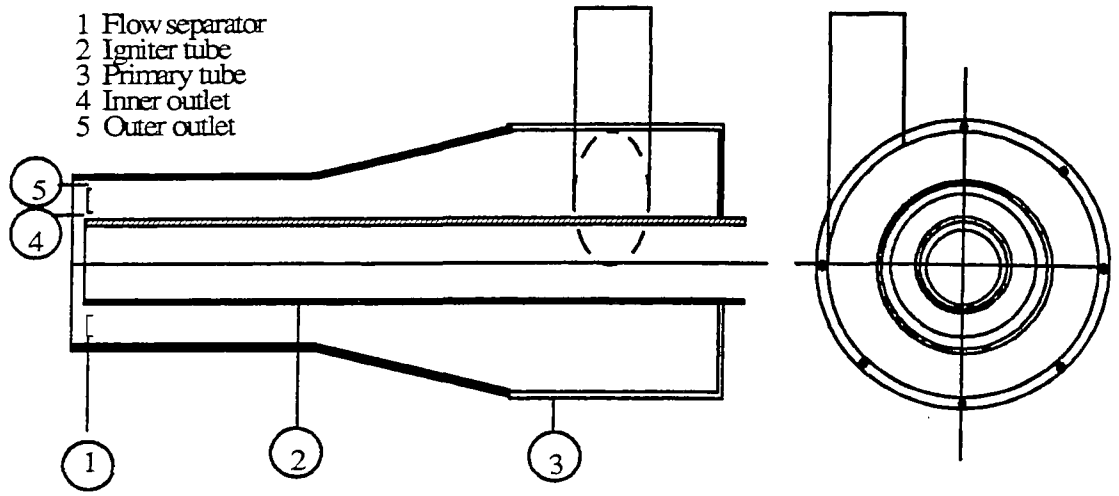


FIGURE 3 Schematic of primary tube modification to ICL burner.
***Note**: The details of the flow separator have been suppressed pending patent.*

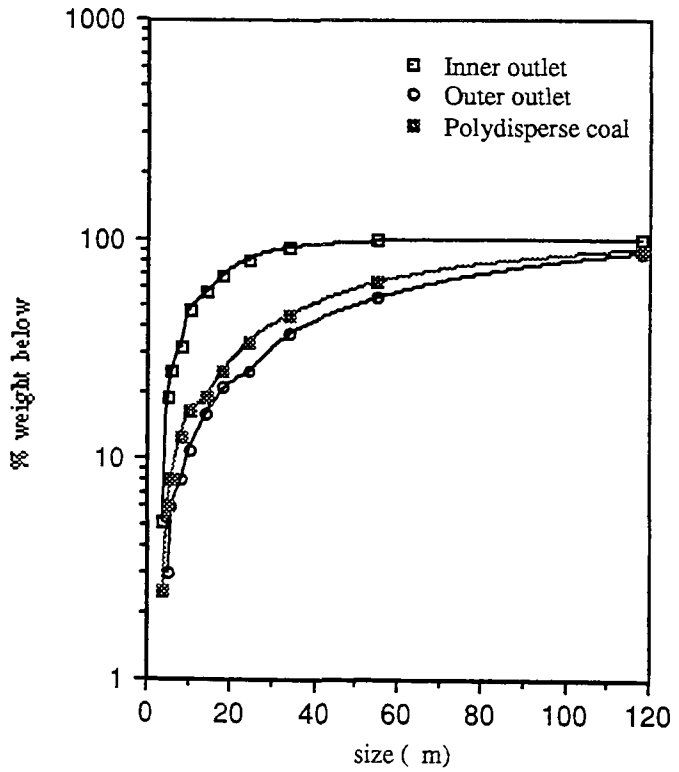


FIGURE 4 Particle size distributions for coal feed and samples from the primary tube outlet.

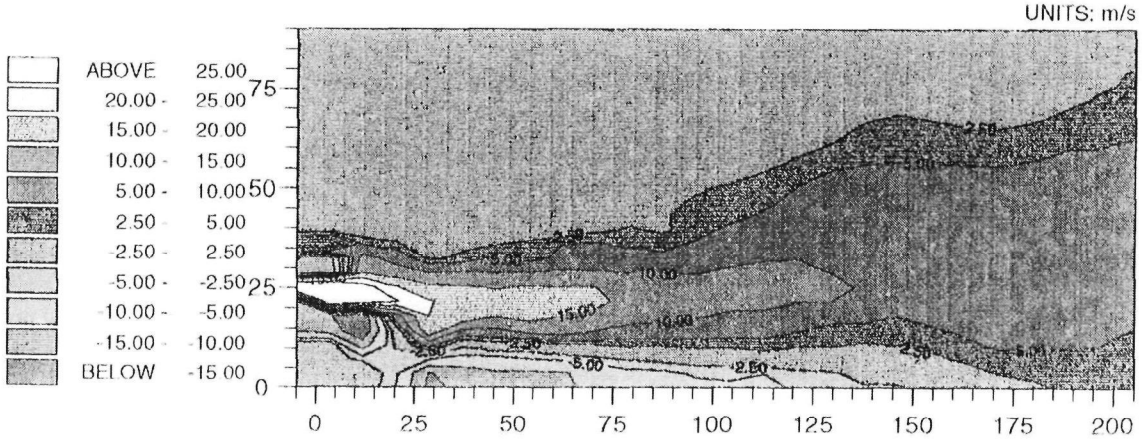
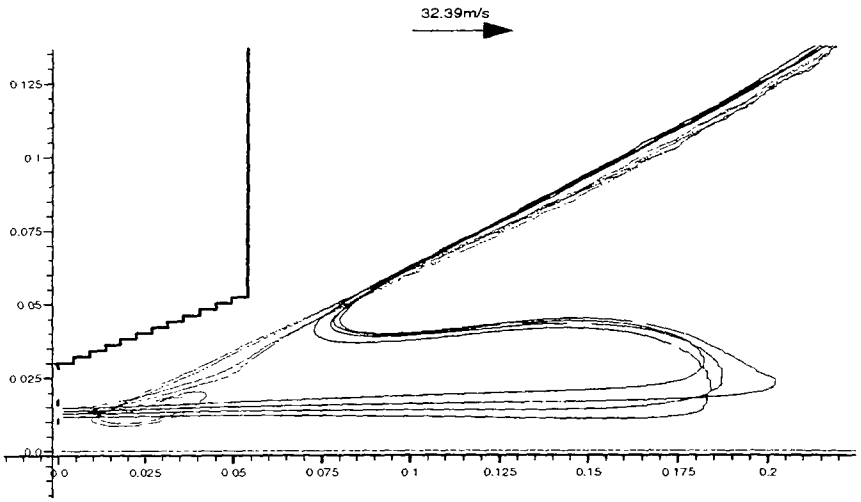


Figure 5 AXIAL VELOCITY CONTOURS OF LOW NOx BURNER

a. Base case



b. Test case

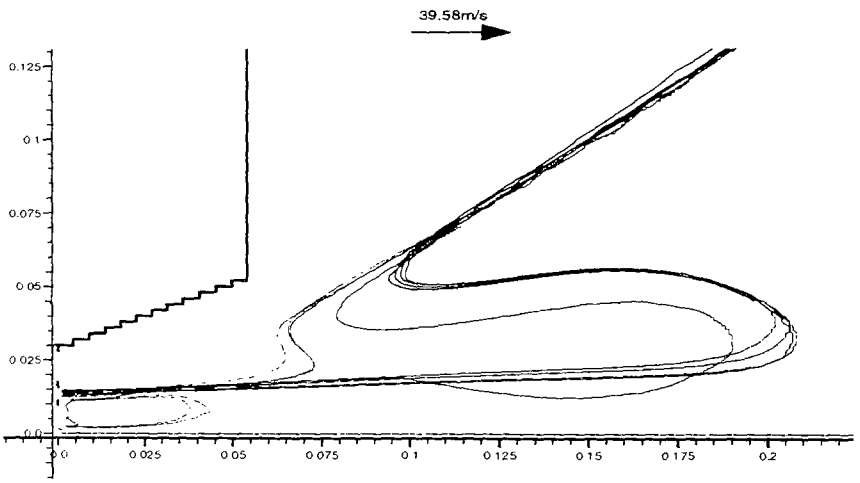
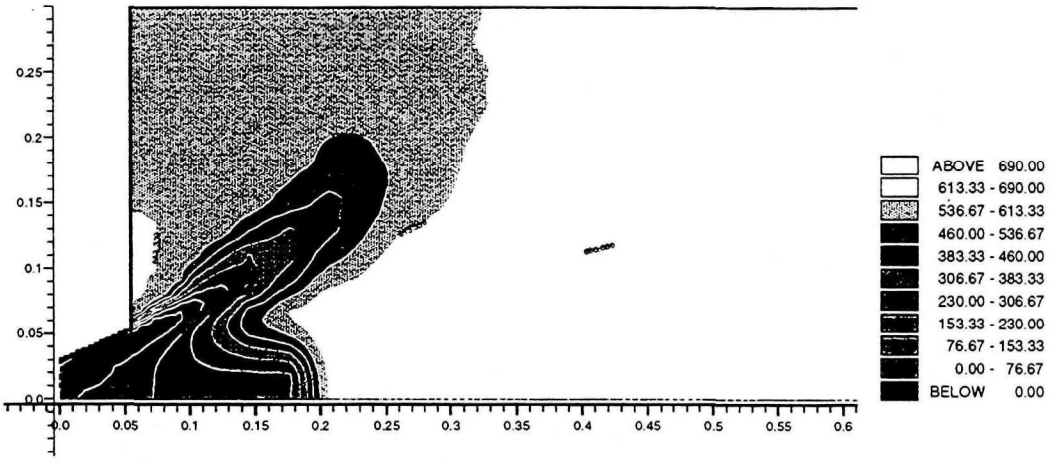


FIGURE 6 Particle trajectories for (a)"base" case and (b)"test" case

a. Base case



b. Test case

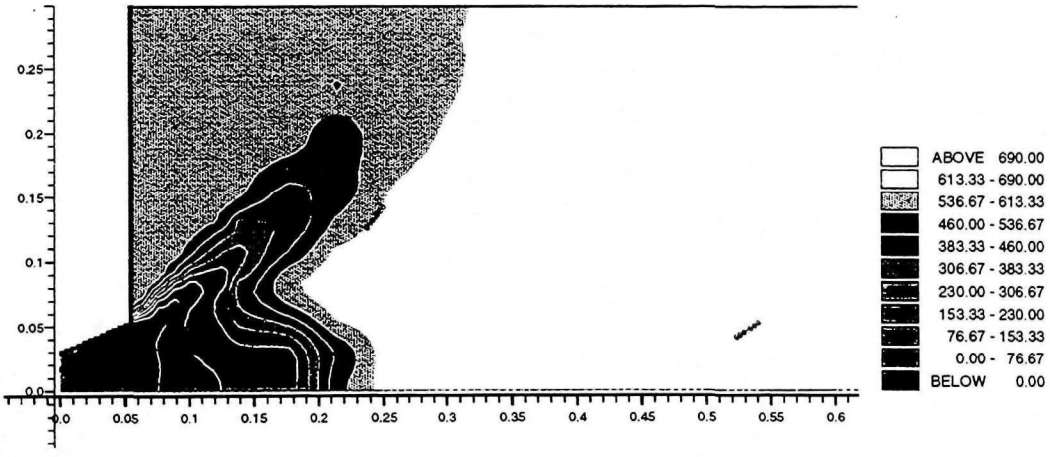


FIGURE 7 NO contours for (a)"base" case and (b)"test" case.

TABLE 1

Cost associated with various NO_x reduction techniques[1]

Technology	Additional Cost £/kWt		NO_x Reduction
	Installation	Operation/year	%
Combustion modification			
1) Low NO_x burners	2	~ 0	30-50
2) Furnace air staging	2.4	~ 0	up to 50
3) Combination of low NOx burners and air staging	3.8	~ 0	up to 70
4) Re-burn	6-14	1.0	up to 60
Flue gas treatment			
5) Selective non-catalytic reduction	2.2-10.6	1.5-2	40-50
6) Selective catalytic reduction	14-18	3 and over	80-90

Table 2

Experimental and computational run conditions

Experimental		
Primary air flow rate, kg/s		0.023
Coal feed rate, kg/s		0.001
Axial primary air velocity, m/s,	INNER OUTLET	40.0
	OUTER OUTLET	25.0
Pulverised coal size, % under 75μm		70.0
% under 25 μm		36.0
Average size, μm		40.0
Cyclone Inlet air velocity, m/s		35
Degree of Swirl, °		25
Computational		
BASE CASE		
Axial primary air and particle velocity, m/s		32.0
Polydisperse coal, 4 entry points, 4 size at each entry, μm	(1)	15
	(2)	33
	(3)	50
	(4)	105
Secondary air swirl number		1.0
TEST CASE		
Axial air and particle velocity, m/s,	INNER OUTLET	39.9
	OUTER OUTLET	27.0
Particle size μm, INNER OUTLET		20.0
OUTER OUTLET		70.0
Secondary air swirl number		1.0

NUMERICAL MODELING INVESTIGATION OF MANAGED NATURAL GAS CO-FIRING IN LOW-NO_X PULVERIZED COAL BURNERS

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Abstract

Combustion and emissions performance of natural gas co-firing in an internally-staged and low-NO_X pulverized coal burner was evaluated by numerical modeling. Numerous permutations in gas velocity, injection angle, injection location, extent of co-firing, and natural gas/air premixing were simulated for comparison. With 15% gas co-firing, NO_X emissions were 16% lower and the unburned carbon levels were 19% less than the corresponding values at 100% pulverized coal firing. Since the NO_X emission levels for the baseline low-NO_X pulverized coal burner are already about 50% lower than the uncontrolled values of conventional burners, total NO_X reduction with 15% gas use would be close to 60%.

* Author of correspondence

Introduction

Many operating coal-fired utility boilers in the United Stated are in the "ozone non-attainment" and "acid rain control" areas identified by the Titles I and IV of the 1990 Clean Air Act amendments. While low-NO_X burner retrofits are considered as "reasonable available control technology" options for reducing the NO_X emissions, they may adversely affect the flame length and carbon conversion in some coal-burning units. Degradations in the combustion and emissions performance are due to the fact that low-NO_X burners operate differently from rapid mixing conventional burners. In a low-NO_X burner, NO_X is reduced by the controlled separation and delayed mixing of the fuel and oxidizer.

Fly ash can be used as a mineral admixture in Portland cement concrete, as long as it meets the chemical requirement of less than 6% loss on ignition [1]. High levels of carbon in fly ash make it unsuitable for recycling and could even pose disposal problems in landfills. Natural gas co-firing of low-NO_X burners offers greater flexibility in overcoming such operational problems. A recent report on this subject [2] reviews the potential benefits of co-firing, including shorter flame length, higher combustion efficiency, higher turndown ratio, lower excess air requirement, and lower SO_2, CO_2, NO_X, and particulate emissions. Since natural gas availability at most coal-fired plants is limited to ignitor use only (< 15% of the combined thermal load), creative and economical ways must be devised for its effective utilization in low-NO_X pulverized coal burners to "leverage" the desired performance.

So far, co-firing has not been incorporated commercially in wall-fired pulverized coal burners, and its application has been limited to natural gas injection through gas ignitors or selected burners. Leveraged natural gas co-firing in low-NO_X pulverized coal burners can be implemented in a number of ways. Independent pilot-scale tests at the International Flame Research Foundation [3] (IFRF) and Babcock and Wilcox (B&W) indicate lower NO_X emissions with centerline introduction of natural gas in the coal nozzle. Gas co-firing at IFRF was accomplished with an externally air-staged burner, while an internally-staged burner was used at B&W. Part of the NO_X reduction by co-firing is due to lower nitrogen content of the combined fuel. Additionally, injection of natural gas influences the local flame environment

and provides a source of hydrocarbon radicals (CH_i; i = 1, 2, and 3) that convert NO to molecular nitrogen.

This paper discusses a numerical modeling investigation of managed natural gas co-firing in the internally-staged DRB-XCL® low-NO_x pulverized coal (PC) burner [4]. Details of the modeling simulations and predictions of temperature, unburned carbon, and CO, O_2, and NO_x emissions for various cases are discussed and compared in the following sections.

Numerical Model Description

Numerical simulations were done with the B&W multi-dimensional COMO (COmbustion MOdel) code. COMO predicts the detailed distributions of velocity, turbulent kinetic energy and dissipation rate, particle dispersion, species concentrations, heat fluxes, and temperature in Cartesian or cylindrical geometries [5-8]. It can simulate the effects of operating conditions and burner and furnace configurations on the combustion and emissions performance. For this purpose, the computational domain is divided into control volumes and the governing differential equations are formulated and solved at each location.

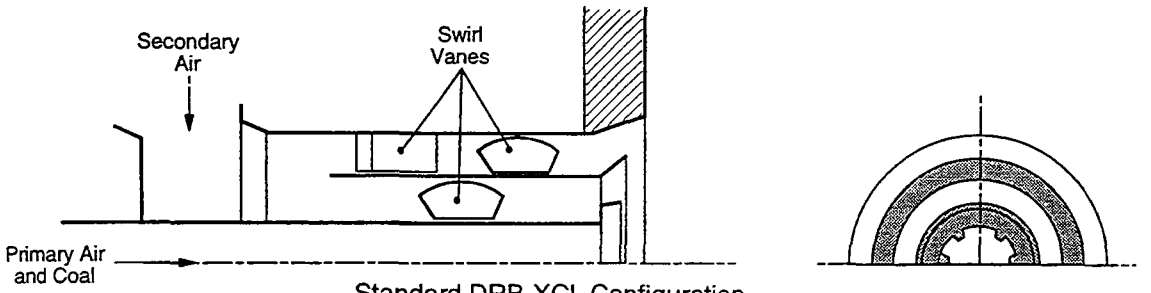
Table 1 summarizes the capabilities of COMO. The model utilizes coal-specific chemical kinetics for devolatilization, char oxidation, and fuel-N transformation. NO_x concentrations are determined with a NO post-processor model that includes the Zeldovich thermal-NO_x reaction mechanism, and global steps for fuel-NO_x formation/destruction [9] and NO/hydrocarbon interaction. The post-processor model is run after the combustion solutions are completed. In recent years, COMO has been utilized as a design tool at B&W in low-NO_x burner development programs.

Table 1 Summary of COMO Burner Model Capabilities	
Flow	■ 2- or 3- dimensional Cartesian or cylindrical ■ Laminar or turbulent flow
Turbulence	■ k - ε, multiple time scale
Chemistry	■ 2 - step : Eddy dissipation model with CO oxidation kinetics ■ Sandia National Lab devolatilization and char oxidation rates
Radiation	■ Discrete ordinates model : 8 or 24 fluxes ■ Gas properties with wide band models
Pollution	■ Zeldovich thermal-NO_X, and global fuel-NO_X and NO/hydrocarbon interaction kinetics ■ Provisions to use reduced fuel-N mechanism
Special Features	■ Coal, oil, and natural gas combustion ■ Particle or droplet trajectories ■ Multigrid solution method to enhance convergence ■ Local grid embedding

Figure 1 shows a schematic of the low-NO_X pulverized coal burner and the main co-firing configurations. Co-firing concepts which were integrated into the design for subsequent modeling include:

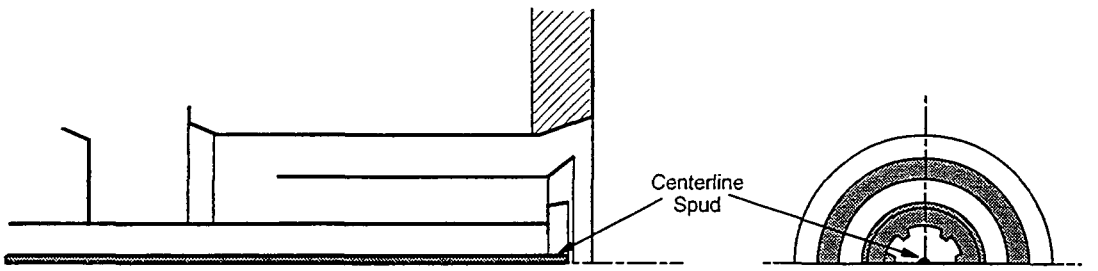
- centerline injection by a single spud centered in the coal pipe and aligned with the coal nozzle exit
- peripheral injection with encircling spuds or gas annuli at various radial locations

In this study, each gas spud was modeled as a circular tube with a single exit. The gas stream directions for individual spuds or annular openings were specified using velocity components (i.e., axial, radial, and tangential). Two-dimensional (2-D) axisymmetric and three-dimensional (3-D) cylindrical models were used to simulate the geometrical representation and operating conditions of the various configurations. These models simulated the entire furnace enclosure including the internal burner components and flow passages, and natural gas injection ports. The combustion chamber was modeled as an idealized cylindrical tunnel furnace, with thermal boundary conditions and a heat loss rate (400 kW_{th}) similar to existing pilot-scale research facilities. The cylindrical furnace geometry was chosen for better numerical compatibility with the cylindrical burner. Dispersion and trajectories of coal particles were modeled using the Lagrangian approach.

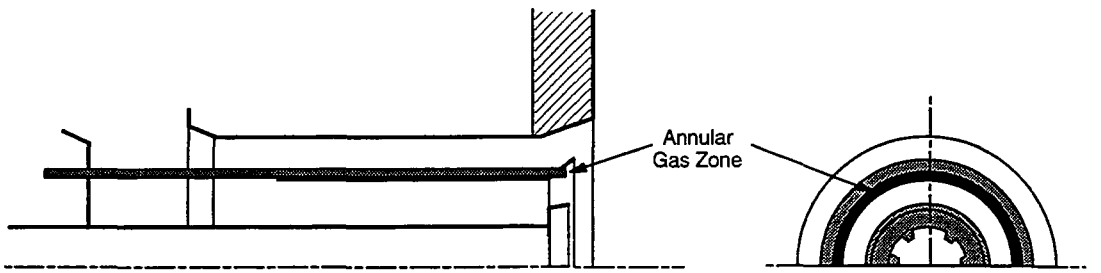


Standard DRB-XCL Configuration

Note: For clarity, vanes shown only for baseline configuration



Centerline Injection



Annular Injection

Figure 1

Typical Burner Configurations for Numerical Modeling Studies of Natural Gas Co-firing

Figure 2 shows the mid-sectional view of the computational grid for the burner/furnace system. For 3-D simulations, a 30° wedge segment of the furnace and the burner was modeled. This permitted a complete representation of the burner when periodic boundary conditions were applied on the circumferential boundary surfaces at 0 and 30 degrees. The 2-D axisymmetric grid modeled only the burner exit area and the furnace. Velocity boundary conditions for the 2-D models were specified using the computed axial, radial, and tangential velocity profiles of the baseline 3-D cylindrical case. The closeup view in Figure 2 highlights the implementation of grid embedding in places where high temperature, high velocity, and high concentration gradients were expected.

In order to have a common basis for comparison, thermal load and burner stoichiometry were held constant at 5 MBtu/hr (1.5 MW_{th}) and 16% excess air, respectively. Fuel-specific kinetics information for high volatile "A" (hvA) Pittsburgh #8 and medium volatile (mv) Lower Freeport bituminous coals was also incorporated in the model. A summary of the operating conditions is given in Table 2.

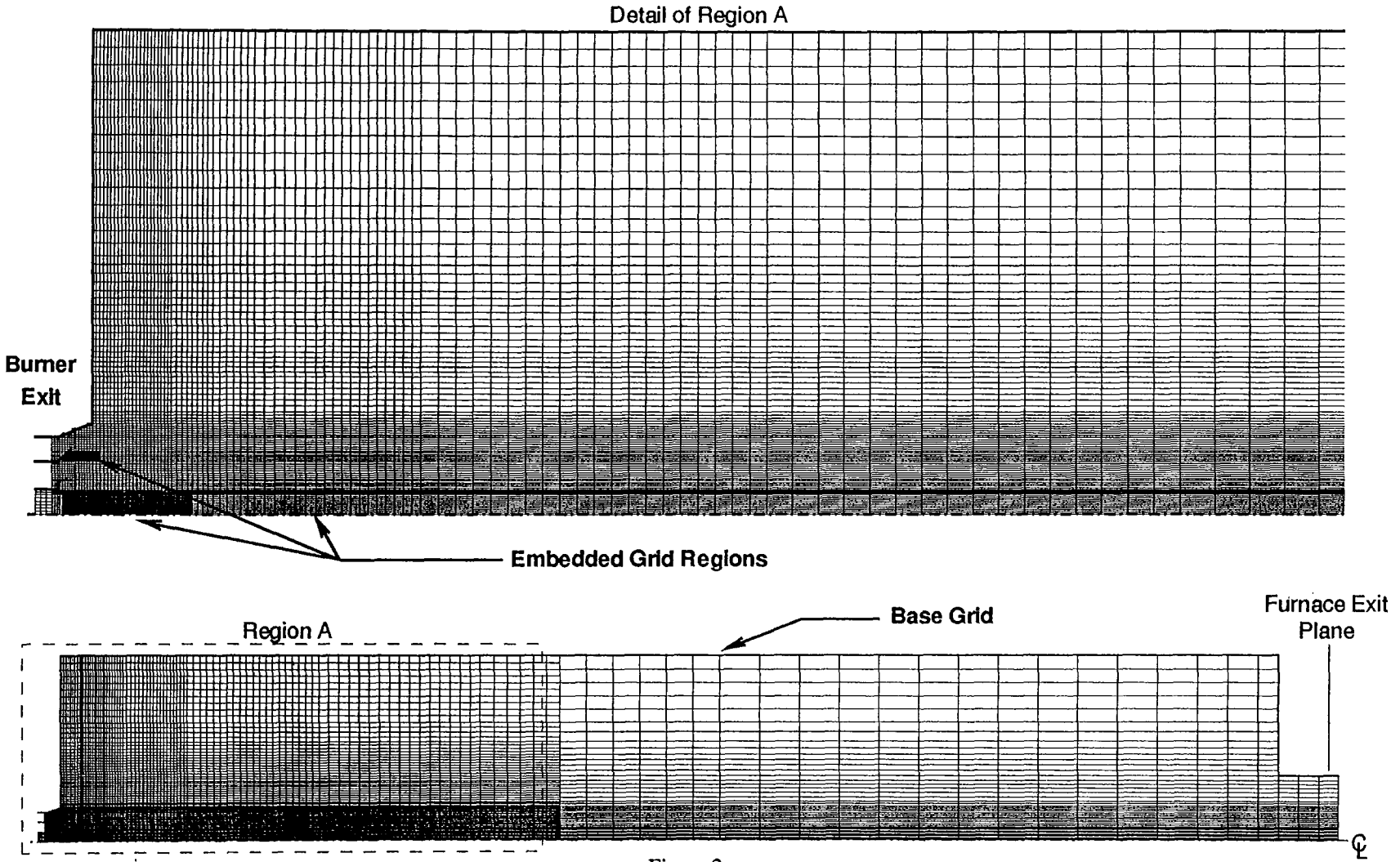


Figure 2

Typical Numerical Grid Arrangement

Table 2
Input Operating Conditions for Numerical Modeling

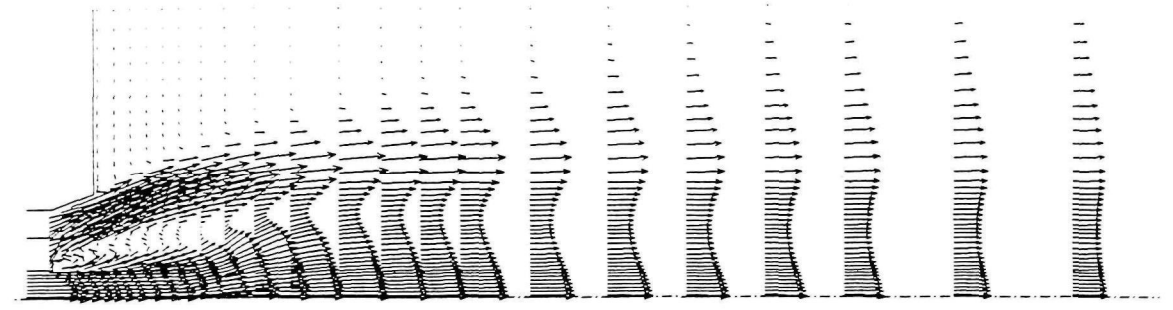
As-Fired Coal Analysis	Pittsburgh #8 hvA bituminous	Lower Freeport mv bituminous
Proximate (weight %) :		
Moisture	1.06	4.85
Volatile Matter	36.61	24.90
Fixed Carbon	55.44	60.62
Ash	6.89	9.63
Ultimate (weight %) :		
Carbon	78.11	75.26
Hydrogen	5.22	4.55
Sulfur	1.12	1.26
Oxygen	6.17	3.00
Nitrogen	1.43	1.35
Ash	6.89	9.73
Moisture	1.06	4.85
Pulverized Coal Size Distribution (weight %) :		
70.65% < 200 mesh (74 microns)		
99.64% < 50 mesh (297 microns)		
Natural Gas Analysis (weight %)		
CH_4 = 93.30		
C_2H_6 = 3.08		
C_3H_8 = 0.27		
C_4H_{10} = 0.35		
CO_2 = 0.80		
N_2 = 2.20		
Thermal Load = 5 MBtu/hr (1.5 MW_{th})		
Excess Air = 16 %		
Primary Air Temperature = 150°F (339 K)		
Secondary Air Temperature = 620°F (600 K)		
Natural Gas Temperature = 80°F (300 K)		

Results and Discussions

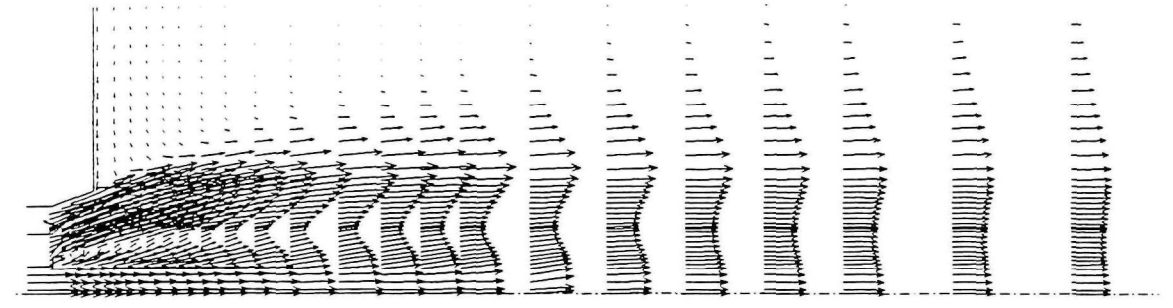
Initial computer runs for 100% PC firing or baseline operation were carried out in both 2-D and 3-D geometries. Numerous permutations in gas velocity, injection angle, injection location, extent of co-firing, and natural gas/air premixing were modeled for comparison and screening. Because of the large number of modeled cases, only selected predictions for annular injection of natural gas are presented. Other key results for the co-firing of natural gas with the high volatile Pittsburgh #8 coal are instead summarized below:

- For a given level of natural gas co-firing, injection velocity had a significant impact on the NO_X emissions but a lesser impact on the unburned carbon loss.
- Increasing the extent of co-firing reduced the NO_X and unburned carbon levels.
- Injection methods that enhanced natural gas/air mixing and consumed oxygen quickly near the burner resulted in lower NO_X but higher fly ash unburned carbon.
- Natural gas/air mixing for peripheral injection cases was dominated by the secondary air flow.
- Introduction of natural gas into an inherently low-NO_X and oxygen-deficient zone demonstrated marginal NO_X and unburned carbon reductions.
- NO_X removal became more effective when natural gas was delivered into flame regions where the NO_X formation rate was high.
- For annular injection of natural gas, NO_X emissions and unburned carbon levels were strongly influenced by the radial placement of the gas stream relative to the coal nozzle.

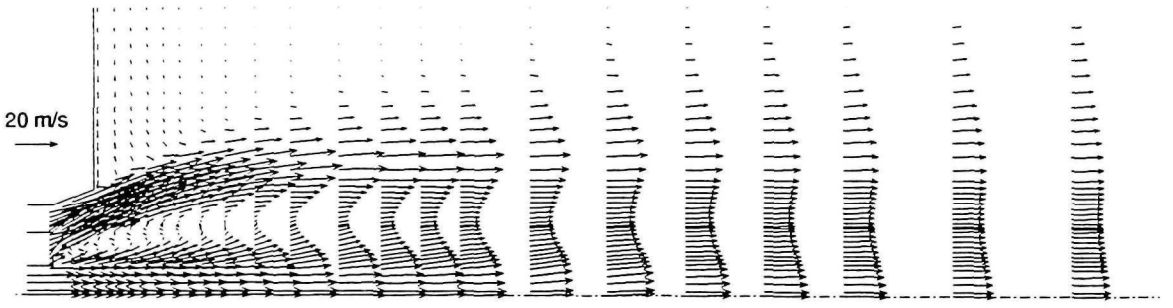
Mid-plane predictions of the gas velocity fields, temperature contours, and O_2 and CO concentration profiles for the baseline and peripheral co-firing of the high and medium volatile coals are compared in Figures 3-6. Here, gas co-firing is equivalent to 15% of the combined thermal load. Burning of natural gas with the fast mixing secondary air alters the combustion gas flow patterns. Particularly, it distorts the low momentum regions near the burner exit as illustrated in Figure 3. Figure 4 shows the differences in the near burner temperature fields among the four cases. With natural gas co-firing, a narrow and high temperature region extends out into the secondary air stream. The longer high temperature lobes are indicative of near stoichiometric combustion zones.



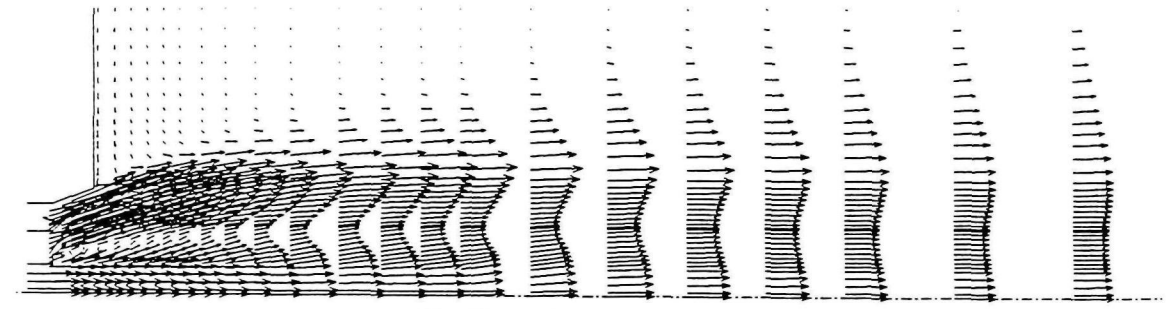
High Volatile Coal Baseline



High Volatile Coal Co-firing



Medium Volatile Coal Baseline



Medium Volatile Coal Co-firing

Figure 3
Near-burner Velocity Fields for the Baseline (100% PC-firing) and 15% Annular Gas Co-firing Predictions

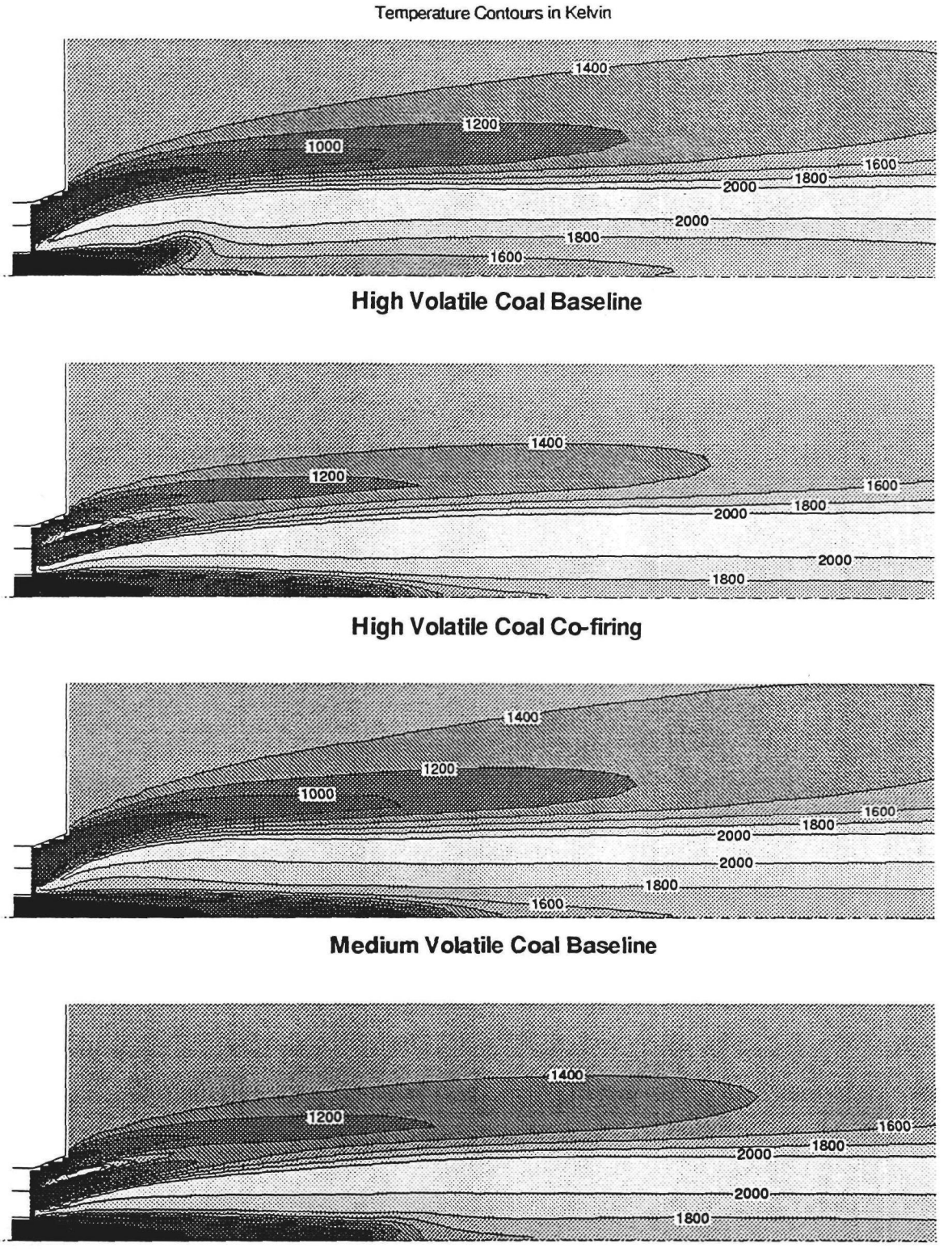


Figure 4

Near-burner Temperature Contours for the Baseline (100% PC-firing) and 15% Annular Gas Co-firing Predictions

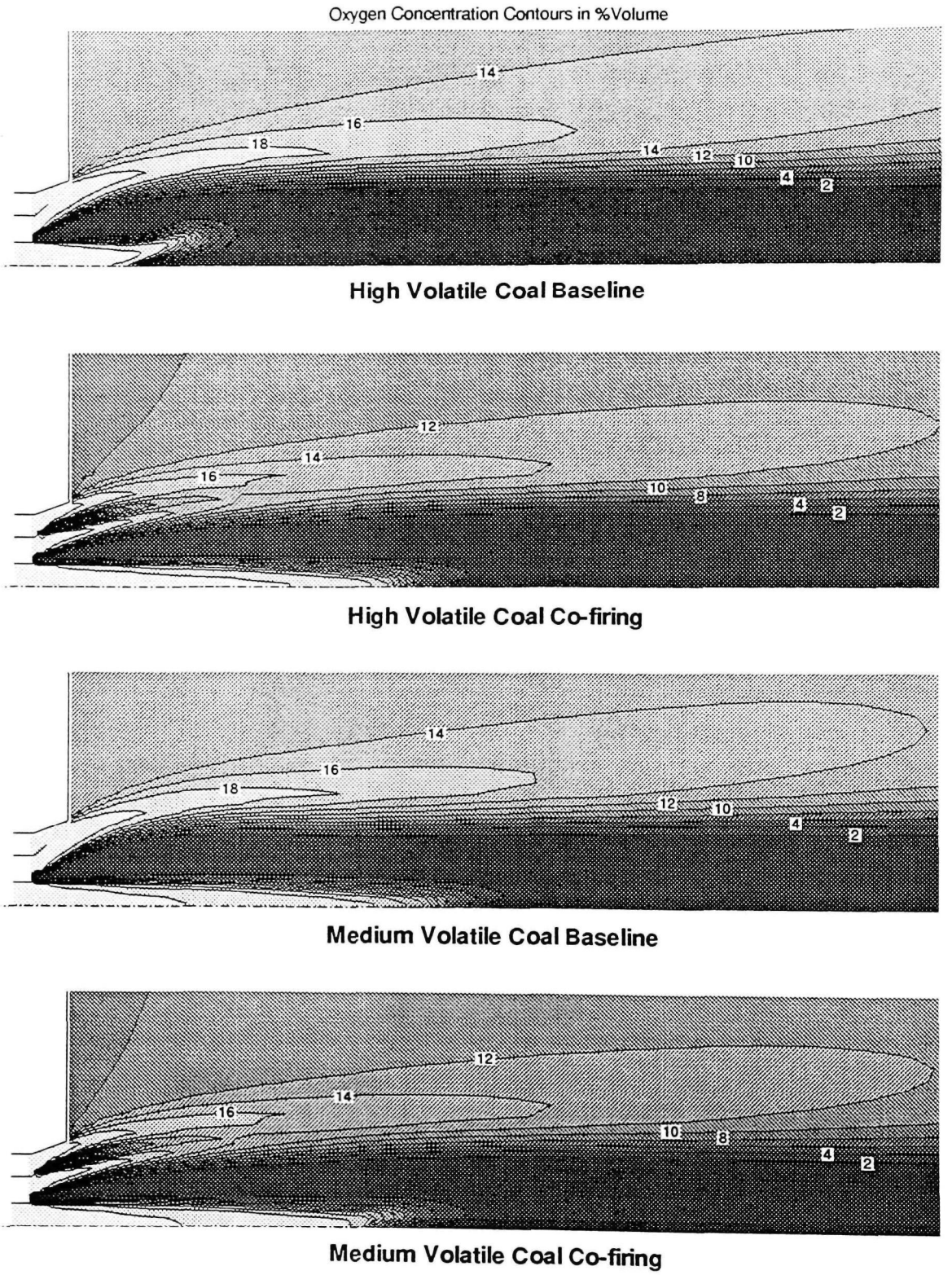


Figure 5

Near-burner Oxygen Concentrations for the Baseline (100% PC-firing) and 15% Annular Gas Co-firing Predictions

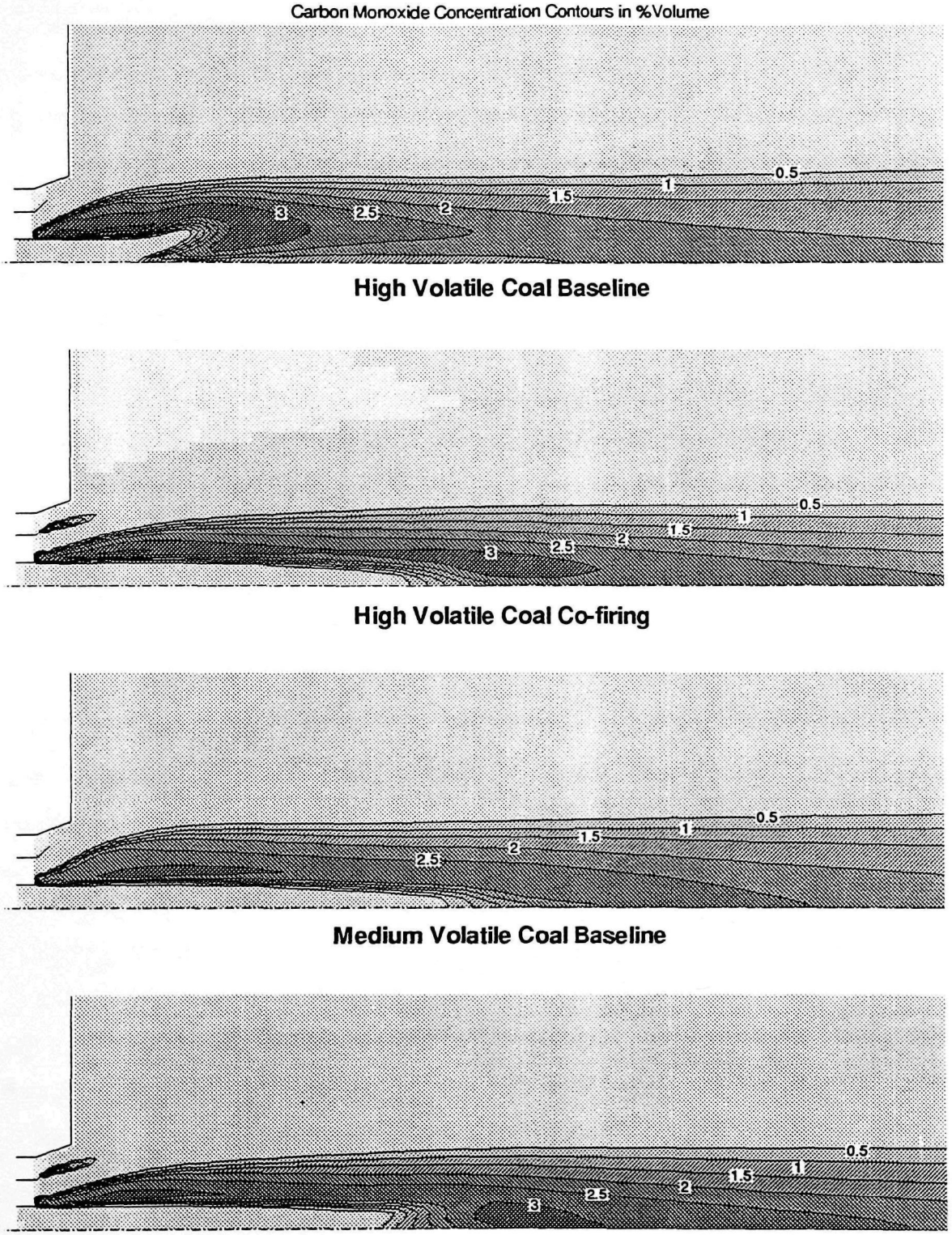


Figure 6

Near-burner Carbon Monoxide Concentrations for the Baseline (100% PC-firing) and 15% Annular Gas Co-firing Predictions

Contours of oxygen concentrations are shown in Figure 5. Relative to the baseline results, the co-firing cases show a more rapid consumption of oxygen in the upstream section of the furnace tunnel where natural gas burns. Carbon monoxide concentrations are also generally high in the vicinity of gas injection and the central primary combustion zones as seen in Figure 6. As natural gas burns, the ensuing flame shields the substoichiometric primary combustion zone (where fuel-N species evolve) from oxygen penetration to reduce NO_x formation.

Table 3 lists the normalized furnace exit values of NO_x, gas temperature, and unburned carbon in the fly ash (UBCA) relative to the baseline (100% PC firing of low-NO_x burner) results for two different coals. Flame lengths were also determined qualitatively based on a pre-specified and arbitrary CO concentration. Tabulated flame lengths are for trend indications only. Actual flame length depends on additional factors such as soot radiation, remaining char fraction, and temperature and oxygen distribution within the furnace. As expected, the baseline NO_x and unburned carbon levels for the more reactive high volatile Pittsburgh #8 coal are lower than values predicted for the medium volatile matter Lower Freeport coal. Burning 15% natural gas with Pittsburgh #8 coal reduced the NO_x concentration and unburned carbon level by 16 and 19%, respectively. Furnace exit gas temperature (FEGT) changed by 3% due to co-firing.

Model predictions indicate that a very small fraction of NO_x reduction from managed natural gas co-firing is attributable to fuel-N displacement. The bulk of the decrease is associated with lower local O_2 concentrations and hydrocarbon interactions with NO via the $CH_i + NO \rightarrow HCN$ and $HCN + NO \rightarrow N_2$ and reactions. Considering the fact that NO_x emissions from the low-NO_x PC burner are already about 50% below the uncontrolled levels of conventional burners, the net NO_x reductions with 15% gas use amount to about 53% for the medium volatile coal and 60% for the high volatile coal.

Table 3 Summary of Key Modeling Results					
Coal	Case Description	Relative NO_X*	Relative UBCA*	Relative FEGT*	Flame Length**
hvA Pittsburgh #8	100% PC Firing	1	1	1	1
	15% Natural Gas Co-Firing	0.84	0.81	0.97	1.04
mv Lower Freeport	100% PC Firing (relative to hvA)	1 (1.06)	1 (2.07)	1 (0.96)	1 (1.04)
	15% Natural Gas Co-Firing	0.95	1.04	1.03	1.02

* Relative exit quantities defined as $X_{relative} = X_{cofire}/X_{baseline\ low\text{-}NOx}$

** Flame length defined arbitrarily by contours of 1% CO as a trend indicator. Actual flame length depends on additional factors.

Conclusions

Implementation of natural gas co-firing in a low-NO_X PC burner was studied by numerical simulation of several concepts. Combustion modeling effectively captured the influence of operating conditions and hardware configuration. Natural gas introduction into appropriate locations of the pulverized coal flame showed possible reductions in NO_X emissions and higher carbon utilization. Gas co-firing with a high volatile Pittsburgh #8 coal was more effective for NO_X and unburned carbon reductions in comparison with a medium volatile Lower Freeport coal. Relative to 100% pulverized Pittsburgh #8 coal firing in a low-NO_X burner, the NO_X emissions and unburned carbon levels for 15% natural gas use were 16 and 19% lower, respectively. Since the NO_X emission levels for the baseline low-NO_X pulverized coal burner are already about 50% lower than the unregulated values of conventional burners, total NO_X reduction with 15% gas use would be close to 60%. Future plans include further modeling and optimization of the gas co-firing concept(s), and proof-of-concept verification tests in a pilot-scale boiler facility.

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AN INTEGRATED FULL, PILOT AND LABORATORY SCALE STUDY OF THE EFFECT OF COAL QUALITY ON NOx AND UNBURNT CARBON FORMATION

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Abstract

PowerGen and EPRI have co-funded a two year research project to investigate the effect of coal quality and coal blends on NOx formation and unburnt carbon in low-NOx combustion systems. The objective of the programme was to develop a better understanding of these effects in relation to combustion staging and coal blending in order to derive more reliable predictive methods. Eight bituminous coals covering a wide range of properties and geographical source were included in the study.

The comprehensive work programme comprised full-scale plant trials on a 500 MW_e tangentially fired boiler fitted with LNCFS backed up by smaller scale testing at the Power Technology Centre using a 1MW (3.4 MBtu/h) Combustion Test Facility and a laboratory Drop Tube Furnace. Coal and char samples pertinent to all experimental scales have also been examined at Nottingham University using a recent development of an automated image analysis technique for coal petrology and char characterisation.

The overall results of the programme are summarised in this paper. The value of rig testing for the prediction of NOx emissions from individual coals in full-scale plant has been successfully demonstrated, through calibration with plant data. As expected, NOx emissions do not correlate simply with either nitrogen content or with fuel ratio based on laboratory determined volatile matter; a stronger correlation is found with a function based on both coal volatile matter and nitrogen content. It has been shown that the sensitivity of NOx emissions to coal quality is reduced as the degree of combustion staging increases. All coals tend towards a NOx minimum of 150-200vpm under deep air staging.

The level of unburnt carbon, an often ignored disadvantage of low NOx combustion systems, has been shown to be highly dependent on coal quality, but is much less sensitive to combustion staging. A novel coal characterisation method based on grey scale reflectance shows considerable promise as an index for carbon burnout propensity.

Introduction

Generally throughout the world, emission limits for NOx are becoming more stringent, necessitating the application of one (or more) of a range of NOx reduction technologies appropriate to the legislative requirements of the different countries. One of the more favoured groups of technologies involves the use of combustion staging and includes such concepts as low-NOx burners, overfire air (OFA) and reburn. However, although this type of technology is attractive from a capital outlay perspective there is a potential penalty arising from a tendency for the level of unburnt carbon (UBC) in the flyash to increase. This represents an efficiency loss and, as there is a carbon-in-ash limit for certain uses for fly ash, could also have a deleterious effect on ash sales to the construction industry.

NOx emissions and UBC levels are clearly highly dependent on boiler design and operation, but experience has also shown that coal quality can have a significant effect. Accordingly, the UK electricity generating company PowerGen has collaborated with EPRI in a research project to investigate the effect of coal quality on NOx emissions and unburnt carbon. The project comprised full-scale plant trials supported by smaller scale testing using a 1MW single burner pilot scale combustor (the Combustion Test Facility, CTF) and a laboratory Drop Tube Furnace (DTF). The main objective of the project was to attempt to develop a simple, but reliable, technique for predicting both NOx and UBC levels for a given coal, preferably not involving combustion or the use of expensive or complex testing equipment. To this end, automated image analysis techniques for both coal petrology and char characterisation recently developed at Nottingham University have been applied to the coal and char samples produced at the different scales of combustion.

Project Outline

The project centred on, and began with, controlled full scale test burns of three coals and a 50:50 blend of two of these coals on one of four 500 MW_e boilers at PowerGen's Kingsnorth Power Station. Extensive flue gas and particulate analyses and performance data were collected during the trials and samples of the test coals of sufficient size for the other phases of the project were withdrawn to Power Technology.

The CTF programme included the characterisation of the three power station coals and the blend, plus five additional coals and two more blends thereby extending the range of coal qualities investigated in the project. Finally, the opportunity was taken to investigate the effect of the particle size distribution of the ground pulverised coal on both NOx emissions and UBC; for this exercise the three primary coals featuring in the plant trials were used.

The Drop Tube Furnace is becoming more widely accepted as a coal characterisation technique that can closely approach the high heating rates and high temperatures experienced by coal particles in a utility boiler, but on much smaller samples and without the need to establish a stable flame. Characterisation of the devolatilisation is carried out under much more representative conditions than produced by standard ASTM (D3175-89) or other equivalent international laboratory methods. In addition, the DTF provides a better method of evaluating the reactivity or burnout propensity of the chars remaining after devolatilisation and of determining the denitrification profile of both coals and chars. The suite of eight coals tested on the CTF were fully characterised on the DTF during this programme.

A major objective of the work programme was to try to calibrate the two pilot scale facilities with the power plant in order to determine the reliability of such techniques for making predictions of the propensity of different coals to produce NOx and unburnt carbon. However, it is recognised that although the CTF, and to a much lesser extent the DTF, provide more realistic conditions under which to test and evaluate coals and should provide credible performance predictions of these two combustion bi-products, such techniques are still too complex to be considered as standard for routine application. Each suffers from a combination of the following disadvantages; capital and revenue intensive, requires expert operation and data interpretation, needs large coal samples and must be pre-calibrated with the boiler under consideration. Consequently it was decided to add to the project an evaluation of the potential of a new image analysis based technique being pioneered in the Chemical Engineering Department at Nottingham University to provide accurate performance predictions of unburnt carbon.

The Nottingham technique, which was initially developed under the co-sponsorship of PowerGen and the Engineering and Physical Sciences Research Council in the UK, combines both automated and semi-automated petrographic/reflectance analysis of the coals with detailed automated analysis of the morphology of chars produced both under full and pilot scale combustion conditions. Should analysis of small samples of the coal in this way prove to correlate well with combustion data, the technique will provide a non-combustion and less subjective method of assessing the reactivity of different coals.

Plant Trials at Kingsnorth Power Station

Description of Plant

Kingsnorth Power Station comprises 4 x 500 MW_e split furnace, tangentially fired boilers. There are five elevations of coal burners within each corner arrangement (Figure 1), the eight burners on each elevation being fed by a Lopulco type mill (A mill feeds the top burner elevation and E mill the bottom). The full burner configuration is of the LNCFS type with large offset secondary air nozzles intermixed with the coal and light up oil burners and a separated overfire air arrangement. Kingsnorth was originally designed for mixed firing (representing a 100% capability on either residual fuel oil or coal) and is therefore effectively one mill/burner group short of ideal in the now routine coal only operational mode.

Selection of Coals

The coals for the four separate trials were:

(1) Ashland - a typical US coal that has been a major supply to PowerGen.

(2) El Cerrejon a Colombian coal that has been associated with high carbon-in-ash levels. This coal also has a low fuel ratio (fixed carbon to volatile matter) and might, therefore, be expected to give relatively low NOx emissions.

(3) Tyne Blend a typical UK coal that is routinely supplied to Kingsnorth.

(4) A 50:50 blend of El Cerrejon and Tyne Blend - this trial was designed to quantify the effects of blending.

Test Schedule

Each trial consisted of 2 days testing during which the Unit was operated exclusively with the chosen coal. Although 5 in service mills are always preferable, the boiler often operates with 4 mills, ideally B,C,D, and E. Previous experience has shown that whether the top mill (A) is in or out of service affects NOx formation. Accordingly, the operational variables selected for investigation were excess oxygen, overfire air, and mill grouping as follows:

Day 1 - top mill (A) out-of-service, four tests each of about 1h duration. Three tests at the normal overfire air (OFA) setting, ie 18%, covering a range of excess O_2 levels, about 1.5, 2.5, and 3.5%. Fourth test again at about 3.5% excess O_2, but with the OFA reduced to about 10%.

Day 2 - as for Day 1, but with A mill in service and the bottom mill (E) out-of-service.

As far as possible, each test was designed to be at full load with operating conditions held constant and no sootblowing or ashing out. The burner tilts were fixed at +8°.

Measurement and Sampling Schedule

Gas Analysis. The concentrations of NO, SO_2, CO, and O_2, were continually monitored by extractive sampling at the ID fan discharges on the A and B sides of the boiler.

Carbon-in-Ash. Representative flyash samples were obtained by multi-point, isokinetic sampling on each side of the boiler at the air heater inlet to give separate A and B side samples.

Coal and PF Samples. 20 tonne samples of each coal were collected for subsequent testing on the CTF and DTF.

On each test day, samples of pf were taken from each operating mill.

Results

Test Record and Basic Data. In setting up the boiler to the desired excess O_2 concentration for each test, it was impossible to achieve a balance between the A and B sides. The difference was generally small, up to about 0.3%, when A mill was out of service but, with A mill in service, varied from 0.5 to 1.1%.

The overfire air (OFA) percentage was reduced to 10% by shutting the appropriate OFA damper. No attempt was made to adjust the windbox pressure which was, therefore, higher during these tests than the equivalent test (3.5% nominal O_2) at 18% OFA.

Coal Analysis. The results of the analysis are presented in Table 1.

Pulverised Fuel (pf) - Size Grading. All of the pf samples taken from the operating mills during each test were size graded. Single point, non-isokinetic sampling is notoriously unreliable and these results were treated with the usual caution. Despite this reservation, it can be stated with some confidence that the proportion of coarse material, ie % >300μm, produced by the mills is quite low, i.e. less than 2%. However, beyond this, it is impossible to say whether there was any significant difference in pf grading between coals.

Carbon-in-Ash. A summary of the residual carbon-in-ash results for the three individual coals plotted against excess O_2 is presented as Figure 2. The data obtained during the blend trial were quite scattered, but in general they were found to lie between those obtained in the individual El Cerrejon and Tyne trials. There were differences between the results for the A and B sides of the boiler with and without A mill in service, but no consistent patterns could be detected.

Whilst the results for Ashland and Tyne Blend can reasonably be represented by a single curve, El Cerrejon coal clearly gave higher results. However, for coals with different ash contents, the direct comparison of carbon-in-ash values is potentially misleading as an indicator of residual combustible matter or efficiency loss. Accordingly, by taking account of the ash contents of the different test coals (Table 1), the carbon-in-ash values have been converted into unburnt carbon or carbon loss values as a percentage of the dry coal. All of the carbon loss data for the individual coals are plotted against excess O_2 in Figure 3. As the ash contents of the Ashland and Tyne Blend coals were very similar, the results for these coals can again be represented by a single curve. For El Cerrejon, the ash content was lower (about 10% compared to 13-14%) and, compared to Figure 2, the results are, therefore, closer to those for Ashland and Tyne Blend. However, on average, they are still higher by about 50%.

NO Measurements. The relationships between NO and excess oxygen for each individual coal for the two firing configurations are shown in Figures 4 and 5. No major differences are apparent between the different coals and, with A mill in service, all the points can reasonably be represented by a single line (Figure 4). With A mill out of service (Figure 5), the NO emissions were slightly higher, about 25vpm, when firing Tyne Blend coal compared to Ashland and El Cerrejon which can again be represented by one line. As expected, the NO emissions are consistently higher with "high firing", i.e. the top burners (A mill group), in service. At 6.5% O_2, equivalent to about 3.2% wet at the air heater inlet, the difference is between 50 and 75vpm, depending on the coal. There is also a suggestion that NO is slightly more sensitive to O_2 with A mill in service, ie the line is slightly steeper.

With A mill out of service, the decrease in overfire air to about 10% had little or no effect on NO emissions and so the points have not been plotted on Figure 5. In contrast, however, it can be seen in Figure 4 that, with A mill in service, the decrease in overfire air from the normal 18% to 10% caused the expected increase in NO emissions. However, the magnitude of the effect appears to be coal dependent and varies from about 75vpm for Tyne Blend down to 10-30vpm for Ashland.

Discussion

NO Emissions. For the three coals chosen for the plant trials very little difference in NO emissions was found under normal operating conditions (18% OFA). Overall, the results for Tyne were slightly higher than for Ashland and El Cerrejon, but only by about 25vpm with A mill out of service (Figure 5). Under normal operating conditions with A mill out of service, 18% OFA and 3.5% (wet) excess O_2 measured at the air heater inlet, NO emissions of 330-360vpm (3% O_2 dry) were measured, which is within the 390vpm target for the Station.

In view of the similarities of the nitrogen contents and fuel ratios for the three coals (Table 1), comparable NOx emission levels would perhaps be expected. However, previous trials of South African coals on Unit 3 at Kingsnorth PS showed that NO emissions can be very sensitive to coal quality, with one of the coals giving about 530vpm NO at 5.5% O_2 dry (A mill in service), compared to about 330vpm for these trials under similar conditions. This is illustrated in Figure 6 where the data obtained from these previous trials are plotted alongside the results of the current study. It was the test of this coal that revealed that it would not be possible to operate the boilers below the agreed NOx emission limit on certain coals and which gave impetus to this study. The small difference between the two sets of Tyne Blend data may be associated with different levels of air heater inleakage between the tests on Units 1 and 3. It is the furnace O_2 concentration that actually determines the level of NO formation.

As expected, the burner firing pattern had an effect on NO emissions. With the top (A) mill in service, NO emissions were up to 75vpm higher than when it was out of service. This is in line with previous findings and is believed to be due to the longer residence time of the pf in the sub-stoichiometric zone with E mill in service compared to A mill in service.

The investigation of the effect of OFA on NO emissions produced some interesting differences, depending on mill combination. With A mill in service, a reduction in OFA from a nominal 18% down to 10% did result in an increase in NO emissions, but the effect was coal sensitive. Tyne Blend gave the greatest increase, about 75vpm, with El Cerrejon and Ashland some 20-55vpm lower. In contrast, with A mill out of service, no measurable effect was observed. The effect of mill combination is probably related, in part at least, to differences in the true proportions of OFA and combustion air. With A mill out of service, the cooling air passing through A group burners has been calculated to be equivalent to about 4% OFA. In effect, therefore, the tests with A mill out of service were conducted with 22% and 14% OFA, compared to 18% and 10% with A mill out in service. The effect of the proportion of OFA on NO emissions is not linear, but is an asymptotic curve approaching a minimum attainable NO level or maximum reduction. The 18% OFA operational norm was chosen as any further increase would simply increase the unburnt carbon without any significant reduction in NO being achieved. This is fairly generally accepted and has been confirmed by the tests on the CTF (see below) which indicate that NO emissions are relatively insensitive to OFA above about 15%. It follows, therefore, that little effect would be expected between 14 and 22% OFA (A mill out of service), compared to the change between 10 and 18% OFA (A mill in service). The other contributory factor is the lower baseline NO level with A mill out of service which will make it intrinsically less sensitive to changes in OFA.

The coal sensitivity of the effect of changing OFA with A mill in service is more difficult to explain, but may relate to the relative amounts of char and volatile nitrogen produced by the different coals. The similarity in the basic NO levels obtained with the three coals suggests comparable char nitrogen levels. It is possible, therefore, that, as the coal nitrogen content of Tyne Blend is a little higher than El Cerrejon and Ashland (Table 1), the amount of volatile nitrogen released would be greater. Under these conditions, the NO formed when burning Tyne Blend would be more sensitive to changes in OFA.

Unburnt Carbon. El Cerrejon coal is widely reported to give rise to high carbon-in-ash levels and this has been found during trials at other PowerGen plant. This is confirmed by these trials and, after taking the difference in ash contents into account, the mean increase compared to Tyne Blend and Ashland was about 50% - see Figure 3. It is assumed that this difference is related to the nature of the char produced by devolatilisation of El Cerrejon coal, ie differences in intrinsic

char reactivity, or, as would seem more likely from the results of other phases of this work, the porosity of the char.

The relationship between unburnt carbon and excess oxygen found for Tyne Blend (and Ashland) was in line with that found during recent thorough boiler characterisation tests. Under normal operating conditions of 3.5% (wet) excess O_2 measured at the air heater inlet (equivalent to 6.7% dry at the ID fan discharge), the unburnt carbon-in-ash value was around 5% - see Figure 2.

Combustion Test Facility Tests

Description

The 1 MW_t (3.4 MBtu/h) rated CTF has been described in detail elsewhere[1] and so only an outline description is given here; the general layout of the CTF is shown in Figure 7. The facility is horizontally fired by a single burner and has been designed with a high degree of fuel flexibility; it is capable of firing on pulverised coal, residual fuel oil (RFO), Orimulsion and natural gas. The furnace and high temperature convective pass are refractory lined and water cooled. The basic design philosophy was to provide an accurate simulation of the time/temperature profile of full-scale plant, i.e. residence time scaling.

For the tests reported here, the rig was fired with a single low-NOx PF burner with a combustion air system comprising primary, secondary, tertiary, and furnace overfire air. When overfire air was being used the flow was adjusted by diverting proportions of the secondary and tertiary air flows such that the ratio of secondary to tertiary air flow remained constant.

Continuous measurement was made of O_2, CO, SO_2, and NO in the flue gas.

Test Coals

The additional five coals introduced to broaden the range to be tested at this scale were typical coals from Australia (Drayton) and South Africa (ATC), two further US coals previously fired on the Southern Research Institute pilot scale facility in Birmingham, Alabama (Shoal Creek and North River) under a complementary EPRI supported programme[2], and a high rank UK coal (Leasingthorne). The proximate and ultimate analyses of all eight of these coals are shown in Table 2. The size of the PF is also shown as the percentage less than 75μm.

Three 50:50 blends were also tested; Tyne Blend/El Cerrejon (as per the plant trials), Tyne Blend/Leasingthorne and Ashland/Shoal Creek. Finally, the effect of changing the particle size distribution of the ground coal for three of the coals, Tyne Blend, Ashland and El Cerrejon has been assessed.

Test Schedule

Each coal was fired on three separate days to cover a range of OFA levels 0, 15, and 25% of total air flow. On each day the effect of excess oxygen concentration was studied by making measurements under steady operating conditions at about 4.5, 3.0 and 1.0% O_2. At the beginning and end of each test day the repeatability of the NO v. O_2 characteristic and the CO breakpoint was checked by ramping the O_2 from 4.5 to about 1.0% over approximately 30 minutes.

NO Measurement Results

The NO and O_2 concentrations in the flue gas were monitored throughout the experiments to provide mean NO results for each test condition. Early in the programme the NO_2 concentration in flue gas was also checked and found to be extremely low, typically 3vpm, maximum 7vpm. Accordingly, this measurement was discontinued and measured NO concentrations are assumed to equate to NOx emissions.

The NO results obtained are shown in Table 3 and are quoted in volume parts per million (vpm) on a dry basis, corrected to 3% O_2 (measured dry). Throughout these experiments the relationship between excess O_2 and NO was found to be linear. These relationships are shown for the 8 coals at the zero and 15% OFA settings in Figures 8 and 9: the results at 25% OFA were only marginally different from those at 15% OFA.

Unburnt Carbon Results

To study carbon burnout, flyash samples were taken from 5 sampling ports from the flame zone through to the end of the top pass where combustion is reasonably assumed to be complete. The samples were extracted using a wet sampling system.

The unburnt carbon content or loss-on-ignition values for the final flyash samples obtained from the final sampling point (gas temperature about 800°C) are presented in Table 4. To correct for the differences in ash content and derive an equivalent boiler efficiency penalty, the results are also shown as a carbon loss as a percentage of the original coal, ie ash fraction multiplied by carbon-in-ash.

Discussion

NO Measurements - General Observations. A number of general trends are apparent from the results presented in Table 3 and Figures 8 and 9. First, over the range tested, the level of NO formation is linearly dependent on excess O_2 concentration, but the slope of the line is variable depending on the degree of combustion staging. At zero OFA the mean slope for all the test coals is 46 vpm/% O_2, falling to 37 at 15% OFA; at 25% OFA it was 26 vpm/% O_2. Second, a dependence on coal quality is also clearly apparent with a spread in the NO results reducing from about 70vpm at zero OFA to about 50vpm at 15% OFA (and 35vpm at 25% OFA). Some variation in the O_2 dependence is also apparent. Third, the reduction in NO that can be obtained by the use of OFA is also clearly apparent, with a mean reduction of 20% at 3% O_2 obtained with 15% OFA. The gains obtained from 25% OFA are more modest with a further 7% reduction relative to zero OFA. Again, the level of reduction obtained is somewhat coal dependent.

Overall, it appears that the NO level obtainable from any coal tends towards a minimum value in the range 150 to 200vpm. This minimum will be fixed by the nitrogen content of the char obtained by devolatilisation of that coal. The NO formed by combustion of the char will not be influenced by combustion staging.

The Relationship between NO Emissions and Coal Quality. It is generally agreed that >80% of NO is formed from the fuel nitrogen and it is to be expected, therefore, that NO emissions will be related to a coal's nitrogen content. It is also recognised that the high temperature volatile matter release is an important coal property, in that the higher the proportion of the coal nitrogen that is released during devolatilisation, the lower will be the NO formation by the subsequent oxidation of the char nitrogen. Standard analysis provides a measure of volatile matter content under laboratory conditions that is lower than the volatile release under high heating rate conditions. Although it will be seen in the section below on the DTF that the ratio of high heating rate to laboratory derived volatile matter covers the range 1.31 to 1.79, if this factor was assumed to be constant then proximate volatile matter content should also relate to NO formation. This is conventionally indicated by the fuel ratio which is defined as the ratio of fixed carbon to volatile matter.

On the above basis, NO emissions would be expected to show some relationship with the product of coal nitrogen content and fuel ratio. Such a plot for the 8 test coals is shown in Figure 10. A rough trend is apparent, but there is considerable scatter in the data and there is too great a concentration of data points at low fuel ratio x nitrogen values and a sparsity at higher values.

In a highly staged combustion system it is probable that the vast majority of the NO is formed by oxidation of the char nitrogen and the contribution from the volatile nitrogen species approaches zero. Assuming that (1) fuel ratio is a reasonable guide to volatile release and, (2) the partitioning

of nitrogen between the volatiles and char matches the overall volatile/char ratio, then a high correlation coefficient would be expected. In practice, though the second assumption is normally true, the first does not necessarily apply. It is probable that variations in the high heating rate volatile release relative to the proximate analysis volatile matter content accounts for some of the scatter in the data shown in Figure 10. (This is the subject of further discussion in the section below on the DTF). The lower correlation coefficient for the 0% OFA results is probably due to a variable contribution of volatile nitrogen to the higher NO levels under less staged combustion conditions.

Unburnt Carbon Results. Table 4 shows that for each of the coals the expected trend of combustibles loss with operational excess oxygen was observed, but there is no apparent effect of overfire air. Plant experience with overfire air generally would suggest that the deeper the staging, the higher the residual combustibles, but this was not observed on the test facility. A reasonable explanation for this is that being a small, well mixed reactor, the CTF still provides the necessary conditions (i.e. sufficient residence time at high temperature and high local oxygen partial pressures) for good burnout even under deeply staged conditions.

Marked differences between the coals are apparent. Looking, for example, at the carbon loss results at 1% excess O_2 (similar trends are apparent at other O_2 levels), there is about a factor of 2 difference between the least reactive coal, ATC and the more reactive coals. During the Kingsnorth trials the results (on a carbon loss basis) for El Cerrejon were found to be about 50% higher that those for Ashland and Tyne Blend. The CTF results for El Cerrejon were generally higher than those for the other two coals, but not on average by as much as 50%. Again, this difference could be due to the better control of the fuel/air stoichiometry on the CTF.

Comparison of CTF and Plant NO Results

The CTF has been designed to provide a realistic simulation of combustion conditions in real plant and should, therefore, give similar NO emissions. In this respect the results obtained are very encouraging. As an example, some comparative results for Ashland are shown in Figure 11. The plant results are those obtained with A mill out of service and a nominal 15% OFA, equivalent to about 22% when the air from the standing burner is included. The CTF results included are those for 0 and 15% OFA.

Agreement between the CTF 0% OFA data and the plant data is excellent. The results obtained on the CTF at 15% OFA (or 25%, the two sets of data are very similar) were much lower and at first sight would appear to show poor agreement under notionally similar conditions. However, it is believed that the 0% OFA case on the CTF actually represents the more accurate simulation of the LNCFS at Kingsnorth. In this system, pf is simply blown into the furnace from each corner with offset secondary air and there are no rooted flames. The overall system, therefore, effectively behaves as a single large burner, with the OFA acting in a similar manner to the tertiary air in the single rooted flame obtained in the CTF. The greater staging that can be obtained from a rooted flame with OFA results in the lower NO levels obtained on the CTF with 15% and 25% OFA.

This theory is supported by results obtained with International Combustion's (ICL) recently developed FAN (flame attached nozzle) burner. In this tangentially fired system the corner flames are rooted by the inclusion of a flame holder and, therefore, simulate the single low NOx burner in the CTF. This has been shown on plant to give a reduction in NO formation of about 50-80vpm[3]; if applied at Kingsnorth, for Ashland coal this would give a similar result to that obtained on the CTF at 15% OFA.

As part of a separate study, the S.African ATC coal has also been fired at Kingsnorth and the results obtained there and on the CTF (0% OFA) are also included in Figure 11. Again agreement is fairly good. It is also encouraging that the sensitivity to excess oxygen on the CTF, ie the slope of the NO v. O_2 line, is similar to that obtained at Kingsnorth PS.

Drop Tube Furnace Tests

Description of Equipment

The DTF (Figure 12) comprises an electrically heated work tube which is approximately 1.5m in length with an inside diameter of about 50mm. Coal particles are fed through a screw feeder into the DTF with a nitrogen purge via a water-cooled feeder probe. The coal devolatilises in a controlled, high-temperature atmosphere and the resultant char material is collected by a water-cooled probe. From the collector probe the char particles and waste gases pass through a cyclone, where the solid particles are removed. When char material is re-fired for reactivity determinations, it is fed in the same way as coal but the resultant residue is collected using a millipore filter rather than the cyclone. The separation between the feeder and collector probes can be varied up to 60cm to provide residence times up to 600ms depending on the temperature and gas flow. The DTF can be operated at up to 1400°C and air and nitrogen are premixed to give a known O_2 concentration in the reaction zone.

Experimental Programme

The test conditions chosen for the devolatilisation experiments were a temperature of 1300°C and an atmosphere of 1% O_2 in nitrogen. It is normal practice to test size graded coal for ease of feeding and to eliminate variations in size distribution between coals. 53-75μm and 106-125μm size cuts were prepared from a nominally ≤212μm sample of each coal taken from the Combustion Test Facility feeder, the former being typical of the mean for pulverised fuel. For these size grades devolatilisation is complete in under 100ms and, at only 1% O_2, little change occurs thereafter. Using a longer residence time gives a more even temperature distribution, ie minimises end effects, and 200ms was adopted for these tests. The chars were collected and refired at a temperature of 1300°C and in an atmosphere of 5% O_2 in nitrogen, at residences times of 200, 400 and 600ms.

A proximate volatile test was carried out for each coal feed and the coal feed and resultant chars and residues were analysed for carbon, hydrogen and nitrogen. The loss on ignition (LOI) at 800°C was determined for each coal feed sample.

Results

Devolatilisation. Table 5 gives a summary of the devolatilisation data from the DTF experiments. The percentage loss of raw coal determined on the DTF ie the high heating rate volatile matter on a dry basis (VM_{hhr}) and the "R" factor, the ratio of the volatile matter content (dry basis) of the feed coal determined by proximate analysis to VM_{hhr}, are shown for each coal. The combustibles loss, ie the proportion on non-ash material lost, was determined by the ash tracer method. The nitrogen loss was calculated from the percentage of nitrogen in the feed coal and resultant char in the same way as combustibles loss. The relative amounts of nitrogen in the char and volatile phase were determined from this nitrogen loss.

Figure 13 compares the combustible and nitrogen loss results. There is some scatter in the data but for devolatilisation the fractional nitrogen loss is on average slightly less than the combustible loss.

Char Refiring. Table 6 gives a summary of the char refiring data. The average measured NO was calculated from the NO measured during the char refiring phase. The char conversion was calculated by the ash tracer method in the same way as the combustible loss for the devolatilisation stage. To compare all the coals on the same basis the NO figure was converted to a value representing the NO produced if 100% of the carbon-rich fraction of the char had been converted and then normalised for a net CV of 29,000 $kJkg^{-1}$. Figure 13 also includes a comparison of the combustible and nitrogen loss results under char refiring. The fractional nitrogen loss is very close to the combustible loss for the char refiring stage.

Figure 14 shows the fractional nitrogen loss as a function of residence time for three of the coals

tested. As expected the fractional nitrogen loss increases with residence time for all coals. ATC is markedly different from the other coals in that the fractional nitrogen losses are lower and the curve does not approach an asymptote as residence time increases.

A prediction of the amount of NO produced can be generated for each coal. Figure 15 illustrates the actual NO produced as a function of the predicted value which assumes full conversion of all evolved nitrogen to NO. A linear regression of the data gives an NO conversion efficiency of approximately 50%.

Discussion

The VM_{hhr} values for the suite of coals range from 41% for Shoal Creek (106-125μm) to 61% for El Cerrejon and Ashland (53-75μm). The "R" Factor for the suite of coals varies between 1.31 and 1.79, with a mean of 1.57. In general, therefore, the proximate volatile content can only be considered as an approximate guide to the high heating rate volatile release. These "R" factors have been used to create a high heating rate fuel ratio and the result has been used to modify the plot of CTF NOx emissions v. fuel ratio x nitrogen content shown in Figure 10 to produce Figure 16. It can be seen that the correlation is much better, thereby emphasising the limited value of some data derived under the conditions incorporated into current standard procedures.

Table 5 shows a range of partitioning for the coals tested which reflect the differences in both percentage nitrogen loss (35-58%) and the absolute levels of nitrogen in the original coal feed (1.43-1.81%). The hypothesis described above that the level of NOx will be primarily related to the nitrogen content of the char in a low NOx system is broadly supported by the trend shown in Figure 17, in which NOx measurements from the CTF trials are plotted as a function of the char nitrogen g/100g coal from DTF. ATC, which gave the highest NOx levels on the CTF, has a relatively high char nitrogen g/100g coal and Ashland, which gave a lower NOx reading, has a much lower char nitrogen g/100g coal. The same trend is evident when char nitrogen g/100g coal is replaced with NO produced from the DTF (figure 18).

The unburnt carbon, as represented by LOI, shows the same trend for the DTF and CTF experiments (Figure 19). ATC, which has a high level of unburnt carbon when burnt on the CTF also gave a high value for the residue collected after char was refired at 600ms. Shoal Creek and Leasingthorne appear mid-range and the other coals all have relatively low LOIs, data which are borne out by plant experience.

Application of Image Analysis

At Nottingham University, image analysis programmes have been developed, specifically aimed at coals for pulverised coal combustion, which can carry out petrographic/reflectance analysis both automatically and semi-automatically (interactively) as well as semi-automated and automated analysis of char morphology[4]. This novel technology, which enables grey scale histograms of the type shown in Figures 20 and 21 (for the eight coals in this study) to be produced, was supplemented by the application of Thermogravimetric and Differential Thermal Analyses (TGA/DTA) to measure intrinsic reactivity. The bar at the lower grey-scales represents liptinite which is determined in a second automated run using blue light to distinguish it from the mounting resin used in the block preparation. In the image analysis technique, a Reactives Number (R_N) is defined which represents the percentage of the total reflected light below a certain grey scale level.

The Shoal Creek histogram shows a very broad and indistinct peak compared with the other coals and demonstrates that the selection of the cut-off point between reactives and unreactives can be very critical for some coals. The profile may also be indicative that the coal is a blend, but this is not as clear as in the Leasingthorne histogram in which two distinct peaks occur in the 60-160 R_N range.

The main objectives of this work were to (1) analyse chars found in fly-ash samples obtained from

plant trials to determine the nature of the char, (2) investigate correlations between the total reflectance (represented as a standardised grey-scale histogram) of a coal and its overall burnout in the CTF, and (3) determine correlations between char morphology and intrinsic reactivity of chars from the CTF and burnout. Comparisons were also made between the char types observed in the Drop-Tube Furnace (DTF), the CTF and from the plant trials.

The first finding was that the intrinsic reactivity of chars decreases significantly throughout the combustion process and that, irrespective of the original coal, chars found in fly-ash have similar intrinsic reactivity values as determined by the TGA/DTA method. This appears to concur with the observations of other workers and indicates that the reactivity of chars is reduced the longer they remain in the furnace.

The percentage of material with a value of less than 190 in the grey-scale histogram (the 190 Reactives Number, R_{N190}) correlates well with the percentage combustibles remaining in the char sample taken furthest from the burner in the CTF. An R-square factor of 0.93, showing an excellent linear correlation, is found for the results at 15% overfire air and 1% O_2 excess air (Figure 22), conditions which were found to be fairly representative of plant and at which there was a high value of residual combustibles. Correlation of percentage combustibles remaining with other more commonly used factors such as fuel ratio, rank and reactive macerals is in all cases not as good as for R_{N190}.

In addition to determining the intrinsic reactivity of the chars, their morphology was also assessed using an image analysis technique which calculates the pixel point "thickness" of all the char particles present in an image. Figure 23 shows representations of the type of chars that can be produced during the combustion process; given a similar intrinsic reactivity, those with an open, lacy structure would be expected to burn out to completion more readily that the chars of more solid appearance.

Good correlation was found between the measured morphology of char particles sampled from the port nearest the burner with both 190 reactives and percentage combustibles remaining. There is also a similar but not so strong correlation with intrinsic reactivity. The decrease in intrinsic reactivity as combustion proceeds is confirmed for chars produced in the CTF. Data linked to morphological changes as combustion proceeds were limited and further work would be required in this area to produce samples from which sufficient carbon can be isolated for a meaningful analysis. Char samples produced in the DTF are similar to those found near the burner in the CTF.

General Discussion

Although each phase and scale of this project has produced a valuable data set which, on their own, can be used to understand and compare the different behaviours and characteristics of a range of coals, the integrated nature of the work has enabled a certain amount of calibration across the scales and approaches to be made in order to establish the credibility of certain methodologies for NOx and unburnt carbon prediction. There are still gaps in the data, due to limitations caused by the scope of the study and the choice of the coal suite, but major advances have been made in defining the usefulness of both established and new laboratory scale methods for predicting full scale plant behaviour.

Although the combustion arrangement in the 1 MW_t pilot scale combustor did not simulate the tangential firing of the 500 MW_e boiler, it was possible to reproduce some of the subtle differences seen between the NOx and unburnt carbon figures produced on what turned out to be three coals of quite similar behaviour on the plant. It vindicates the rationale that residence time scaling is important when investigating the effect of coal quality on the products of combustion, but the findings also emphasise how difficult it is to simulate the behaviour of a large multi-burner furnace with a single burner system which is, by comparison, a very well stirred reactor. This is borne out by much better burnout being achieved on the pilot scale than on plant even when the NOx emissions are very similar and the coal grind is coarse, and suggests that the pilot scale could

provide more reliable absolute predictions for NOx than for unburned carbon.

It should be stressed, however, that the calibration obtained (i.e. pilot scale operating conditions which produce good similarity to plant NOx and unburnt carbon results) only holds for the plant in question and that predictions on the basis of pilot scale data for other plant could only be made after a similar calibration exercise. Similarly, if the combustion equipment or operational regime on the calibrated plant were to change significantly, then a re-calibration would be needed.

The expanded range of coals tested on the CTF (and subsequently on the other small scale facilities) did exhibit some differences both in NOx and, to a lesser extent, unburnt carbon. For NOx, a rough correlation was found with the product of coal nitrogen content and fuel ratio especially when the latter incorporated high heating rate volatile matter content, but for unburnt carbon there was no similar relationship. The dominant mechanism for NOx formation in deeply staged combustion is understood to be via oxidation of the residual nitrogen in the char and would suggest that, because of the difference in the rate and quantity of volatile (and hence nitrogen) release at the high heating rates experienced in a furnace, the incorporation of proximate volatile matter in the fuel ratio will result in some inaccuracy.

Coals expected to give high unburnt carbon figures, i.e. the higher rank coals such as ATC and Shoal Creek, did, but there was no pattern that could substantiate anything but the weakest correlation.

The DTF results indicated an approximately constant conversion rate of 45% of the char bound nitrogen to NO for the suite of coals tested. It was also observed that the actual NO measured during the char refiring stage on the DTF correlated reasonably well with NO measured on the CTF, even though the partitioning of nitrogen between the volatiles and the char varied by a factor of two. This tends to confirm that volatile nitrogen does indeed play only a very minor role in the formation of NOx in staged combustion systems. The conclusion is that once calibrated against a combustion system (either full or small scale) to establish a coefficient linking the measured NO on that system with the char bound nitrogen a reasonably accurate prediction for the NOx formation propensity of different coals will be possible.

The situation with respect to LOI prediction from DTF measurements is also quite encouraging in that a reasonable correlation has been observed with final unburnt carbon figures from the CTF. However, the limited range of coal qualities tested makes the correlation too reliant on a single coal with poor burnout characteristics and more data are required.

One of the most encouraging findings of the current work has been the good correlation found between the R_{N190} number generated on the Nottingham University automated image analysis technique and unburnt carbon. It would appear that the level of unreactive material in the coal that this index represents also correlates reasonably well with the morphology of the chars, thereby substantiating the observation that the intrinsic reactivity of fully developed chars is constant, i.e. coal quality independent.

Conclusions

1. The propensity to form NOx in a utility boiler is dominated by plant factors, whereas the levels of unburnt carbon are controlled by coal quality considerations.

2. For the prediction of NOx emission levels from different coals, the use of a calibrated pilot scale combustor is the preferred option currently. However, the residual nitrogen in chars formed on a Drop Tube Furnace which must also be calibrated with the combustion system concerned shows some promise. There is a need for the correlations already established to be tested on a wider range of coals.

3. For the prediction of unburnt carbon there is less need to resort to expensive pilot scale testing. The Drop Tube Furnace has been shown to provide a reasonable indication of burnout

propensity, but automated image analysis of the pattern of reflectance from coal samples is beginning to show promise as a simpler method of predicting coal reactivity without the need to heat the coal samples. Again, the range of coals needs to be expanded before confidence in the concept can be further substantiated.

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Table 1

Composite PF Sample Analyses

Coal Date	Ashland		El Cerrejon		Tyne Blend		El Cerrejon/Tyne	
	6/10/93	7/10/93	12/10/93	25/10/93	20/10/93	21/10/93	22/11/93	23/11/93
Moisture, % Total	2.0	2.1	2.6	2.8	2.4	2.1	2.9	2.8
Ash, %ar	13.2	13.6	10.4	10.3	14.5	14.3	13.3	12.6
%dry	13.5	13.9	10.7	10.6	14.9	14.6	13.7	13.0
Volatile Matter %ar	31.4	31.3	33.8	35.1	30.1	29.9	31.4	31.7
CV, kJ/kg ar Gross	29,200	29,100	29,510	29,180	28,630	28,850	28,630	28,950
Nett	28,165	28,178	28,367	28,065	27,580	27,845	27,570	27,876
Sulphur, %ar	0.92	0.86	0.81	0.81	1.50	1.46	1.05	1,03
Chlorine, %ar	0.15	0.10	0.05	0.04	0.31	0.31	0.14	0.14
Carbon, %DAF	83.66	84.60	82.62	80.97	83.60	84.65	82.28	82.69
Hydrogen, %DAF	5.27	4.65	5.68	5.51	5.39	5.15	5.35	5.40
Nitrogen, %DAF	1.71	1.61	1.68	1.69	1.81	1.78	1.69	1.70
Oxygen, %DAF (Diff)	8.10	8.00	9.03	10.85	7.02	6.30	9.26	8.83
Volatile Matter, %DAF	37.0	37.1	38.9	40.4	36.2	35.8	37.5	37.5
CV, kJ/kg DAF Gross	34,434	34,520	33,920	33,579	34,452	34,510	34,165	34,220
Fuel Ratio	1.70	1.69	1.57	1.48	1.76	1.80	1.67	1.67
SiO2	58.74	58.99	64.30	62.03	50.30	51.17	57.71	57.48
Al2O3	27.41	27.93	20.92	19.38	27.83	27.93	25.04	24.82
Fe2O3	7.39	6.55	7.63	10.01	11.25	11.28	9.05	9.56
CaO	1.70	1.35	2.21	3.25	4.24	3.62	2.76	2.90
MgO	0.96	0.95	1.44	1.99	1.74	1.43	1.42	1.46
K2O	2.29	2.48	2.05	2.03	1.92	1.98	2.03	2.00
Na2O		0.17	0.36	0.37	1.57	1.42	0.84	0.63
TiO2	1.51	1.57	1.09	0.95	1.15	1.16	1.16	1.15
BaO	0.13	0.12	0.11	0.11	0.12	0.16	0.17	0.17
Mn3O4			0.05	0.06	0.08	0.07	0.06	0.07
P2O5	0.19	0.17	0.14	0.15	0.24	0.21	0.25	0.26

Table 2

Composition of Test Coals

Coal	Tyne Blend	ATC	Drayton	Cerrejon	Ashland	Shoal Creek	North River	Leasing-thorne
Moisture, % Total	3.4	3.2	0.1	3.4	2.4	1.7	2.4	2.7
Ash, %ar	12.1	13.4	14.2	10.0	12.7	10.0	10.8	7.4
Volatile Matter, %ar	31.2	28.5	33.9	33.9	32.1	25.6	34.8	29.5
CV, kJ/kg ar Gross	28,960	28,100	29,300	29,670	29,460	31,550	29,640	30,960
Nett	27,908	27,124	28,259	28,539	28,387	30,523	28,583	29,896
Sulphur, %ar	1.18	0.68	0.76	1.06	0.82	0.86	2.09	1.30
Chlorine, %ar	0.27	0.07	0.04	0.14	0.12	0.06	0.01	0.04
Carbon, %DAF	83.70	83.95	83.08	83.05	85.49	88.02	81.87	85.36
Hydrogen, %DAF	5.21	4.86	5.49	5.54	5.43	5.10	5.25	5.11
Nitrogen, %DAF	1.76	2.01	1.76	1.70	1.62	1.90	1.70	1.71
Oxygen, %DAF (Diff)	7.61	8.28	8.74	8.32	6.35	3.94	8.76	6.33
Volatile Matter, %DAF	36.9	34.2	39.6	39.1	37.8	29.0	40.1	32.8
Fuel Ratio	1.71	1.93	1.53	1.55	1.64	2.45	1.49	2.05
CV, kJ/kg DAF Gross	34,272	33,693	34,189	34,261	34,700	35,730	34,147	34,438
Ash Comp								
Si02	51.33	50.67	68.71	60.61	61.36	51.81	44.10	50.94
Al203	27.69	30.82	19.02	22.16	27.10	29.32	26.35	27.99
Fe203	10.77	5.25	6.34	9.52	4.93	9.71	18.05	13.97
Ca0	4.14	7.94	2.53	1.91	1.09	2.59	5.88	2.29
Mg0	1.55	2.06	1.01	1.42	0.94	1.39	1.48	1.12
K20	2.07	1.02	0.78	2.37	2.67	2.32	2.15	2.17
Na20	1.24	0.39	0.40	0.94	0.26	1.18	0.62	0.31
Ti02	1.21	1.85	1.21	1.07	1.71	1.66	1.37	1.21
Ba0	0.13	0.30	0.14	0.11	0.18	0.30	0.22	0.18
Mn304	0.08	0.06	0.05	0.05	0.07	0.05	0.05	0.11
P205	0.21	1.56	0.90	0.19	0.22	0.69	0.74	0.24
Size, %-75 μm	66.1	87.8	67.0	66.3	71.6	73.2	75.8	65.1

Table 3 **Summary of NO Measurement results, vpm Dry (Corrected to 3% Oxygen Dry)**

OFA %	O_2 % Dry	Tyne (TY)	ATC (AT)	Drayton (DR)	El Cerrejon (EC)	Ashland (AS)	Shoal Creek (SC)	North River (NR)	Leasing -thorne (LS)	Mean	Min	Max
0	1.0	220	260	195	225	215	240	205	230	224	190	260
	3.0	325	355	305	315	300	340	290	360	324	285	360
	4.5	410	420	380	370	360	410	360	440	394	360	440
	SLOPE	54	46	53	41	41	48	46	60	49	41	60
15	1.0	165	205	205	180	180	220	160	175	186	170	220
	3.0	250	290	250	235	230	285	235	255	254	235	290
	4.5	305	360	280	280	265	330	290	320	304	265	360
	SLOPE	40	42	22	30	24	31	37	41	33	22	42
25	1.0		200		165	175	195	155	160	175	165	200
	3.0		255	235	215	215	260	210	225	231	210	260
	4.5		300	270	255	245	315	245	285	274	245	315
	SLOPE		29	23	27	20	34	26	36	28	20	36

Note: SLOPE is in vpm per percent oxygen derived from best fit straight line

Table 4 **Summary of Unburnt Carbon Results**

OFA %	O_2 % Dry	TYNE	ATC	DRAYTON	EL CERREJ	ASHLAND	SHOAL CREEK	NORTH RIVER	LEASING-THORNE
Loss-on-ignition at 800°C, % Dry									
0	1.0	5.2	10.5	7.0	7.0	4.9	9.7	8.0	8.4
	3.0	1.4	3.4	2.6	2.1	1.6	3.4	2.8	2.4
	4.5	0.8	2.9	1.4	1.7	0.8	1.5	1.9	1.3
15	1.0	3.5	14.6	4.0	7.5	5.0	10.2	4.0	11.3
	3.0	2.1	6.1	3.4	3.0	2.0	5.4	2.6	3.4
	4.5	1.5	3.7	3.1	2.3	1.6	2.6	1.8	2.3
25	1.0		8.5		6.7	5.1	10.6	5.3	13.5
	3.0	3.3	4.5	3.7	4.0	1.8	3.4	2.5	2.9
	4.5	2.0	3.7	3.0	2.5	1.7	2.2	2.5	1.7
Ash (%Dry)		12.5	13.8	14.2	10.4	13.0	10.2	11.1	7.6
Carbon Loss, % Dry									
0	1.0	0.65	1.45	0.99	0.73	0.64	0.99	0.89	0.64
	3.0	0.18	0.47	0.37	0.22	0.21	0.35	0.31	0.18
	4.5	0.10	0.40	0.20	0.18	0.10	0.15	0.21	0.10
15	1.0	0.44	2.01	0.57	0.78	0.65	1.04	0.44	0.86
	3.0	0.26	0.84	0.48	0.31	0.26	0.55	0.29	0.26
	4.5	0.19	0.51	0.44	0.24	0.21	0.27	0.20	0.17
25	1.0		1.17		0.70	0.66	1.08	0.59	1.03
	3.0	0.41	0.62	0.53	0.42	0.23	0.35	0.28	0.22
	4.5	0.25	0.51	0.43	0.26	0.22	0.22	0.28	0.13

Table 5 **A Summary of the Devolatilisation Data from the Drop Tube Furnace EPRI Contract Work**

Devolitisation at 1300°C, 200 ms & 1% O_2

Coal	Feed Size μm	% Nitrogen Content		Coal LOI or C+H+N		% Volatile Matter (Dry Basis) Coal	%VMhhr (Dry Basis) Char	"R" Factor	Combustibles Loss	Nitrogen Loss	gN/100g Coal	
		Coal	Char	Coal	Char						In Volatiles	In Char
ATC	53-75	1.76	1.98	88.25	78.43	29.98	46	1.52	0.52	0.39	0.68	1.08
	106-125	1.81	2.02	89.02	80.63	31.34	43	1.38	0.49	0.37	0.66	1.15
Tyne Blend	53-75	1.62	2.15	91.25	82.13	34.21	51	1.49	0.56	0.35	0.57	1.05
	106-125	1.65	1.97	92.05	81.66	34.36	57	1.65	0.62	0.48	0.80	0.85
El Cerrejon	53-75	1.54	1.91	93.20	82.68	37.11	61	1.64	0.65	0.51	0.79	0.75
	106-125	1.59	1.92	95.47	90.88	38.43	50	1.31	0.53	0.40	0.63	0.95
Leasingthorne	53-75	1.64	2.12	94.30	88.52	30.79	50	1.64	0.53	0.36	0.58	1.05
	106-125	1.73	2.07	95.87	92.13	32.83	48	1.45	0.50	0.37	0.64	1.09
Ashland	53-75	1.43	1.52	88.25	70.03	34.00	61	1.79	0.69	0.58	0.84	0.59
	106-125	1.45	1.74	91.64	79.47	36.59	59	1.62	0.65	0.51	0.74	0.71
Shoal Creek	53-75	1.73	1.82	90.56	82.43	26.54	46	1.74	0.51	0.43	0.75	0.98
	106-125	1.65	1.74	89.83	82.70	26.62	41	1.55	0.46	0.38	0.63	1.02
North River	53-75	1.68	1.91	89.05	73.80	36.31	58	1.60	0.65	0.52	0.88	0.80
	106-125	1.71	2.37	90.53	77.77	36.77	57	1.56	0.63	0.41	0.70	1.01
Drayton	53-75	1.47	1.62	84.85	68.98	33.20	51	1.54	0.60	0.46	0.68	0.79
	106-125	1.61	1.74	89.03	75.07	35.37	56	1.58	0.63	0.52	0.84	0.77

Table 6

A Summary of the Char Refiring Data for Drop Tube Furnace EPRI Contract Work
Chars refired at 1300°C, 5% O_2

Coal	Feed Size μm	Res Time ms	Average Measured NO mg/g coal	Char Conversion	Combustibles Loss	Nitrogen Loss	NO Produced if 100% Char Conversion mg/g coal	Average NO Produced if 100% Char Conversion mg/g coal	Average NO Produced if Char Conversion 100% (Normalised for Net CV of 29000 kJ/kg) mg/g coal
ATC	53-75	200	4.31	0.40	0.71	0.67	10.70	10.44	11.17
		400	7.16	0.69	0.85	0.80	10.41		
		600	8.63	0.84	0.92	0.91	10.21		
	106-125	200	4.19	0.32	0.65	0.60	13.11	12.49	13.35
		400	8.15	0.61	0.80	0.76	13.34		
		600	8 29	0.75	0.87	0.88	11.02		
Tyne Blend	53-75	200	7.30	0.78	0.90	0.87	9.37	8.85	9 19
		400	8 48	0.97	0.99	0.97	8.75		
		600	8.28	0.98	0.99	0.97	8.41		
	106-125	200	4.71	0.49	0.80	0.75	9.70	8.51	8 84
		400	6.79	0.84	0.94	0.92	8.07		
		600	7.06	0.91	0.97	0.95	7.75		
El Cerrejon	53-75	200	5.79	0.80	0.93	0.90	7.27	7.71	7.84
		400	7.54	0.96	0.99	0.97	7.88		
		600	7.87	0.98	0.99	0.99	7.99		
	106-125	200	5.63	0.60	0.81	0.76	9.32	10.55	10.72
		400	10.23	0.88	0.95	0.93	11.56		
		600	10.18	0.94	0.97	0.97	10.77		
Leasingthorne	53-75	200	6.19	0.69	0.86	0.83	8.94	10.10	9.80
		400	9.55	0.92	0.96	0.95	10.41		
		600	10.63	0.97	0.99	0.98	10.94		
	106-125	200	6.02	0.44	0.72	0.66	13.63	12.17	11.81
		400	8.74	0.76	0.88	0.85	11.49		
		600	10 13	0.89	0.94	0.92	11.39		
Ashland	53-75	200	4.86	0.55	0.86	0.75	8.83	6.92	7.07
		400	5.54	0.94	0.98	0.94	5.89		
		600	5.91	0.98	0 99	0.96	6.03		
	106-125	200	5.24	0.63	0.87	0.82	8.31	7.24	7.40
		400	6.18	0.91	0.97	0.95	6.78		
		600	6.28	0.95	0.98	0.97	6.63		
Shoal Creek	53-75	200	9.38	0.78	0.89	0.89	11.98	11.32	10.76
		400	10.09	0.94	0.97	0.95	10.73		
		600	10.88	0.97	0.98	0.98	11.25		
	106-125	200	7.88	0.56	0.76	0.72	14.10	12.10	11.50
		400	9.34	0.82	0.90	0.91	11.40		
		600	9.31	0.86	0.92	0.92	10.81		
North River	53-75	200	5.14	0 81	0.94	0.92	6.33	7.00	7.10
		400	6.66	0.94	0.98	0.95	7.09		
		600	7.30	0.96	0.99	0.97	7.57		
	106-125	200	5.22	0.67	0.88	0.82	7.85	7.80	7.92
		400	6.37	0.86	0.95		7.44		
		600	7.60	0.94	0.98	0.97	8.12		
Drayton	53-75	200	5.04	0.65	0.86	0.81	7.74	8.82	9.05
		200	5.12	0.62	0.85	0.80	8.31		
		400	7.50	0.87	0.95	0.91	8.63		
		400	7.67	0.83	0.93	0.89	9.21		
		400	7.82	0.85	0.94	0.91	9.22		
		400	8.24	0.85	0.94	0.91	9.70		
		600	8.28	0.94	0.98	0.95	8.84		
		600	8.47	0.95	0.98	0.96	8.93		
	106-125	200	3 37	0.46	0.80	0.75	7.26	8.97	9.20
		200	3.57	0.40	0.78	0.73	8.93		
		400	6.73	0.75	0.91	0.89	9.01		
		400	6 99	0.74	0.90	0.88	9.44		
		600	8.16	0.85	0.95	0.93	9.56		
		600	8.31	0.87	0.95	0.94	9.61		

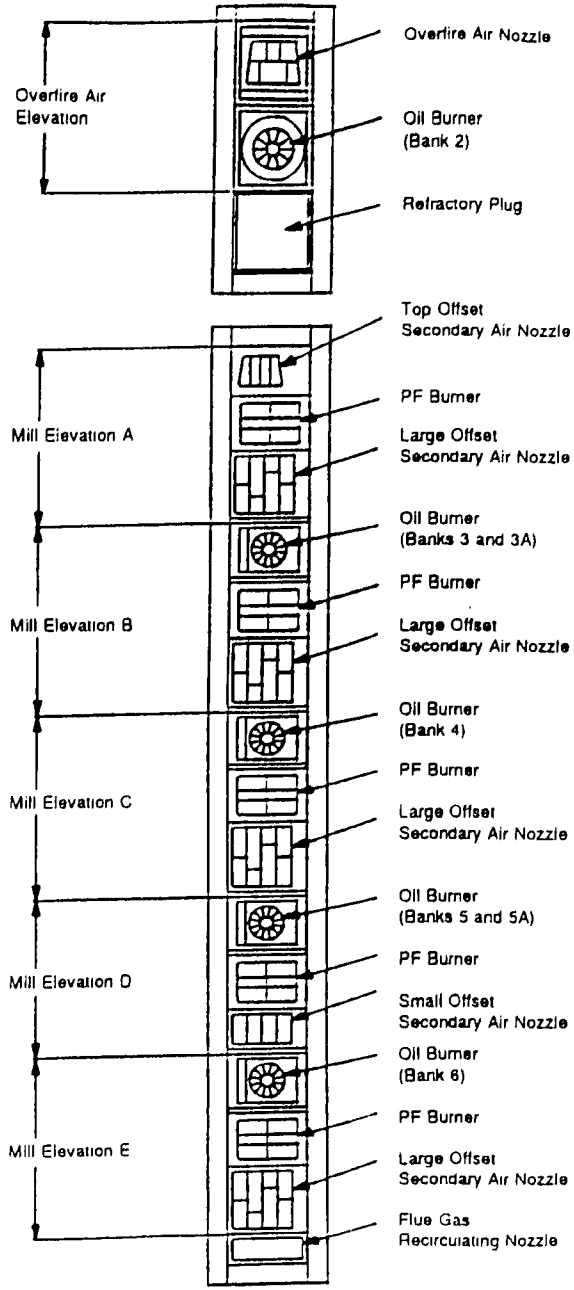


Figure 1 Kingsnorth Burner Arrangement

Figure 2 **KINGSNORTH PS: the Relationship between Carbon-in-Ash and Excess O2**

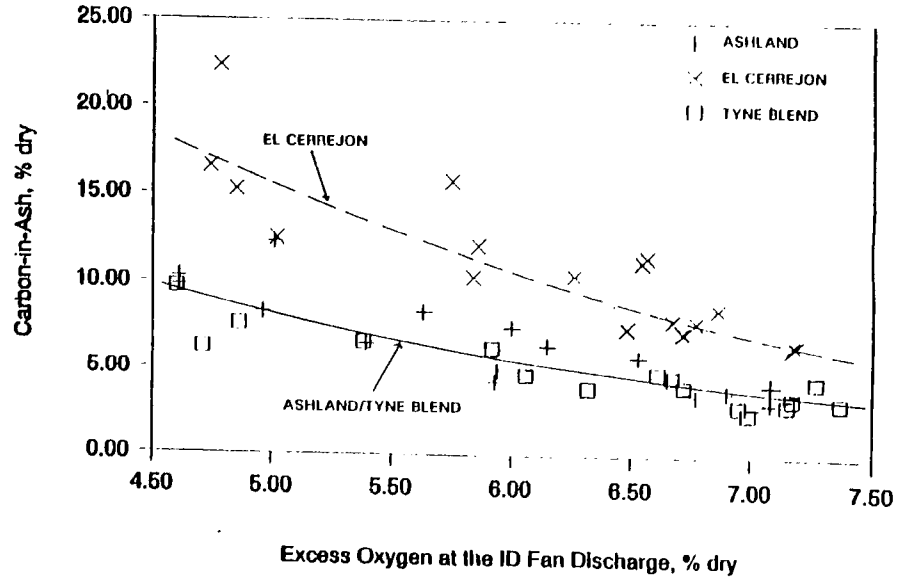


Figure 3 **KINGSNORTH PS: the Relationship between Carbon Loss and Excess Oxygen**

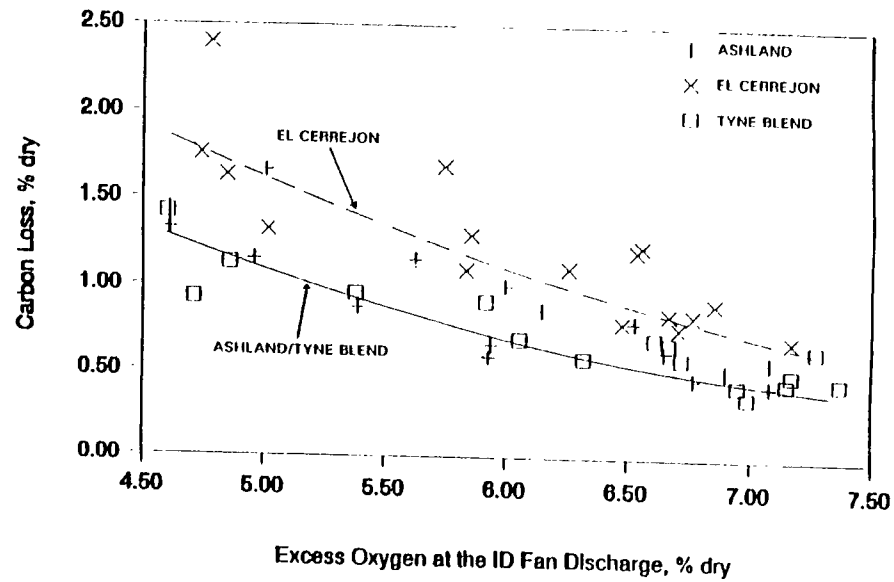


Figure 4 **KINGSNORTH PS: the Relationship between NO Emissions and Excess O2**
A MILL IN SERVICE

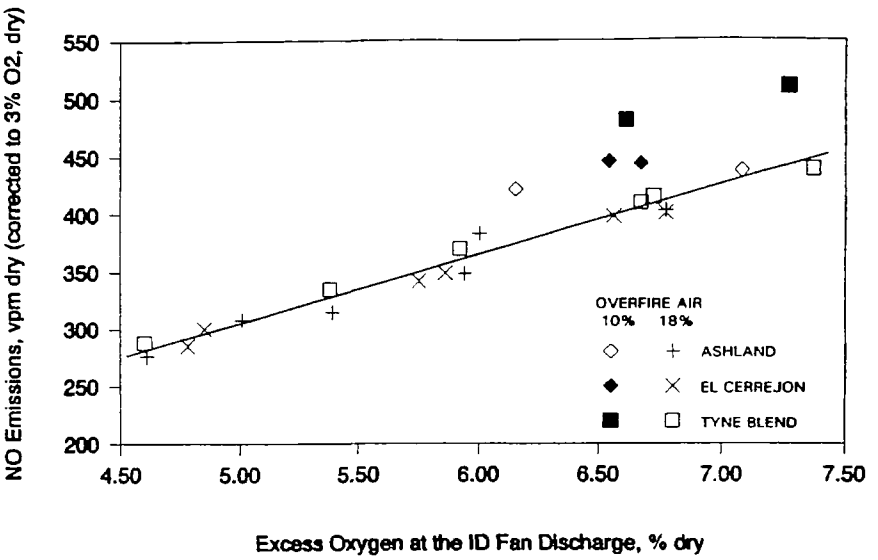


Figure 5 **KINGSNORTH PS: the Relationship between NO Emissions and Excess O2**
A MILL OUT OF SERVICE

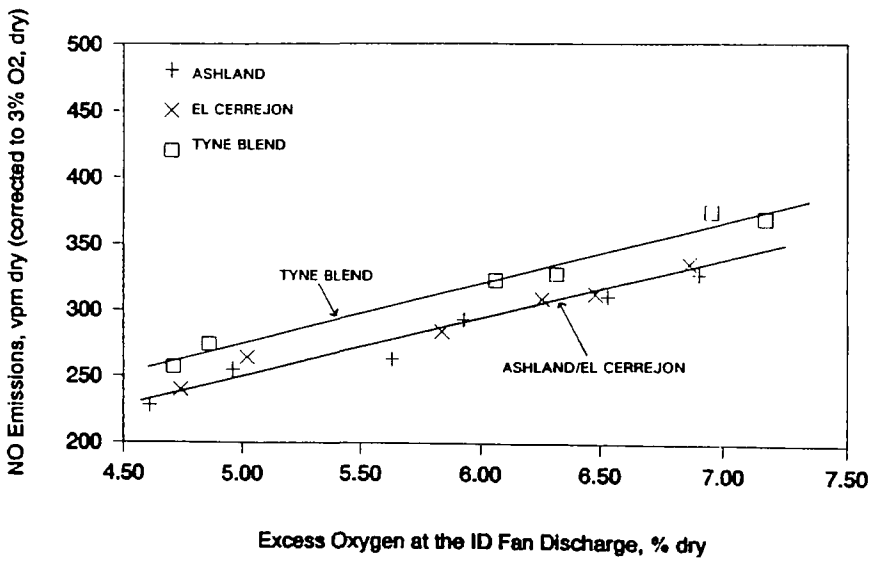


Figure 6 **KINGSNORTH PS NO Emissions: Comparison with Result for a S.African Coal**
A MILL IN SERVICE

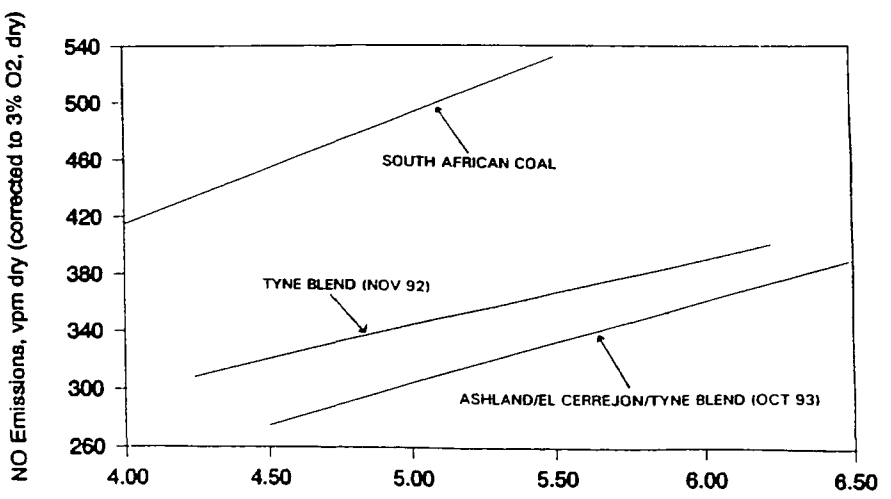


Figure 7 COMBUSTION TEST FACILITY

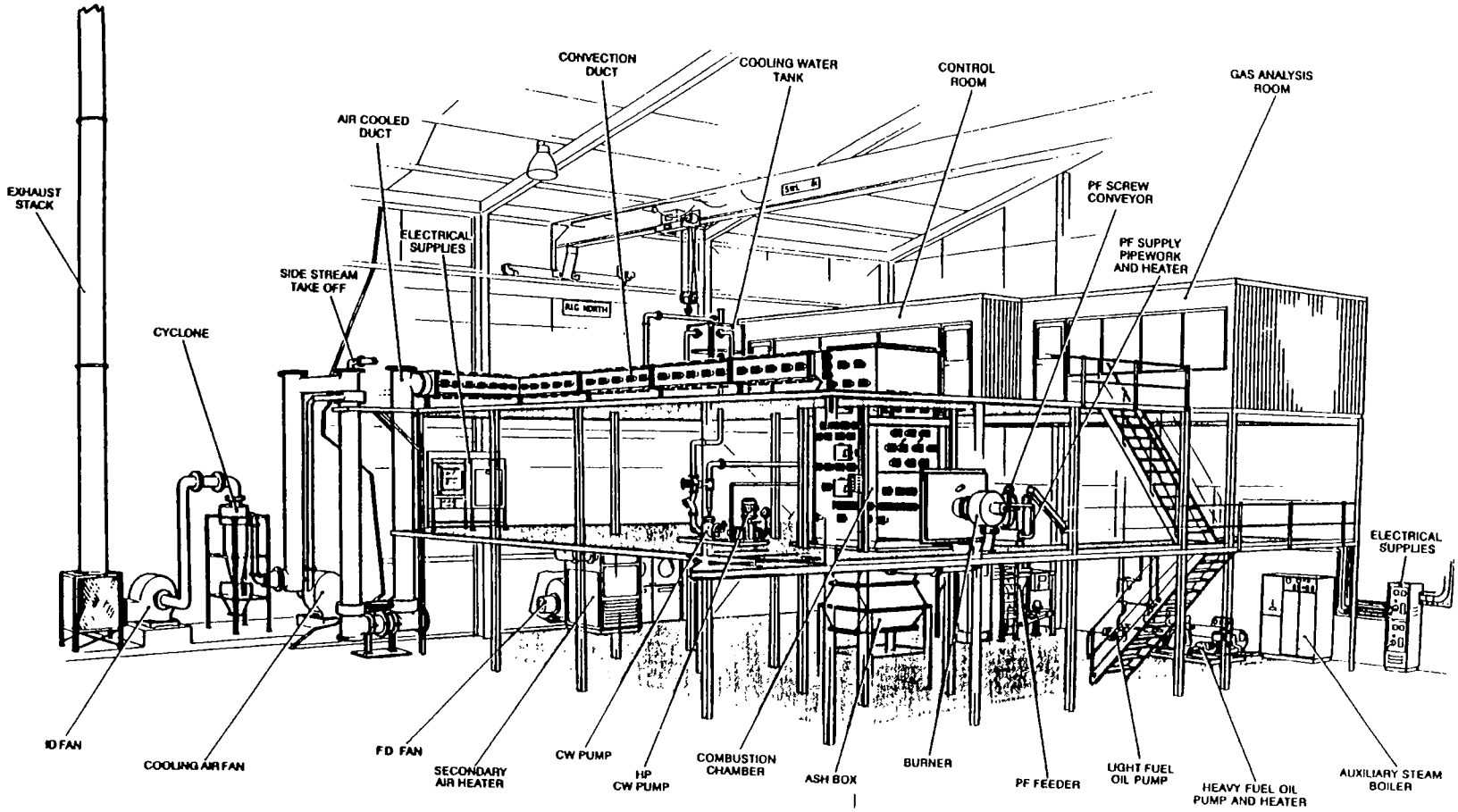


FIGURE 8 NO CONCENTRATION V. EXCESS OXYGEN
0% OVERFIRE AIR

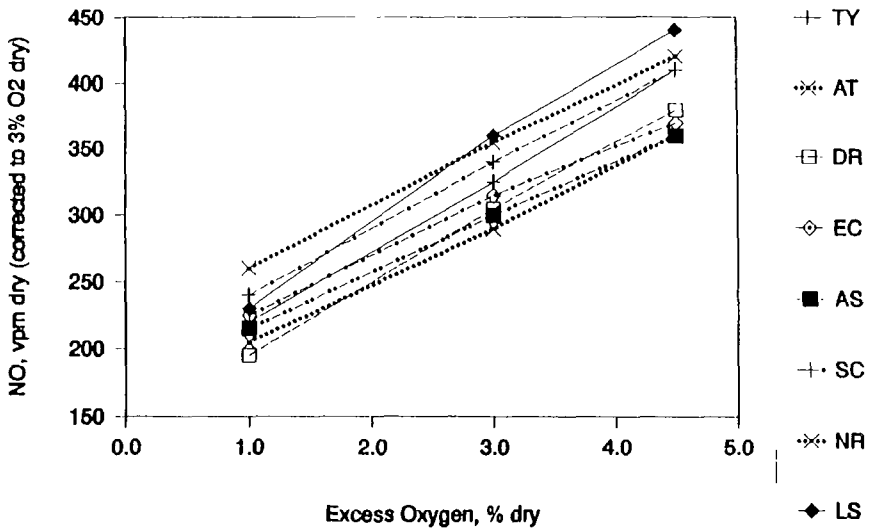


FIGURE 9 NO CONCENTRATION V. EXCESS OXYGEN
15% OVERFIRE AIR

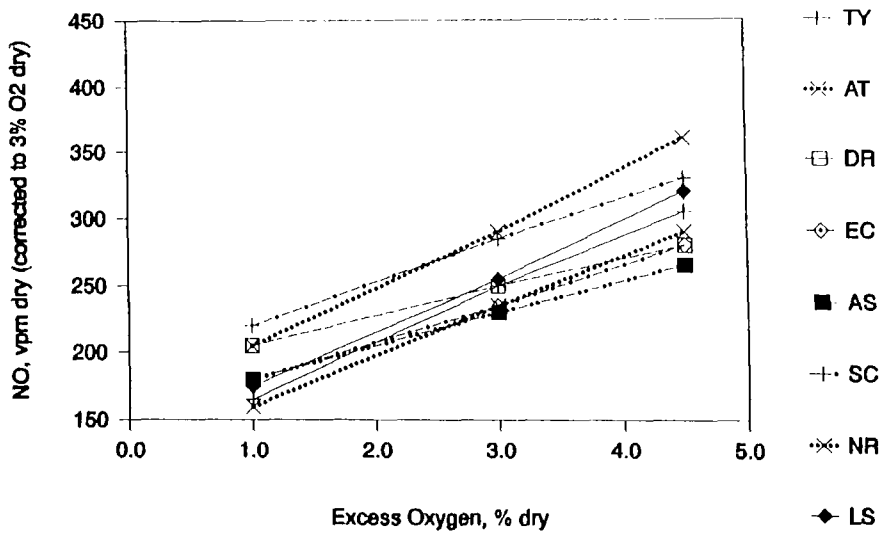


Figure 10 The Relationship Between NO Emissions and Fuel Ratio (VM) x Fuel Nitrogen

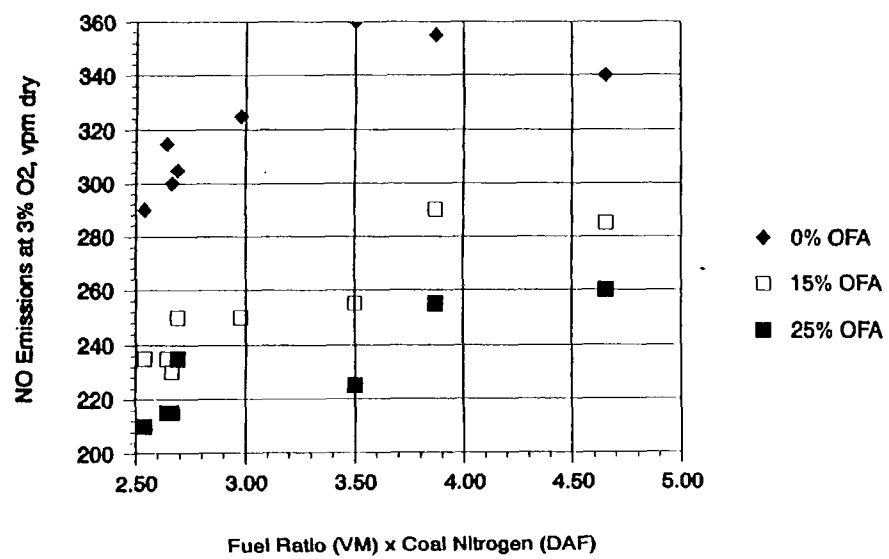
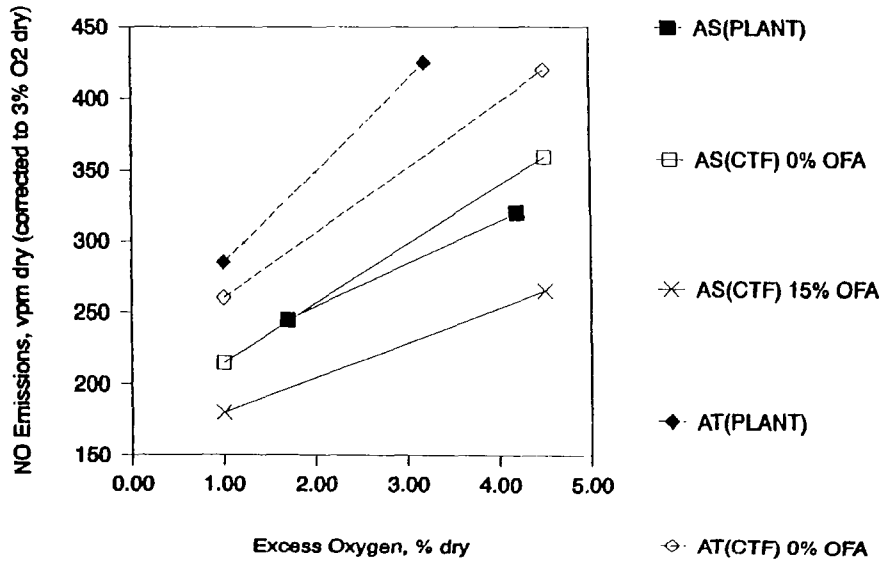
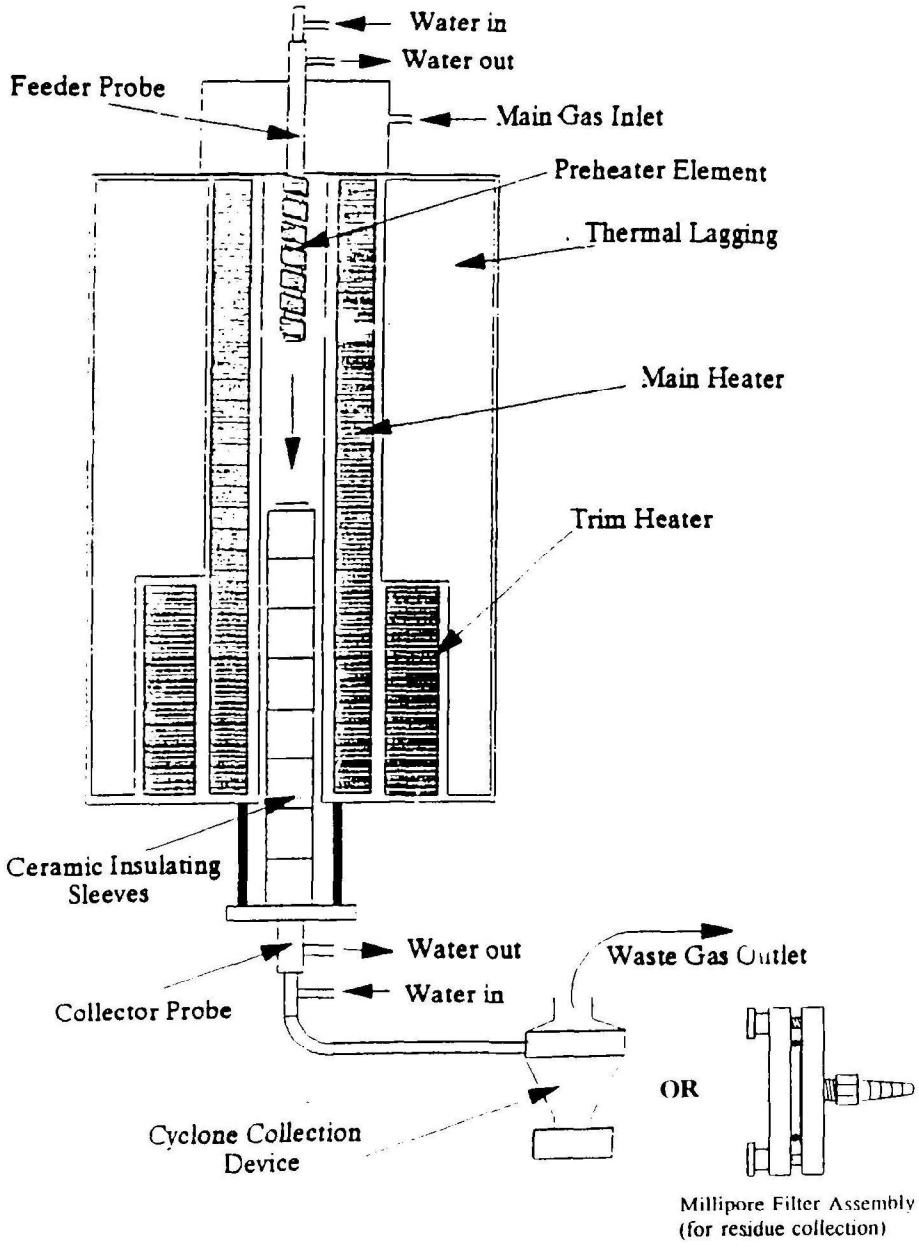


FIGURE 11 NO Emissions: Comparison of Plant and CTF Results





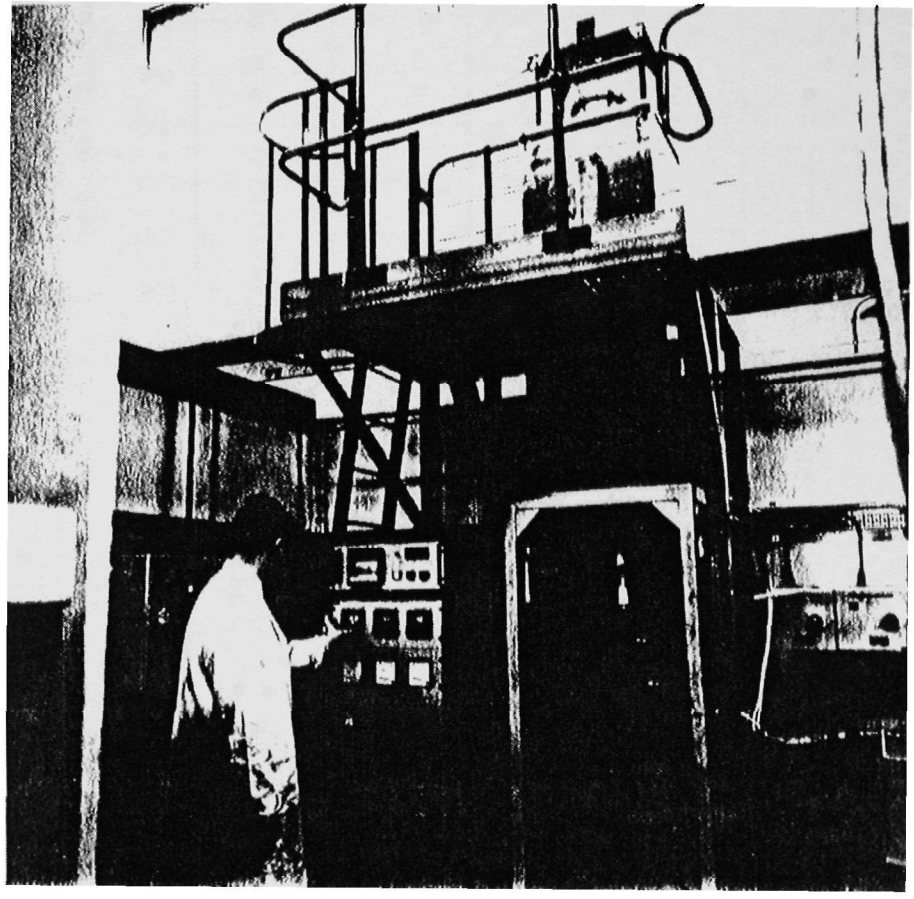


Figure 12 Schematic and Photographic Representation of the Drop Tube Furnace

Figure 13

Comparison Between Fractional Combustibles Loss and Fractional Nitrogen Loss.
(Chars produced @ 1300oC, 1%O2 & 200ms, refired at 1300oC, 5%O2 & 200,400 & 600ms)

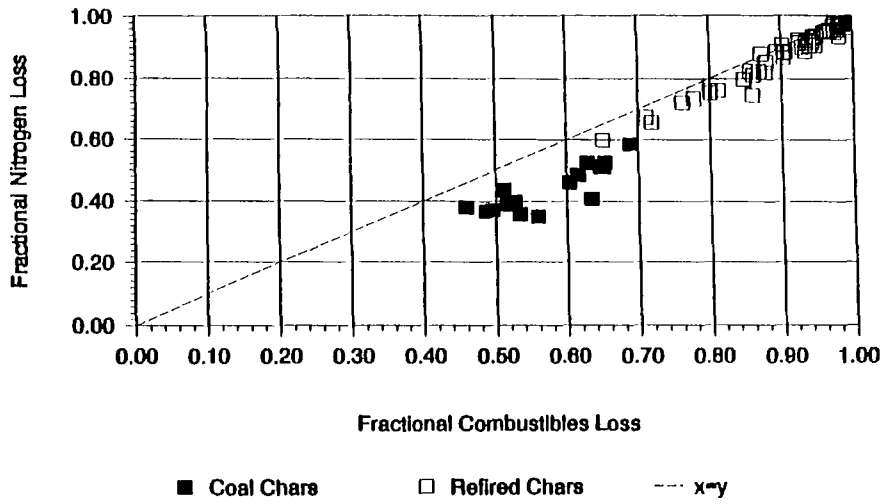


Figure 14 **Fractional Nitrogen Loss as a Function of Residence Time Chars Refired at 1300oC and 5%O2.**

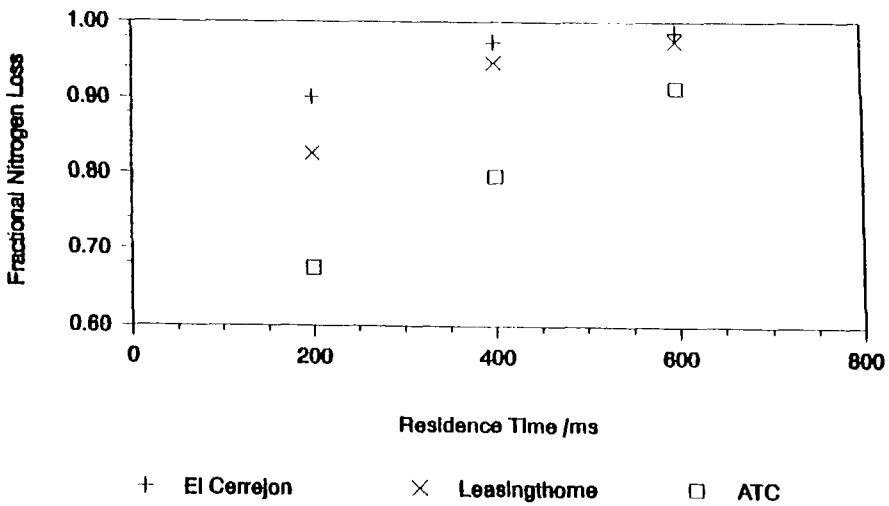


Figure 15 **DTF - NO Produced as a Function of Predicted NO.**
(Devolatilised Chars Refired at 1300oC, 5%O2 & 200, 400 & 600ms)

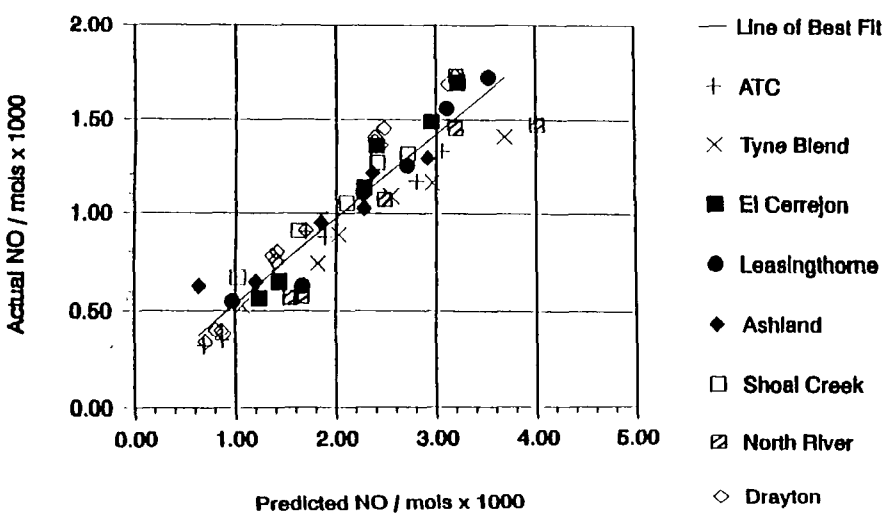


Figure 16 The Relationship Between NO Emissions and Fuel Ratio (VMhhr) x Fuel Nitrogen

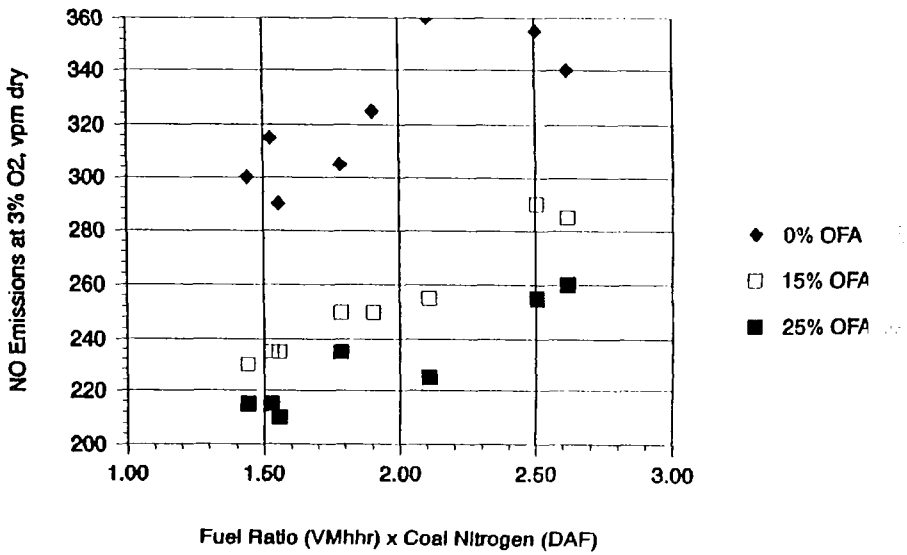


Figure 17 **CTF NOx as a Function of Char Nitrogen / 100g Coal**

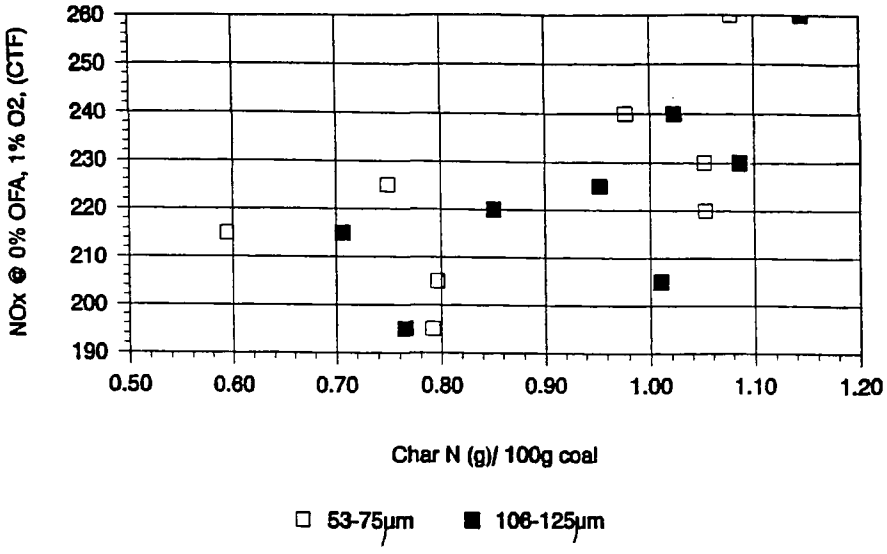
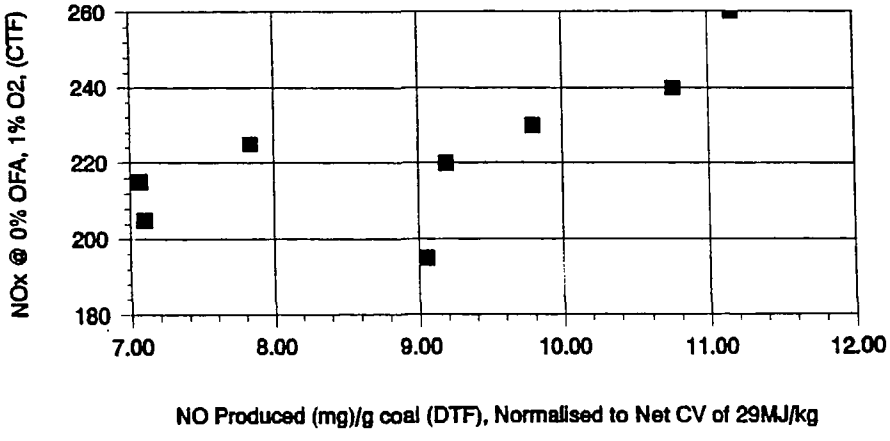


Figure 18 **CTF NOx as a Function of NO Produced (mg) / g Coal (DTF)**
Assuming 100% Char conversion.



Devolatilisation @ 1300oC, 200ms & 1%O2. Char refiring @ 1300oC, 200, 400 & 600ms & 5%O2. 53-75µm original coal feed.

Figure 19 **A Comparison Between DTF and CTF LOIs**

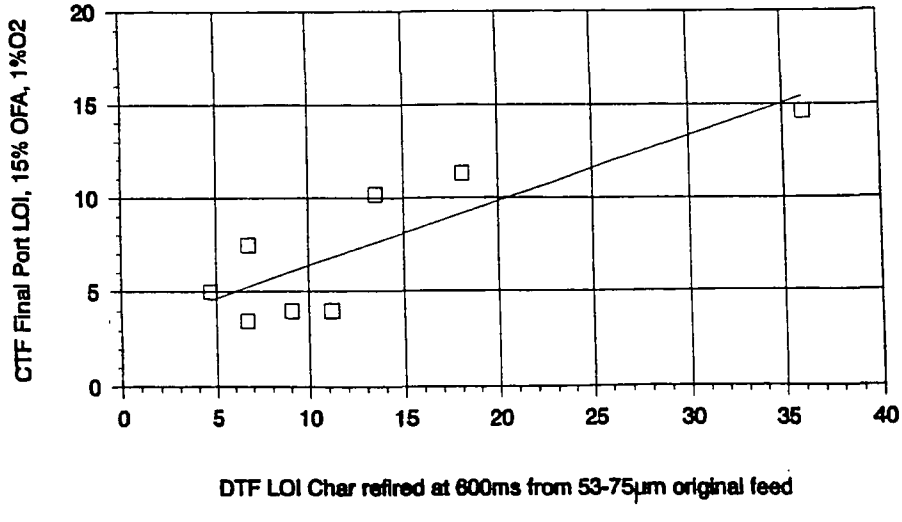
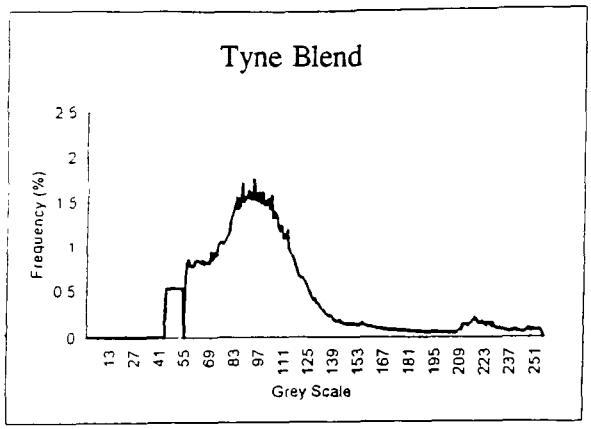
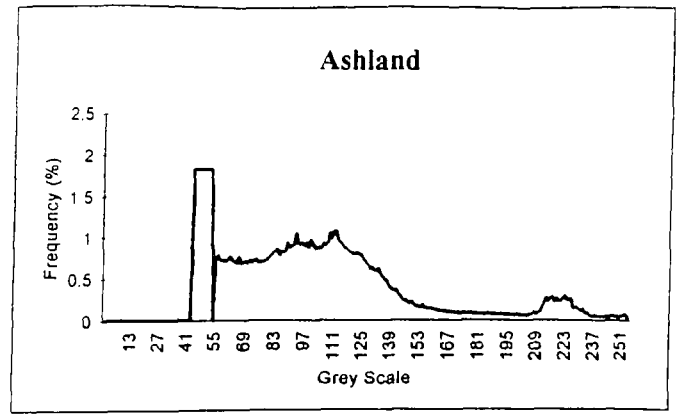
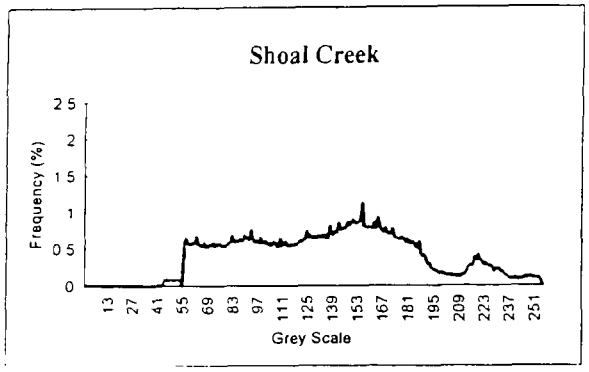


Figure 20 Grey Scale Histograms







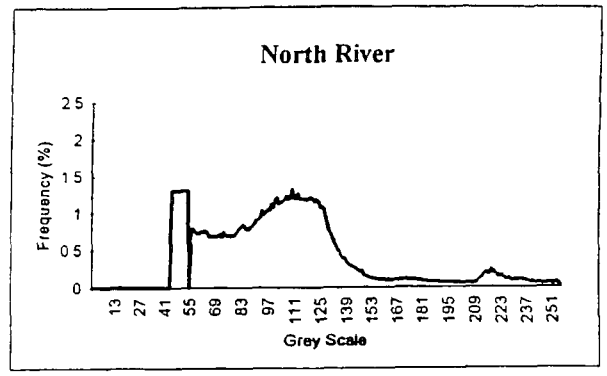
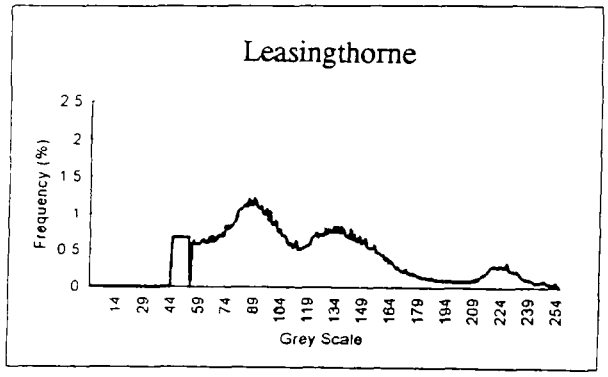
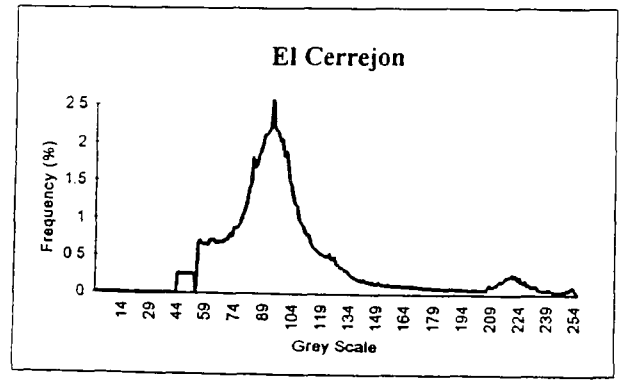
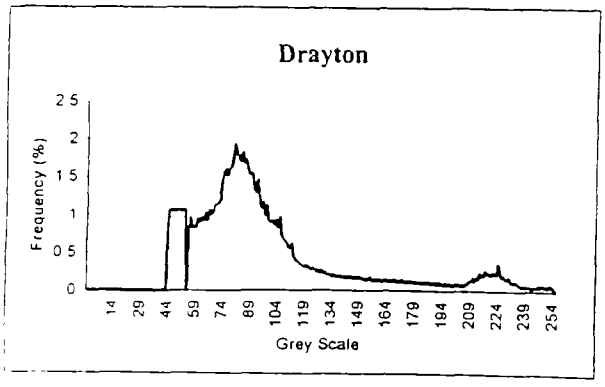


Figure 21 Grey Scale Histograms







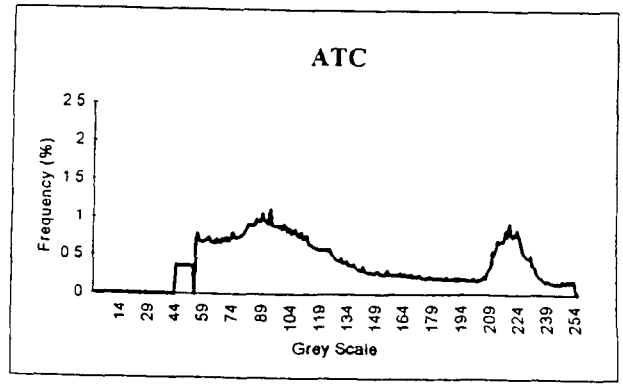
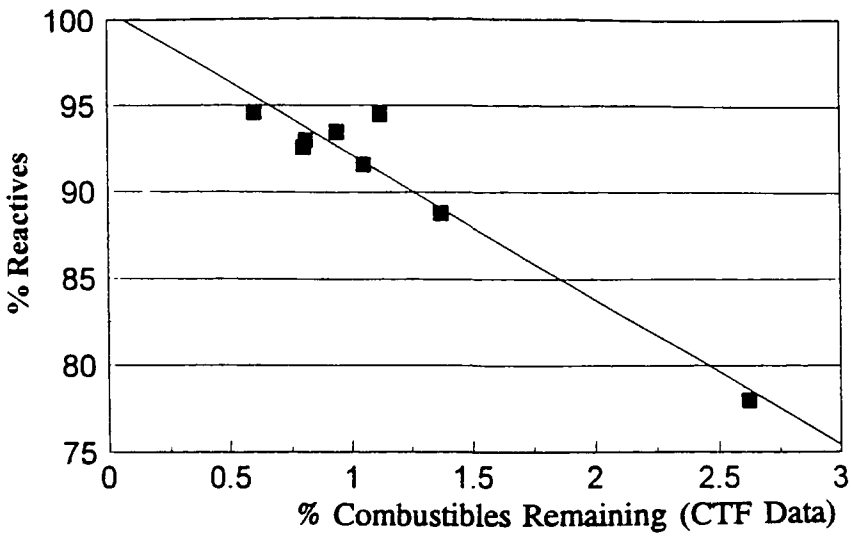
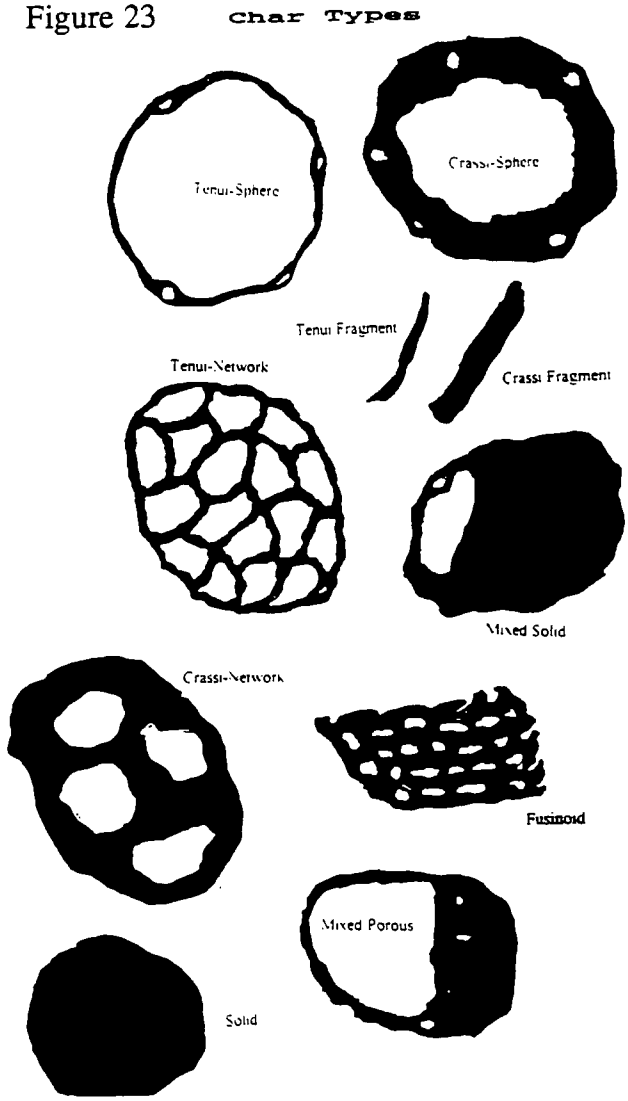


Figure 22 Plot Of % Combustibles Remaining (CTF Data) vs 190 Reactivity



Reactivity is % less than 190 on coal grey-scale
R-square value is 0.93

Figure 23 char Types



COMPARISON OF PILOT-SCALE FURNACE EXPERIMENTS AND PREDICTIONS TO FULL-SCALE BOILER PERFORMANCE OF COMPLIANCE COALS

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Abstract

A series of compliance coals have been fired in the Southern Company Services and Southern Research Institute pilot-scale Combustion Research Facility, with the goal of predicting NO_x emissions, unburned carbon levels, and other operating parameters. The research was financed by a tailored collaboration between Alabama Power and the Electric Power Research Institute. The coals included a South American coal, a Powder River Basin coal, and several local Alabama bituminous coals. They were fired in conventional and low NO_x firing modes, where some of the coals were fired in a tangential-fired simulation, others as wall-fired, and some in both types of firing. Two of Alabama Power's boilers were used as the plants simulated in the pilot-scale experiments, along with a Gulf Power boiler. The results of the NO_x emissions and unburned carbon from the pilot furnace testing are presented and compared to the limited full-scale boiler data available on these coals.

Introduction

Most utilities are pursuing a strategy of fuel switching to low sulfur coal, in order to meet the sulfur dioxide (SO_2) emission limits of the first phase of the 1990 Clean Air Act Amendments. However, since this legislation also requires reductions in the level of emissions of nitrogen oxides (NO_x), utilities must predict the ability of alternative low-sulfur coals to meet emission limits on NO_x, along with the ease of boiler operation of these low sulfur coals. Typical strategies employed to meet NO_x emission limitations, installation of low NO_x burners, are usually designed for a particular coal or coal type, and may not perform well on other fuels.

Alabama Power, in a tailored-collaboration with the Electric Power Research Institute, has funded a research program to predict the performance of certain low sulfur coals in their full-scale plants with low NO_x combustion modifications. This work is focused mainly on NO_x emissions and unburned carbon. Since most of the full-scale units that will burn low sulfur coal did not have low NO_x firing modifications at the time this work was started, it was not possible to test burn these coals in the actual boilers. The predictive ability of the pilot-scale facility for the important issues of NO_x emissions and unburned carbon was to be used to aid in the coal acquisition process. Also, any insight from the pilot-scale testing on boiler operations affected by coal quality, such as milling capacity, slagging, fouling, and particulate control, would also assist in coal procurement.

This paper reports some of the results of this research program. The Alabama Power plants simulated in this study are Unit 5 of Plant Gaston (880 MW CE tangentially-fired, retrofitted with LNCFS II) and Unit 3 of Plant Miller (660 MW B&W w/ dual-register burners), along with Plant Crist Unit 7 (500 MW Foster-Wheeler w/ F-W dual-register burners) belonging to Gulf Power. Baseline unit operating data have been collected at each of the three plants and it was used to help in the setup of the simulations. In the case of Gaston Unit 5, the field data collection effort of the DOE and EPRI Coal Quality Expert (CQE) during September 1991 was available to help in evaluating the pilot-scale simulations.

Description of the Pilot-Scale Research Facility

The Southern Company and Southern Research Institute Combustion Research Facility is located on the Birmingham, Alabama, campus of Southern Research Institute (SRI). The facility was designed and constructed by SRI under contract to Southern Company Services (SCS). SCS and SRI jointly own the facility, and SRI operates the facility under contract to SCS. However, SRI can also perform confidential research for outside parties, including other utilities, with the payment of a facility usage fee to SCS. The facility is designed for up to six million Btu per hour firing on natural gas or coal, which is equivalent to 1.75 MW thermal or about 0.6 MW electric.

The design of the facility was carefully chosen to provide a close simulation of the physical processes that occur in a full-scale utility boiler. The facility, shown in Figures 1 and 2, consists of a coal crushing and milling area, a coal feeding system, a vertical refractory-

lined furnace, a single up-fired burner, a horizontal convective section pass with air-cooled tube banks, a series of heat exchangers, an electrostatic precipitator, a pulse jet baghouse, and a packed column scrubber.

Coal Preparation

The coal handling part of the facility is shown in Figure 1. The coal preparation area includes covered on-site storage bins, a rotary drum coal crusher, a CE Raymond bowl mill, and pulverized coal storage. Eight storage bins, each capable of holding up to 25 tons of coal, are located adjacent to the crusher. As-mined coal is loaded into the crusher with a small front-end loader and is crushed to a size of minus 3/8 inch. From the crusher, the coal is pneumatically transported to a storage bin located above the coal mill. The coal mill is a refurbished 1937-vintage CE bowl mill, which has a rated capacity of 2 tons per hour. This type of mill should give representative milling simulations of the different air-swept table and roller mills normally used in power plant service. The mill air is preheated with a dedicated natural gas burner, which also helps inert the mill. The pulverized coal is captured in a pulse-jet baghouse and transported to an inerted pulverized fuel storage hopper. In such small-scale pulverized coal firing, it is normal practice to use indirect firing to help eliminate coal surges which can cause flame stability problems. The mill circuit can also be used to perform stand-alone milling experiments.

Coal Feeding

The coal feeding system is designed to deliver a constant feed of pulverized coal to the burner. The storage hopper is equipped with an orbital-motion live bottom to prevent bridging or rat-holing of the fuel. An Acrison loss-in-weight auger-type feeder weighs a controlled amount of fuel into the primary air through a rotary airlock valve. This gravimetric feeder is suspended on a weighing mechanism, and the feed rate is controlled by varying the auger speed to maintain the desired fuel feed rate.

Radiant Furnace

The furnace is a vertical, up-fired 28 foot high cylinder, with an inner diameter of 3.5 feet. (See Figure 2). The body of the furnace is built of seven 4 foot tall sections, each being a water-cooled jacket with 4 inches of refractory lining cast on the fireside. This refractory lining is used to limit the heat extraction and to ensure the proper simulation of the radiation environment found inside full-scale furnaces. Small furnaces have much higher ratios of inside surface area to total volume than full-scale furnaces, and the flame would be quenched if the entire interior was lined with heat exchange surface. The furnace diameter and height were chosen to best match the velocities and residence times found in full-scale units, with a velocity of 20 feet per second and a residence time to the furnace exit of 1.2 seconds chosen, at full firing rate.

Burner Assembly

The burner is mounted coaxially on the bottom of the furnace and is up-fired using natural gas, pulverized coal, or any combination of the two. It is equipped with a flow control system for secondary air flow and a set of registers which impart swirl to the secondary air, separate from the flow control. The secondary air and the primary air-coal mixture enter the furnace through a refractory quarl with a 25° half angle. Two cleanout ports are provided in this section, to allow bottom ash to be periodically removed from the furnace. A closed-circuit television camera with a control-room monitor allows constant monitoring of the view of the flame from the top of the furnace. A low NO_x firing system, consisting of the a generic dual-register burner and an overfire air system, can be installed to simulate several combinations of low NO_x firing.

Convective Sections

The combustion gases exit the vertical furnace through a horizontal convection pass which is designed to remove a substantial part of the heat from the flue gases. the extraction of heat is designed in order to simulate the time-temperature profile found in a utility boiler. A series of three air-cooled tube banks are installed in the convective pass, and the air-cooling is used to control either the temperature profile of the flue gases or the tube metal surface temperatures for fouling/ash deposition studies. A cross-flow tubular air preheater follows the convective tube banks and is used to preheat the primary and secondary air. Finally, two air-to-flue-gas recuperators are used to cool the flue gas down to a nominal 300°F before the flue gas enters the pollution control devices, which are described in a following section.

Computer Data Acquisition and Control System

The facility is controlled and monitored by a combined process control and data acquisition computer, a Yokogawa Yewpack system. This system performs all the process control and allows complex feed-forward and calculated variable control. This computer control also performs the safety monitoring needed for safe operation of combustion equipment, including flame scanning and interlocks. The Yokogawa computer also has the software to allow automatic startup and shutdown of the entire facility. Process data acquisition and storage is accomplished by sending data from the Yokogawa computer to an IBM-compatible computer, which stores the process data and allows on-line and historical data examination.

CEM System

A complete extractive continuous emissions monitoring (CEM) system is installed in the facility, and it is also interfaced to the computer control system. A set of gas analyzers, which analyze the flue gas for concentrations of O_2, NO_x, SO_2, CO_2, and CO, receives the dry flue gas sampled from a set of three extractive lines. Flue gas is sampled from the facility stack, the ESP/baghouse inlet, and a multi-purpose spare line. The flue gas is dried before the analyzers by a sample conditioning system which uses an ice bath to condense water from the sampled gases.

Pollution Control Equipment

The pollution control devices installed in the facility include an electrostatic precipitator, a pulse-jet baghouse, and a packed column scrubber. The pulse-jet baghouse and the scrubber were required for the air quality permit of the facility issued by the Jefferson County Board of Health, and therefore are always on-line. The ESP can be valved into the system before the pulse-jet baghouse to give a series ESP-baghouse combination.

Coals Studied

The suite of coals used in the study are listed, along with their analyses, in Table 1. The North River coal, which served as the baseline fuel for Gaston Unit 5, is the only non-compliance coal in the set. The other coals include local Alabama bituminous compliance coals, South American coal, and Powder River Basin coal. This set represents a wide range of coal type, especially regarding NO_x emission potential and unburned carbon. The Shoal Creek coal has a fuel ratio (proximate fixed carbon divided by volatiles) of 2.5, while the Belle Ayr coal has a fuel ratio of 1.2. The set also includes the Gusare coal from Venezuela, one of several regional coals notorious for producing large amounts of unburned carbon in low NO_x combustion.

Pilot-Scale Simulation of Full-Scale Boilers

Useful results from any small-scale testing requires the ability to predict performance in the full-scale equipment from the experimental results. The pilot-scale experiments must be designed to replicate the controlling mechanisms that occur in the real boiler. Fortunately, a great deal of previous work on scaling is available to guide the experiments. Because of the focus of this testing on the NO_x emissions and unburned carbon, the intensity of combustion in the radiant furnace of the full-scale equipment must be matched in the pilot furnace. In this work, the most common scaling parameter of combustion intensity, the volumetric heat release ratio (VHRR), from each full-scale plant was matched in the Combustion Research Facility. The VHRR is computed by dividing the fuel heat input, as Btu per hour, by the radiant furnace volume to yield Btu per hour per cubic foot.

The formation of NO_x from the combustion of a particular coal is influenced by the flame temperature, the amount of combustion air, the mixing of air with the fuel in the flame, and the flame structure. In performing pilot-scale simulations for NO_x emissions, each of these factors are taken into account. The flame temperature, and the subsequent furnace temperature profile, are fixed by the amount of heat removed from the furnace, given that the VHRR is matched. It is then a simple matter to match the amount of combustion air provided to the furnace to provide equivalent oxygen concentrations. Matching the mixing of fuel and combustion air requires a minimum size of the pilot furnace, greater than 3.5 million Btu per hour firing rate, due to the relative rates of macro- and micro-mixing. Beyond this limitation, it is impossible to reproduce the large-scale mixing patterns of a normal boiler in a research furnace, especially since the pilot furnace has only a single burner. However, the single burner can be configured to match the setting of an average

burner in a multi-burner wall-fired furnace. It is then assumed that inter-burner effects are small, and do not greatly influence the NO_x emissions. But any mal-distributions of coal or secondary air are also ignored. For tangentially-fired units, the overall furnace flame structure is simulated in the pilot furnace, by setting the single burner to produce a long, slowly swirling flame intended to represent the fireball found inside a single chamber of a tangentially-fired furnace. The diameter of the pilot furnace is so small that an attempt to use tangential fuel injection would result in coal particle deposition on the walls of the furnace.

The secondary scaling rules that are used include matching burner velocities, coal grind size, primary and secondary air temperatures, and the time-temperature history through the furnace. The burner is scaled to deliver the relative amounts of secondary and primary air, at the same velocities found in the full-scale burner. The time-temperature history of the pilot furnace should match the full-scale. For this experiment, it was necessary to replace some of the radiant furnace section refractory to maintain furnace exit temperatures, which is necessary to produce similar time-temperature histories. In general, it is possible to control heat extraction in a pilot furnace to match almost any time-temperature history.

Low NO_x firing simulation in the pilot furnace is achieved by the use of a generic dual register burner for wall-fired units and separated overfire air injection for tangentially-fired units. The dual-register burner can simulate the common aerodynamics of most commercial low NO_x burners. There are many different designs of fuel gun tip and coal spreader designs for commercial burners, which could be adapted to the pilot burner.

The final scaling rule used is the most useful and important. A baseline fuel, which has been fired in the full-scale boiler, is tested in the simulation of the pilot furnace, followed by the proposed compliance coal. Predictions of compliance fuel performance are then made relative to this baseline fuel testing. By using a relative measurement technique, not only the coal quality, but the NO_x emissions and unburned carbon, of the two fuels can be compared directly, while using the best possible pilot simulation of the full-scale furnace.

NO_x Emissions and Unburned Carbon Results

Tangential-Fired Boilers in Conventional Mode

Results of the testing are summarized in Table 2. The first seven test runs were conducted in a simulation of Alabama Power's Gaston Unit 5. Initially, the baseline fuel used in the previous CQE test program, North River, was burned in the Combustion Research Facility. The NO_x emissions and unburned carbon results from the full-scale testing and the CRF are displayed graphically in Figure 3. As can be seen in this figure, the pilot furnace gave lower NO_x emissions than that seen in the full-scale plant data, although the results are much closer at high furnace exit oxygen. For LOI, the bottom graph shows that the pilot furnace did an excellent job of simulating the LOI performance of Gaston 5.

Tangential-Fired Boilers in Low NO_x Mode

Further testing on possible compliance coals for Gaston 5, under both conventional and low NO_x modes, are listed in Table 2. These results have been previously reported.[1,2,3] The only pilot-scale testing of compliance coals that can be compared to the full-scale at Gaston 5 is the low NO_x testing of the Shoal Creek coal. This coal, produced by Drummond, is from the Blue Creek seam in west-central Alabama and is very similar to the Jim Walter #3 and #5 mines. A full-scale test burn of a blend of Jim Walter #3 and #5 has been performed in Gaston 5 after the installation of the low NO_x firing system. Therefore, the full-scale results on this very similar coal are compared to the CRF testing in Figure 4. Unfortunately, there is no reliable full-scale information on LOI yet available, so only the NO_x emission comparison is shown. As can be seen in the graph, the pilot furnace results correctly predicted the NO_x emissions seen in the Gaston 5 system.

Wall-Fired Boilers in Low NO_x Operating Mode

Two wall-fired boilers were simulated in the testing program: Alabama Power's Plant Miller Unit 3 and Gulf Power's Plant Crist Unit 7. The last two lines of Table 2 summarize the pilot testing results for these two wall-fired simulations. The pilot furnace results are compared to Gulf Power's Plant Crist Unit 7 in Figure 5, for the Gusare coal from Venezuelan. The pilot results were obtained for five different size grinds of the same coal[3], but the graph only presents the largest and smallest sizes. The coarse size, which matched the full-scale plant coal grind, was 50 percent through 200 mesh and the ultra-fine grind sized at 99.9 percent through 200 mesh and 80 percent passing 325 mesh. The pilot testing results underpredicted the actual NO_x emissions and also underpredicted the LOI as compared to the actual full-scale data. It is a good possibility that the pilot scale burner, being very well-controlled for fuel and air flows, does not adequately represent the maldistributions of fuel and air typically seen in full-scale plant.

The comparison of the CRF testing results to Plant Miller Unit 4, a sister unit to #3, is presented as Figure 6 for the Belle Ayr coal, an AMAX Powder River Basin coal. This plot compares the NO_x emissions from the pilot furnace and the full-scale boiler, but only shows LOI for the pilot rig. The NO_x emissions predicted from the pilot testing agreed very well with the full-scale data. Since the pilot furnace produced almost unmeasurable quantities of unburned carbon, it is expected that there would be no problem in the full-scale. Unfortunately, the LOI data from the full-scale test burn is not available.

Boiler Operation Results

A secondary purpose of the pilot testing program was to also predict the existence and severity of any boiler operating problems that one of these compliance coals might cause in a full-scale boiler. These would include such coal quality issues like milling behavior, slagging and fouling, ESP operation, etc. Several observations of coal behavior were made in the testing program which mirrored or, in some cases, predicted operational problems in the full-scale plant. The Shoal Creek coal, being finely-sized and wet from a washing plant, caused considerable coal stoppages in the coal bunkers in the pilot testing. This

behavior has also been seen in the Gaston 5 test burn of Jim Walter #3 and #5 blend, which is very similar in nature to the Shoal Creek. The other notable observation in the pilot testing was the tendency of the ash from the Belle Ayr coal to suddenly deposit in the convective tube bank sections. As the temperature of the flue gas slowly increased over the testing period, a critical temperature (which was not measured) would be reached and rapid deposition of ash would occur. Alabama Power's Plant Miller did experience convective section fouling while burning this same coal.

Conclusions

Pilot-scale testing of a wide range of coal types in both tangential and wall firing modes show remarkable similarity to the limited full-scale data available for both NO_x emissions and unburned carbon levels. Because of the database established by this research program, there is a high degree of confidence that pilot testing in the SCS/SRI Combustion Research Facility can predict full-scale NO_x and unburned carbon levels. Other coal quality parameters, which will be reported on in the future, can also be measured which predict behavior in full-scale plants.

Acknowledgements

The authors would like to thank Drummond Coal for providing coal and transportation for this research program. Alabama Power's Central Analytical Facility is providing the necessary analytical support for the project. The SRI staff responsible for the operation and maintenance of the combustion facility are Wim Marchant, Sam O'Neal, Ken O'Neal, Bill Page, and Dave Smith.

References

1. L. Monroe, R. Clarkson, and J. Stallings, "Pilot-Scale Evaluation of Compliance Coals for NO_x Emissions and Boiler Operations," presented at the EPRI/EPA 1993 Joint Symposium on Stationary Combustion NO_x Control, Bal Harbour, Florida (May 1993).

2. L. Monroe, R. Clarkson, and J. Stallings, "Pilot-Scale Evaluations of Compliance Coals for NO_x Emissions and Unburned Carbon Levels," presented at the EPRI 1994 NO_x Workshop, Scottsdale, Arizona (May 1994).

3. L. Monroe, R. Clarkson, and J. Stallings, "Predictions of Full-Scale NO_x Emissions and LOI from Coal and Coal Blends in Pilot Combustion Experiments," presented at the EPRI Fourth International Conference on The Effects of Coal Quality on Power Plants, Charleston, South Carolina (August 1994).

Table 1. Pilot Furnace Coal Analyses

Analysis	North River	Cedrum	Shoal Creek	Cedrum-Shoal Creek Blend (80/20)	Gusare (Venezuelan)	Belle Ayr (PRB)
Proximate, as received						
% Moisture	5.53	6.10	6.95	10.22	5.66	26.32
% Volatile	34.28	31.12	24.28	27.78	33.87	31.02
% Fixed Carbon	47.95	49.07	60.18	49.52	54.37	36.54
% Ash	12.23	13.71	8.59	12.48	6.10	6.12
Total	100.0	100.0	100.0	100.0	100.0	100.0
HHV, Btu/lb	12098	11647	13104	11420	13115	8682
MAF HHV, Btu/lb	14711	14525	15515	14776	14864	12851
Fuel Ratio	1.40	1.58	2.48	1.78	1.61	1.18
Ultimate, dry basis						
% Sulfur	2.72	0.65	0.77	0.69	0.60	0.59
% Carbon	72.32	72.71	80.76	72.34	78.90	68.57
% Hydrogen	4.84	4.60	4.67	4.53	5.15	4.84
% Nitrogen	1.67	1.73	1.78	1.74	1.49	0.87
% Oxygen (diff.)	5.50	6.25	2.79	6.80	7.39	16.82
% Chlorine	0.03	0.08	0.02	0.05	0.02	0.01

Table 1. Pilot Furnace Coal Analyses (continued)

Ash Analysis	North River	Cedrum	Shoal Creek	Cedrum-Shoal Creek (80/20)	Gusare (Venezuelan)	Belle Ayr (PRB)
% Al_2O_3	25.00	22.04	30.45	25.59	23.97	16.99
% CaO	7.35	4.62	2.45	3.74	3.25	20.22
% Fe_2O_3	19.84	4.97	9.58	5.89	6.75	6.18
% MgO	1.35	1.44	1.38	1.44	3.69	3.62
% MnO_2	0.05	0.06	0.04	0.05	0.06	0.06
% K_2O	2.07	4.22	2.76	3.02	4.01	0.39
% SiO_2	38.90	62.55	50.29	54.58	56.03	32.21
% Na_2O	0.60	0.73	1.12	0.71	0.69	2.44
% TiO_2	1.09	1.25	1.52	1.21	1.04	1.37
Ash Fusion Temp.						
Reducing, °F						
Initial	2096	2450	2565	2489	2205	2135
Softening	2289	2570	2692	2593	2315	2170
Hemispherical	2320	2605	2743	2636	2355	2180
Fluid	2490	2740	2795	2736	2500	2210
Oxidizing, °F						
Initial	2365	2360	2250	2523	2260	2195
Softening	2427	2590	2783	2664	2300	2220
Hemispherical	2480	2635	2822	2713	2360	2229
Fluid	2587	2750	>2824	2773	2525	2272
Grindability	52.3	47.0	76.5	54	45.5	50.5

Table 2. Alabama Power/EPRI Pilot Furnace Testing Summary

Date	Coal	Firing Mode	Plant Simulated	NO_x and LOI 3.5% O_2 Furn. Exit	Full-Scale Comparison
Sept. 92	N. River	T-fired Conventional	Gaston #5	0.48 lb/MBtu 1.25 %	0.60 lb/MBtu[1] 1.0 %
Oct. 92	Shoal Creek	T-fired Conventional	Gaston #5	0.68 lb/MBtu 2.6 %	
Dec. 92	Shoal Creek	T-fired Low NO_x	Gaston #5	0.47 lb/MBtu 2.3 %	0.50 lb/MBtu[2] NA
May. 93	Cedrum	T-fired Conventional	Gaston #5	0.58 lb/MBtu 0.6 %	
Jun. 93	Cedrum	T-fired Low NO_x	Gaston #5	0.37 lb/MBtu 1.9 %	
Jul. 93	Gusare (Venezuelan)	T-fired Conventional	Gaston #5	0.49 lb/MBtu 10.3 %	
Nov. 93	Cedrum-Shoal Creek	T-fired Conventional	Gaston #5	0.56 lb/MBtu 1.0 %	
Jan. 94	Cedrum-Shoal Creek	T-fired Low NO_x	Gaston #5	0.30 lb/MBtu 1.9 %	
Jun. 94	Gusare (Venezuelan)	Wall-fired Low NO_x	Crist #7	0.46 lb/MBtu 14.4 %	0.59 lb/MBtu[3] 22 to 41 %
Nov. 94	Belle Ayr (PRB)	Wall-fired Low NO_x	Miller #3	0.34 lb/MBtu < 0.1 %	0.33 lb/MBtu[4] NA

[1] EPRI/DOE CQE testing at Gaston Unit 5 with furnace exit oxygen of 3.6%.

[2] Fuel test burn with Jim Walter #3/#5 blend, a coal from the same seam and very similar to Shoal Creek with furnace exit oxygen of 3.7%.

[3] Fuel test burn at Crist Unit 7 with furnace exit oxygen of 3.2%.

[4] Fuel test burn at Miller Unit 4 with furnace exit oxygen of 3.5%.

SCS and SRI Combustion Facility

COAL HANDLING

-INDIRECT FIRED

-ON-SITE STORAGE, 10 EACH 20-TON BINS

-CRUSHING AND PULVERIZING EQUIPMENT

-GRAVIMETRIC FEED TO FURNACE

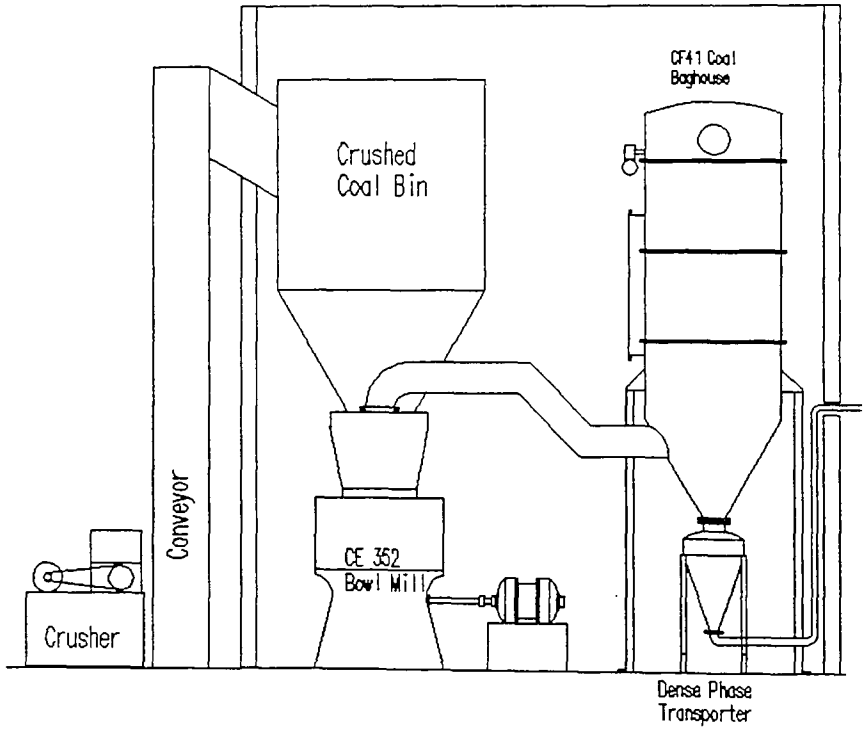


Figure 1. Coal handling equipment of the SCS and SRI Combustion Research Facility.

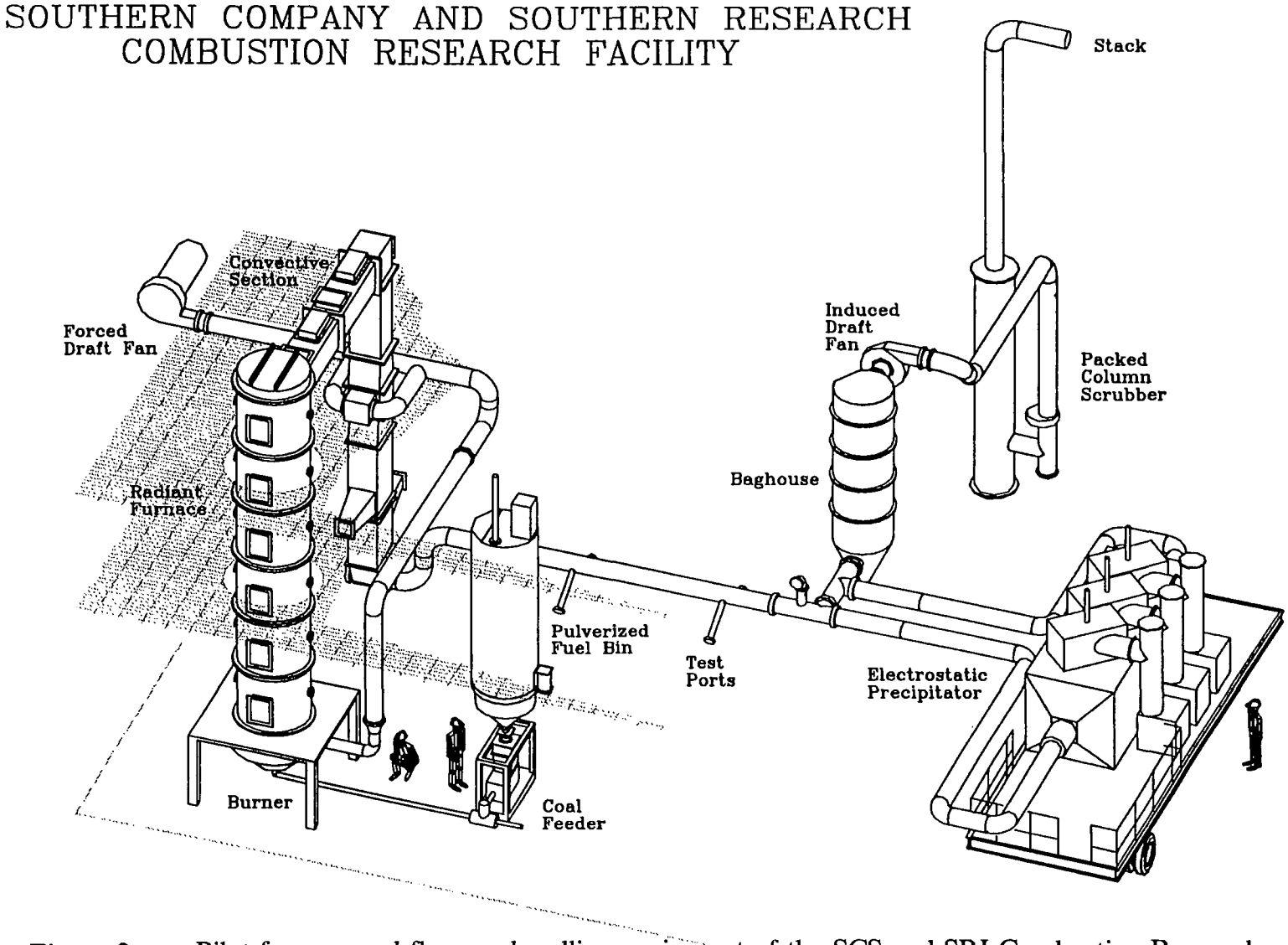
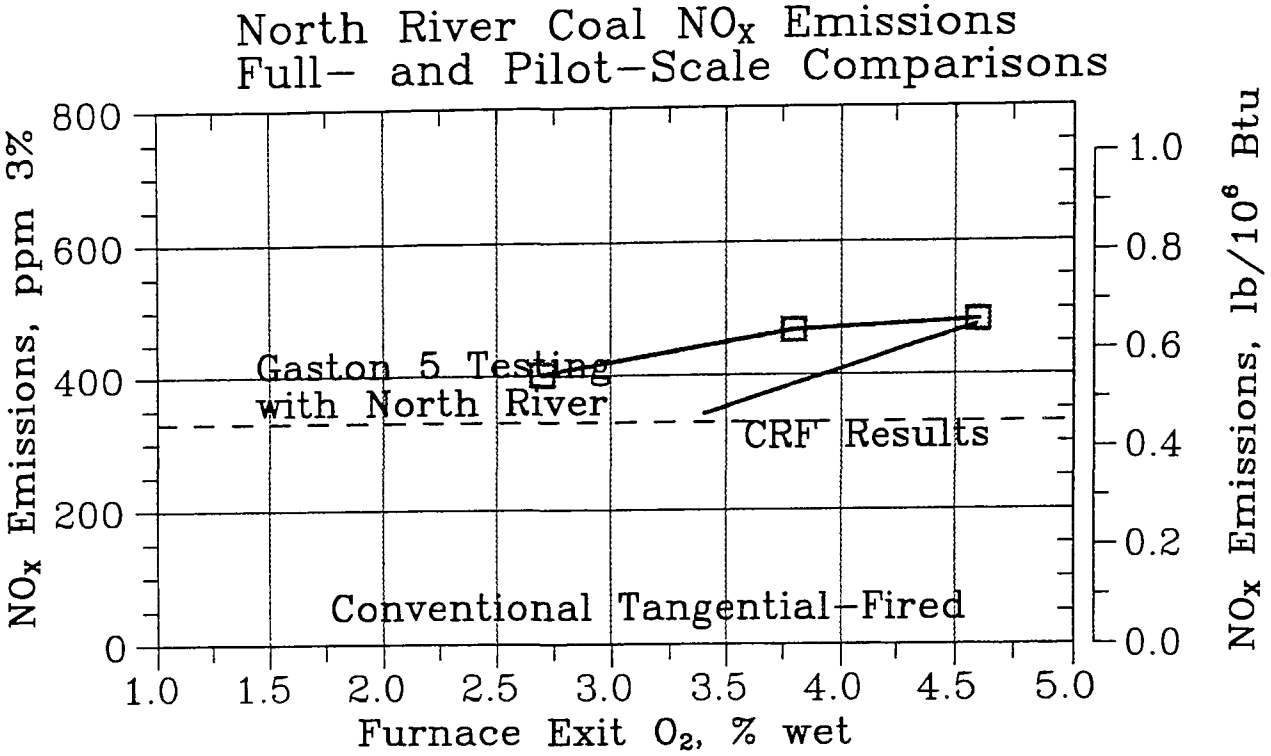


Figure 2. Pilot furnace and flue gas handling equipment of the SCS and SRI Combustion Research Facility.



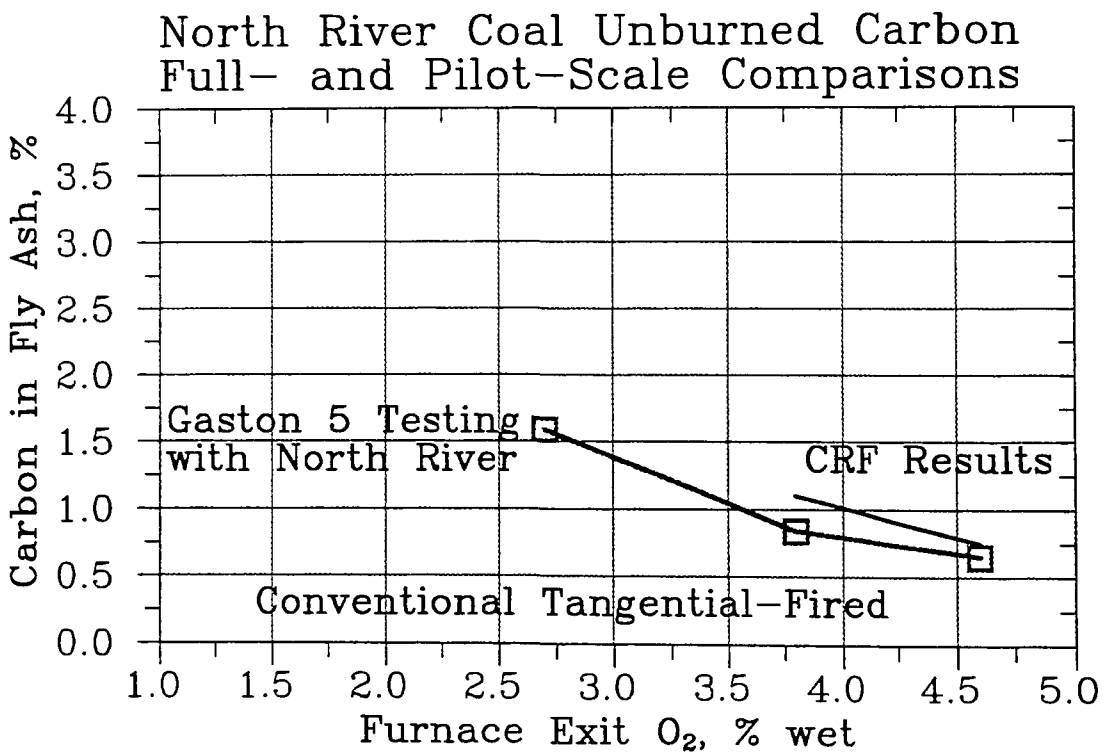


Figure 3. NO_x emission and unburned carbon comparison of Alabama Power's Gaston Unit 5 and the SCS and SRI Combustion Research Furnace for the North River coal, with conventional tangential firing.

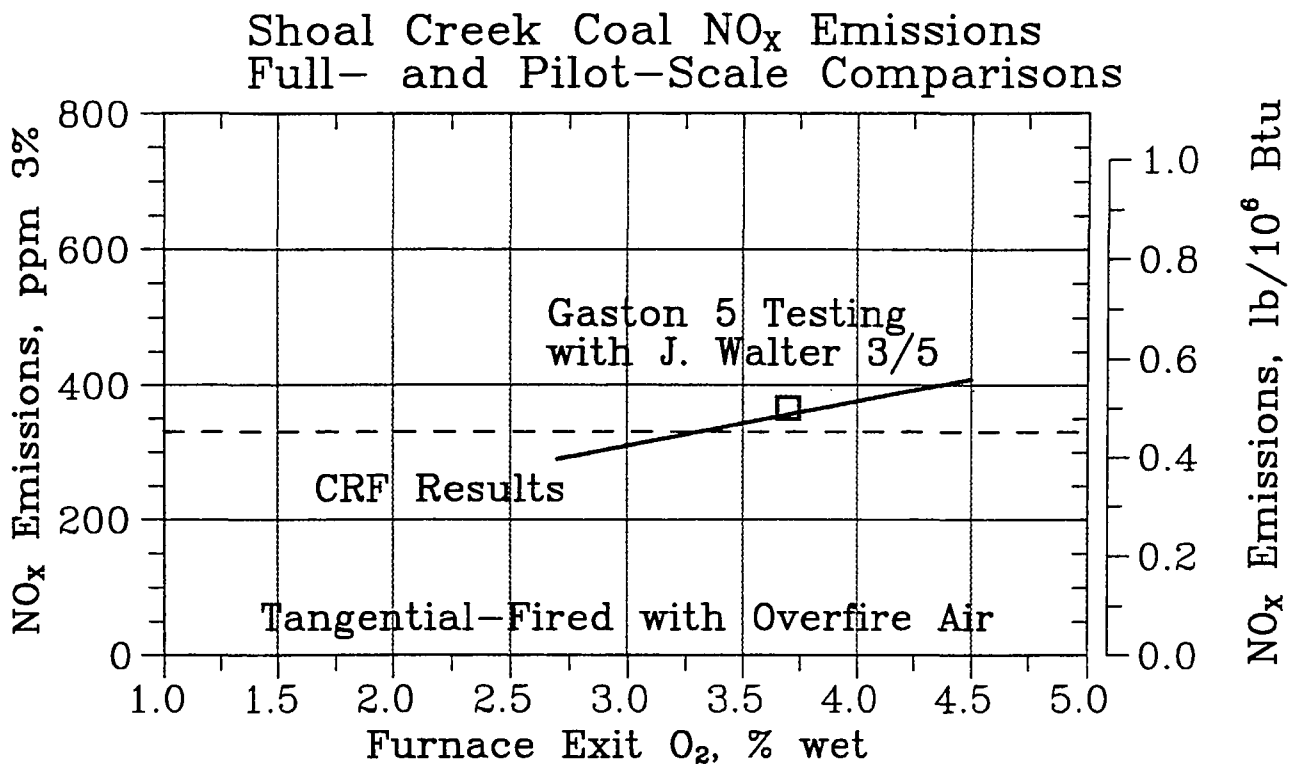


Figure 4. NO_x emission comparison of Alabama Power's Gaston Unit 5 on Jim Walter #3 and #5 coal and the SCS and SRI Combustion Research Furnace for the Shoal Creek coal, with low-NO_x tangential firing.

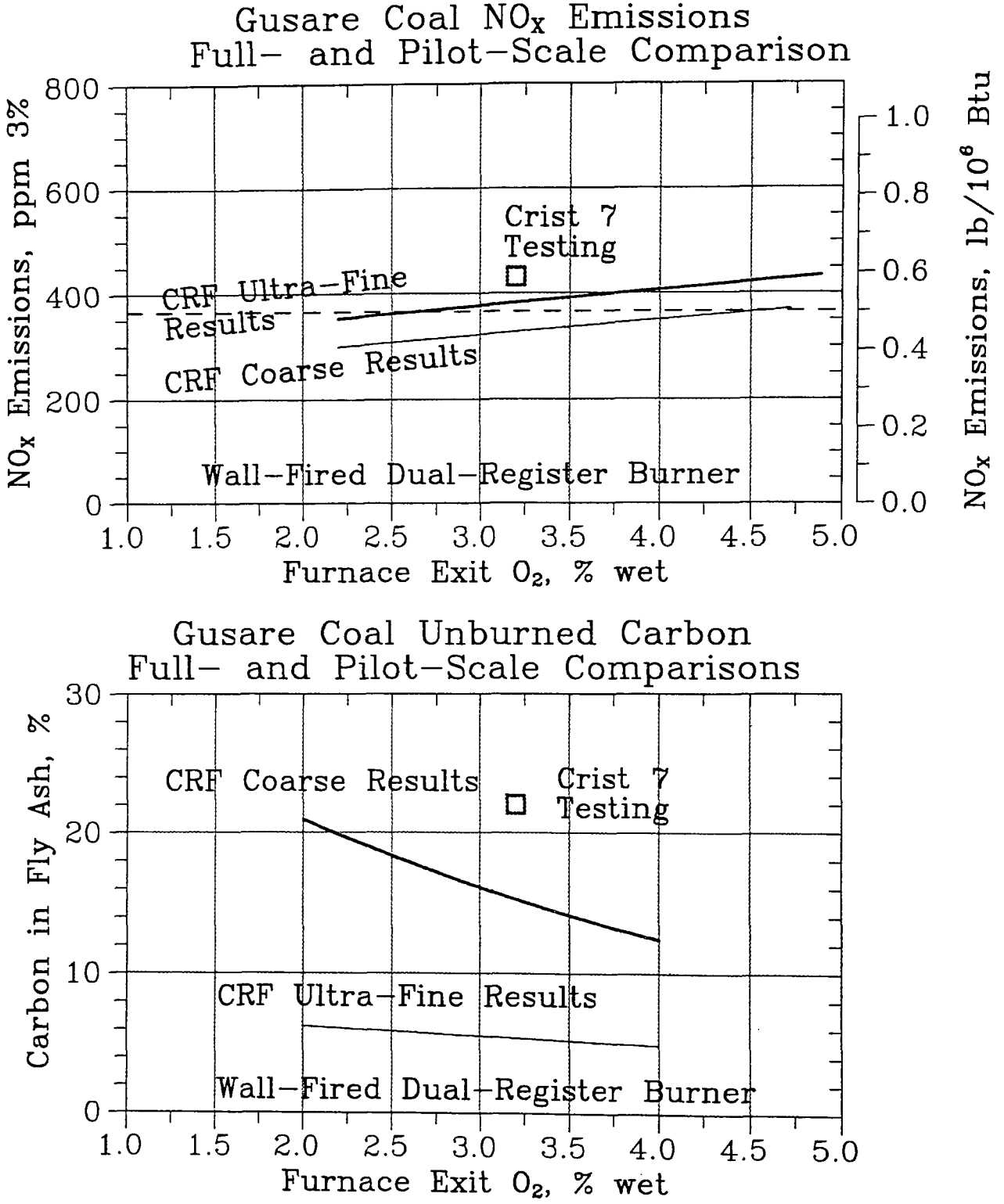
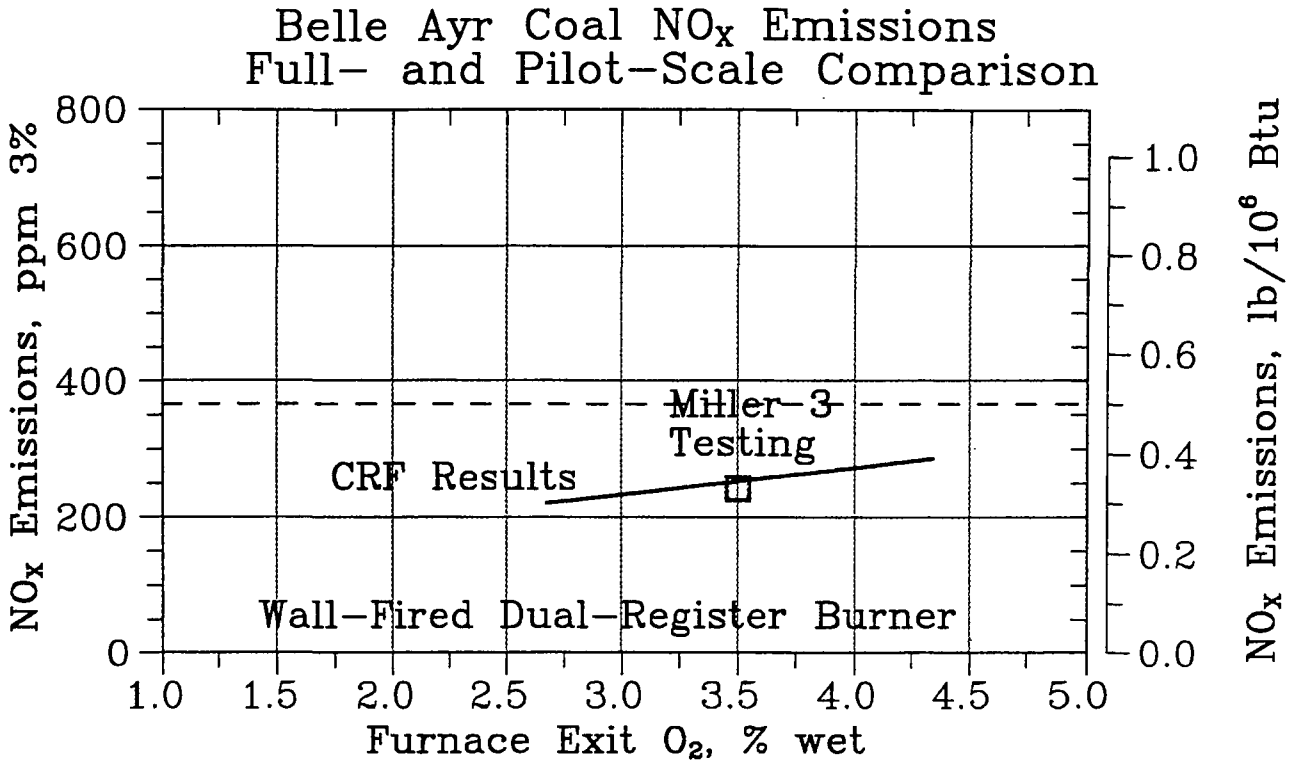


Figure 5. NO_x emission and unburned carbon comparison of Gulf Power's Crist Unit 7 and the SCS and SRI Combustion Research Furnace for the Gusare coal, with dual-register low-NO_x burner firing.



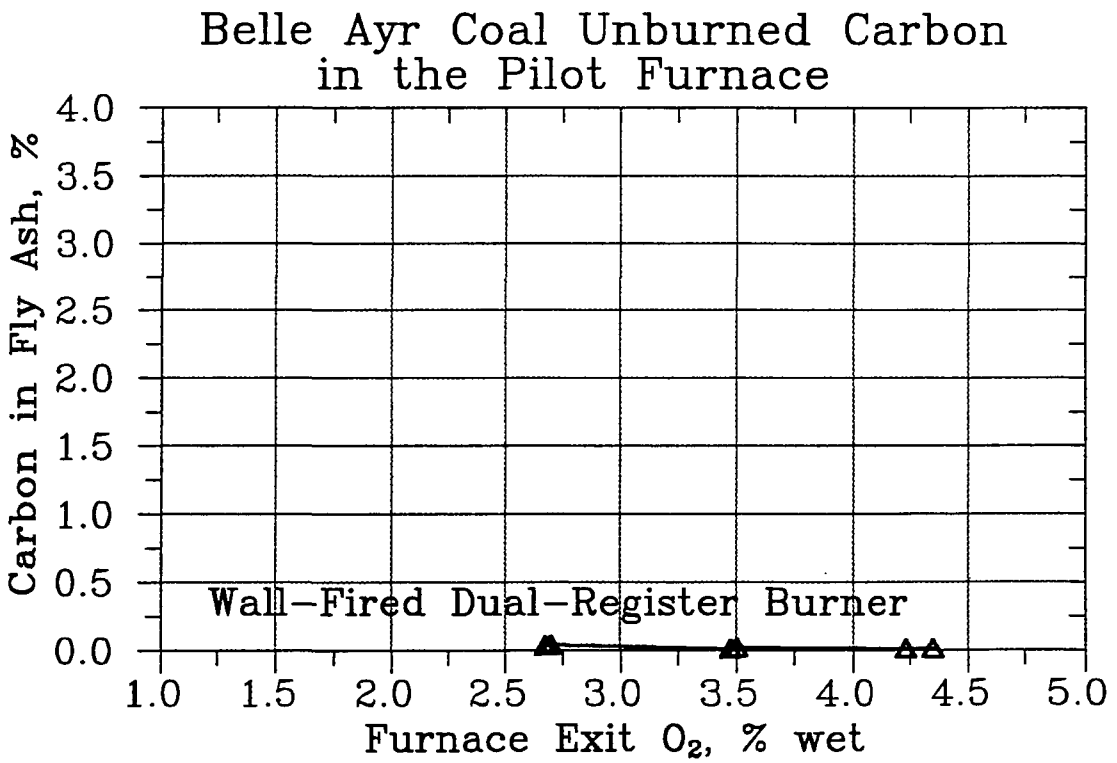


Figure 6. NO_x emission and unburned carbon comparison of Alabama Power's Miller Unit 4 and the SCS and SRI Combustion Research Furnace for the Belle Ayr coal, with dual-register low-NO_x burner firing.

1995 EPRI/EPA Joint Symposium on Stationary Combustion NO_x Control

Methodology for the Selection of Low NO_x Firing Alternatives

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Abstract

Combustion Modification (CM) is an often used technique for reducing NO_x formation from utility boilers. CM can be implemented by changing boiler operation parameters using the existing firing equipment, or through hardware modifications. Achievable potential NO_x reduction levels and long term operational effects of either approach are boiler specific. Implementation of either methodology can lead to changes in boiler operating efficiency, reliability, availability, and maintenance costs. An economic analysis of alternatives must accurately reflect associated costs. Proper implementation of the selected NO_x reduction technology is essential in preventing the transformation of an emission problem into an operational problem.

In a joint effort by ComEd, EPRI, and Radian Corporation, a technique has been developed to facilitate the comparison of the long term cost effectiveness of each means of implementing combustion modification. The approach begins with Baseline and Exploratory Combustion Modification (ECM) testing. Baseline tests identify critical operational design parameters. ECM testing determines the flexibility of the existing furnace to accommodate operational and/or hardware changes. The resultant test data is integrated into the Radian Furnace Simulation Model (FSM).

The Radian FSM is state-of-the-art three dimensional, 2 phase flow, fluid dynamic model of the boiler furnace. The model simulates furnace conditions by drawing extensively from experience and combustion data. The model computes NO_x values and operational rating factors used to compare CM approaches. These factors may include emissions concentrations, slagging, waterwall corrosion potential, and flue gas temperature. These calculated factors can then be used to evaluate long term effects of CM alternatives.

This paper outlines the key features of the FSM and its successful application to a boiler on the ComEd system.

INTRODUCTION

ComEd operates one of the largest coal-based power systems in the U.S. Earlier decisions to convert to Powder River Basin, low sulfur coal resulted in only one of ComEd's 9 coal fired plants being affected by the 1990 Clean Air Act Amendments Phase I requirements. ComEd may be required to implement boiler hardware or operational changes on its existing coal fired furnaces by the year 2000. Currently, ComEd is evaluating NO_x compliance options for a matrix of potential emission limit scenarios. An important aspect of all the possible scenarios will be NO_x controls for ComEd's ten tangentially fired boilers. As such, ComEd needed an approach to evaluate the impacts of low NO_x operations on this class of boilers.

This need evolved into a joint effort by ComEd, EPRI and Radian to employ a three dimensional fluid dynamic model to evaluate various boiler optimization and hardware alternatives. ComEd selected Will County unit 3, a twin furnace CE boiler rated at 278 gross megawatts, for evaluation. The unit currently burns Powder River Basin coal. The unit is equipped with off-set auxiliary air nozzles which lower NO_x and provide an air curtain along the walls. The Unit 3, reheat boiler study is focused on the development of a specific Furnace Simulation Model (FSM) that fully characterizes the unit's operating performance. The prime use of the model is to provide information necessary to evaluate NO_x emissions and the potential of immediate and long term operational problems associated with use of combustion tuning or hardware changes as the means to affect NO_x reductions. The FSM for Will County unit 3 is verified with existing data.

The FSM couples modern computational fluid dynamics (CFD) capabilities with the practical boiler knowledge and experience of Radian and ComEd. The model is designed to run within the PHOENICS Navier-Stokes equation solver developed by CHAM Limited, London, England. The model predicts the following boiler parameters and operating conditions:

- furnace temperature profile
- furnace flow field velocity profile
- furnace coal combustion rate profile
- furnace NO_x generation profile
- unburned carbon at the furnace exit
- O_2 at the furnace exit
- furnace exit gas temperatures
- furnace slagging potential
- furnace tube corrosion potential.

APPROACH

The approach is broken down into four steps: (1) model development, (2) model validation for baseline conditions, (3) assessment of tuned configuration and (4) assessment of hardware options. These steps are discussed in detail next.

Model Development. This step involved the gathering of information necessary for input and verification. Pertinent data included the current burner configuration and general boiler geometry as well as boiler operating parameters (e.g., excess air levels, burner tilt angles). Input variables are then modified to reflect geometric and operational changes.

The model input requirements include burner details sufficient to simulate coal, primary air, and secondary air flow areas and velocities at each region. Ultimate and proximate fuel analyses, ash fusion temperature data, mill fineness test results and fuel air balancing data were provided by ComEd for input into the model.

The result of this phase was a detailed uncalibrated model incorporating the specific geometry characteristics of Will County Unit 3.

Model Validation - Baseline Conditions. In this step, the previously developed boiler model is run at specific conditions for which Baseline test data has been acquired. Comparisons are made of predicted NO_x levels, unburned carbon levels, furnace exit gas temperature and economizer O_2 levels to field measured values. Any operating anomalies noted during the testing (e.g., impinging flame patterns and slag buildup) are compared with model results for verification.

Combustion relationships reflecting the properties of Powder River Basin coal were used in this work. These properties include high moisture and volatiles. The setting of combustion parameters is based upon constructing a separate model of a test burner for which detailed flame probing data is available for this coal. Parameters are adjusted to "fine tune" the combustion to obtain the best match with the subscale data. These parameters are then used in the full 3D boiler model.

Assessment of "Tuned" Configurations. The next step in the model development is to assess the suitability of various "tuned" firing configurations for long term operation. The present model data includes flue gas furnace exit temperature, profiles of excess oxygen, NO_x and unburned carbon. The model also provides indices reflecting possible concerns of slag buildups and tube wastage. Resultant values for these correlation indices are compared with values established for the "as found" firing condition (baseline task) to provide a comparison between previous long term observations of these problems with any future problems that could result from "tuned" low NO_x configurations. Model predictions are also compared to actual field data taken while operating Will County unit 3 under low NO_x firing conditions.

The tuned conditions are examined over a load range with excess air varied to reflect test results. Mills-out-of-service conditions are also examined as to their influence on the calculated parameters.

Assessment of Hardware Options. The final step in the model development focuses on "generic" configurations of various firing hardware changes; specifically close coupled, separated, and close coupled and separated OFA installation. Model simulations are made to examine the immediate and long term effects of these hardware changes. Current efforts are concentrated on close coupled OFA. The results of these calculations are compared with those from the base validation effort and those developed in the evaluation of the tuned alternatives.

RESULTS

Model Validation. Combustion parameters obtained from correlation with subscale data were used to predict NO_x and operating factors for validation with a series of Baseline tests. The results of the comparison are presented in Table 1. The test data are presented in terms of actual species concentration by "uncorrecting" the reduced test data for a direct comparison with model output. The model output data are a flux weighted area average over the exit plane (nose region) of the model.

All the cases were computed with the same combustion and kinetic parameters as correlated from the subscale model. Predictability of NO_x was very good at the full load conditions.

Indices of the relative changes in unit availability and reliability are also computed. These are computed in terms of the local gas equivalence ratio. This parameter is a measure of the fuel richness of both the gas species and the solid particle environment. Fuel richness affects both corrosion and slagging potential. These relative indices are:

- Average Equivalence Ratio
- Maximum (Fuel Rich) Equivalence Ratio
- Minimum Equivalence Ratio
- Fuel Rich Area Fraction
- Slag Index
- Low Temperature Corrosion Index
- High Temperature Corrosion Index

The wall patches (computational areas) for which the indices are computed are presented in Figure 1. To illustrate a practical application of the modeled results, the index computations are presented for a base case at baseline operation (Case

3) and for a simulation of OFA (Case 10). The relative indices for the two cases are shown in Tables 2 and 3. A compilation of the slag and corrosion indices are shown in Boiler Diagram 1. These computations show that, although significant NO_x reduction can be achieved with SOFA, the environment in the vicinity of the water walls is much more fuel rich, thus conducive to excessive slag and corrosion.

Boiler Optimization. Comparison of NOx predictions with a second test series of tuned configuration cases is shown in Table 4. Typically, very good agreement was found. Where discrepancies were noted, potential flow maldistribution effects associated with fuel and air biasing are being evaluated.

A summary of the slag and corrosion index values is shown in Boiler Diagram 2. The simulation results show, for cases where fuel was negatively biased in the furnace, increased slag and corrosion potential in the lower third part of the boiler and the right side of the boiler. Even with "concentric air" positioning of the auxiliary air flow, the model indicates that coal and primary air velocities are high enough to penetrate the air curtain yielding a fuel rich environment adjacent to the walls.

The expected impact utilizing LEA by reducing O_2 is shown in Figure 2. The results show that the NO_x levels are reduced with reduction in excess oxygen. However, the wall fuel rich fraction increased with decreased O_2, thus increasing the likely potential for slag and corrosion. High and low temperature corrosion indices behavein an inverse parabolic relationship to the excess oxygen. The corrosion is expected to be minimum at excess oxygen of about 3.0%.

Hardware Modification. Potential hardware changes to Unit 3 include Low NO_x Control Concentric Firing System, Level I (LNCFS1) hardware. Boiler Diagram 3 and Figure 3 shows the calculated impact of varying excess O_2 on NO_x generation levels and wall slag and corrosion indices. The results of the simulation suggests that NO_x emission can be further reduced with lower O_2. However, as was shown with the optimization cases, high temperature and low temperature corrosion increases with reduced O_2 levels.

ONGOING WORK

Work continues on the study of fuel richness and its impact on the water walls in low NO_x firing situations. Of particular importance is the correlation of absolute index values to actual observations of furnace slagging/emission conditions. Proper evaluation and correlation will lead to confidence that long term low NO_x firing modes will be acceptable in terms of changes in heat rate, reliability, availability, maintainability, and operability.

CONCLUSIONS

Modern computers and CFD analysis techniques have advanced to the point where the complex processes of combustion in boilers can be described quite well. Combining computational capability with extensive theoretical and hands-on field experience in combustion NO_x control can result in a combined modeling and testing approach which satisfies the unit-specific requirements for evaluating low NO_x firing approaches. Application of this model has given confidence that low cost optimization approaches will work on Will County #3 without significant deleterious long term effects. Further correlation of absolute index values to actual observations of furnace slagging/corrosion conditions is critical. Work will continue to strengthen this confidence.

Table 1: Baseline Configuration

Test Information--Baseline Configuration Tests				Test Results	Calculations
ID#	Load (Mw)	0_2 (% v)	Notes	NO_2 (ppmv)	NO_x (ppmv)
3	278	Normal		243	242
4	278	Low		125	133
5	278	High		259	245
6	260	Normal	Tilts - 28 deg.	280	228
7	260	Normal	Tilts + 12 deg.	236	221
8	278	Normal	Damper Var	232	393
10	190	Normal	OFA Simul	111	110
11	190	Normal	MOOS.Dmp	209	364

Table 2: Case 3. Baseline Condition Index Computation

Wall	Avg. Eq. R.	Max. Eq. R.	Min. Eq. R.	Fuel Frac.	Slag	LT Corr.	HT Corr.	Area, ft2
6	.690	.919	.483	.000	.00	.000	.000	706
12	.815	1.230	.423	.170	.00	2.390	1.350	524
10	.774	1.070	.464	.054	.00	.727	.425	1310
7	.799	.819	.728	.000	.00	.000	.000	706
13	.695	.800	.541	.000	.00	.000	.000	524
11	.798	.859	.597	.000	.00	.000	.000	1310
5	.710	.819	.506	.000	.00	.000	.000	397
15	.580	.840	.348	.000	.00	.000	.000	584
9	.765	.823	.566	.000	.00	.000	.000	1470
3	.762	.862	.651	.000	.00	.000	.000	397
14	.703	.933	.402	.000	.00	.000	.000	584
16	.858	1.040	.685	.058	.00	26.100	.454	1190

Table 3: Case 10. Separated OFA Index Computations

Wall	Avg. Eq. R.	Max. Eq. R.	Min. Eq. R.	Fuel Frac.	Slag	LT Corr.	HT Corr.	Area, ft2
6	1.170	1.250	.963	.987	74.10	12.90	10.400	706
12	.942	1.260	.067	.691	.00	9.97	6.140	524
10	.470	.753	.165	.000	.00	0.00	0.000	1310
7	1.130	1.200	1.020	1.000	20.52	2.70	10.100	706
13	.938	1.130	.077	.671	13.00	0.12	5.640	524
11	.458	.604	.174	.000	.00	0.00	0.000	1310
5	1.110	1.140	.875	.909	18.60	11.70	8.810	397
15	.634	1.110	.021	.135	3.22	423.00	1.180	584
9	.473	.583	.222	.000	.00	0.00	0.000	1470
3	1.120	1.260	.942	.938	10700.00	12.50	9.000	397
14	.675	1.200	.020	.073	288.00	28100.00	>100,000	584
16	.433	.697	.245	.000	.00	0.00	0.000	1190

Table 4: Tuned Configurations

Test Information--Tuned Configuraiton Tests				Test Results	Calculations
ID#	Load (Mw)	O_2 (% v)	Notes	NO_x (ppmv)	NO_x (ppmv)
35	278	Normal	Top FDR bias	238	230
36a	278	Normal	Same, Aux Air Bias	190	229
36b	278	Normal	Same, Aux/FA Bias	181	218
36d	278	Low	Same, Aux/FA Bias	162	476
37b	278	Normal	Same, Aux/FA Bias	190	231
37e	278	Normal	Top FDR bias, top Aux/FA 100%	170	295
38a	278	Normal	Top FDR bias, POOS	217	219
38b	278	Normal	Same, Aux 100%	209	232
38c	278	Normal	Same, Aux/FA 100%	197	219

Designation of Wall Patches

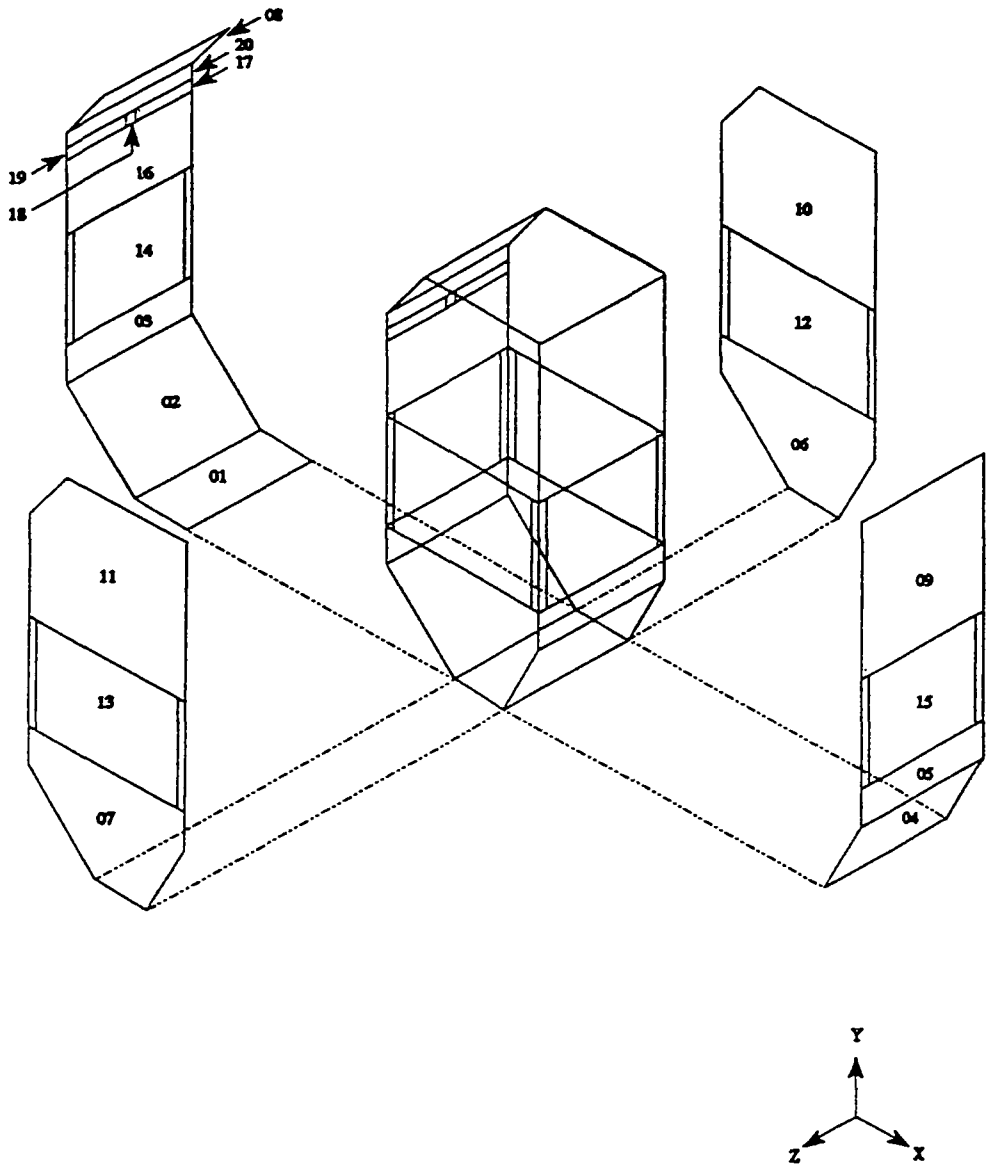
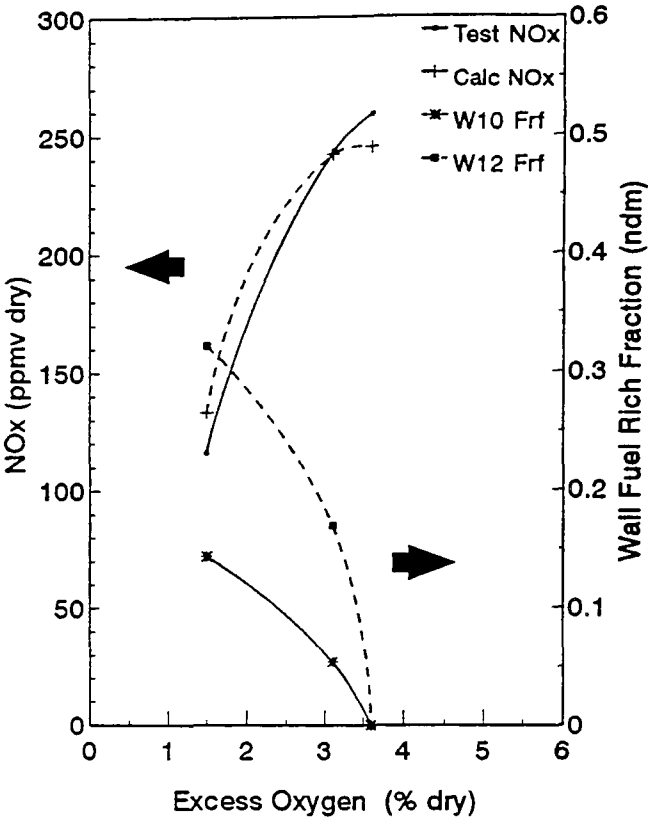


Figure 1

Effect of Excess Air on
NOx and Wall Fuel Rich Fraction



Effect of Excess Air on
High and Low Temperature Corrosion Indices

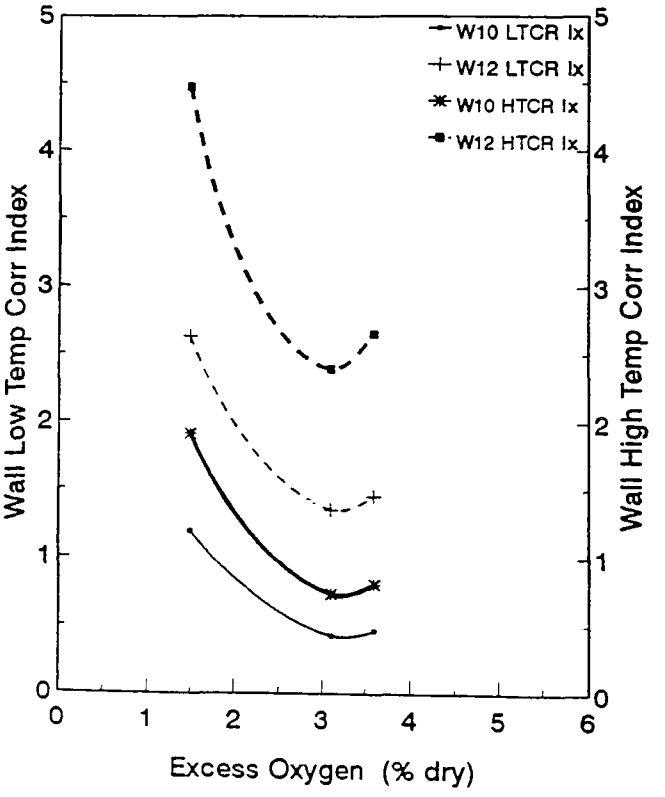
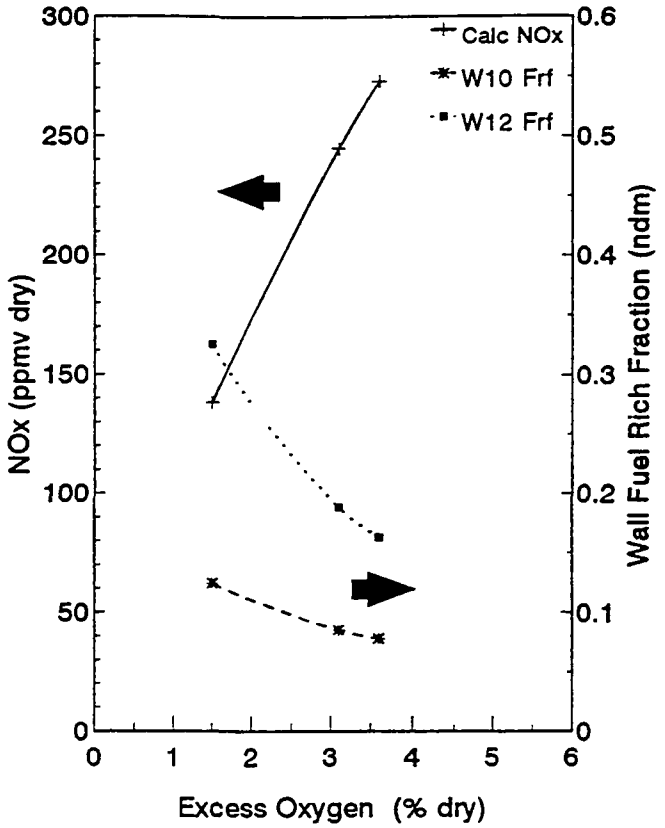


Figure 2

Simulation of LNCFS#1 Hardware
Effect of Excess Oxygen on NOx and Wall Fuel Rich Fraction



Simulation of LNCFS#1 Hardware
Effect of Lowered Excess Air Corrosion Indices

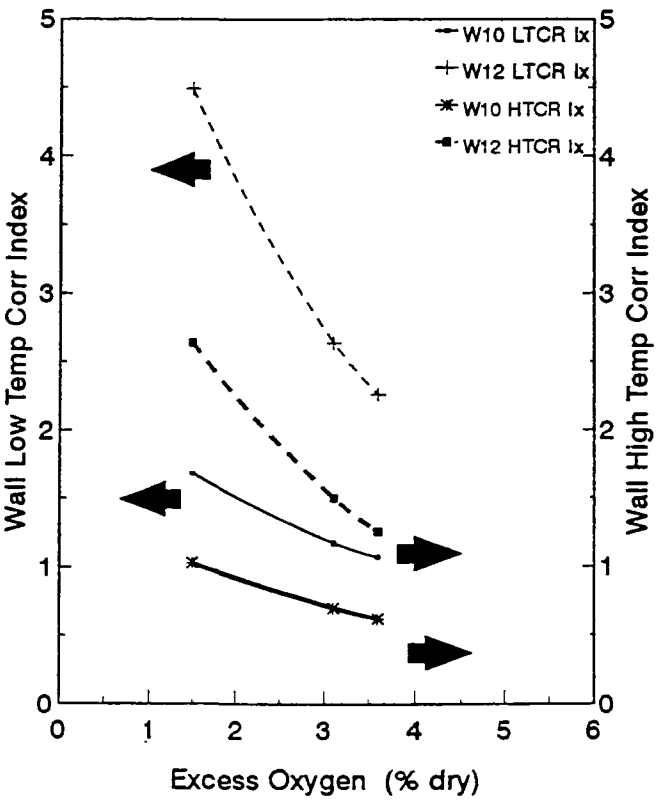
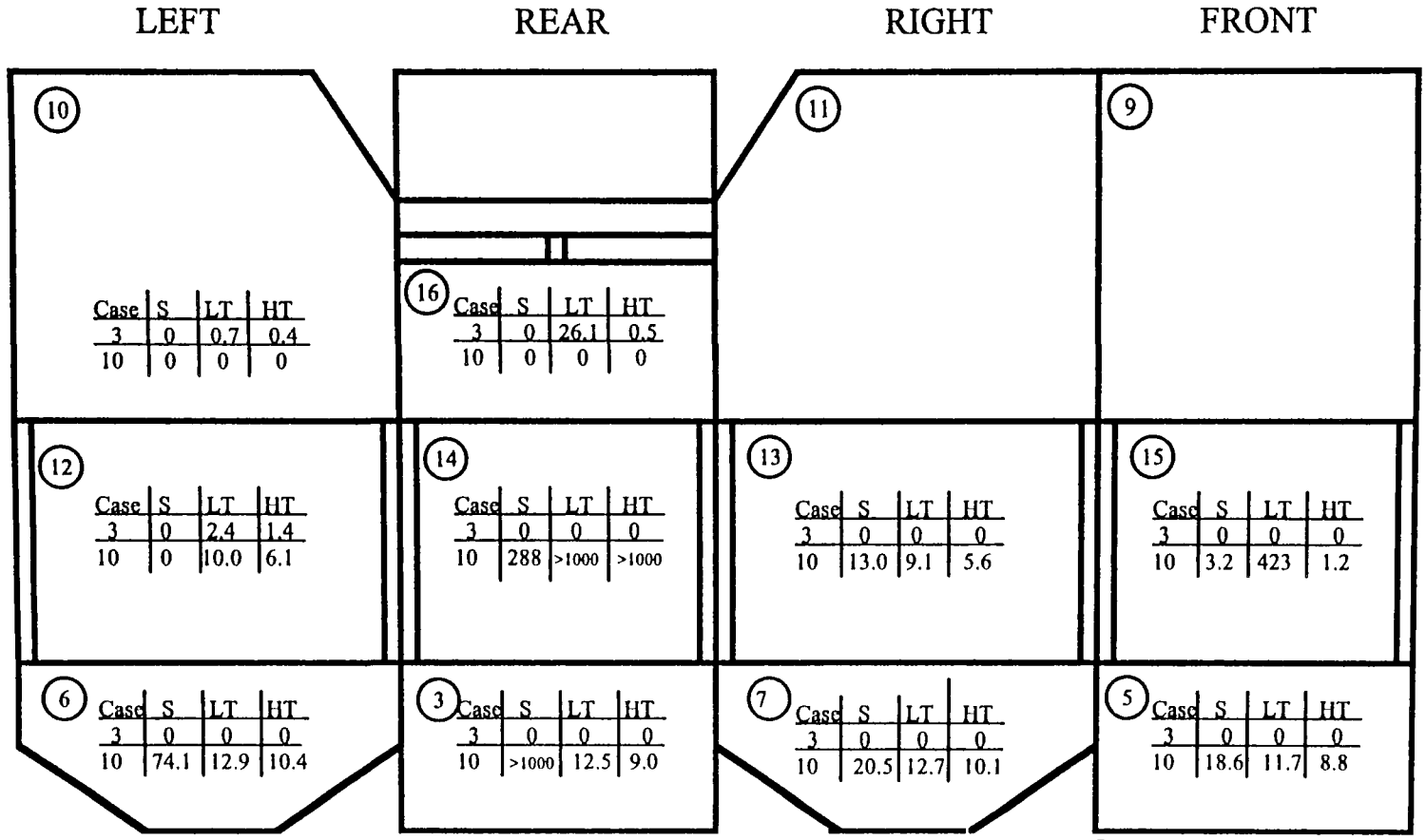


Figure 3

Diagram 1

WILL COUNTY 3

Slagging and Corrosion Indices for Baseline and Simulated Overfire Air Test

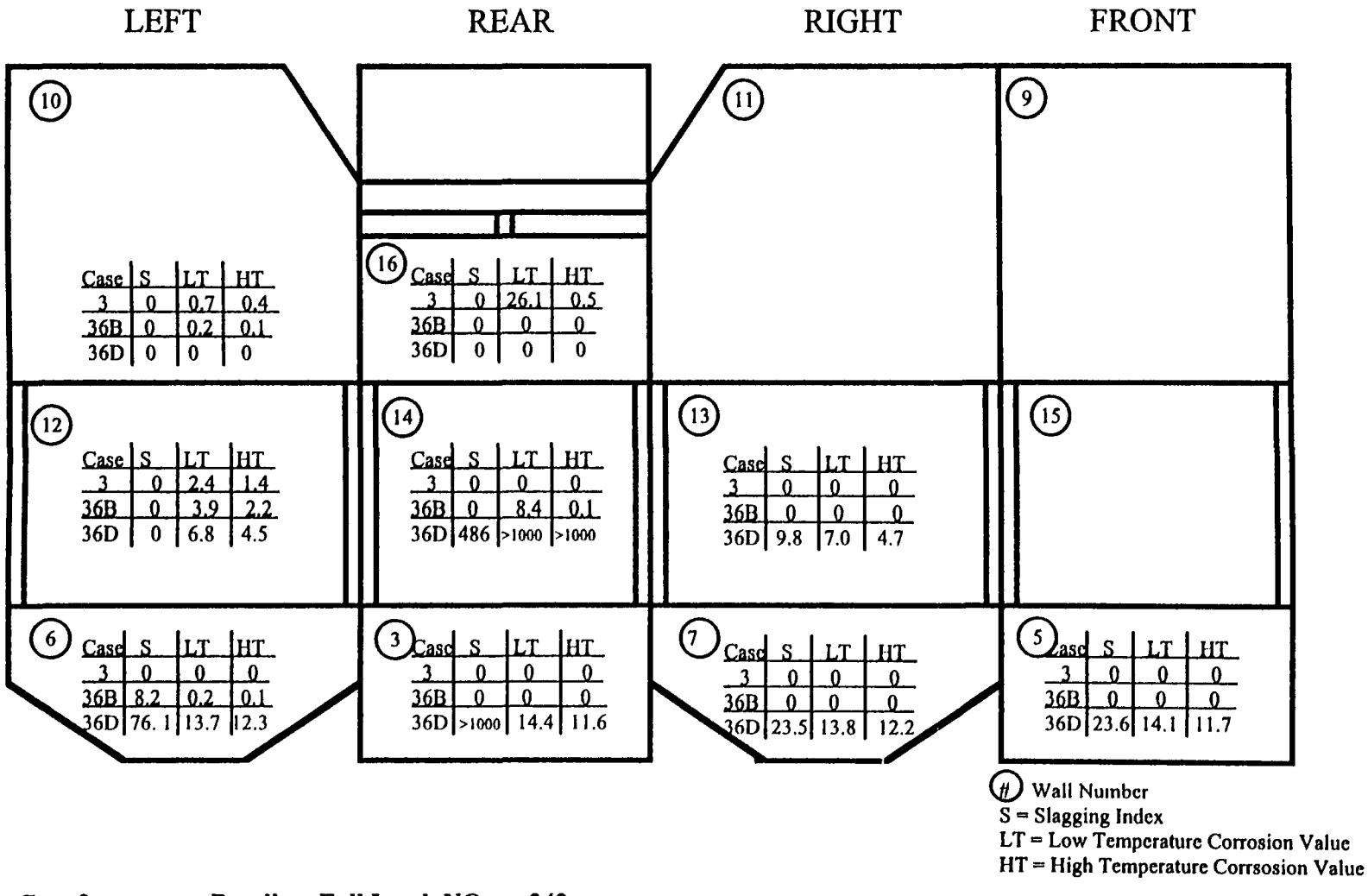


(#) Wall Number
S = Slagging Index
LT = Low Temperature Corrosion Value
HT = High Temperature Corrsosion Value

Case 3: **Baseline, Full Load, NO_x = 242 ppm**

Case 10: **SOFA Variation, NO_x = 110 ppm**

Diagram 2
WILL COUNTY 3
Slagging and Corrosion Indices for Baseline and Optimization Using Low O_2 or Feeder Bias

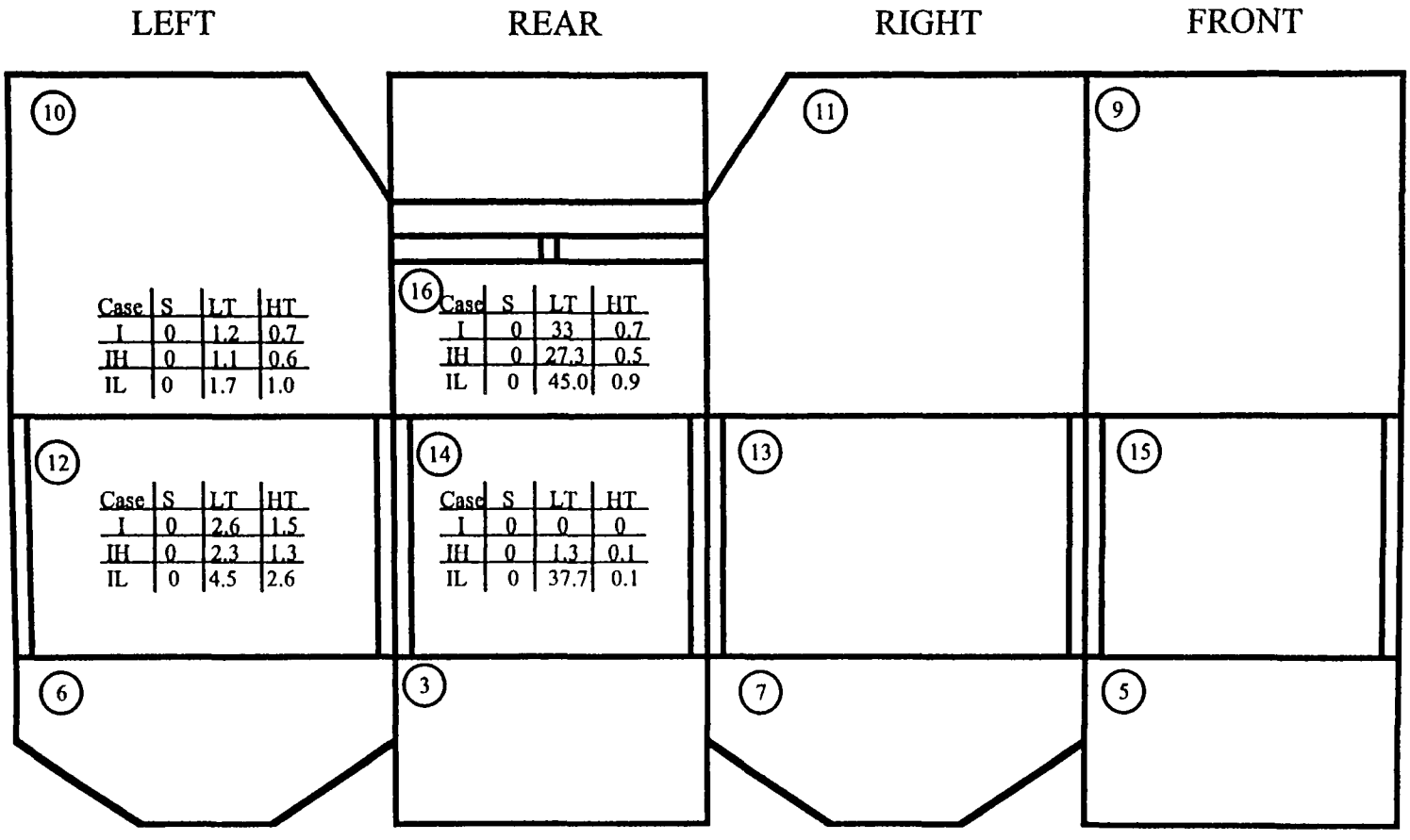


Case 3: **Baseline, Full Load, NO_x = 242 ppm**

Case 36B: **Full Load, A/E Feeder = -50%, Aux. Air = 75%, Fuel Air = 46%, NO_x = 202 ppm**

Case 36D: **Full Load, A/E Feeder = -50%, Aux. Air = 75%, Fuel Air = 100%, O_2 = 2.0%, NO_x = 161 ppm**

Diagram 3
WILL COUNTY 3
Slagging and Corrosion Indices for Typical Hardware
LNCFS I at Full Load and Normal, High, and Low O_2



Case	S	LT	HT
I	0	1.2	0.7
IH	0	1.1	0.6
IL	0	1.7	1.0

Case	S	LT	HT
I	0	33	0.7
IH	0	27.3	0.5
IL	0	45.0	0.9

Case	S	LT	HT
I	0	2.6	1.5
IH	0	2.3	1.3
IL	0	4.5	2.6

Case	S	LT	HT
I	0	0	0
IH	0	1.3	0.1
IL	0	37.7	0.1

(#) Wall Number
S = Slagging Index
LT = Low Temperature Corrosion Value
HT = High Temperature Corrsosion Value

Case I: LNCFS I Simulation, NO_x = 164 ppm
Case IH: LNCFS I High O_2 Simulation, NO_x = 273 ppm
Case IL: LNCFS I Low O_2 Simulation, NO_x = 138 ppm

Development and Testing of the MIT-RSFC Low-NO_x Burner for Coal Combustion

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Abstract

MIT as a partner with Babcock & Wilcox, a McDermott Company in the USDOE sponsored Low Emission Boiler System (LEBS) program has performed an experimental analytical and experimental effort to refine and define the operational characteristics of the MIT Radially Stratified Flame Core (RSFC) Burner for the combustion of pulverized coal. The RSFC has been previously developed to a very promising state for natural gas[1] and for heavy fuel oil[2] firing. In the present effort the burner was operated at a nominal firing rate of 5 MM Btu/hr on Illinois #6 coal. NO_x CO and unburned carbon emission data have been collected and analyzed for a wide range of conditions. Burner performance has been evaluated as a function of burner settings, firing rate, air equivalence ratio, air preheat, transport air to coal ratio and excess air. For selected conditions , detailed in-flame measurements of gas composition, temperature and velocity as well as solids concentration and composition have been performed. These data can be used for detailed analysis and modeling studies. The performance of the burner was found to be superior to the expected results. The burner was extremely stable, and relatively insensitive to operating conditions. NO_x levels below 100 ppm with carbon burnout greater than 99% were achieved.

The present contribution is based on a paper presented to the Joint AFRC-JFRC Conference in Maui, Hawaii, November 19, 1994.

Introduction

The purpose of this paper is to describe work performed for the U.S. DOE under subcontract to Babcock & Wilcox on the development of a Low NO_x pulverized coal burner (< 0.15 lb/MM Btu fired) for the DOE Low Emission Boiler System (LEBS) program. In this paper we will describe data and analysis obtained at the MIT Energy Laboratory's Combustion Research Facility (CRF) on the development of the Low NO_x pulverized coal burner.

For the purposes of this development an existing Low NO_x research burner called the RSFC (Figure 1) and described in patent application No. 771,739 was used. An unique feature of the RSFC burner is its capability of reducing air-fuel mixing in the near field while promoting it further downstream of the burner. Radial density stratification, brought about by the combination of rotating burner air flow and the density increasing with increasing radial distance near the burner, damps turbulence and hence reduces air-fuel mixing, but further downstream due to vortex breakdown the mixing is vigorous, sufficiently, to ensure complete burnout of the fuel. Standard grind Illinois No. 6 coal, a widely used high volatile bituminous coal, was fired at about 400 lb/hr for a variety of operating conditions and a large body of data collected. Burner parameters, staging, excess air and turndown were investigated.

The RSFC burner was used in two operating modes. In the first of these all the combustion air was supplied through the burner. (No overfire air.) In this mode of operation radial density stratification, necessary for the internal air staging of the flame, required that increasing fractions of air be introduced at increasing radial distances through the burner. This resulted in a minimum NO_x emission of 217 ppm (3% 0_2). A very low value for a low NO_x burner but not sufficiently low to satisfy the LEBS emissions goal of 120 ppm.

In the second mode of operation less than the stoichiometric amount of air was passed through the RSFC burner with the rest supplied as overfire air, downstream of the burner. A hot, fuel rich flame was established by fast mixing of the pulverized coal-transport air and the primary burner air flows. The fuel gun was recessed into the primary air conduit to permit ignition upstream of the burner face. A central recirculation zone was formed due to the strongly swirling primary air flow just downstream of the burner exit. For the "with overfire" operation the RSFC burner was operated with a burner air flow

distribution to give decreasing air flow fractions with increasing radial distance from the burner axis. The residual combustion air was introduced as "overfire air" at $x/D=10.5$ or 13.5, where D is the burner diameter: 0.3 m.

Very low NO_x emissions were obtained even for the mildly staged flame: 107 ppm for SR=0°92, with continued reduction at deeper staging, 70 ppm for SR=O°82 while maintaining high carbon conversion efficiencies, >99%. These NO_x emissions are low enough to satisfy the LEBS emissions goal without additional NO_x reduction techniques.

When the burner is operated with substoichiometric air flow and overfire air, a larger fraction of the air can be fast admixed to the pulverized coal without the risk of providing fuel lean flame zones. Under these conditions both the volatile evolution and the heat release are fast and consequently the residence time in the high temperature, density stratified region of the fuel rich flame is increased resulting in lower NO_x concentrations.

It is noteworthy that the mode of burner operation to give decreasing fraction of the burner air flow with increasing radial distance is not suitable for fuel-lean burner conditions, it produces high conversion of the fuel-N to NO and increases the NO emission.

Theoretical Considerations

The effort of achieving low NO_x emission (100-150 ppm) from coal flames involves the study of many chemical and physical processes taking place during coal combustion, such as flame ignition, volatile evolution and combustion, char particle burnout, chemical reactions of N-bearing species and the physics of turbulent, reacting, particle laden flows. Among these processes, only a few, can be considered in this paper. Our attention will be focused on engineering design solutions which are based on the following generalizations:

- the rate of devolatilization and the volatile yield are strong functions of the flame temperature,
- the rate of fuel nitrogen release from coal particles is proportional to the rate of volatile evolution and char combustion,
- the conversion of fuel nitrogen to molecular nitrogen is kinetically controlled in fuel rich regions of flames, and its rate is a strong function of local stoichiometry and temperature,

As it is reflected in the above mentioned principles, the fate of fuel

nitrogen is the primary focus of study since fuel nitrogen is known to be the major source of NO_x emission from coal flames. When the objective is to reach ultra low NO_x concentration (e.g., 40-60 ppm) attention also has to be paid to the processes involving the production and destruction of thermal and "prompt" NO_x. For the purpose of this study it has been sufficient to focus on the fuel nitrogen processes since prompt and thermal NO can be controlled to a large extent by the same methods as fuel NO, e.g., staged combustion.

Early ignition of coal flames is important for establishing a high temperature core to effect the maximum rate of release of fuel nitrogen from coal particles into a sub stoichiometric gas phase, and consequently to lengthen the residence time for fuel nitrogen conversion. The ignition process is controlled by the rate of volatile matter evolution, which increases exponentially with increasing temperature.

Gradual addition of oxygen to the evolved volatile matter is also important for burning the fuel as it becomes available during the ignition process. With increasing air to coal mass ratio one can generate more heat. elevating the local gas and particle temperature. Part of this heat will be fed back to the flame front by flame radiation and gas recirculation and will enhance the igition process. If, however, too much oxygen is provided, the mean stoichiometric ratio increases which, in turn, can increase the amount of fuel nitrogen converted to NO_x. It follows, that there exists an optimal air to coal mass ratio, at which the evolution of fuel nitrogen is fast due to high temperature, while the surrounding gas phase is sub-stoichiometric, and favorable for the conversion of fuel nitrogen to molecular nitrogen. This air to coal ratio is undoubtedly coal dependent. The RSFC burner is capable of approaching this optimum coal-air mixing history by the suitable combination of the radial distributions of the combustion air mass flow and of the swirl velocity.

In the second mode of operation described above, the well stirred mixture of transport and primary air with coal creates a nearly uniform distribution of stoichiometric ratio and temperature, which can be optimized to obtain the highest rate of fuel nitrogen conversion to molecular nitrogen. Both the heat release and volatile evolution are fast, and, consequently, the residence time and temperature in the stratified region of the fuel rich zone are increased resulting in lower NO_x concentrations.

Parametric Studies

The experiments were conducted with standard grind Illinois #6 coal (43% vol., 9% ash, 12148 Btu/lb). The Radially Stratified Flame Core (RSFC) burner (Figure 1) was used in the studies. It has three annuli (primary, secondary and tertiary) and independent adjustable swirl generators can be used to set different input conditions such as the amounts of air flow through the primary, secondary and tertiary annuli with variable axial and tangential velocity to obtain optimum radial stratification for a sufficient axial distance or residence time. Experiments were performed to investigate the effect of burner settings and burner air distributions on the NO_x formation and carbon conversion.

The effect of turndown on exit NO_x emission was investigated by increasing the thermal input from 1.0 MW to 1.8 MW for the same burner air distribution. The overall and local stoichiometries were changed by adding air through the burner or the over fire section of the combustion tunnel. The transport air to coal mass ratio was increased from 1.38 to 1.82, for a constant thermal input. Different flame characteristics were also investigated by varying the primary, secondary and tertiary air fractions and swirl numbers, respectively. External staging was achieved by introducing overfire air into the combustion chamber at two axial positions, 10.5 and 13.5 burner diameters downstream of the flame. (The burner diameter is 0.30m.)

The measured exit NO_x concentrations are plotted in Figure 2 as a function of the first stage stoichiometric ratio for the two modes of operation: unstaged and staged. The uncontrolled non-staged flame produced about 700 to 900 ppm exit NO_x concentration. This was reduced to 217 ppm in the radially stratified flame. The minimum exit NO_x concentration, 55 ppm NO_x at 3% O_2, was obtained by a staged RSFC burner flame with a stoichiometric ratio in the primary stage of 0.6.

A mildly staged flame was chosen as the baseline for investigating the effect of transport air to coal mass ratio, thermal input, exit O_2 concentration, air preheat and the location of the over fire air injection (first stage residence time) on NO_x emission. The details of the baseline flame are given in the second data column of Table 1.

The heat input was 1.5 MW and the overall and burner stoichiometric ratio was measured as 1.13 and 0.92, respectively. The total air flow rate was 1460 Nm^3/h. The transport air fraction of the total air was 14.7%.

The coal gun was recesses from the flush position by 5 inches.

High primary air flow rate and swirl number were used to promote early and uniform mixing between coal particles and primary air. The secondary air flow rate was set at the minimum necessary to cool the hot primary barrel and to avoid fly ash deposition on the primary and secondary barrels. The tertiary air fraction was se to 18.5% of the total air and was physically separated from the initial flame front by the secondary air annulus.

The over fire air was introduced X/D=13.5 burner diameters downstream of the flame through an injection bank consisting of two, triple air nozzles. The inner diameter of each of the six nozzles was 1 inch, producing a high momentum air jet which enhanced mixing between the over fire air and the combustion products of the first stage.

Measurements of exit concentrations of the main gas components, O_2, CO_2, CO and NO_x were carried out in steady state conditions. Ash samples were also taken in isokinetic conditions at the exit of the furnace. For the base case, the CO concentration was low, about 33 ppm. The high CO_2 concentration indicates good carbon conversion with low excess air. The carbon conversion was determined to be 99.52% (from the carbon content of the fly ash) by using a Perkin-Elmer analyzer. The exit NO_x concentration was 107 ppm at 3% O_2.

A parametric study was carried out using the flame described above as a base case. The effect of transport air to coal mass ratio (T/C) was determined by increasing the fraction of transport air. The primary or tertiary air flows were adjusted accordingly to keep a constant burner stoichiometric ratio. The exit NO_x concentrations were measured after reaching steady state conditions in each case (Fig. 3). With no compensation on the burner air, the exit NO_x concentration increased from about 105 to 150 ppm, when the T/C was increased from 1.4 to 1.8. The higher NO_x concentrations are due to the higher burner stoichiometric ratio. This considerable increase of NO_x concentration (50%) indicates that the burner stoichiometric ratio has a very strong impact on NO_x formation. When the tertiary air flow was adjusted (reduced) to compensate for the additional transport air, the exit NO_x concentration decreased from about 105 ppm to 80-85 ppm. These results indicate that NO_x exit concentrations are very sensitive to the mass flow rate of tertiary air. On the other hand, when the compensation was made on the

primary air flow, a negligible difference was measured in the exit NO_x concentration. There was little change in the flame shape during these experiments, even when the primary to transport air mass ratio was changed significantly, indicating stable flame characteristics.

The available residence time for the fuel nitrogen conversion in the fuel rich zone is an important design parameter for low NO_x boilers. The effect of the first stage residence time was investigated by injection of air at two separate axial positions -10.5 and 143.5 burner diameters downstream of the flame (Fig. 4). By changing the distribution of air between the injection ports, the mean gas residence time in the first stage of the combustion chamber is easily varied. The maximum variation of residence time achievable with this configuration is estimated at about 600 msec.

During these experiments, the over fire air split between the two inject stations was varied for 5 different cases and the exit gas composition was determined. Introducing all the over fire air at the upstream axial location caused the exit NO_x concentration to increase from 110 ppm to 200 ppm. The correlation between the two variables is linear. This experiment underscores the importance of the mean gas residence time in the fuel rich zone in obtaining low NO_x emission.

The effect of the thermal input was investigated by increasing the coal flow rates while maintaining a constant (transport + primary air)/coal mass ratio. The fraction of tertiary air was allowed to change. The burner SR decreased from 0.92 to 0.82 when the thermal input was increased from 1 MW to 1.8 MW. The variation of exit NO_x as a function of thermal load and burner stoichiometry is shown in Figure 5. At lower load (1 MW) the burner stoichiometric ratio was 0.92 and at this condition, the exit NO_x concentration was 94 ppm corrected to 3% O_2. At maximum load (1.8 MW), the NO_x concentration decreased to about 70 ppm, but the stoichiometric ratio was also lower (0.82). Thus, the exit NO_x concentration is relatively insensitive to thermal load within the range of investigation.

The exit O_2 concentration was also varied between 2.0% and 4.5%, both at the staged (SR=0.92) and non-staged conditions. In thge non-staged flame, the exit O_2 concentration was varied by increasing the fraction of tertiary air. For the staged case, the over fire air fraction was increased, and consequently the burner stoichiometric ratio was

constant. The results show an increase of exit NO_x concentration from 217 ppm to 310 ppm for the non-staged flame (Fig. 2). No change was detected for the staged condition. For the latter case, the exit NO_x concentration was found to be between 103 and 107 ppm, (corrected to 3% O_2). The air preheat was also increased at the staged condition (SR=0.92) from 500 K to 562 K but negligible change (96 to 98 ppm) was detected in the exit NO_x concentration.

Detailed Flame Characteristics

Detailed flame measurements were performed in order to gain insight into the governing mechanisms of NO_x formation. Only a few of those results will be discussed here. The characteristics of the three flames studied are given in Table 1. Two of these flames were staged flames and one was an unstaged flame. The study of staged flames would help determine the effect of staging on the NO_x formation and the comparison between staged and unstaged flames would help explain the difference in the exit NO_x concentrations for these flames . The flame designated as W OFA II in Table 1. has been chosen as the baseline staged flame. The flame has been discussed in the preceding section.

Visual observation of the flames provided valuable qualitative information about the flame length, the radial flame stratification and the depth of intrusion of the transport air jet. The visual features of the two staged cases are similar, the quarl is filled with combustion products. There is no transport air/coal jet visible on the center line of these flames.

The non-staged flame W/O OFA, had different attributes. The burner was operated with a different air distribution for this flame. There was a smaller amount of primary air, a higher amount of secondary air an especially high fraction of tertiary air (see Table 1.). The shape of the flame is different as it does not fill the quarl and the tertiary air is clearly separated from the flame front (radially stratified). At a distance of about three burner diameters downstream of the burner the flow diverges as it approaches a stagnation zone in the flow. This stagnation zone marks the end of the radially stratified core region. Downstream of this, vigorous mixing between the tertiary air and the combustion products occurs.

The different flame shapes indicate that the interaction of the coal laden transport jet with the primary air jet is very important. For the

staged cases, the angular momentum of the primary air is much higer than that of the non-staged stratified case. It follows that the interaction of the transport air jet is signifcantly different in the two cases. Both the visual observation and concentration measurements of solid samples along the axis support this conclusion. In the staged cases, the angular momentum of the primary air was sufficiently powerful to create a stagnation point within the primary barrel, which, in turn, resulted in a well mixed fuel rich primary air-coal flow to emerge from the burner. In the non-staged case, on the other hand, stagnation did not occur as early due to the weak angular momentum of the primary air (low swirl number and low air fraction).

The exit gas composition was measured for each case and the results are summarized in Table 1. Low CO concentrations were detected even for the higher staged flame. The carbon conversion was calculated from the carbon content of the sampled fly ash. The NO_x concentrations are also given in Table 1. The lowest NO_x emission of these flames (70 pm at 3%O_2) was obtained from the staged flame with SR=0.82. This increased to 107 ppm at 3%O_2 for the mildly staged flame and to 217 ppm @3%O_2 for the unstaged flame.

Gas temperature measurements were made for each flame. The measurements were made using a water cooled suction pyrometer equipped with a triple shield and operated at a high gas velocity, so as to minimize errors due to radiation. Axial temperature profiles taken for the two staged and one unstaged test flames are shown in Fig. 6. The advantages of the burner operating mode for staged combustion (W OFA), i.e., the increased fraction of the combustion air as strongly swirling primary burner air flow, can be seen by the steep temperature rise near the burner. This is consistent with high volatile yield and fast decay of fuel-N content of the solid samples along the distance from the burner (Fig. 7).

Early ignition and partial combustion in a well stirred zone close to the burner produces a faster initial decay of the solid carbon as shown in Fig. 8. The carbon burnout continues along the flame to reach high carbon burnout efficiency, in excess of 99.5% for both the staged and unstaged flame.

The NO and NO_x concentrations measured along the flame W/O OFA are shown in Fig. 9. The large differences between NO and NO_x levels close to the burner are in the flame zone where fuel bound nitrogen, HCN, NH_3

show up as NO_x. As the NO and NO_x values merge at around x/D=10 an NO_x level very close to that of the emission from the flame > 217 ppm ($3\%O_2$), is established. Under staged combustion conditions (W OFAII) the NO_x decays faster to an axial distance from the burner of x/D=6 and there is no further change to that level, 107 ppm ($3\%O_2$), to the end of the flame (Fig. 10).

Conclusions

Experimental work was carried out at the MIT Combustion Research Facility with the MIT Radially Stratified Coal Burner with an Illinois #6 coal. The highlights of the results are:

- 217 NO_x ($@3\%O_2$) concentration with unstaged stratified flame at 1.08 stoichiometric ratio and 99.54% carbon conversion,
- 107 NO_x ($@3\%O_2$) concentration with staged flame at 0.92 first stage stoichiometric ratio and 99.52% carbon conversion,
- 70 NO_x ($@3\%O_2$) concentration with staged flame at 0.82 first staged stoichiometric ratio and 99.3% carbon conversion.

The parametric study showed relative insensitivity of exit concentrations to thermal load (between 1 and 1.8 MW) air preheat (510-560 F) and exit O_2 concentrations (2.5-4.5%). The effect of transport air on exit NO_x concentration depended on where the air compensation was applied at staged conditions.

It is noteworthy that the very low NO_x emissions form both the mildly staged and more deeply staged flames were obtained with a radial air flow distribution in which more than 50% of the total air flow was introduced through the transport plus primary air conduits. This has resulted in the maximum yield of volatile matter, stable ignition and a steep initial temperature rise, factors which are conducive to the production of a hot flame core, a prerequisite of low-NO_x operation.

The in-flame measurements point to engineering solutions leading to further improvement of the RSFC burner performance. It seems that reduction of ignition delay resulting in steep temperature rise near the burner are the key parameters of low NO_x emission.

Acknowledgments

The authors are grateful for the stimulating discussions with the B&W visiting team, Jennifer Sivy and John Koslosky. They thank Tony

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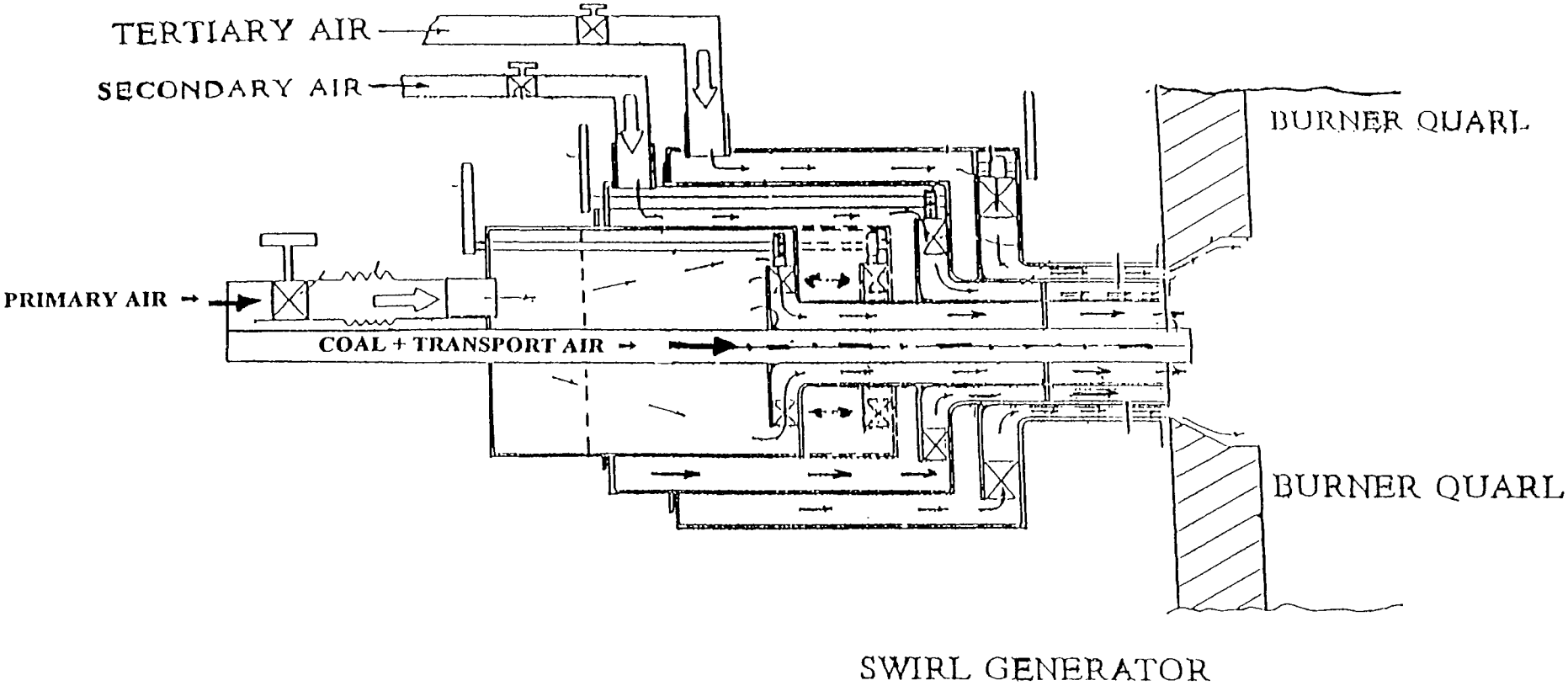
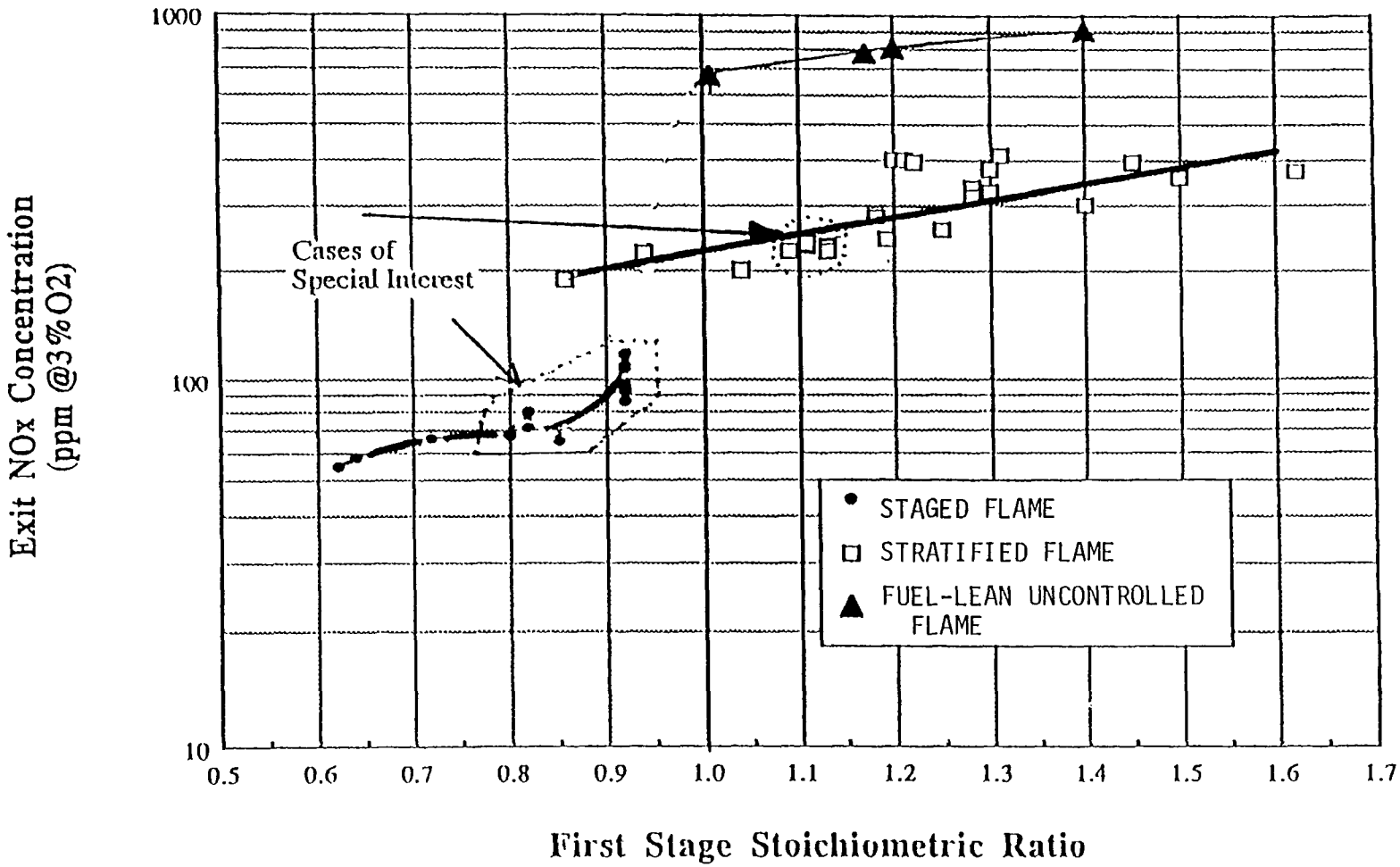


Fig 1. Schematic of the MIT-RSFC Burner (U.S. Patent Pending)

Fig. 2 **Effect of Stoichiometric Ratio on Exit NOx**
All Experimental Data



Characteristics of the Three Reference Cases

	w OFA I	w OFA II	w/o OFA
Heat Input (MW)	1.5	1.5	1.5
Overall SR	1.21	1.13	1.08
First Stage SR	0.82	0.92	1.08
Total Air (Nm3/h)	1628	1460	1446
Transport/Coal Ratio	1.5	1.5	1.5
Transport %	13.5	14.7	15.1
Primary % / Swirl No.	37.2 /2.35	44.8/2.35	12.7 /1.21
Second. % / Swil No.	5.3/2.52	6.9/2.52	27.2 /1.63
Tertiary % / Swirl No.	11.6/1.66	18.5/1.66	45 /1.66
OFA %	32.4	15.6	0
CO (ppm)	45	33	30
CO2 (%)	15.5	15.9	16.7
O2 (%)	3.7	2.0	1.5
NOx (ppm @3% O2)	70	107	217
Carbon Conv. (%)	99.3	99.52	99.54

Fig. 3 **Effect of Transport Air to Coal Mass Ratio on Exit NOx Concentration With OFA, SR=0.92**

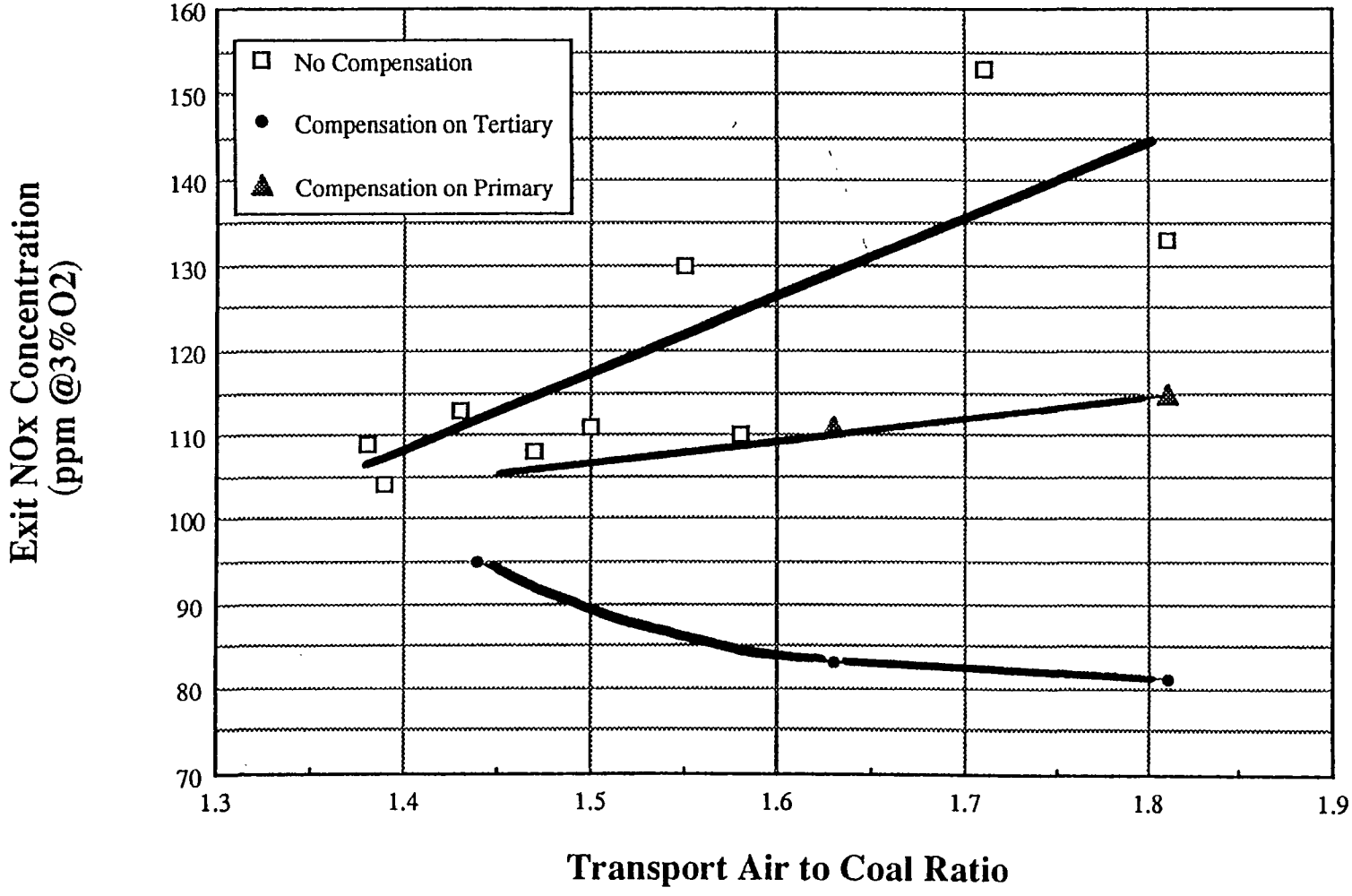


Fig. 4 **Dual Overfire Air Injection x/D=10.5 and x/D=13.5 (SR=0.92)**

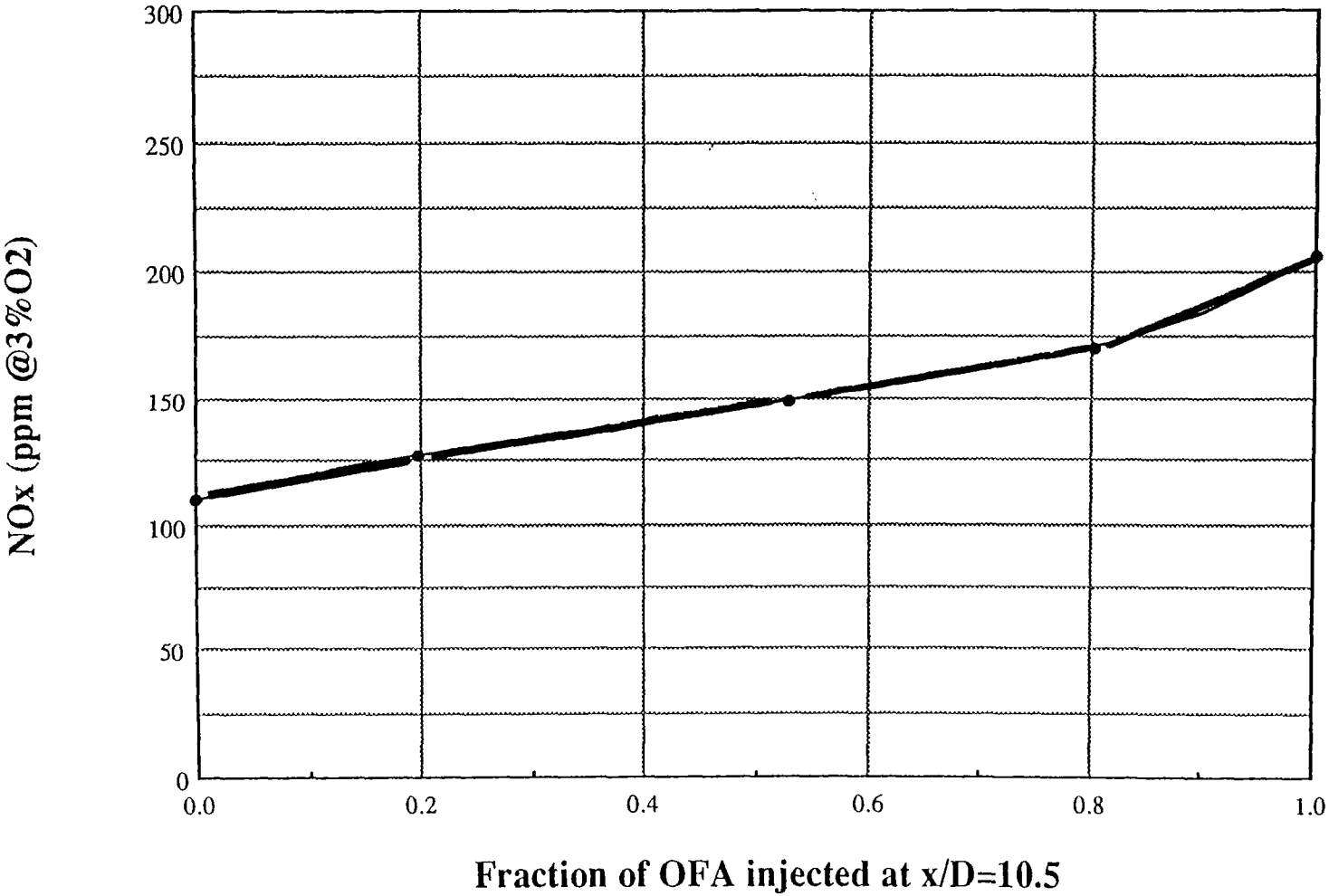


Fig. 5 **Effect of Thermal Input on Exit NOx**

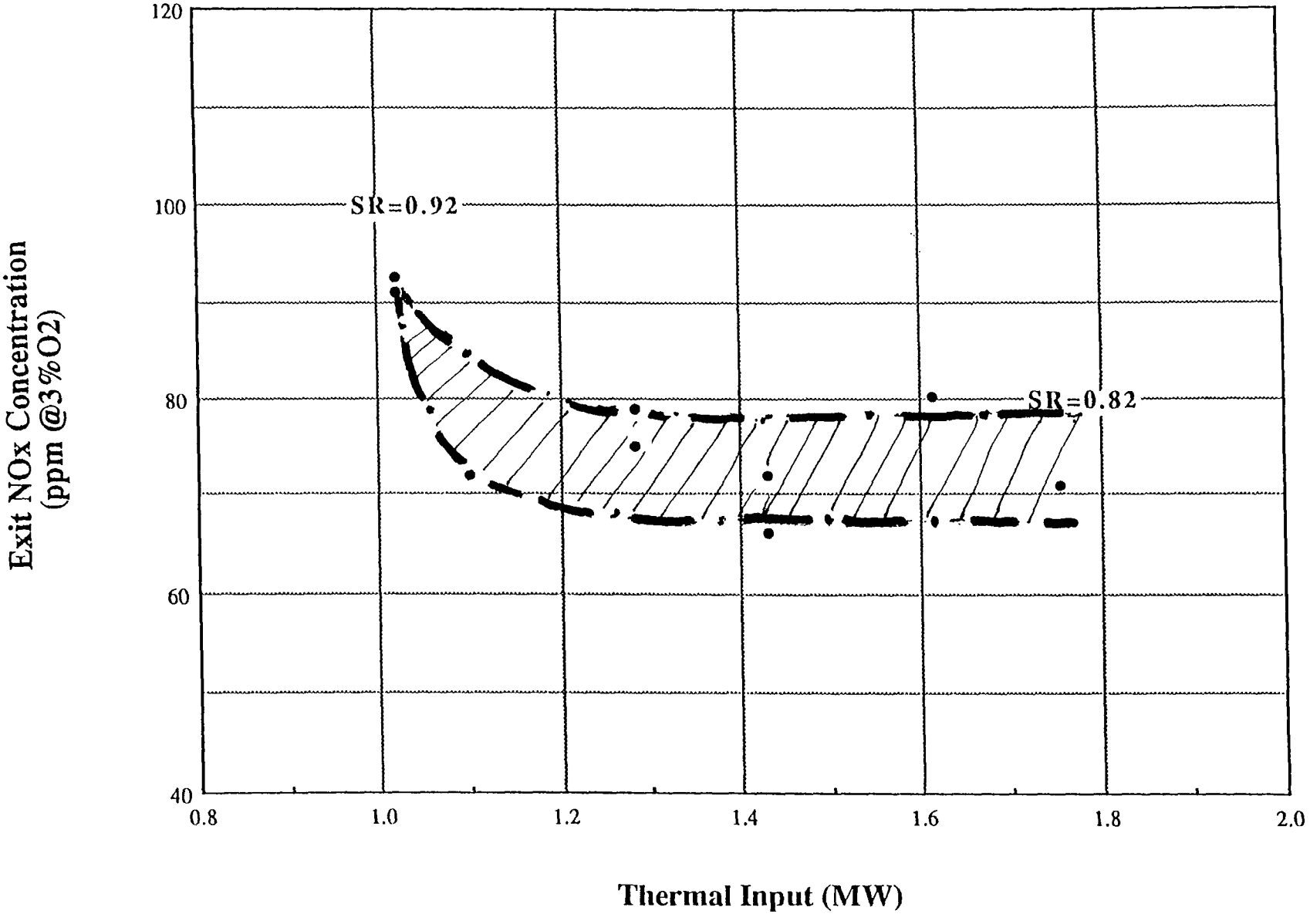


Fig. 6 **Axial Profiles of Temperature**

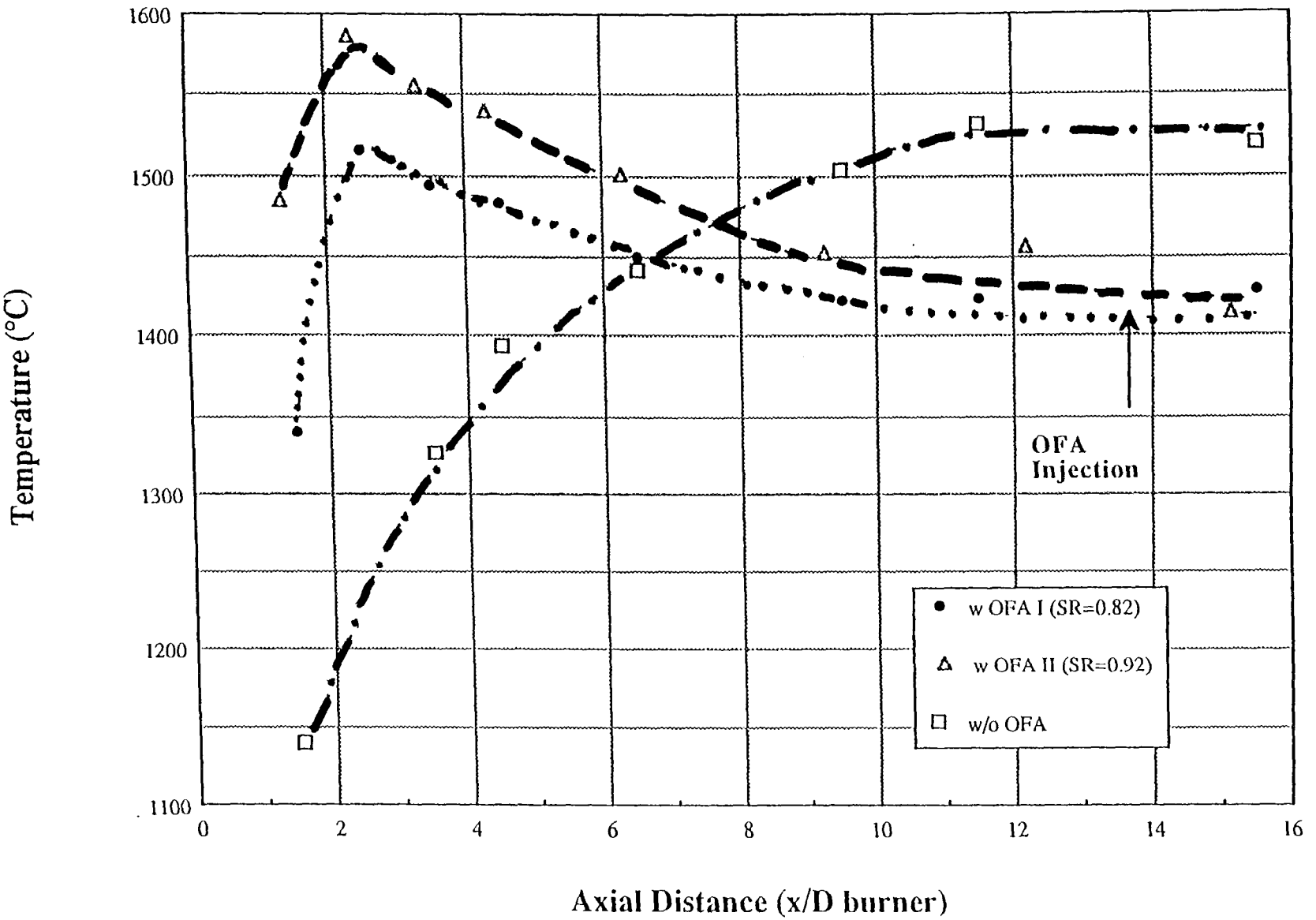
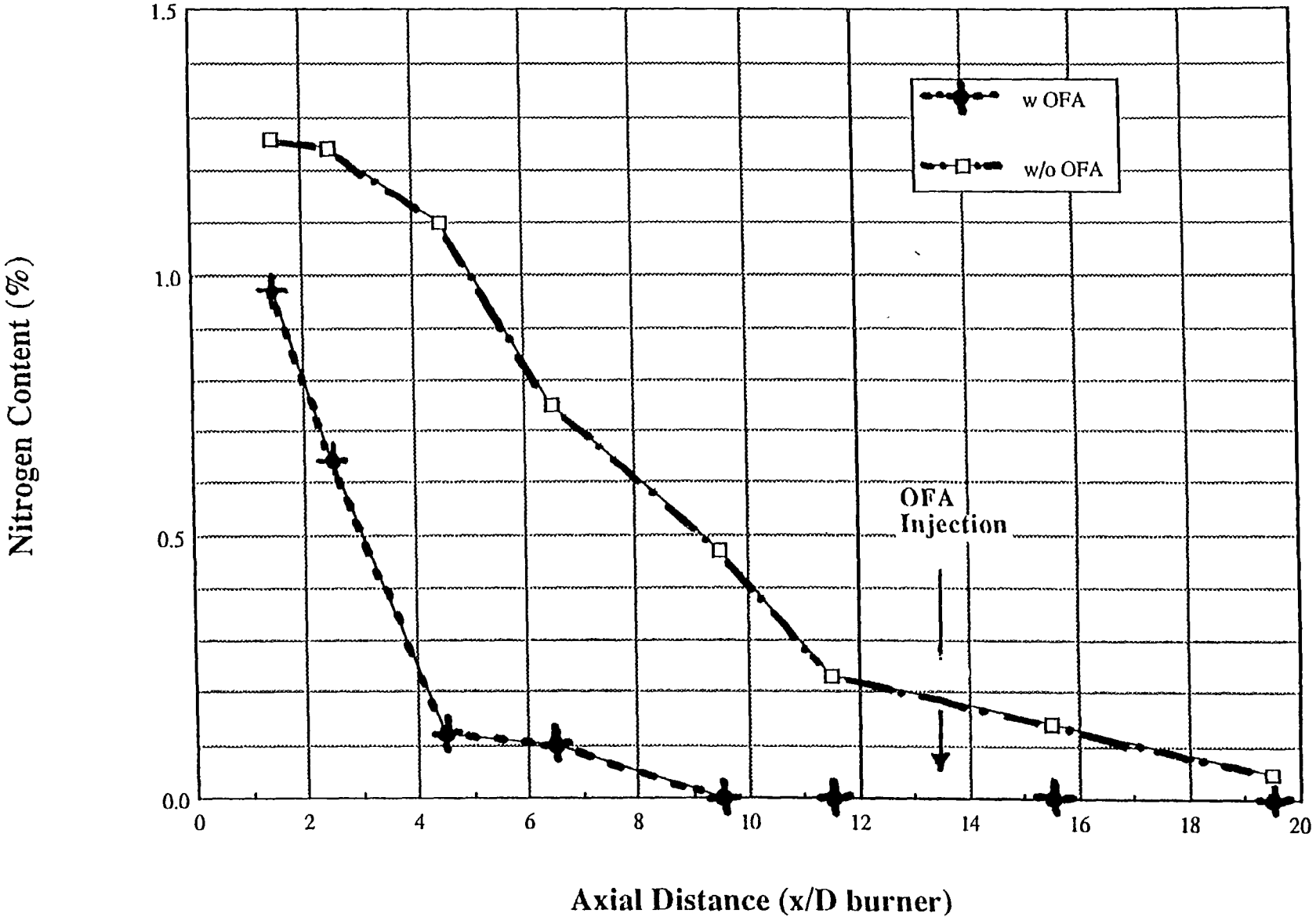


Fig. 7 **Axial Profiles of Nitrogen Content of Solid Samples**



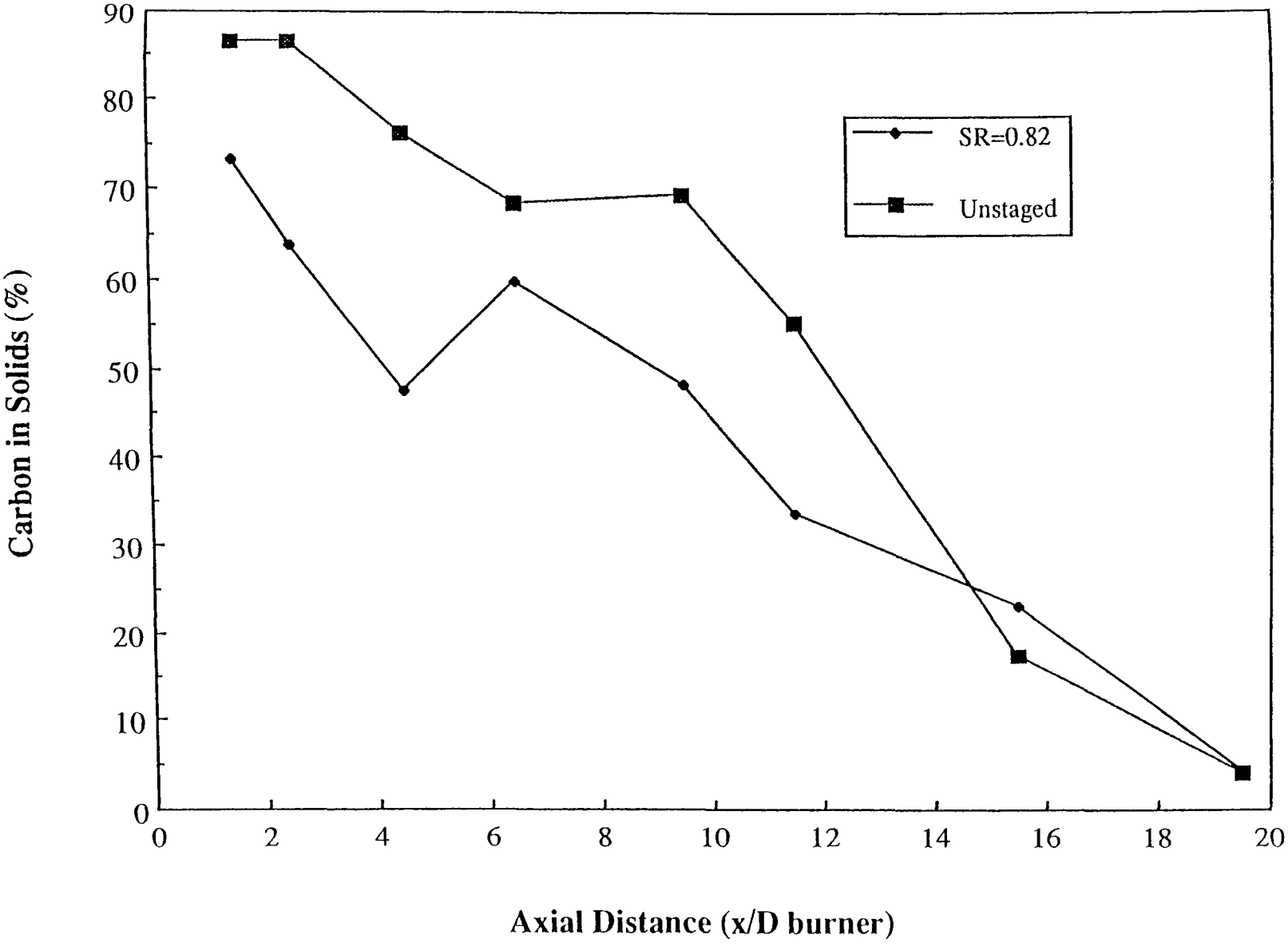


Fig. 8 Axial Profile of Carbon Burnout

Fig. 9 **Axial NO and NOx Profiles Without OFA**

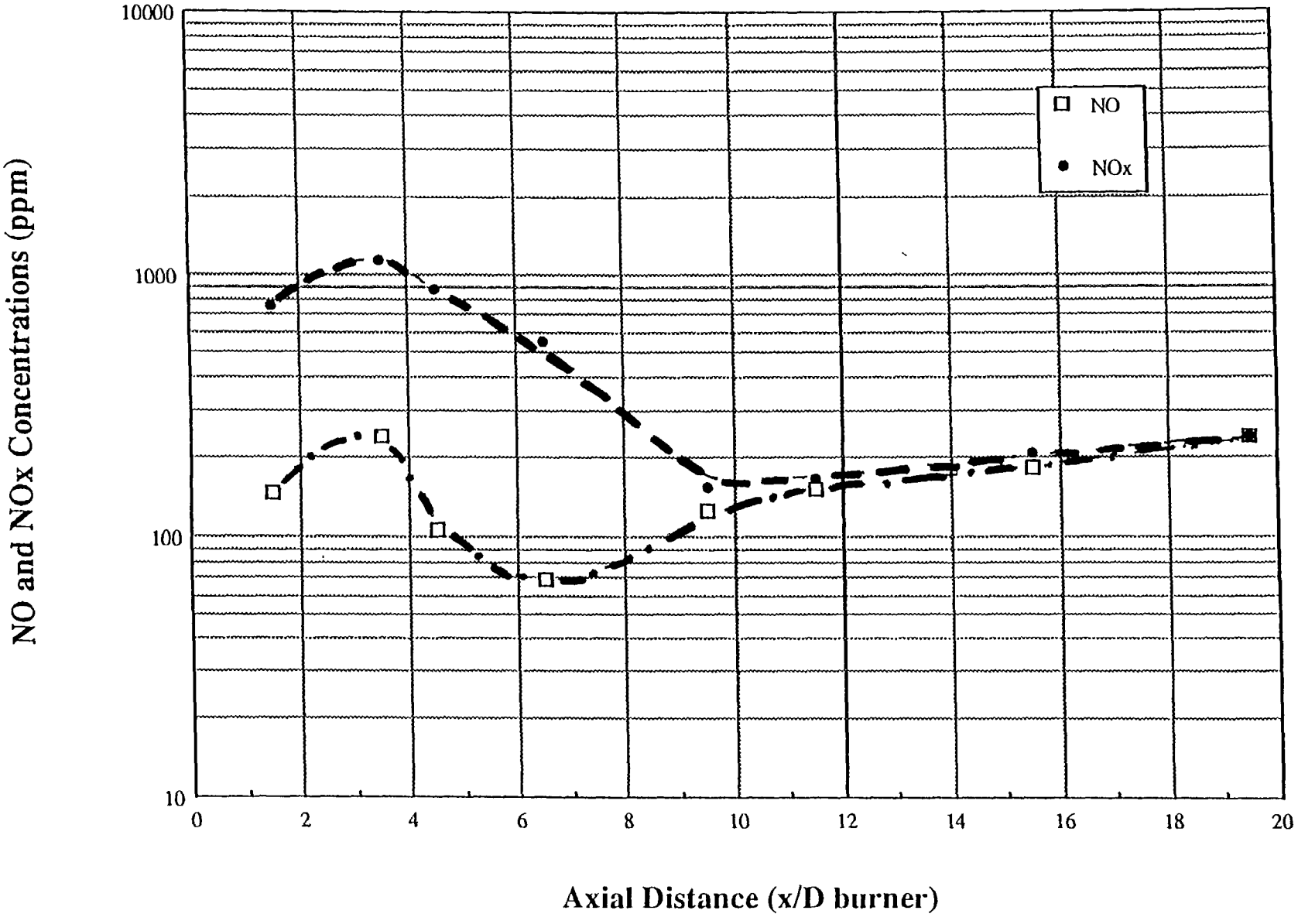
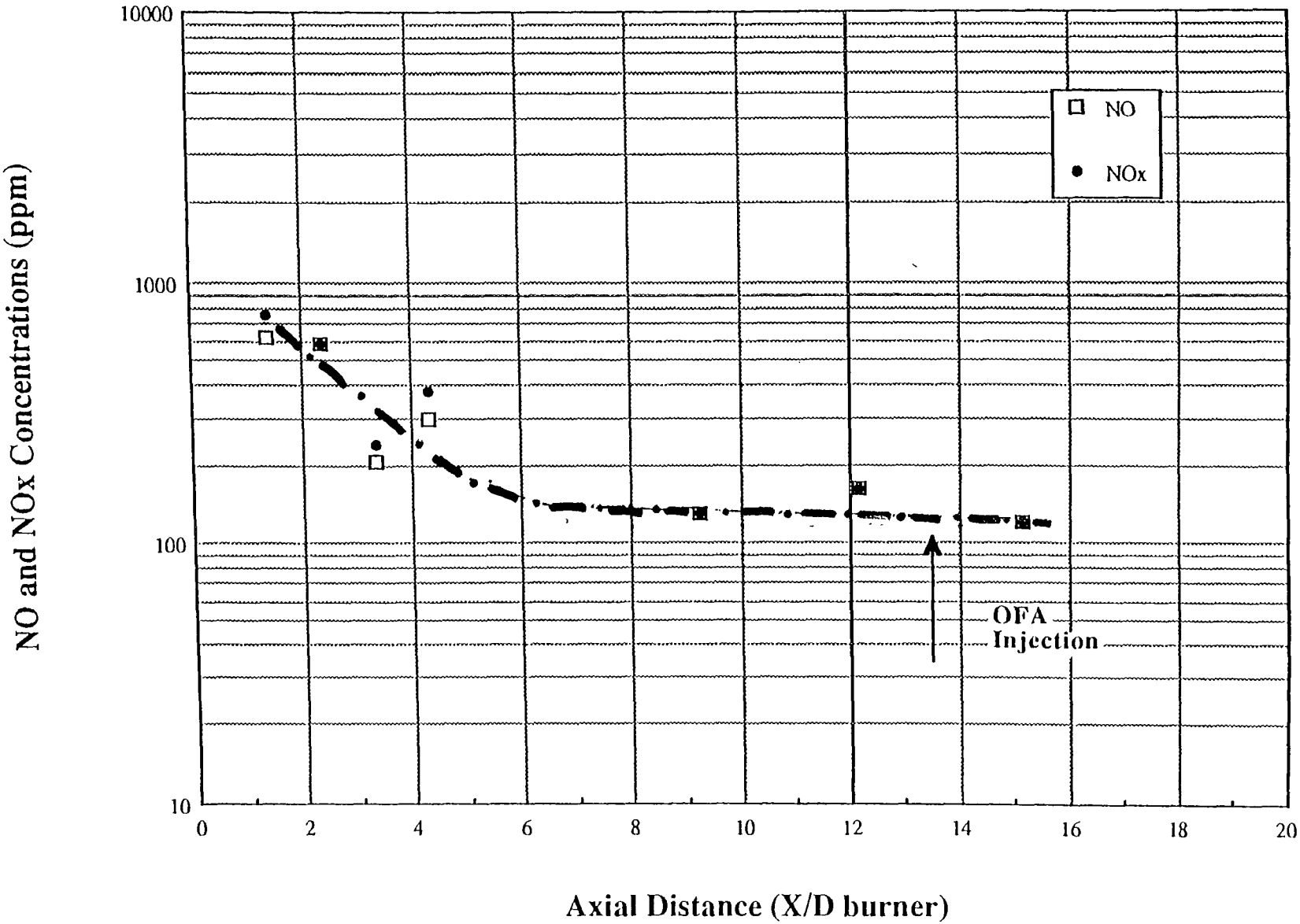


Fig. 10 **Axial NO and NOx Profiles With OFA, SR=0.92**



DEVELOPMENT AND TEST OF DIFFERENT METHODS TO IMPROVE THE DESCRIPTION OF NO_X EMISSIONS IN STAGED COMBUSTION

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Abstract

Two methods to improve the modeling of NO_x emissions in numerical flow simulation of combustion are investigated. The models used are a reduced mechanism for nitrogen chemistry in methane combustion and a new model based on regression analysis of perfectly stirred reactor simulations using detailed comprehensive reaction kinetics. The applicability of the methods to numerical flow simulation of practical furnaces, especially in the near burner region, is tested against experimental data from a pulverized coal fired single burner furnace. The results are also compared to those obtained using a commonly used description for the overall reaction rate of *NO*.

Introduction

Air and fuel staging are effective techniques to reduce the NO_x formation. These techniques rely on the fact that under fuel rich conditions and suitable temperature *NO* previously formed is reduced to molecular nitrogen. Consequently, to model staged combustion successfully, the fuel and nitrogen chemistry must be well described also at fuel rich conditions.

In applications where computational fluid dynamics (CFD) is used to model combustion, an often used approach is to neglect reaction kinetics and assume that the combustion rate is limited by the mixing rate. The prediction of the NO_x level is, however, dependent on a correct description of the

fuel chemistry. The simplest level of reaction kinetics for fuel oxidation is to assume that the reaction rate can be described with a single irreversible reaction. The exclusion of the intermediates CO and H_2 formed in the hydrocarbon breakdown leads to an overprediction of the heat release[1] In fuel lean cases, it is possible to obtain a good description of the fuel oxidation assuming a mixing limited breakdown of the hydrocarbons to CO and H_2O, followed by a mixing or kinetically controlled oxidation of CO. In the literature, more complicated descriptions of the fuel oxidation process can be found[2,3]. Several expressions for the global reaction rate of the CO oxidation have been proposed[1].

For simple hydrocarbons, e.g. methane, comprehensive schemes based on elementary reactions has been proposed. Elementary reactions leading to the formation of NO can also be included in these schemes. In CFD analysis of practical applications the direct use of such full mechanisms, consisting of hundreds of elementary reactions, is computationally too demanding. Using the strategy outlined by Peters and co-workers[4,5], it is possible to reduce the comprehensive mechanism to only a few formal reactions. One such reduced mechanism is the reduced four-step mechanism for nitrogen chemistry in methane combustion derived for the perfectly stirred reactor environment[6] The use of a reduced mechanism restricts the number of species for which transport equations must be solved to a number that can be handled in CFD applications.

The effective rates of the homogenous gas phase reactions in combustion are controlled mainly by the mixing caused by turbulence. For reactions with a rate comparable to or slower than the mixing rate, it is essential that the turbulence-chemistry interaction is taken into account. Several methods for this have been put forward. One such model is the eddy dissipation concept[7]. Here, the combustion is assumed to take place in isolated regions of the fluid, so called fine structure. Within the fine structure the reacting components are mixed on a molecular level. In the model, it is assumed that these parts can be modeled as perfectly stirred reactors.

In pulverized coal combustion, not only the homogenous gas phase reactions are of importance. For successful modeling, detailed information about the pyrolysis stage and the ignition of the char particles, among others, are also required. Best results are obtained if empirically determined data are available. Unfortunately, the detailed composition of the pyrolysis gas is usually unknown and the necessary data is difficult to obtain. Consequently, approximate expressions describing these processes are often applied[8-12].

The purpose of this work was to improve the description of homogenous fuel and the nitrogen chemistry used in CFD simulations of staged combustion. Two models are presented. As a first test case for their applicability to CFD simulations, a 2.5 MW pulverized coal fired single burner furnace was simulated.

The homogenous gas phase models

A reduced mechanism for nitrogen chemistry

A reliable description of the NO_x chemistry is dependent on a good description of the fuel chemistry, including the heterogeneous processes. For the homogenous chemistry, among the

most complete descriptions possible to incorporate into CFD are reduced mechanisms. The formal reactions obtained by Glarborg *et al.* for methane combustion in a perfectly stirred reactor are given below[6],

$$CH_4 + 2H + H_2O \rightleftharpoons CO + 4H_2$$
$$CO + H_2O \rightleftharpoons CO_2 + H_2$$
$$2H + M \rightleftharpoons H_2 + M$$
$$3H_2 + O_2 \rightleftharpoons 2H_2O + 2H$$

where M represents any third body. The reaction rates for the formal reactions can be calculated from the sum of the rates of certain important elementary reactions. Using steady-state balances for N, NH, NCO and CN, Glarborg *et al.* obtained the reaction rates for HCN and NO in methane combustion. The mechanism was developed for high temperatures ($T \geq 1500$ K) and for fuel lean to moderately fuel rich conditions, $0.7 \leq \lambda \leq 1.4$, where λ is the air to fuel ratio.

A simplified model for hydrocarbon oxidation

In the literature, a number of proposals on simplified fuel oxidation descriptions can be found. In fuel rich regions, a two-step description with only CO as an intermediate leads to an overprediction of the local temperature. Omitting H_2 also affects the predicted air to fuel ratio. One of the difficulties encountered when including H_2 is to decide the proportion between H_2 and H_2O formed during the hydrocarbon breakdown. Here, a new approach has been used were the proportions have been calculated from the global equilibrium[13]. Using this approach, the oxidation of a hydrocarbon can be described by

$$C_nH_m + (\frac{n}{2} + \alpha\frac{m}{4})O_2 \longrightarrow nCO + \left(\frac{m}{2} - \alpha\frac{m}{2}\right)H_2 + \alpha\frac{m}{2}H_2O$$
$$CO + \frac{1}{2}O_2 \longrightarrow CO_2$$

where α is the ratio between the mole fraction of H_2O to the sum of the mole fractions of H_2O and H_2 obtained from the equilibrium composition. If H_2 and H_2O are present in the fresh mixture, these too are assumed to form H_2 and H_2O in the same proportions. The oxidation rate for CO given by Howard *et al.* is used to describe the second step[14]. The oxidation of H_2 at fuel lean condition is assumed to be mixing limited.

A regression-based NO_X model

The models mainly used for calculating the overall reaction rates for the NO formation do not contain mechanisms accounting for the reduction of NO to molecular nitrogen at fuel rich conditions[15,16]. Numerical modeling of the destruction and formation of NO in a perfectly stirred reactor with the scheme given by de Soete reveled surprisingly large deviations from the results obtained using the comprehensive mechanism. This can also be seen in **Fig 1**. Therefore, the development of a completely new model capable of modeling the NO_x chemistry under conditions encountered in staged combustion and applicable to CFD simulations was motivated.

From the calculations with the full mechanism, an abrupt change in the concentration profile was obtained for short residence times. This abrupt change was due to the ignition of the system. Much of the conversion of the nitrogen containing species took place during the ignition. Longer residence times did not affect the conversion grade to the same extent. Since it was obvious that the radical chemistry played a significant role, accurate predictions of the radical concentrations would be needed in the Arrhenius-type rate expressions. Since the fuel oxidation process usually is described in a very simplified manner, it was realized that a regression based model for the NO_x chemistry would be a way to overcome this problem.

The results obtained from calculations on a perfectly stirred reactor with the comprehensive mechanism were used as reference data for the regression model. From the simulations with the comprehensive scheme it was noticed that some simplifications could be made. First, the conversion grade for the nitrogen containing species included in the model. ,i.e. *NO*, *HCN* and NH_3, are to a great extent independent of the incoming concentrations. It was also noticed that the only important synergistic effect in the reactions could be found between *NO* and *HCN*. Using these approximations it was possible to apply regression analysis to construct expressions for the conversion of *NO* to NH_3, *HCN* and N_2, and comparable expression for the conversion of *HCN* and NH_3 separately. A regression expression was included to account for the synergetic effect noticed. This procedure was repeated for five discrete residence times, i.e. 0.1 ms, 1 ms, 10 ms, 100 ms and 1s. For intermediate residence times the conversions can be obtained by means of interpolation. The regression analysis was done using the backward elimination technique requiring a 95 % significance for the independent variables retained. As independent variables were the air to fuel ratio, the temperature, second order terms of these, as well as products tested. Best results were obtained excluding the point related to non-ignited systems from the reference set and using a separate regression equation for estimating the ignition time during the prediction step. An example of the accuracy of the regression based NO_x model is shown in **Fig 1**.

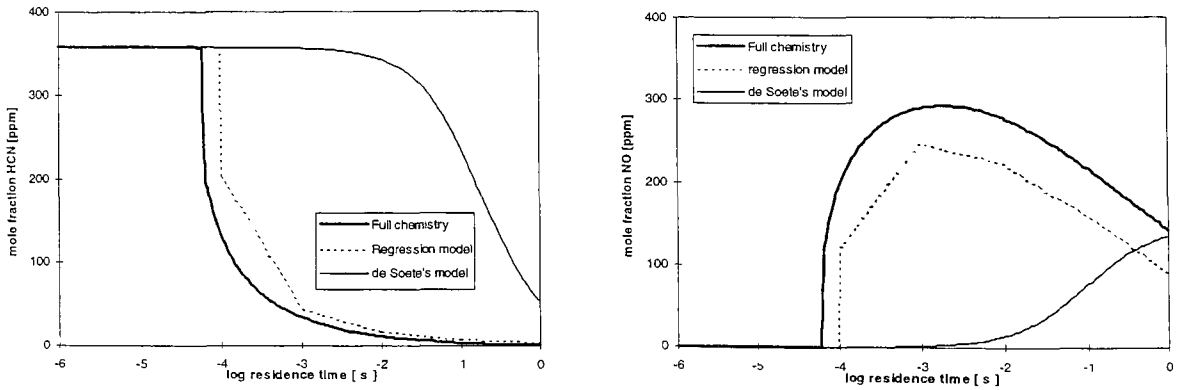

Figure 1. To the left, the mole fraction of *HCN* remaining as a function of residence time. To the right, the mole fraction of *NO* formed. The incoming gas mixture contained 360 ppm *HCN*. The temperature was 1560 K and the stoichiometric air to fuel ratio, λ,was 1.1. The fuel consisted mainly of methane.

The reference data were calculated varying the temperature from 1300 to 1850 K and the air to fuel ratio from 0.5 to 1.5. The fuel was assumed to consist of mainly methane. Later, additional studies were made to investigate the significance of the fuel composition. The findings were incorporated to the model through a parameter describing the fraction of the fuel consisting of hydrocarbons.

When using the regression based NO_x model, it is important that the conditions respond to those the model was set up for. This restricts the use of the regression-based NO_x model to systems were the main part of the NO_x is formed from fuel-*N*, and the thermal-*NO* can be neglected.

The applicability of the chemical schemes to flow simulations

The chemical models presented above have proven to model the gas phase reactions satisfactory for perfectly stirred reactor conditions[6,13]. It is difficult to assess the accuracy of models for the fuel and nitrogen chemistry from a multi-dimensional flow simulation of a pulverized coal combustion. The near burner fields of temperature and concentrations of the main species depend strongly on e.g. the pyrolysis and the char ignition which in practical applications usually are inadequately know. The combustion of the pyrolysis products and the formation of *NO* from the nitrogen released from the fuel occur mainly in the near burner region. Consequently, the role of the present simulation in the assessment of the chemical reaction models is in the first place to see the applicability of the models to CFD analysis. Key questions are then the influence of the chemical submodels on the convergence and computing time, and their range of validity with respect to prevailing conditions in the furnace.

For the flow simulations a computational environment ARDEMUS, developed by VTT and Imatran Voima Oy for furnace analysis, was used[17]. The turbulence-chemistry interactions were modeled using the eddy dissipation concept[7]. With this model it is possible to describe the influence on the mean reaction rate from mixing, and from chemical reactions described with a multi-step reaction mechanism. A multiple-time scale turbulence model was used for the flow simulations[18]. A closer description of the submodels used can be found elsewhere[13,17]. As a first test case a 2.5 MW furnace located at the International Flame Research Foundation (IFRF) was used. Extensive measurements made by the IFRF are available as well as a detailed description of the furnace and the operational parameters[19,20].

Results from the simulations

The experience from the present study indicates that the investigated models can be used in CFD simulations. With the reduced mechanism for nitrogen chemistry in methane combustion, the main species can be calculated separately from *NO* and *HCN*. For the main species a six-fold increase in the computational time compared to a two-step description of the fuel oxidation was obtained. The improved fuel oxidation mechanism with H_2 and *CO* as intermediates resulted in a only slight increase of computational time. In **Fig 2** are the oxygen and temperature distributions obtained with the overall mechanism shown. In **Fig 3** is the sensitivity with regard to the pyrolysis models and ignition model demonstrated. Here the fuel chemistry was described with a two-step

irreversible reaction assuming that CO is the only intermediate. The sensitivity of the near burner predictions to the description of the pyrolysis has been discussed by Costa *et al.*[21].

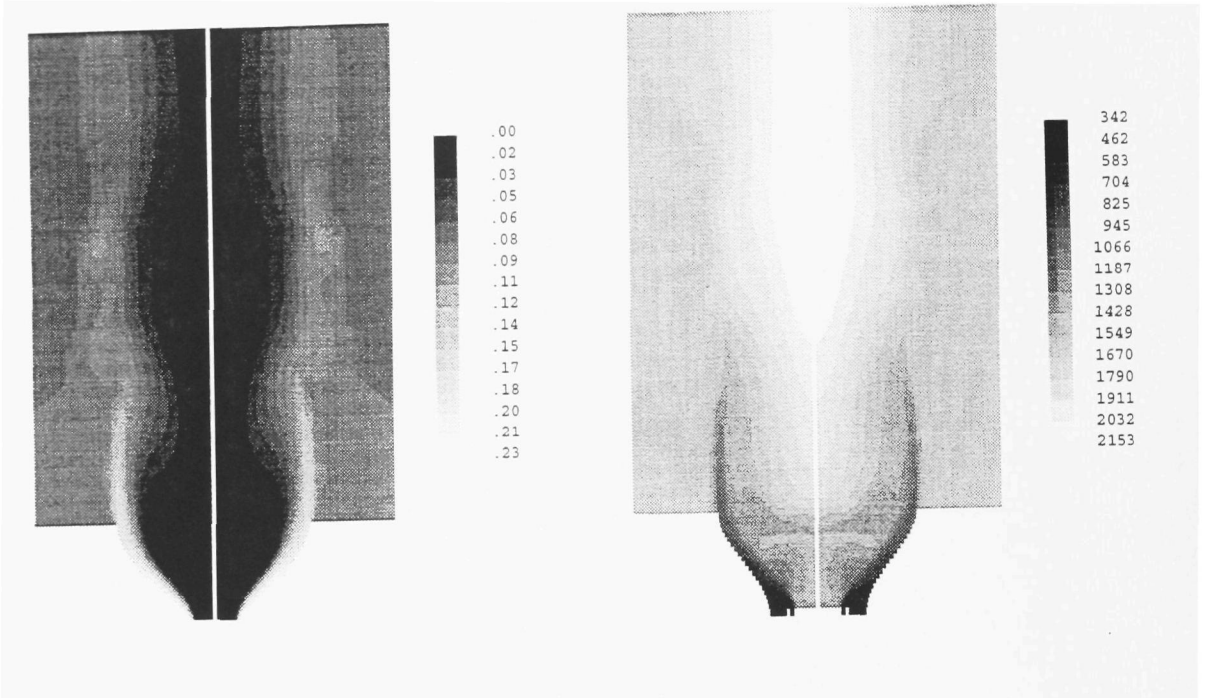

Figure 2. To the left, the mass fraction of oxygen in the near burner region obtained with the global mechanism including H_2 and CO. To the right, the temperature distribution in Kelvin.

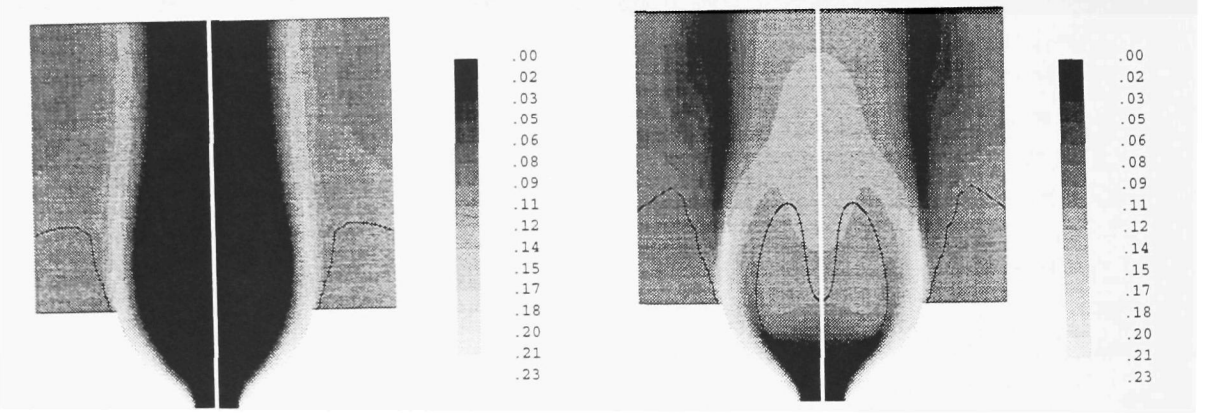

Figure 3. To the left, the oxygen distribution in the near burner region obtained using a two-step irreversible oxidation scheme with CO as the only intermediate. The description suggested by Solomon *et al.* for the pyrolysis rate was used[8]. The particles were assumed to ignite as they reach 1000 K. To the right, the same description of the gaseous fuel chemistry has been used, but the pyrolysis was described according to Fu *et al.*[9], and Zhang *et al.*[10]. For the ignition, a more detailed model was also used[11,12]. The concentrations are expressed as mass fractions. The boundary of the internal recirculation zone is indicated.

Since *NO* and *HCN* can be calculated subsequently to the flow, temperature and species distribution, the calculation times needed for the nitrogen containing species did not cause any concern. In the flow simulations, the local conditions were such that the stoichiometry range for the reduced mechanism was partly exceeded. With the regression-based model, similar problems were encountered. Especially the local *HCN* concentrations were outside the range for the regression-based NO_x model. In **Fig 4**, the *NO* and *HCN* levels calculated with the regression-based model are shown.

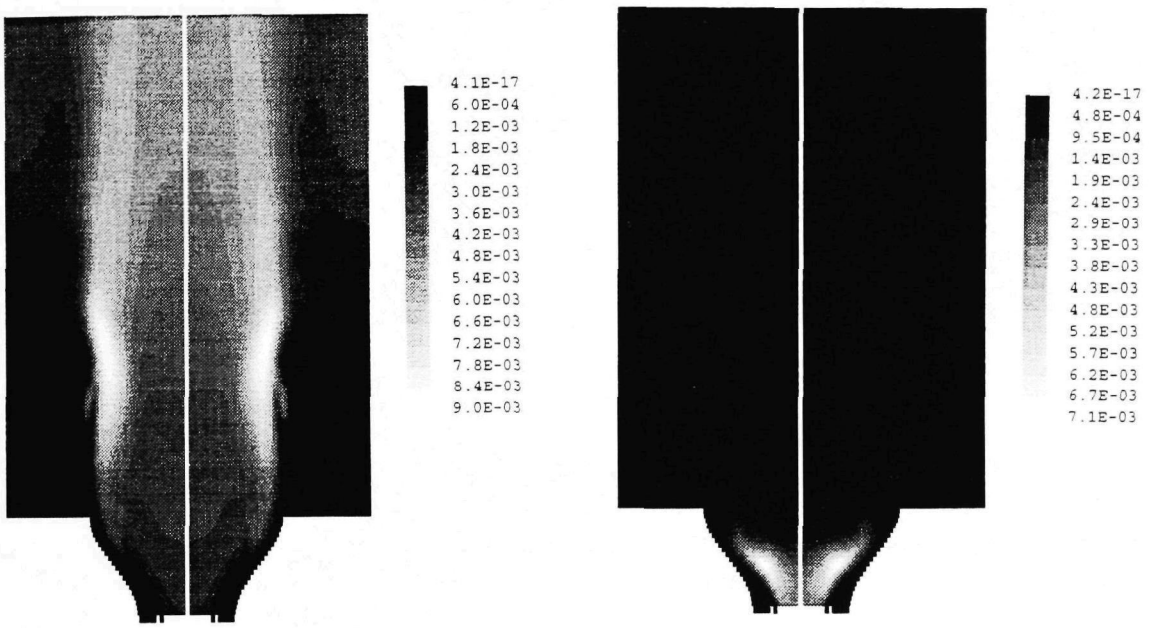

Figure 4. To the left, the *NO* distribution, and to the right the *HCN* distribution obtained with the regression-based NO_x models. The concentrations are expressed as mass fractions. The combustion of the pyrolysis gases has been modeled with the global model including H_2 and *CO* as intermediate. The pyrolysis rate has been modeled according to Solomon *et al.*[8].

The same case was also simulated with the de Soete model. To account for the thermal *NO* production, a model for the Zeldovich mechanism was added. Depending on the assumptions for pyrolysis, char ignition and influence of turbulence on the mean reaction rate, calculated values at the furnace outlet of 460 ppm, 745 ppm and 1040 ppm by volume in dry gas were obtained. These values can be compared to the measured value of 820 ppm reported by IFRF. The *NO* and *HCN* distribution obtained using the de Soete model is shown in **Fig 5**. The oxygen distribution used with the de Soete model is shown to the right in **Fig 3**.

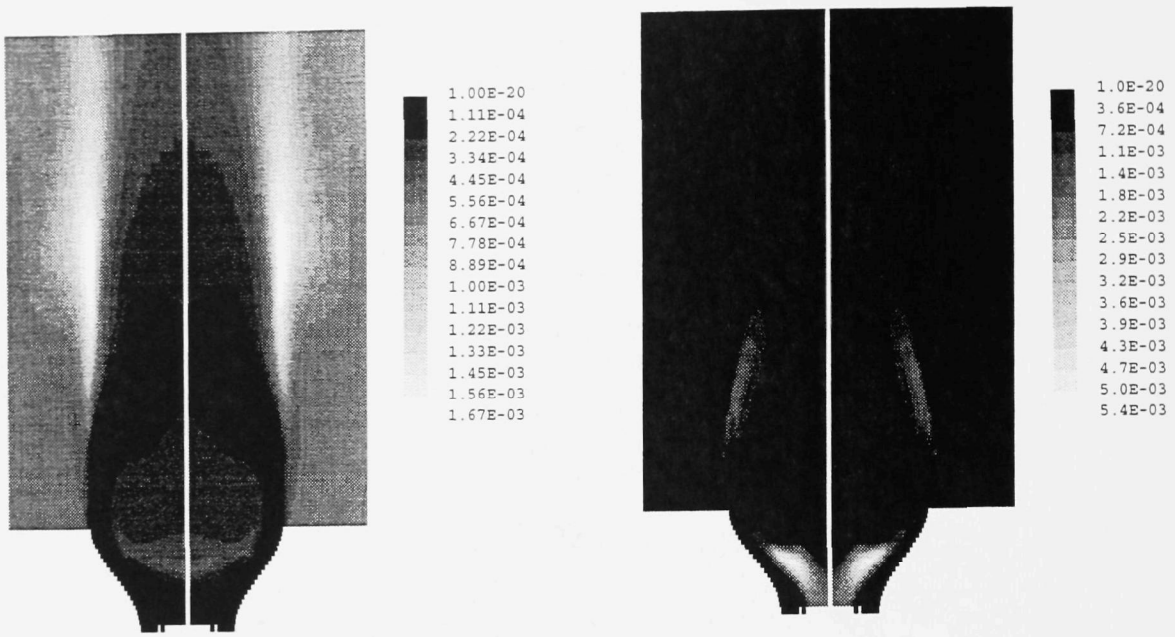

Figure 5. To the left, the *NO* mass fractions obtained with the de Soete mechanism. To the right, the *HCN*. To account for the thermal *NO* production, the Zeldovich mechanism has been added. The fuel oxidation was modelled using a two-step irreversible mechanism including *CO* as the only intermediate. Only the near burner region is shown.

Discussion

The aim with this work was to improve the description of homogenous fuel and nitrogen chemistry used in multidimensional flow simulations of pulverized fuel combustion including staged combustion. The pulverized coal fired single burner furnace used as a first test case could not reveal the applicability of the two models to reburn conditions. There was a fuel rich internal recirculation zone in the near burner region, but the conditions here were not comparable to those in staged combustion. Both models studied could be used without numerical complications. Since most of the reactions governing the NO_x release occur in the near burner region, the use of the computationally more demanding reduced mechanism can preferably be restricted to this region.

The influence the description of the pyrolysis stage and the char ignition has on the predicted distribution of the main species makes it difficult to draw conclusions regarding the accuracy of the gas phase chemistry from the test case. When compared to a detailed reaction mechanism in a well defined environment, the chemical models were found reasonably accurate. The local *HCN* mass fraction close to the injection point of the pulverized coal was outside the range the regression-based NO_x model has been tested for. It can, however, be anticipated that an effective conversion of *HCN* to *NO* occurs under the prevailing temperatures and air to fuel ratio encountered in parts of this zone. The result obtained for *NO* with the de Soete model were in the near burner region in closer agreement with the measurements made at IFRF. This simulation was, however, done with a fuel oxidation model inaccurate for rich conditions.

Although the simulations predicted rather low H_2 concentrations and in a limited region, the effect of incorporating H_2 was greater than expected. The formation of H_2 in the fuel rich near burner region have influence on heat release, the local stoichiometry, the local absorption coefficient and temperature, and finally on the behavior of the coal particles and on the flow field. This influence is enhanced by the internal recirculation zone. Unfortunately, no information about the H_2 levels was available. Regardless of how the fuel chemistry or the heterogeneous processes were modeled, the temperature and concentrations of the main species predicted at the outlet were similar in all cases.

Conclusions

This paper shows that the reduced model for nitrogen chemistry in methane combustion as well as the improved combustion and the new regression-based NO_x models can be used in flow simulations. The effect of the models on convergence was small and the computational times tolerable. Due to the sensitivity of the results with respect to how pyrolysis and char ignition are modelled, an improvement in the results from the flow simulation of the pulverized coal combustion studied was not obtained. The addition of H_2 as an intermediate combustion product had a clear influence on the results for the near burner region. The results are, however, encouraging, and further studies including staged combustion will be undertaken.

Acknowledgments

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Session 7B
SCR/Hybrid

SCR PERFORMANCE EVALUATION AND TROUBLESHOOTING ON A NATURAL GAS-FIRED SOURCE

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Abstract

This paper presents an engineering evaluation of SCR performance on a natural gas-fired 110 MW gas turbine/heat recovery steam generation plant. The purpose of the evaluation was to mitigate excessive ammonia slip from the system. This paper presents results of the engineering assessment of the system design which culminated in a decision to relocate and redesign the ammonia injection grid. The program entailed development and execution of field performance tests for gas flow distribution, process temperature, and NH_3/NO_x uniformity. An evaluation conducted by Carnot included: a technical assessment of anhydrous ammonia injection nozzles in a custom-designed cold flow test apparatus, SCR efficiency at varied NH_3/NO_x mole ratios, and assessment of injection grid design. The modifications were implemented by AES, and the system currently operates within compliance limits of 7 ppm NO_x and 20 ppm NH_3, corrected to 15% O_2. This is a "lessons learned" paper including useful insights for SCR design and troubleshooting, emphasizing design features necessary to control ammonia slip. The results and methodology are applicable to any SCR system, particularly on gas-fired sources.

Introduction

Applied Energy Services (AES) operates a natural gas-fired 110 MW gas turbine/heat recovery steam generation (HRSG) plant in Placerita, California. The plant has two HRSGs, each equipped with a selective catalytic reduction (SCR) system. The combustor utilizes steam injection to control turbine outlet NO_x to approximately 30 ppmc. In July, 1992 the SCR unit on one of the boilers was shown to require excessive NH_3 injection to maintain compliance with 7 ppm @ 15% O_2 ("ppmc") NO_x emissions from the stack. The NH_3 slip was measured at 47 ppmc, while the permitted limit for the plant is 20 ppmc. This paper presents results of the

engineering assessment of the system performance conducted by Carnot which culminated in a decision to relocate and redesign the ammonia injection grid. The modifications were implemented by AES, and the system currently operates within compliance limits for both NO_x and NH_3 emissions.

Figure 1 is a diagram of the plant showing primary equipment and full load parameters typical of plant operations at the time the SCR was evaluated. The unit consists two ABB Type 8 turbine/generator sets, and two HRSGs linked with a turbine exit crossover duct. The Steam Electric (SE) boiler generates superheated steam for the steam turbine, and the Steam Injection (SI) boiler is used to generate 85% quality steam for enhanced oil field recovery. The proportion of exhaust gas directed to the SI boiler is varied to match demand for oil field injection. During the time the SE boiler SCR system experienced problems, operating requirements dictated that the SE boiler receive 40% more exhaust gas than the SI as shown in Figure 1.

Figure 2 shows the layout of the anhydrous NH_3 injection sparger piping. There were six discrete cells referred to as North and South top/middle/bottom. The effect of separating the SCR system into six cells was to create six semi-independent SCR systems. It should be noted that there was no mechanism for biasing NH_3 flow to various zones, thus NH_3 distribution was controlled by piping design and the injection nozzle orifice pressure drop characteristics.

Scope of Problem

The plant started up in 1988 and operated within compliance for both NO_x and NH_3. After approximately two years, SCR performance deteriorated to the point where AES decided to solicit bids for catalyst replacement. New catalyst was then installed within the existing reactor space by a supplier other than the original manufacturer. Concurrent with this retrofit, attempts were made to improve the ammonia injection grid (AIG) design, although the modifications were not thoroughly documented. All cells contained 178 ft^3 of NO_x catalyst, except the bottom south cell which was modified to hold 202 ft^3. Source test results in mid-1990 showed the system performing well with 8.4 ppmc NH_3 slip from the SE stack. Over the course of the next two years the NH_3 slip became excessive and measurements in mid-1992 showed the average slip to be 47 ppmc.

The diagnostic program began on a fast track due to regulatory concerns with the following corrective action approach:

1. Perform "screening" tests on the SCR system to identify gross maldistributions of NH_3/NO_x ratios. Concentrations of NH_3 and NO_x at available test ports were to be measured. Simultaneous with the NH_3 slip investigations, consideration would be given to removing the CO catalyst as part of any potential design modifications. Removing the CO catalyst was seen as a method of making space available for additional NO_x catalyst layers and to increase the mixing length between the AIG and the catalyst.

2. Based upon the results of the screening tests, a "detailed" test plan would be developed and implemented. This was envisioned to entail more detailed mapping of NH_3 and NO_x

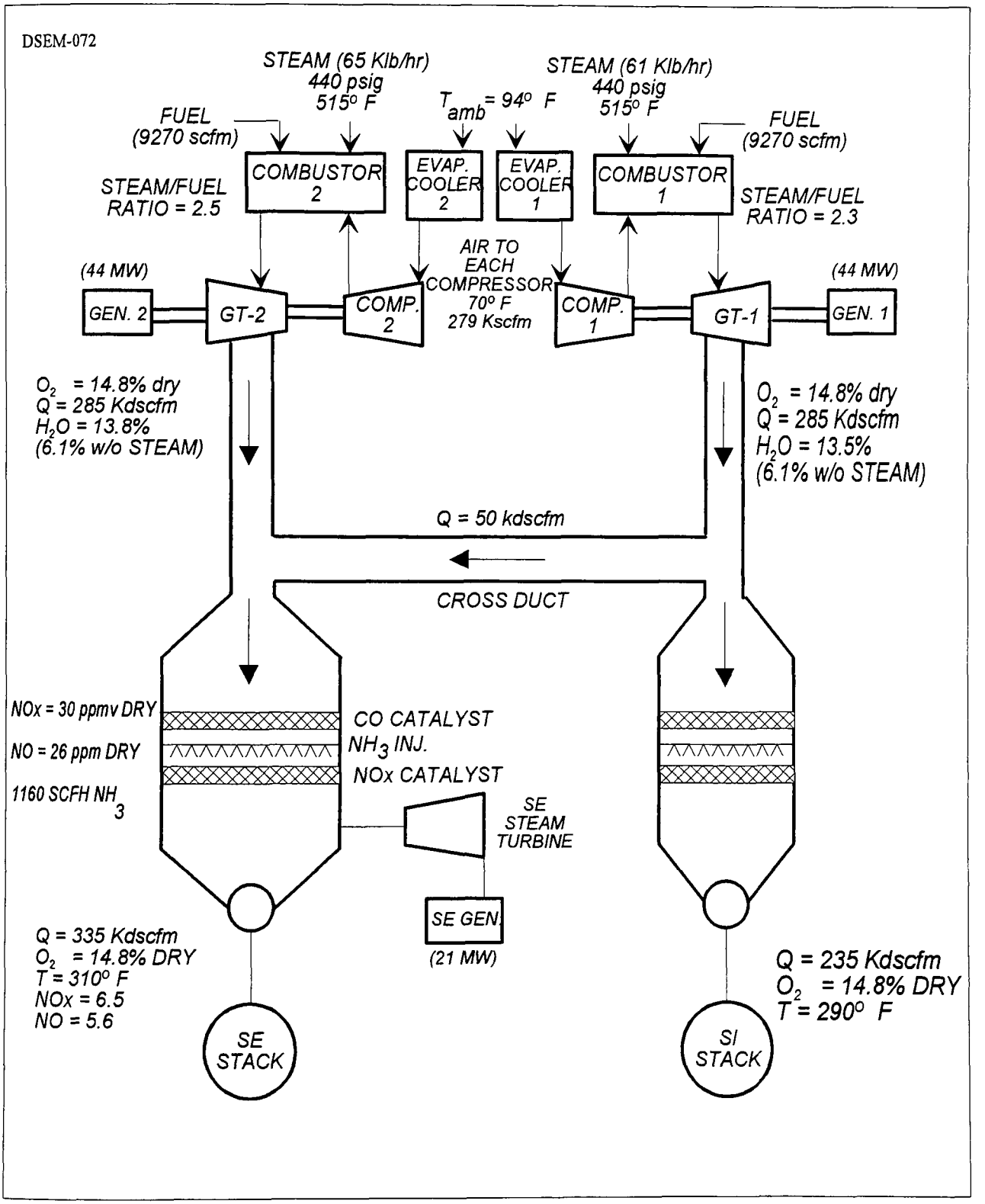

Figure 1. Plant Block Diagram with Approximate Full Load Operating Parameters at AES, Placerita

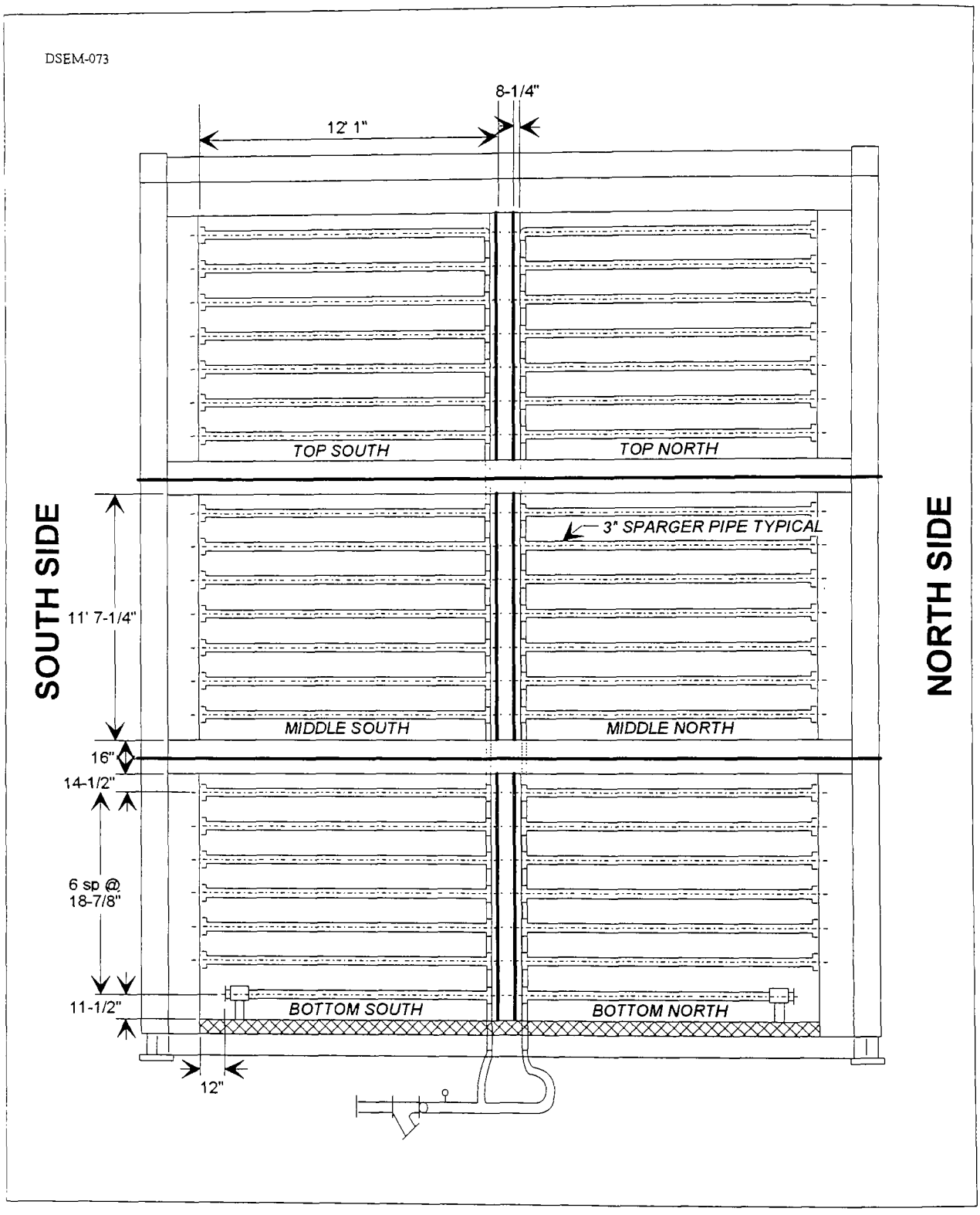

Figure 2. Ammonia/Dilution Air Injection Piping Network as Viewed from NOx Catalyst (all dimensions typical)

concentrations, exhaust gas flow distribution measurements, temperature measurements, catalyst activity core samples, and off line inspections.

3. Using the data from testing and inspections, it was anticipated that a design modification could be developed and implemented.

There were other regulatory issues and time constraints involved with resolving the slip problem, however only the technical investigation is described in this paper.

Screening Diagnostic Tests

Because information regarding the design modifications to the SE boiler's SCR system were sparse, "screening" diagnostic tests were conducted by Carnot. The primary objective of the screening tests was to measure gross cell-to-cell distribution of NO_x and NH_3 using available sample ports. Measurements performed included:

- NO, NO_x and NH_3 concentration stratification upstream of the SCR catalyst.
- NO and NO_x concentration at the available test port downstream of the SCR.
- CO concentration upstream and downstream of the CO catalyst.
- Total Hydrocarbons (THC) upstream and downstream of CO catalyst, and at the stack. (The CO catalyst was designed to oxidize hydrocarbons as well.)
- Stack NO, NO_x, CO, and NH_3 slip.

The tests were conducted over a two-day time period, and results are summarized in Table 1. Figure 3 shows the location and labeling of sample ports utilized for the testing. There were no available sample ports upstream of the SCR for the middle cells, thus measurements were made at the top and bottom cells only. The NH_3 injection rate was set by the operator to maintain constant stack NO_x emissions just below 7 ppmc. In addition to verifying the previous source test showing there was a slip problem, the screening tests showed the following results:

- Upstream of the SCR, the north bottom cell was relatively NH_3 lean with a NH_3/NO_x mole ratio of 0.75, while the other three cells measured were relatively NH_3 rich with an average ratio of 3.0.

- Turbine exhaust CO levels were already very low at 0.6 ppmc, with a comfortable margin relative to the 2 ppmc stack limit. This demonstrated that removing the CO catalyst could possibly be pursued to make room for additional NO_x catalyst and/or moving the AIG further upstream.

- The CO catalyst provided no benefits to reducing THC emissions. The flame ionization detection (FID) measurements actually indicated an increase from 1.2 to 2.9 ppm. This difference is within the scatter of data expected for this instrument's accuracy.

Table 1
*Screening Test Results**

Location	Species	Measured Value
GT2 turbine exhaust	NO, ppmv	25.7
	NOx, ppmv	31.7
	CO, ppmv	0.58
upstream of CO catalyst (downstream of cross duct)	THC, ppm CH4	1.2
upstream of NOx catalyst, bottom north cell (ports A-E)	NH3, ppmv	20.9
	NO, ppmv	24.6
	NOx, ppmv	27.7
	NH3/NOx ratio	0.75
	CO, ppmv	0.13
upstream of NOx catalyst, top north cell (ports F-J)	NH3, ppmv	92.6
	NO, ppmv	24.3
	NOx, ppmv	28.7
	NH3/NOx ratio	3.23
	CO, ppmv	0.06
upstream of NOx catalyst, bottom south cell (ports K-O)	NH3, ppmv	86.0
	NO, ppmv	27.0
	NOx, ppmv	30.2
	NH3/NOx ratio	2.85
	CO, ppmv	0.09
	THC, ppm CH4	2.9
upstream of NOx catalyst, top south cell (ports P-T)	NH3, ppmv	90.2
	NO, ppmv	27.8
	NOx, ppmv	31.4
	NH3/NOx ratio	2.87
	CO, ppmv	0.08
downstream of NOx catalyst, bottom south cell (port U)	NO, ppmv	4.8
	NOx, ppmv	5.2
	CO, ppmv	0.08
stack of SE boiler	NH3, ppmv	43.1
	NO, ppmv	5.6
	NOx, ppmv	6.5
	CO, ppmv	0.21
	THC, ppm CH4	2.4

* All units are raw concentrations on a dry basis. O2 measured at approximately 14.8% dry at all locations.

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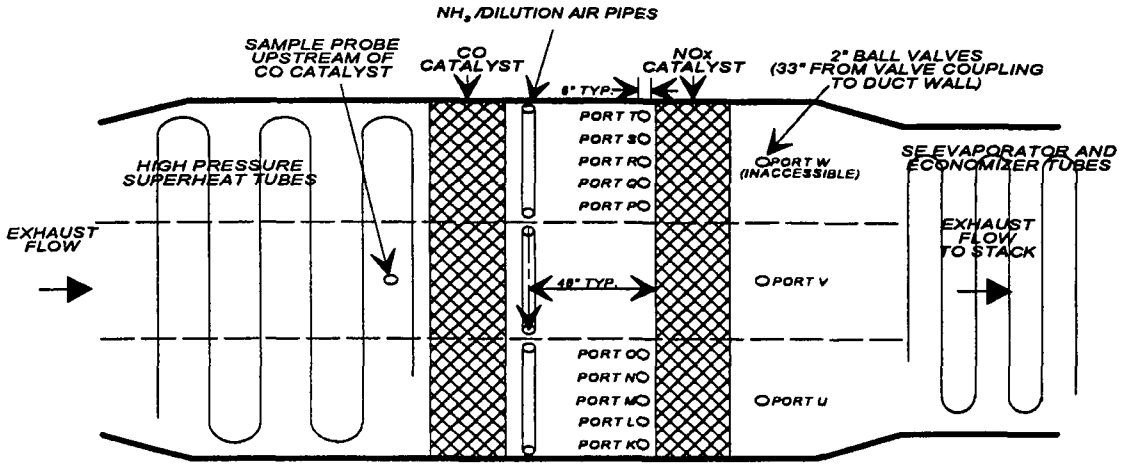

Figure 3a. SE Boiler Sample Port Locations and Side/Sectional View Looking at the South Side as Viewed from Outside

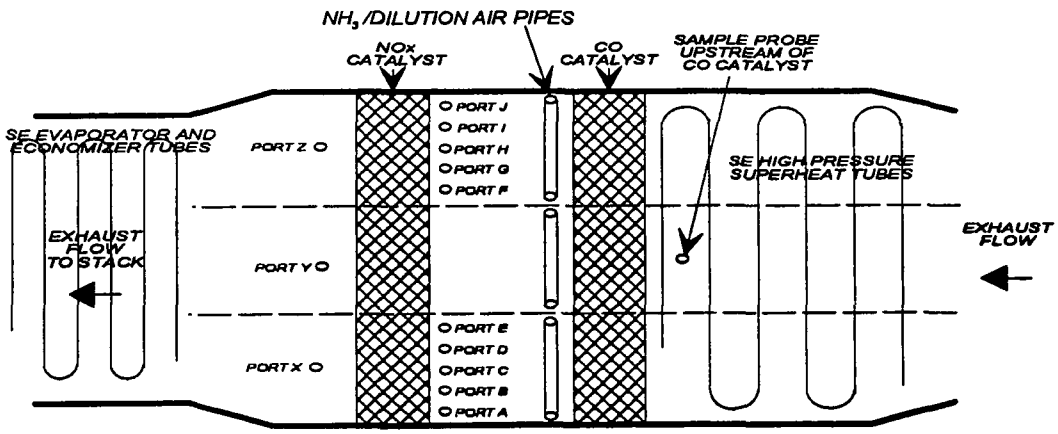

Figure 3b. SE Boiler Sample Port Locations and Side/Sectional View Looking at the North Side as Viewed from Outside

Detailed Diagnostic Tests

Building upon the information garnered from the screening test effort, a "detailed test" plan was developed and implemented by Carnot. The basic idea was to obtain flow and NH_3 distribution within a cell, adding to the gross cell-to-cell data obtained previously. These tests were designed to utilize the 20 available sample ports located upstream of the NO_x catalyst, and the 5 available ports located downstream of the catalyst as illustrated in Figure 3. To the extent possible, these tests were to include multiple-point composite traverses in each port for NH_3, NO_x, velocity, and temperature. Further measurements of CO and THC emissions were not conducted.

Table 2 contains a summary of the NO_x and NH_3 measurements collected over a three-day time period. Refer to Figure 3 for sample port locations and labeling. The SCR inlet NO_x was approximately constant due to constant combustor operating conditions including steam injection rate. The NH_3 injection rate was set by the operators to maintain constant stack NO_x emissions just below 7 ppmc. The SE boiler flue gas flow variations ranged from +4% to -9% from the average of 5,507 scfs, while the NH_3 flow varied from +4% to -13% from the average of 1,107 scfh.

Since measured values of NO_x and NH_3 were obtained over three days with the variations noted above, values were normalized to the average SE boiler gas flow and NH_3 injection rate. This normalization, shown in the right columns of Table 2, was performed using equations (1) and (2) as follows:

(Inlet NO_x or Outlet NH_3) * (actual scfs / 5,507) * (1,107 / actual scfh NH_3) (1)
= Normalized Inlet NO_x or Outlet NH_3

(Outlet NO_x) * (5,507 / actual scfs) * (actual scfh NH_3 / 1,107) (2)
= Normalized Outlet NO_x

This was done to normalize all data to approximately constant operating conditions including stack NH_3 slip. A more exact method would be to perform a NO_x and NH_3 mass balance on each cell including the effects of varied space velocity through each cell and accounting for NO_2 effects (i.e. 1.33 moles NH_3 per mole NO_2). However, all required parameters were not measured for each cell and applying the above correction was better than none.

The emissions and flow measurements were then used to generate SCR performance parameters shown in Table 3. As seen in the Table 3 data, different cells operated at different mole ratios due to cell-to-cell NH_3 maldistributions. Figure 4 illustrates the performance of different cells graphically with NO_x removal efficiency versus NH_3/NO_x mole ratio. A perfect system would remove 100% of the NO_x at a mole ratio of 1.0 (excluding NO_2 effects) shown with the upper curve in Figure 4. The system at AES was designed to operate at a mole ratio of 1.0-1.6, depending upon space velocity and catalyst age, as shown with the "design" curve in Figure 4. The "post-mod test" curve shown in this figure is explained later in this paper.

Although the screening tests identified cell-to-cell NH_3 maldistribution, these detailed tests showed that NH_3 distribution within each cell was unacceptable as well. The point by point data showed NH_3 variations by factors ranging from 1.6 to 3.6 in the three cells where measured.

Table 2
Detailed Diagnostic Test Results

Cell/Location	Port(s)	Meas NOx ppm	Meas NO2 ppm	NH3 ppm	Exhaust Flow dscfs *	Total NH3 scfh *	Stack NOx ppm *	Normalized ** Meas NOx ppm	Normalized ** Meas NH3 ppm	Normalized ** Stack NOx ppm
SE (upst CO catly)	Inlet	30.8	5.8		5,600	942	6.4	30.8		5.5
SE (upst CO catly)	Inlet	28.3						28.3		
SE Boiler Exhaust	Stack	6.5			5,532	1,092		6.6		
SE Boiler Exhaust	Stack			47.1	5,153	1,146	6.4		41.4	7.3
SE Boiler Exhaust	Stack			58.1	4,978	1,146	6.4		49.3	7.6
SE Boiler Exhaust	Stack			44.4	5,310	1,115	6.3		41.3	6.8
Bottom South	U	5.9	0.1		5,419	1,104	5.9	6.2		6.1
Bottom South	U			50.1	5,515	1,083	6.1		49.9	6.1
Middle South	V	3.1	0.0		5,419	1,104		3.2		
Bottom North	X	18.9	0.2		5,663	1,126		19.2		
Bottom North	X			4.0	5,314	1,125	6.1		3.7	6.6
Bottom North	X	13.9	1.3		5,663	1,015		12.7		
Middle North	Y	2.7	0.1		5,663	1,126		2.8		
Middle North	Y	5.1	0.2		5,663	1,015		4.7		
Top North	Z	4.3	0.1		5,663	1,126	6.4	4.4		6.5
Top North	Z			75.9	5,606	1,109	6.1		75.0	6.2
Top North	Z	4.7	0.4		5,667	1,036	6.5	4.4		6.1
Bottom North	A,B,D,E			23.2	5,600	942	6.4		26.9	5.5
Top North	F,G,H,I,J			82.5	5,600	942	6.4		95.9	5.5
Bottom South	K,L,M,N,O			76.2	5,700	1,066	6.4		79.7	6.1
Top South	P,Q,S,T			78.8	5,700	1,066	6.4		82.4	6.1
Bottom North	A			65.5	5,532	1,092			64.9	
Bottom North	B			33.2	5,532	1,092			32.9	
Bottom North	C			48.7	5,532	1,092	6.0		48.3	6.1
Bottom North	D			19.0	5,532	1,092			18.8	
Bottom North	E			27.7	5,685	1,033			29.8	
Top North	F			78.5	5,685	1,033			84.5	
Top North	G			134.8	5,580	1,078	6.2		136.5	6.1
Top North	H			99.8	5,580	1,078			101.0	
Top North	I			77.8	5,201	1,120			70.7	
Top North	J			111.4	5,201	1,120	6.3		101.2	7.0
Bottom South	K			79.9	5,685	1,033			85.9	
Bottom South	L			88.4	5,580	1,078	6.2		89.5	6.1
Bottom South	M			109.5	5,580	1,078			110.8	
Bottom South	N			96.4	5,201	1,120			87.5	
Bottom South	O			39.8	5,201	1,120	6.3		36.1	7.0

* Data from Plant Instrumentation

** Normalized to: 1077 average scfh NH3 (measured by plant instrument)

5507 average dry scfs exhaust (measured by plant instrument)

Table 3
*SCR Performance Measurements**
(Using Screening and Detailed Test Data)

Cell	Catalyst Volume cubic feet	Average Measured Velocity ft/s	Measured Temp. °F	Measured Exhaust Flow dscfs	Space Velocity (dry) 1/hr	Inlet NOx ppmvd	Inlet NH3 ppmvd	Inlet NH3/NOx Mole Ratio	Outlet NOx ppmvd	NH3 Slip ppmvd	NOx Removal Efficiency
Top North	178.0	21.1	689	1,135	22,959	28.7	96	3.3	4.4	75	85%
Middle North**	178.0	19.8	690	1,064	21,515	30.8	N/A	N/A	3.8	N/A	88%
Bottom North	178.0	18.5	691	993	20,075	27.7	27	1.0	16.0	4	42%
Top South ***	178.0	17.0	695	909	18,388	31.4	80	2.5	7.0	N/A	78%
Middle South**	178.0	18.2	690	977	19,767	30.8	N/A	N/A	3.2	N/A	90%
Bottom South	202.0	19.4	692	1,040	18,530	30.2	83	2.7	6.2	50	79%

* All NH3 and NOx measurements normalized to 5507 scfs (60°F) exhaust and 1077 scfh inlet NH3 -- see Table 2 and text).
** There were no inlet sample ports to these middle compartments; inlet NOx same as turbine exhaust; inlet velocity = average of top & bottom
*** The outlet sample port was not accessible; outlet NOx was estimated considering inlet NH3/NOx ratio and space velocity.

Figure 4.
NOx Reduction Efficiency vs. NH3/NOx Mole Ratio
SE Boiler SCR

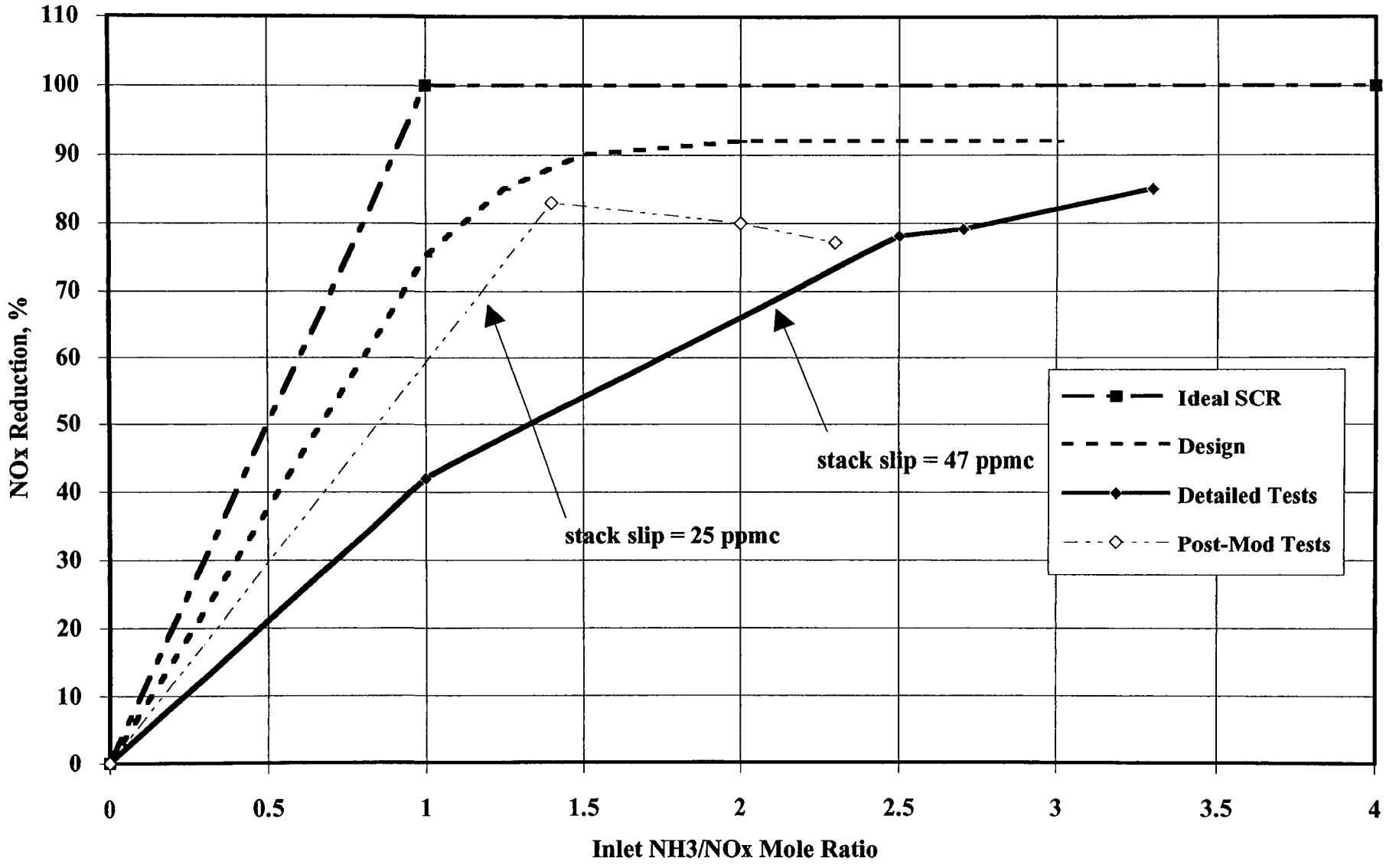

Figure 4 also supported this conclusion since cells with mole ratios over 2.5 could not achieve the removal efficiency expected for a mole ratio of only 1.5. Lastly, it is noted that the true catalyst efficiency cannot be extracted from this data since the effects of non-uniform NH_3 and NO_x mixtures entering the catalyst are not separable from the catalyst activity effects.

Off-Line Inspection and AIG Nozzle Replacement

An inspection of the SCR modifications was planned for an upcoming outage scheduled to begin within one week of completion of the detailed tests. In terms of the SCR performance evaluation project, the purpose of the outage was to:

1. inspect the AIG to confirm the extent of design modifications made; the system was converted from using threaded nozzles to a pattern of holes drilled in the sparger pipes; the details of this modifications were not well documented.

2. make any modifications which appeared technically feasible and cost-effective.

Since the system originally had passed NH_3 slip tests, it was tentatively planned to re-install nozzles of the original design and spacing unless something during initial inspection dictated otherwise. It should be noted that, besides the AIG modification, two other key changes had taken place (1) the catalyst had been replaced with a new design by a different manufacturer, and (2) the catalyst had aged.

An inspection on the first day of the outage by Carnot revealed that the AIG modifications made were asymmetric. The bottom south cell had 1" pipe tees mounted on the 3" sparger pipes, with three 1/16" holes spaced vertically in each tee as shown in Figure 5a. All other cells had a spaced pattern of 1/8" holes drilled directly into the 3" sparger pipes as shown in Figure 5b.

The cell-to-cell AIG design variations explained the cell-to-cell maldistribution of NH_3 measured during the screening tests. The bottom north cell was NH_3 lean most likely due to partial plugging of the 1/16" holes. Although there were more 1/16" holes, the plugging created a higher resistance than the 1/8" holes drilled in the spargers of the other five cells.

Furthermore, the point-by-point NH_3 data shown in Table 2 indicates that this pipe tee design did not distribute NH_3 any better than the 1/8" hole design. Consequently, it was decided to proceed with re-installing nozzles of the original design and spacing regardless of the catalyst design changes and aging which had occurred.

Core samples of catalyst were also extracted during this outage by the manufacturer for laboratory determination of activity. The tests conducted by the manufacturer showed catalyst activity was nearly the same as the retained fresh samples from the original installation.

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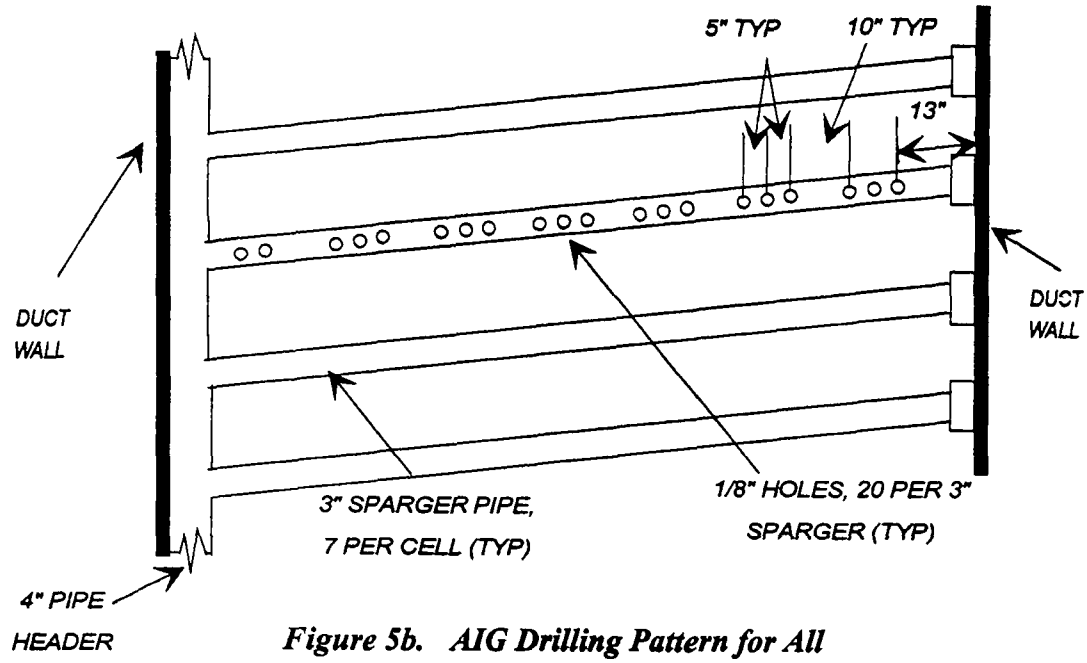

Figure 5b. AIG Drilling Pattern for All Except Bottom North

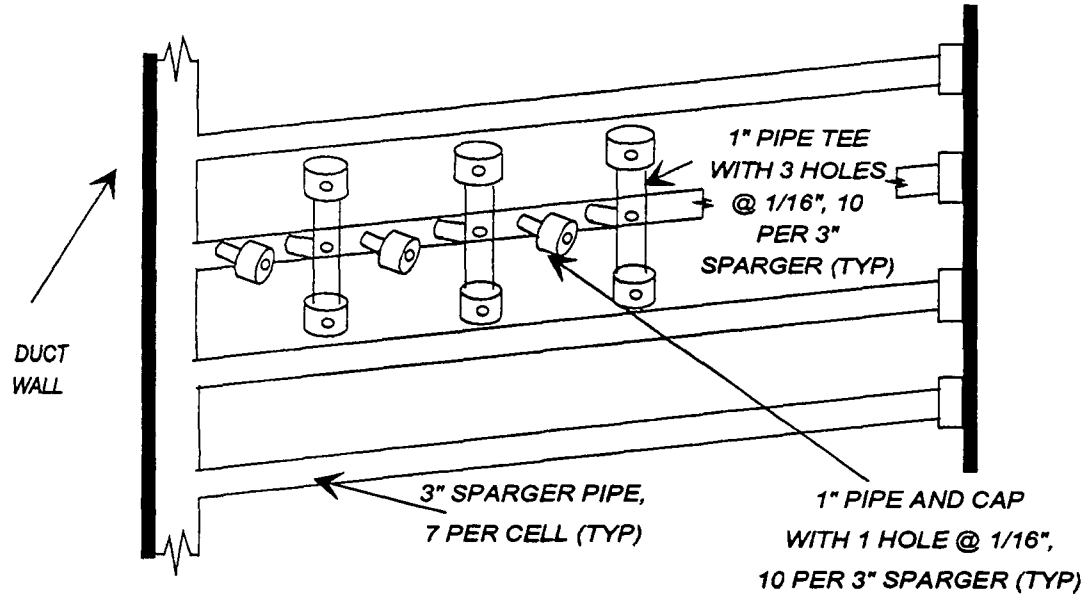

Figure 5a. AIG Pipe Tee and Hole Drilling Pattern for Bottom North Cell Only

Post-Modification Diagnostic Tests and Analyses

During the aforementioned outage, the original design nozzles were re-installed by AES, which replaced the hole-in-pipe design shown in Figure 5. Tests of the modified system were conducted by Carnot shortly after the outage and showed the stack NH_3 had been reduced to 25 ppmc from the pre-modification level of 47 ppmc.

From the measurements performed both upstream and downstream of the NO_x catalyst, system performance parameters were again estimated as done with the detailed tests. The NO_x removal efficiency versus inlet NH_3/NO_x ratio is plotted in Figure 4 along with previous results. Re-installing these nozzles substantially improved the SCR removal efficiency. For example, at an inlet mole ratio of 1.5 for a given cell the removal efficiency was approximately 78%. With the hole-in-pipe design, 78% removal efficiency required an inlet mole ratio of 2.5.

Although significantly improved, the stack NH_3 slip still exceeded the 20 ppmc regulatory limit. Single point NH_3 measurements at various ports upstream of the catalyst ranged from approximately 50% to 200% of the port average NH_3 concentration. These data and the manufacturer's catalyst activity tests indicated that non-uniform NH_3/NO_x ratios entering the catalyst was still causing excessive slip.

Engineering calculations were performed by Carnot to gain an understanding of the problem. Four items were addressed:

1. Since the basic objective was to properly entrain the injected ammonia with the exhaust gases, a calculation of the entrainment of the actual mass flows of jet and bulk (turbine exhaust) gases was performed. It showed that for a simple hole-in-pipe orifice design, the minimum distance required to entrain the required amount of bulk gas is approximately 138 inches, whereas the average available length as designed is 48 inches.

2. A calculation of the heat transfer between the exhaust flow and the air/NH_3 inside the sparger pipe was performed to determine heating (density change) effects on the distribution of the mixture through the AIG. It helped simplify flow distribution calculations if the air/NH_3 mixture could be considered to be isothermal at the same temperature as the turbine exhaust gases. This assumption was shown to be acceptable as the air/NH_3 flow would approach the exhaust gas flow temperature within several feet.

3. Proper distribution of air/NH_3 through the AIG was calculated by comparing the frictional and kinetic energy characteristics of the air/NH_3 mixture relative to pressure drop across the nozzle (or orifice). A design criterion to achieve +/- 5% flow distribution through the pipe header is that the ratio of kinetic energy and (AIG) pipe frictional losses be less than 10% of the nozzle/orifice pressure drop. Calculations showed that this ratio was less than 1%, indicating maldistribution was not due to the sparger piping design.

4. Entrainment of a jet with bulk flow is enhanced with increased velocity differential between the jet and bulk flow. To check the possibility that pipe pressure losses could limit nozzle exit velocity, frictional losses were estimated for the actual AIG design from

the air/NH_3 mixing point to the point where the mixture enters the 3" spargers at the boiler bottom. The typical mixing point pressure was measured at 36 psia, which exceeded the critical pressure requirement of 28 psia required for sonic velocity. This meant that an 8 psia pressure drop would be allowable. An analysis of the pipe network demonstrated that frictional losses were negligible relative to the 8 psia allowable.

In parallel with Carnot's evaluations, an assessment of the modified system was performed by the catalyst manufacturer using a computational fluid dynamic code. This analysis showed the system would require a path length of 15 feet from the AIG to the catalyst to achieve proper NH_3 mixing. However, the system only had 3 to 4 feet of space.

Thus the diagnostic tests, engineering calculations, and catalyst manufacturer analyses all indicated an NH_3 distribution problem. Before proceeding with a complete re-design of the AIG, it was decided to evaluate a low-cost installation of various nozzles which could utilize the existing sparger piping arrangement.

NH_3 Injector Nozzle Evaluation

In order to inexpensively evaluate different nozzle types, a bench-scale cold flow test was designed and built by Carnot as shown in Figure 6. Only co-flowing designs were to be evaluated in order to maintain the cost of designing, constructing, conducting and evaluating the cold flow tests within project time and cost constraints. Similarity of the bench-scale test to the actual installation would be gained with respect to actual bulk velocities, mixing length, utilization of actual nozzles (rather than scaled nozzles), and maintenance of critical pressures to achieve sonic velocity at the choke point for each nozzle.

The bench-scale test consisted of a 20" x 20" plexiglass duct with a blower pulling air at approximately 20 ft /sec. The duct was sized so as to negate boundary layer effects. A compressed cylinder of 30,000 ppm CO gas in N_2 was injected through test nozzles coflowing with the bulk air. A high concentration of CO was required to produce a measurable CO concentration in the cold flow duct. A 20-point traverse measurement of CO concentration was made 48" downstream of the injection plane to evaluate the relative dispersion effectiveness of NH_3 injector nozzles. Layers of screen material were installed until the bulk air velocity distribution was within +/- 5% of the average velocity.

Since the nozzle (or orifice) injector pressure drop was the controlling factor for NH_3 distribution through the grid, the use of nozzles was preferred over pipe holes in the interest of long-term maintenance of discharge coefficients (C_D). It was anticipated that holes would create significant variations in the C_D due to field drilling imperfections or hole plugging. Four different nozzles were evaluated as shown in Figure 7:

- 180° deflector head hollow cone (originally designed, and re-installed during the outage)
- 0.013" slotted axial nozzle - flat fan pattern
- 0.023" slotted axial nozzle - flat fan pattern
- 75° deflector nozzle - flat fan pattern

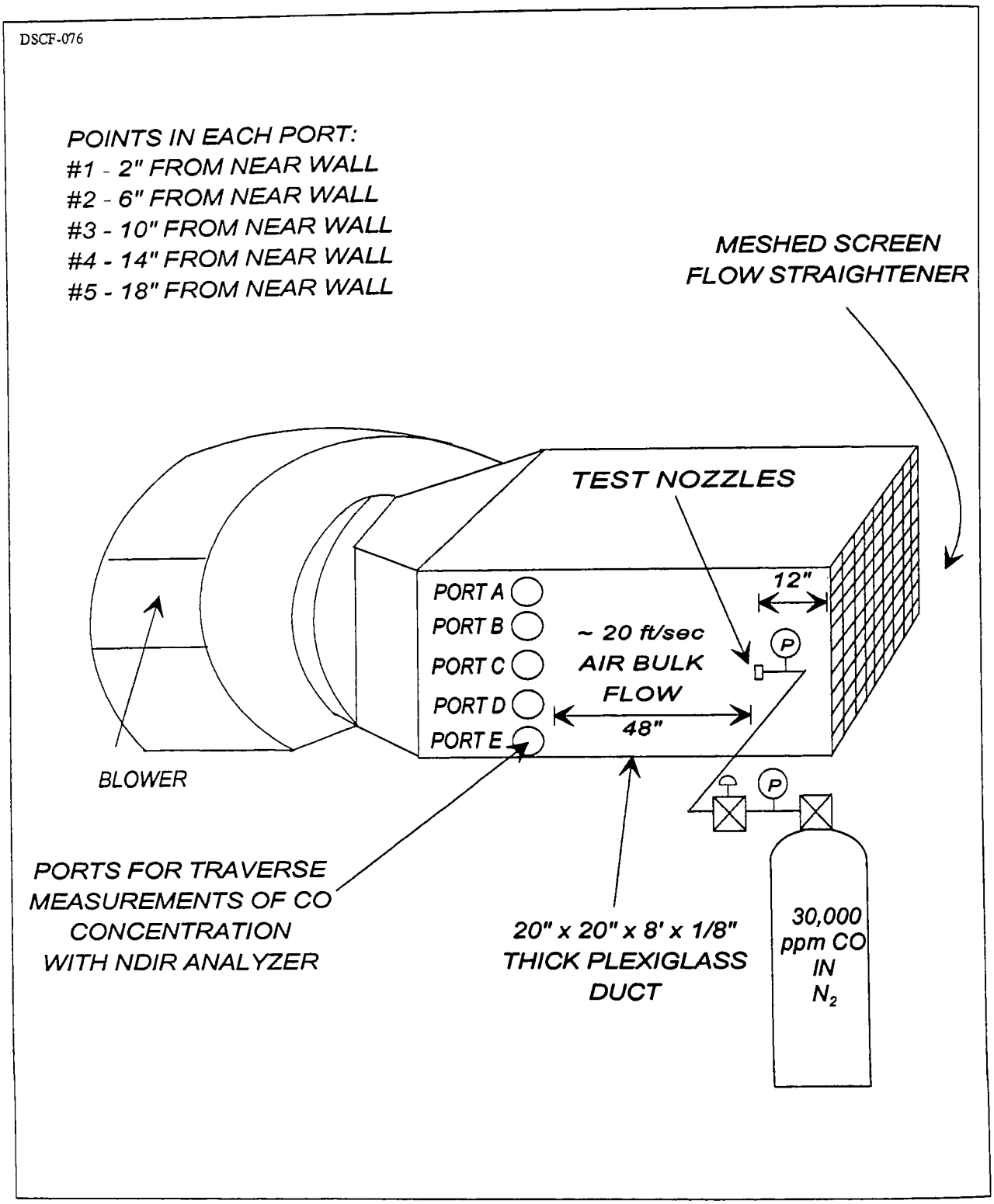

Figure 6. Test Set-up for Cold Flow Nozzle Evaluation

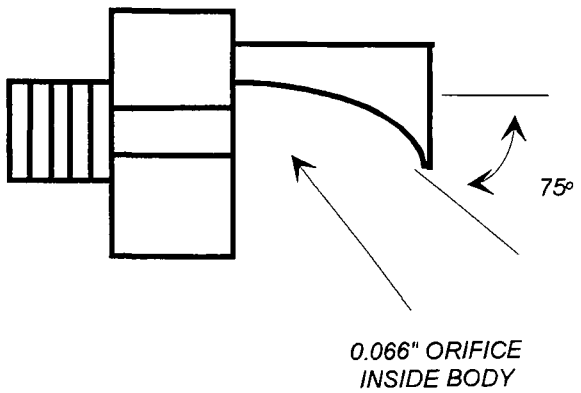

Figure 7c. 75° Deflector Flat Fan Nozzle

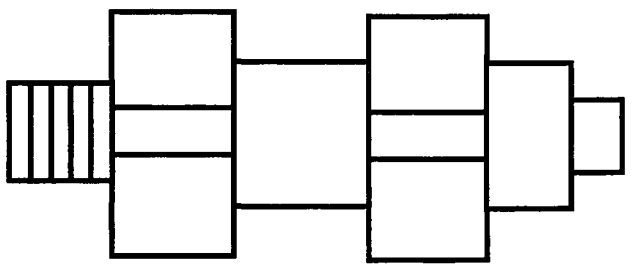

Figure 7b. Flat Fan Nozzles (0.013" and 0.023" slots evaluated)

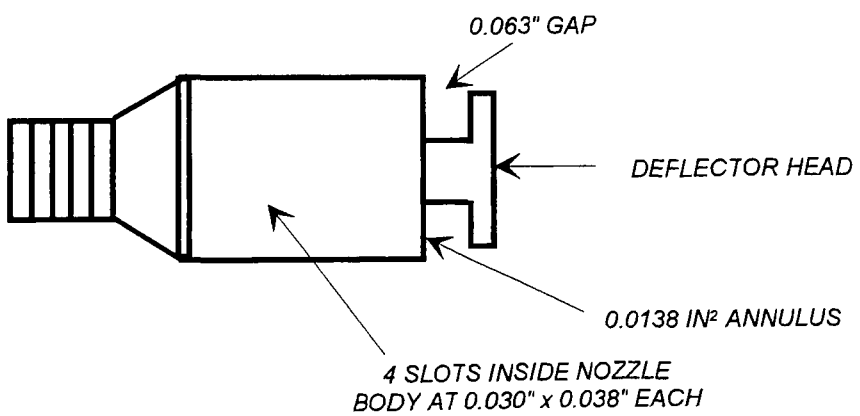

Figure 7a. Deflector Nozzels (original design for SE boiler SCR: also reinstalled during outage)

Important dimensions for each nozzle were measured, and the critical pressure to achieve sonic velocity at the choke point was calculated. A tap and gauge were installed just upstream of the nozzle connection to measure and maintain the minimum (critical) pressure during each cold flow test. Since the pressure requirement was high, large CO/N_2 gas flows were required and thus CO traverses had to be made rapidly. In order to assure good readings, initial response time tests were run to quantify the system response prior to testing nozzles.

The results of the cold flow model are contained in Table 4. As a measure of dispersion, the standard deviation percentages of mean for downstream CO concentrations were calculated. The 0.013" slotted nozzle produced the best results at 56%, compared to 61-65% for the 180° deflector nozzle installed in the SE boiler grid. The 0.023" slot and 70° deflector flat fan nozzles performed the worst. However, the total range of this dispersion parameter was only 56-71% indicating there were no major differences in dispersion effectiveness within the tested path length. The data was viewed in different ways, such as ratio of minimum or maximum to mean and number of points near the mean, but did not reveal any different results as seen in Table 4.

Although the NH_3 slip needed to be reduced only an additional 5 ppmc from 25 to 20 ppmc, it was decided that new nozzles would not be installed. The results of the cold flow evaluation indicated that new nozzles would only slightly improve the SCR performance, if at all. Regulatory time constraints influenced this decision as well.

Conclusions

The results of the screening tests, detailed tests, and post-modification tests performed after nozzle re-installation indicated NH_3 maldistribution both cell-to-cell, and within each cell. It was never clear why the system could maintain NO_x at 7 ppmc with less than 10 ppmc slip initially, but only 25 ppmc slip after the nozzle re-installation. The possible reasons were that a combination of process changes such as inlet NO_x level or moisture content (both influenced by combustor steam injection rate for example), plugging of holes in the drilled pipe or injection nozzle bodies, or minor catalyst aging effects combined to inhibit the SCR performance. Measurements showed that process temperature, boiler flow distribution, and inlet NO_x concentration distribution were all within design guidelines. However, the engineering calculations, computer modeling, and cold flow nozzle evaluations collectively indicated that the 4 foot path length was inadequate to achieve proper mixing of NH_3 and NO_x.

As a result of the evaluation, it was decided that the AIG be completely re-designed and located further upstream in the SE boiler. In order to make space in a location with proper gas temperatures, the CO catalyst was removed. This was justifiable based upon initial CO and hydrocarbon measurements. The new grid was installed several feet further upstream, but still downstream of the low pressure superheat tubes (see Figure 3). The new grid was designed with the capability to bias NH_3 to different zones. Flow-straightening louvers were also required as the gas flow became stratified without the CO catalyst. Currently the system operates within compliance for both NH_3 slip and NO_x emissions.

Table 4
Cold Flow Nozzle Test Data Summary

port	point	180° Deflector 4 @.03" x .038" ppmv CO	(repeat test) 180° Deflector 4 @.03" x .038" ppmv CO	Slotted Flat Fan .013" slot ppmv CO	Slotted Flat Fan .023" slot ppmv CO	70° Deflector Flat Fan .066" orifice ppmv CO
A	1	22.5	7.5	10.0	20.0	15.0
A	2	22.5	7.5	25.0	25.0	7.5
A	3	10.0	50.0	25.0	40.0	7.5
A	4	25.0	67.5	42.5	15.0	12.5
A	5	22.5	35.0	20.0	15.0	20.0
B	1	27.5	25.0	10.0	--	17.5
B	2	20.0	65.0	17.5	--	40.0
B	3	30.0	85.0	40.0	--	22.5
B	4	22.5	90.0	55.0	--	60.0
B	5	30.0	55.0	50.0	--	15.0
C	1	50.0	35.0	7.5	15.0	12.5
C	2	70.0	45.0	17.5	80.0	45.0
C	3	70.0	125.0	30.0	100.0	80.0
C	4	35.0	80.0	40.0	115.0	72.5
C	5	50.0	15.0	20.0	105.0	25.0
D	1	80.0	30.0	15.0	--	35.0
D	2	80.0	72.5	40.0	--	85.0
D	3	95.0	120.0	40.0	--	100.0
D	4	105.0	85.0	40.0	--	45.0
D	5	40.0	25.0	10.0	--	27.5
E	1	35.0	17.5	15.0	45.0	20.0
E	2	90.0	35.0	65.0	100.0	40.0
E	3	125.0	55.0	50.0	115.0	67.5
E	4	70.0	30.0	30.0	70.0	30.0
E	5	50.0	17.5	10.0	--	20.0
	mean	51.1	51.0	29.0	61.4	36.9
	std dev	31.1	33.1	16.3	40.5	26.3
	std dev/mean	61%	65%	56%	66%	71%
	Min	10.0	7.5	7.5		7.5
	Max	125.0	125.0	65.0		100.0
	Min/mean	20%	15%	26%		20%
	Max/mean	245%	245%	224%		271%
within 25% mean		4	4	4		6
within 50% mean		13	14	16		12

The SCR investigation described in this paper was appropriate for the unique circumstances involved, although the methods employed are generally applicable to any gas-fired source. However when encountering SCR performance problems on other gas-fired sources, the trouble-shooting approach should be specifically tailored to the combustion device features and SCR design capabilities. Some suggestions follow:

- Verify with repeated measurements that SCR system is malperforming (e.g. confirm good slip measurement, confirm problem not related to CEM or other instrument miscalibration).

- If a performance problem is verified, consider the magnitude of the problem (e.g. NH_3 slip is 12 ppm or 50 ppm versus a 10 ppm limit).

- Design investigative tests considering SCR design features such as NH_3 distribution control, available test ports, inlet NO_x magnitude and distribution, and catalyst age.

- Design investigative tests considering the gas-fired combustion device features such as variations in inlet NO_x or flow; load ramping; variations in O_2, temperature, or moisture levels; and catalyst poisoning potential.

- Investigative tests should also consider other factors such as regulatory deadlines and manufacturer guarantees.

- Tests should be designed stepwise considering costs, and targeted towards the most likely causes of malperformance. This is especially true for troubleshooting situations since the plant owner is usually interested in compliance and minimum operating costs, not in conducting a research project.

- Compliance quality confirmation tests should be performed once the system is optimized or physically modified and tuned.

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SELECTIVE CATALYTIC REDUCTION PERFORMANCE PROJECT AT PUBLIC SERVICES ELECTRIC AND GAS COMPANY'S MERCER GENERATING STATION UNIT NO. 2

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Abstract

PSE&G, Wahlco, and Siemens have successfully demonstrated the technical feasibility of a Post Combustion NO_x Reduction System on a 321 MW net coal-fired utility boiler at PSE&G's Mercer Station Unit 2. The program featured a selective catalytic reduction (SCR) reactor in a horizontal flue gas duct and an air heater SCR (AHSCR). Predicted performance of the In-Duct SCR was confirmed. The AHSCR was shown to be an effective means to enhance overall system performance as compared to an In-Duct SCR alone. Selective Non-Catalytic Reduction (SNCR) was also tested in combination with the SCR/AHSCR system using aqueous ammonia as a common reduction agent. Ammonia based jet mixing SNCR and the use of an ammonia injection grid was found to be the operation of choice in the NH_3 based SNCR/SCR Hybrid NO_x reduction system.

Introduction

Public Service Electric and Gas Company (PSE&G) has completed an assessment of post-combustion NO_x control on a wet-bottom,coal-fired utility boiler. The technologies under study are In-Duct air heater SCR, and ammonia based SNCR/SCR Hybrid performed at PSE&G's Mercer Generating Station.

While SCR and, to a limited extent, SNCR have been used on coal-fired boilers, these processes have not been demonstrated in the U.S.A. on a unit with the same configuration as the wet-bottom, continuous slagging, pulverized coal furnaces operated at Mercer. These furnaces produce NO_x emissions of approximately 1.8 lb/mmbtu in which PSE&G working together with NJ Department of Environmental Protection have determined the need for NO_x Reduction.

PSE&G's interest in NO_x control for the Mercer Unit's is the result of our overall understanding of the role that NO_x plays in formation of ground level ozone. When PSE&G assessed the interim state emission inventory it was determined that PSE&G's power plants comprised 27% of the total NO_x emitted in the State of New Jersey. As a result of this emission profile, PSE&G established a public commitment to reduce NO_x from our power plants by 60% by 1995 and 80% by the year 2000. Following the SNCR demonstration in 1993, it was concluded that a greater level of NO_x reduction would be necessary for New Jersey to achieve attainment of the ozone standard. This is because of the severity of upwind transported ozone into New Jersey at levels far in excess of the National Ambient Air Quality Standard. PSE&G subsequently proposed to EPA and the Ozone Transport Commission that a regional NO_x standard of 0.2 lb/mmbtu be adopted on power plants throughout the Ozone Transport Region. The Mercer project demonstrates that emission reductions of the magnitude proposed can be obtained in a cost-effective manner.

PSE&G determined that installation of a full scale SCR reactor in the congested design and footprint of the Mercer units was not a viable option. The only location with a flue gas temperature appropriate for SCR was in a horizontal duct between the economizer outlet and air preheater. This existing duct was expanded to three times its cross-section to accommodate sufficient SCR capacity. In addition, the In-Duct SCR system included catalyst located in the hot-end of the air preheater to reach the .2 lb/mmbtu goal, a reduction of 88-89%.

The SCR demonstration was conducted on one of our four flue gas ducts with a horizontal shaft Ljungstrom air preheater.

The goals of the program were to demonstrate:

- First time use of horizontal In-Duct SCR on a coal-fired boiler
- First time air heater/In-Duct SCR combination coal application
- First time NH_3 SNCR/SCR Hybrid system while firing coal
- Minimal balance of plant impacts (i.e., air heater fouling, pressure drop)
- Acceptable NH_3 slip and effects on fly ash

Construction of the In-Duct SCR and AHSCR began in January and testing ran from June through October 1994. The aqueous ammonia SNCR/SCR Hybrid system testing was done in December 1994 and January 1995.

Description of Mercer Unit No. 2

Mercer Generating station has two Foster Wheeler continuous slagging, twin-furnace steam generating units. Units Nos. 1 and 2 are identical with a rated capacity of 321 MW net each. Each unit is designed for 2,060,000 lb/hr of superheated steam flow at 2456 psig and 1050°F. Reheated steam flow is 1,760,000 lb/hr at 445 psig and 1050°F. In this twin furnace design, one furnace generates superheated steam and the other reheated steam. Both units were originally designed with pressurized furnaces and have been converted to balanced draft with some modifications to the radiant superheater to allow for operations at 25% load, without adverse effects on boiler operations. Figure 1 shows a side view of the reheat furnace on Unit No. 2.

The Mercer units burn low sulfur Eastern bituminous coal as the primary fuel, with natural gas available for start-up and as a secondary fuel. Fuel enters each furnace through twelve front wall mounted burners arranged in three levels of four burners per level. There are three water tube division walls located above the top burner elevation and centered between each row. During coal firing, the furnaces are wet-bottomed, continuous slagging by design. The burners are arranged close to the furnace floor to keep the slag in a molten state at all loads. While it is possible to introduce coal and gas through the same burner at one time, general plant practice has been to operate each row of burners on either coal or gas. Fuel distribution between the furnaces is controlled to maintain the appropriate superheat and reheat temperatures.

Furnace gases exit the boiler through a convection superheater in one furnace and a convection reheater in the other. Economizers are installed after both the convection reheater and superheater; however, the economizer in the superheat furnace has more surface area. The flue gas splits into two ducts after each economizer to transfer the gases around the slag tank under each furnace. The flue gases pass through four Ljungstrom horizontal shaft regenerative air heaters (two per furnace) and are ducted to a multi-chamber electrostatic precipitator.

Testing Program Description

PSE&G contracted Carnot to conduct a comprehensive test program for evaluation of the SCR and Hybrid SNCR/SCR systems. The test data included NO_x emissions, CO, CO_2, O_2, SO_2, SO_3, NH_3, and N_2O measurements as well as plant operating data at various loads firing either coal or gas.

Measurements were made at several different locations in order to provide gas composition data at the inlet, outlet and intermediate locations of the SCR, as well as to determine the impact of the SCR operation on the remainder of the plant. Figure 2 presents a schematic of the reheat furnace illustrating the SCR system in duct 22B, the SNCR system, and the various test locations. A damper on duct 22A allowed for flow balancing.

Each sample location was equipped with a multi-point probe grid installed by PSE&G prior to the testing program. The grid consisted of a series of stainless steel probes that transported the flue gas sample under vacuum to Carnot's CEMS system for determination of NO_x, O_2, CO, CO_2 and N_2O. NH_3 emissions were determined by extracting a metered volume of flue gas through a dilute sulfuric acid solution. The NH_3 concentration of the sample was then determined by a colorization process via spectrophotometry.

Performance testing was conducted to establish the performance of the SCR system over the normal range of loads and operating conditions. Testing was conducted on both coal and natural gas at four nominal load points, 310, 220, 135, and 80 MW (net).

To determine the effect of NH_3/NO_x mole ratio on NO_x reduction and NH_3 slip, several NH_3 injection rates were tested for each load and fuel. The NH_3 injection rates were selected to best approximate the following conditions:

- 60 percent NO_x reduction
  Maximum NO_x reduction at:
  - 5 ppm NH_3 slip at Location D (ahead of air heater SCR)
  - 10 ppm NH_3 slip at Location D (ahead of air heater SCR)
  - 5 ppm NH_3 slip at Location E (down stream of the air heater SCR)

Ammonia and SO_3 in the flue gas stream can react to form ammonium sulfate and ammonium bisulfate which can foul the air heaters. In addition, since PSE&G is able to sell most of its ash, there was concern that excessive ammonia carryover into the fly ash may inhibit these sales. Therefore, it was decided that the target ammonia slips would be tested cautiously while observing the air heater pressure drop.

For each of these conditions composite measurements of NO_x and O_2 were made at Location A and Location D to determine the NO_x reduction performance of the In-Duct SCR. (NO_x concentration was determined by measuring NO from a composite sample and taking periodic single point NO_2 measurements.) For many of the test points NO_x and O_2 were also measured at Locations C and E. This provided data to evaluate the performance of the first layer of in-duct catalyst and the air heater SCR, respectively.

NH_3 measurements were made for each of the above test conditions at Location D. Additional NH_3 measurements were made at Locations C and/or E depending on the test condition (i.e., measurements were typically made at Location C when the NH_3 injection rate was low and measurements were made at Location E when the injection rate was high). All of these measurements were composite samples used to determine the NH_3 slip, an important factor in determining the performance of the SCR system.

In addition to the NO_x reduction and NH_3 slip, one of the most important design parameters to be considered for an SCR system is the conversion of SO_2 to SO_3 by the catalyst. In order to determine the percentage of the SO_2 converted, the inlet and outlet concentrations of SO_2, SO_3, and sulfate in the ash were measured. These measurements were made prior to injecting NH_3 as well as on a periodic basis during NH_3 injection at full load coal firing.

In order to determine the impact of the catalyst on pressure drop through the duct work, the pressure was monitored at each of the Locations A through E. PSE&G installed differential pressure cell and transmitters to monitor and record this data continuously on the plant DCS. The data were used to determine the overall increase in pressure drop at various loads as well as to determine the increase in pressure drop over time and to evaluate the impact of soot blowing on pressure drop.

SCR System

Description

The SCR system was jointly designed by PSE&G, Wahlco, and Siemens and consisted of an ammonia (NH_3) injection system, air soot blowers, in-duct catalyst and catalytic air heater baskets.

The NH_3 injection system consisted of two NH_3 vaporization skids and an injection grid located in the flue gas duct just downstream of the economizer. Aqueous NH_3, at approximately 27 percent by weight concentration, was vaporized in a hot air stream provided by dilution air fans and electric heaters which heat ambient air. The air/NH_3 mixture was routed to a fifteen zone ammonia injection grid (AIG) through a series of throttle (biasing) valves used to match the NH_3 flow distribution with the flue gas NO_x flow to produce a uniform NH_3/NO_x ratio distribution at the catalyst face. As the flue gas exited the economizer, it passed through a Turbomix® static mixer above the AIG, made a 90° turn, and entered duct 22B. Guide vanes were located downstream of the 90° bend to straighten the flow and to balance the velocity distribution at the catalyst face. This reduced catalyst erosion and pressure drop.

Duct 22B was expanded to approximately three times its original cross sectional area to incorporate the in-duct catalyst banks. Inside the duct were two separate banks (5 feet in depth each) of catalyst elements separated by approximately 5 feet of duct space. The catalyst plates were oriented vertically to minimize fly ash deposition. Metal frames secured the catalyst in place. Previous coal-fired boiler SCR experience has been limited to vertical down flow configurations. **This is the first horizontal In-Duct SCR™ application on a pulverized coal unit in the world.**

The plate type catalyst was manufactured by Siemens and consisted of a metal substrate covered by catalyst. The active catalytic materials are Titanium Dioxide and Molybdenum/Vanadium Oxide.

The success of this SCR system is dependent upon the operation of a complex soot blowing system. Air operated soot blowers are located ahead of the first and second catalyst banks to periodically clean the catalyst plates of ash build-up. In addition, horizontal soot blowers are located on the duct floor (in front and behind each catalyst bank) to re-entrain any settled fly ash into the flue gas flow.

The existing horizontal shaft Ljungstrom air heater was modified by replacing the hot-end baskets with catalytic air heater baskets supplied by Wahlco. The catalyst plates were

identical to those used in the In-Duct SCR except that they were located in the rotating wheel of the air heater. Therefore, slightly less than 50 percent of the air heater catalyst was exposed to the flue gas at any given time.

Model Study

Siemens performed a physical flow model study to confirm and optimize the design and check the feasibility of the horizontal In-Duct SCR study for a coal fired application. The duct cross sectional area was increased threefold to maximize space available for the catalyst. Due to space limitations, a short transition was required for the duct expansion. This steep transition created concerns regarding: flow straightening to achieve uniform velocity profiles; pressure drop increases in addition to the pressure drop over the catalyst; and ash sedimentation. Due to the unique horizontal flow arrangement of the reactor, no well-established guidelines exist for addressing these concerns. Finally, ensuring a uniform NH_3/NO_x ratio at the catalyst face is necessary for best NO_x conversion. The following SCR design objectives were established as idealized targets which were not necessarily achievable in all cases, but provided a goal:

- velocity distribution of +/- 15% maximum deviation over 90% of the catalyst face and +/- 20% over the remainder
- NH_3 distribution of +/- 5% maximum deviation
- pressure drop increase of 5 i.w.g.

A 1:20 scale model was built from the economizer exit through the air heater inlet. It included the AIG, economizer hopper, the flue gas duct with a 90° turn into the horizontal #22B duct, the reactor inlet with guide vanes, and the SCR catalyst banks. With the overall geometry established by the duct expansion, the strategy involved varying the inlet guide vane arrangement with the following results:

- velocity distribution of +/- 15% maximum deviation over 90% of the catalyst face and +34% to -23% over the remainder. RMS[1] values ranged from 13% to 15%. The result was deemed acceptable.
- NH_3 distribution of +8% to -13% maximum deviation and RMS values ranging from 5% to 6%. Since the test was performed without bias of the injection grid which would be done in practice, the result was deemed satisfactory.
- pressure drop increase of 6 i.w.g. which included an allowance for ash deposition.

Ash sedimentation was a concern due to low velocity in the expanded duct of 26 fps at full load and 7 fps at minimum load. Simulations were made using actual ash from Mercer Station. No deposition was observed in the catalyst banks but did occur on the duct floor at

[1]Root Mean Square (RMS)

the lower loads. Floor deposits were observed to re-entrain when the gas flow was increased to full load equivalent and it was concluded that sedimentation could be managed with the provision of sootblowers.

Performance

Table 1 shows the predicted NO_x conversion performance of the In-Duct SCR, for coal firing at the four usual power production levels. At full load with fresh catalyst, the NO_x conversion was predicted to be 88% at 10 ppmvd (at 7 % O2) NH_3 slip. After two years, the full load NO_x conversion is projected to drop to 72% at 10 ppm NH_3 slip.

The pressure drop from the economizer outlet to the air heater inlet while firing coal at full load, was predicted to increase from 1 i.w.g. to 7 i.w.g. which included 5 i.w.g. across the catalyst banks.

Commissioning

Optimization of the NH_3/NO_x ratio at the In-Duct SCR inlet was paramount during the commissioning phase of the program. Early, it was determined that the flue gas velocity profile at the AIG was skewed by a factor of about 3:1 across the duct. This prevented the target NH_3/NO_x distribution from being achieved even with some AIG valves wide open and others nearly closed. Therefore, Siemens recommended addition of a Turbomix® static mixer which substantially reduced the flow bias to $< 2:1$.

Figures 3 and 4 compare the normalized AIG flows and the NH_3/NO_x distributions before and after installation of the mixer. The range of the relative flow settings was greatly reduced and the NH_3/NO_x ratio distribution RMS improved from 6.4% to 5.7%, which compares well with the model study.

The AIG flow settings, established for full load coal firing which is the most stringent condition, were used throughout the program for both coal and natural gas fuels, and all boiler loads.

The pressure drop across the system was unchanged by the mixer. The mixing energy consumption apparently was offset by reducing the velocity gradient at the 90° bend in the duct.

In-Duct SCR Test Results

The Performance of the In-Duct SCR was evaluated with respect to NO_x reduction and NH_3 slip vs NH_3/NO_x ratio at loads from 321 to 80 net MW firing both coal and natural gas. Pressure drop, SO_2 to SO_3 conversion, fly ash effects, ash deposition and various other operating data were collected.

Figures 5 and 6 show NO_x reduction and NH_3 slip vs NH_3/NO_x ratio for all tests. Most tests were made for full load coal, since this is the most critical case for PSE&G with respect to the reduction requirement. Performance firing coal at full load was exactly as predicted, 88% with 10 ppm NH_3 slip, Figure 5.

The catalyst was essentially fresh during the six-month program and, therefore, its performance was at its peak. Since catalyst activity will decline with time, samples are being taken periodically for lab scale activity analysis. Results after six months show the aging is proceeding according to the predicted 72% NO_x reduction efficiency after 2 years. For a commercial installation, total system efficiency could be maintained by additional catalyst and catalyst management subject to pressure drop limitations and SNCR, as discussed later.

The NO_x reduction efficiency, firing natural gas at full load, was less than when firing coal. This is due to the higher gas flow, higher space velocity, inherent to gas firing stoichiometry and heat rate. Also, the AIG was not reset for natural gas since it would be impractical to do so given the frequent fuel switching practiced.

NO_x conversion performance is presented Figures 7 and 8 as outlet NO_x in lb/mmbtu and NH_3 slip for full load coal and gas firing. The 0.2 lb/mmbtu target was achieved at less than 10 ppm NH_3 slip for both fuels; again, the results were with fresh catalyst.

Figure 9 shows performance for both fuels at all loads at an NH_3/NO_x ratio of about 1.0. NO_x conversion at the lower loads is greater than 90% as predicted. This reflects the lower gas flow at low load which more than offsets reduced gas temperature. The variability observed at full load is due to variation in the baseline NO_x.

Overall, the In-Duct SCR was able to achieve NO_x reductions in the range of 85-90% while maintaining the ammonia slip to less than 10 ppm as measured at the air heater inlet. NO_x reductions in excess of 95% are possible, however, ammonia slip would be greater than 30 ppm.

Air Heater SCR Test Results

The AHSCR raised the SCR system full load coal firing NO_x reductions to > 90% with no measurable gaseous NH_3 slip. NO_x reduction above 95% was achieved while still maintaining gaseous NH_3 slip to less than 5 ppm at the air heater outlet. Figure 10 shows AHSCR NO_x conversion vs NH_3 slip for full load coal firing together with the previously shown result for the in-duct SCR alone. The NH_3 slip is the gaseous NH_3 downstream of each. The overall performance improvement as a result of the AHSCR is substantial by comparison to the NO_x conversion at 5 ppm slip: 78% for the In-Duct SCR alone and 96% for the combined system. This is primarily due to the greater performance of the in-duct SCR of higher NH_3 slip. It can operate at this level because the downstream NH_3 utilization of the AHSCR utilizes a portion of the ammonia slip for additional NO_x reduction.

Figure 11 shows a similar result for full load natural gas firing where NO_x conversion at full load is improved from 58% to 94% at 5 ppm NH_3 slip. Again, the improvement is mainly attributed to greater allowable NH_3 slip from the In-Duct SCR.

Overall NO_x reduction in terms of lb/mmbtu for the SCR alone and with the AHSCR are shown in Figures 12 and 13. Again, a significant enhancement of overall system performance is demonstrated.

The foregoing results are encouraging indeed. But, when firing coal, there are potential technical limitations that must be taken into account on a case by case basis. Depending on the disposition of the fly ash, its NH_3 content can become an issue when looking at salability (primarily the concrete industry). During the Mercer project, ash samples were taken by PSE&G's Research and Testing Lab and resulted in no repeatable trend. For 5 to 10 PPM NH_3 in the flue gas, approximately 175 to 300 PPM of ammonia was detected in the fly ash. Ammonium bisulfate fouling of the air heater is another operational concern but there was no indication of this during the entire program. Additionally, established correlations of NH_4HS0_4 deposition considering NH_3 and SO_3 concentrations and the AHSCR temperature profile indicate that it should not be a future problem for Mercer station. This is a site-specific issue that must be carefully addressed for each application.

Pressure Drop

Figure 14 presents the average full load differential pressure measured across the individual SCR system components. The data are weekly averages for all periods firing coal at loads greater than 90% of full load. Although there is some fluctuation, there is no trend in time. An effective soot blowing program was instrumental in minimizing deposits and therefore pressure drop across the catalyst banks. After testing more frequent soot blowing cycles, it was determined that twice per day was sufficient to prevent increased pressure drop.

The average pressure drop from the economizer to the air heater inlet was 3.2 i.w.g. at full load. This is roughly the net increase of the In-Duct SCR catalyst banks, the AIG, and the mixer. It is less than predicted mainly due to successful ash deposition management and some conservatism in the original estimate.

As the current fans are capable of overcoming the additional pressure drop, there is no associated capital cost. The additional power consumption per ID fan is approximately 325 BHP at full load.

SO_2/SO_3 Conversion

One of the important design parameters for an SCR system is the catalytic oxidation of SO_2 to SO_3. Their concentrations and sulfates in the ash were measured at locations B, D, and E with and without ammonia injection firing coal at full load.

The average conversion of SO_2 was 1.7% across the In-Duct SCR. The vast majority of the SO_3 formed, however, was found to react with alkaline constituents in the fly ash forming sulfates and, in this benign, form is unavailable for ammonium bisulfate formation in the air heater.

Hybrid SNCR/SCR System Using Aqueous Ammonia (NH_4OH)

Description

One of the purposes of the SNCR component of the SNCR/SCR Hybrid NO_x Reduction System is to supplement the NO_x conversion capacity of the downstream SCR and especially to compensate for its decreasing capacity as the catalyst ages. Another is to provide NH3 slip to the downstream SCR. In general, both the NO_x conversion and NH_3 slip from the SNCR system increase together. In principle, all of the NH_3 could be supplied as slip from the SNCR, but, mal-distribution of NO_x and NH_3 at the catalyst occur. Since SCR performance is enhanced by uniform spatial distribution of NH_3/NO_x at the catalyst face, the use of an AIG to complement the NH_3 supply is therefore a better design for high-efficiency Hybrid system operation.

The objectives were to 1) demonstrate the performance of aqueous ammonia SNCR, 2) develop and demonstrate overall NH_4OH SNCR/SCR Hybrid System NO_x removal of greater than 90% (this operation simulated the use of SNCR as a catalyst management tool; catalyst age at two years), and 3) obtain data for design of a commercial system.

The SNCR portion of the NH_4OH SNCR/SCR Hybrid NO_x Reduction System was tested using only existing observation ports on the front wall of the upper furnace and openings made in the rear wall above the economizer. The aqueous ammonia injectors used superheated steam as an atomizing and delivery medium by jet mixer nozzles.

Testing focused primarily on full load operation while burning coal only. Limited testing at lower loads was conducted on the front wall to evaluate the effects of cooler flue gas temperatures in the SNCR injection zone.

Hybrid operation was performed at full load using only the four rear wall injectors. NH_3 addition via the AIG was practiced and the proportions of NH_3 supplied to the catalyst was varied from all SNCR NH_3 slip to all from the AIG.

Hybrid SNCR/SCR Test Methods

NO_x and O_2 distribution measurements at Locations B and D with and without ammonia injection were made to determine SNCR NO_x reduction and NH_3 slip. When operating in the Hybrid mode with supplemental NH_3 via the AIG, the NO_x reductions of each stage were determined via baseline vs SNCR/SCR injection-on measurements at B, D, and E. Spatial distributions of NO_x and NH_3 were measured in four and three point composite samples in the 16 point (B&D) and 9 point (E) test grids, respectively. Ammonia concentration was determined by the change in NO_x concentration by composites samples from Locations B and D. Spatial ammonia distributions were also measured by wet chemistry by 4 and 3 point composites samples at Locations D and E respectively.

Hybrid SNCR/SCR Test Results

The NH_4OH SNCR/SCR Hybrid test program was divided into four distinct phases. Since the optimum temperature window for ammonia-based SNCR is narrow, approximately 1600

to 1850°F, the first two test phases were designed to evaluate SNCR performance at two different injection locations. Testing during Phase I evaluated SNCR injection at a location on the front wall of the boiler at the furnace exit. Phase II evaluated SNCR injection at a cooler location on the rear wall of the boiler at the same elevation as the front wall location. Phase III of the test program focused on evaluating the SCR system with injection through the AIG alone to prove the integrity of the SCR system as it was at the end of the SCR Demonstration Project. Phase IV evaluated SNCR/SCR Hybrid System performance with simultaneous SNCR and AIG injection.

A summary of the prominent results of the NH_4OH SNCR/SCR Hybrid demonstration program is provided below:

- An overall NO_x removal of 90% and less than 2 ppmvd NH_3 slip at location "E" with 40% SNCR removal was demonstrated. These results were attained with the AIG settings optimized for the fresh catalyst, and indicate that AIG adjustments would not be necessary as the proportion of required SNCR NO_x removal increases as the catalyst ages.

- NO_x removal by the SNCR subsystem ranged up to half of the total. The tests with Hybrid injection confirm the role of SNCR, for an application such as Mercer with a high NO_x percentage reduction requirement, to be an effective supplementary NO_x removal technique for retrofit situations where space and pressure drop constraints limit the amount of In-Duct SCR catalyst.

- The SNCR optimization tests showed the rear wall to be the preferred injection location for the Hybrid system, where both the NO_x removals and NH_3 slips were higher than those for injection at the front wall location.

- Rear wall SNCR injection is capable of providing 40% NO_x removal with NH_3 slip of 150 ppmvd ahead of the catalyst as shown in Figure 15.

- The overall NO_x removal performance of the rear wall injectors was found to be best at a steam-to-flue gas ratio of 0.6 percent and this condition also provided the most uniform NH_3 slip distribution at the catalyst inlet.

- Sootblowing was shown to have a large effect on baseline NO_x emissions as well as NH_3 slip with rear wall SNCR injection, indicating a significant effect on furnace exit temperature.

- The SCR tests with AIG "only" confirmed that the catalyst performance was similar to that measured during the previous SCR-only demonstration.

The results of the Phase I and Phase II tests indicated that flue gas temperatures at the front and rear wall injection locations were slightly high and slightly low, respectively, for optimal SNCR performance with NH_4OH. Options for further optimization of the SNCR system were identified that can achieve greater than 50% SNCR NO_x reduction.

Conclusions

Evaluation of the In-Duct/air heater SCR and Hybrid SNCR/SCR systems for use on PSE&G's Mercer Generating station has demonstrated sufficient NO_x reduction toward PSE&G's goals. Balance of plant impacts were minimal.

PSE&G's Mercer Duct No. 22B was successfully tested as the first horizontal In-Duct SCR reactor on a coal-fired unit in the world. The vertical orientation of catalyst plates in a horizontal duct arrangement utilizing SCR Sootblowers minimized the pressure drop across the system. There was no measurable ash deposition within the ductwork at all load settings. Pressure drop was less than predicted and within the range originally specified by PSE&G.

The In-Duct SCR achieved the predicted 88% NO_x conversion with <10 ppm NH_3 slip.

The air heater SCR increased the overall system NO_x reduction. At 5 ppm NH_3 slip, the overall NO_x reduction improved by approximately 20% from $\approx$ 75% NO_x reduction at D to $\approx$ 95% NO_x reduction at E. The air heater SCR therefore played an important role in ammonia slip elimination, allowing higher levels of ammonia slip from the stationary In-Duct SCR.

The ability to maintain low NO_x emissions and ammonia slip at all loads and with both fuels, as well as low pressure drop, absence of fouling, and acceptable ash NH_3 content was successfully demonstrated.

The Hybrid SNCR/SCR system tests using aqueous ammonia as a common reagent revealed that the SNCR process can effectively reduce NO_x entering the catalyst system. Operation in this manner can be performed as a means for managing catalyst life because sufficiently high SNCR NO_x reduction can be attained. In addition, these and earlier tests revealed that the SNCR process can furnish NH_3 to the in-duct catalyst for combined performance equivalent to an SCR NH_3 injection system. However, the most efficient SCR performance is achieved when using supplemental ammonia from the ammonia injection grid.[1]

The Mercer project demonstrated that emission reductions of the magnitude required to meet the proposed NO_x limit can be obtained in a cost-effective manner. Overall, PSE&G was able to satisfy its objectives for evaluating these NO_x control technologies.

Acknowledgements

The authors acknowledge and thank the following people for their many contributions to the project: Alex Huhmann, Robert Johnson, Robert Laflesh, James Panacek, Ronald Rauffer, and Peter Tomas of PSE&G; Kerstin Hauenstein, Wolfgang Herr and Horst Spielman of Siemens; William Baker, Kevin Dougherty, James Foley, David Oehley, Felix Spokoyny, and Barry Whyte of Wahlco; Matthew Dugan, Richard Himes, Robert Markoja, Roland Roy, Paul Stahley, and Zachary West of Carnot; Michael Polacek of Stone & Webster; Lawrence Muzio and Randall Smith of FERCO; and Thomas Moskal of Diamond Power.

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1. A. J. Wallace. F.X. Gibbons, R.O Roy, J.H. O'Leary, and E.W. Knell "DEMONSTRATION OF POST COMBUSTION NO_x CONTROL TECHNOLOGY ON A PULVERIZED COAL WET BOTTOM UTILITY BOILER," presented at Air & Waste Management Associations 88th Annual Meeting & Exhibition, San Antonio, Texas, (June 1995).

2. Turbomix is a registered trademark of Siemens.

3. In-Duct SCR is a trademark of Public Service Electric and Gas Company.

TABLES AND FIGURES

Table 1

PSE&G-Mercer 22B Demonstration Predicted Performance
Hybrid NO_x Reduction System Basis: Per Duct Except Boiler Load

		CASE			
BOILER LOAD	WORST	1	2	3	4
Percent (%)	100%	100%	68%	42%	27%
Gross MW	344	344	235	146	92
Net MW	321	321	220	135	80
UNCONTROLLED NO_x EMISSION					
LB NO_x/MMBTU	1.8	1.65	1.36	1.18	0.55
PPMVD @ 7% O_2	1017	931	770	667	311
LB NO_x/HR	1427	1283	710	401	117
IN-DUCT SCR PERFORMANCE					
%NO_x Reduction, Zero Years					
NH_3 Slip = 10 PPMVD7	84	88	95	98	99
NH_3 Slip = 5 PPMVD7	75	82	94	98	98
%NO_x Reduction, Two Years					
NH_3 Slip = 10 PPMVD7	63	72	90	96	99
NH_3 Slip = 5 PPMVD7	46	57	88	95	98

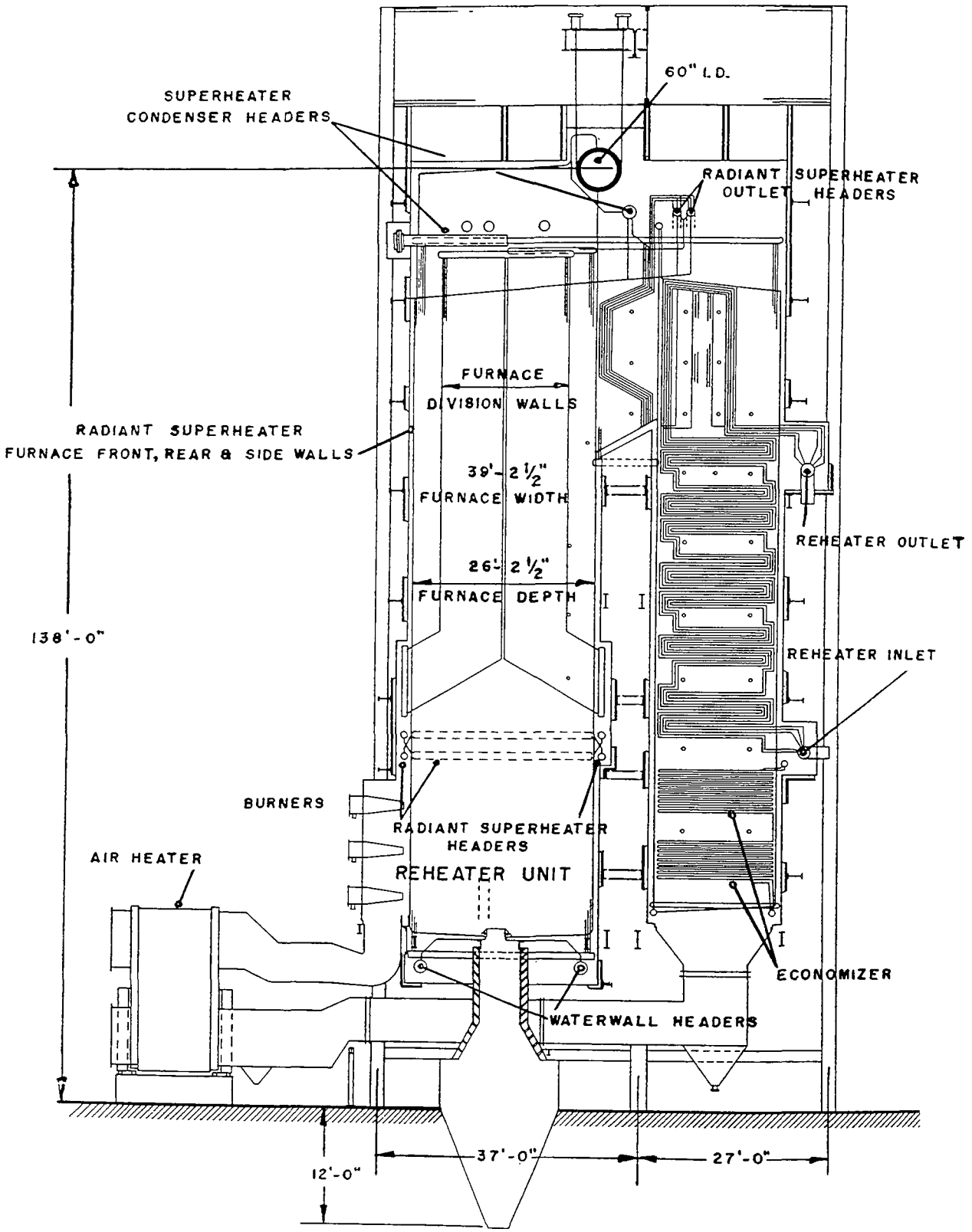
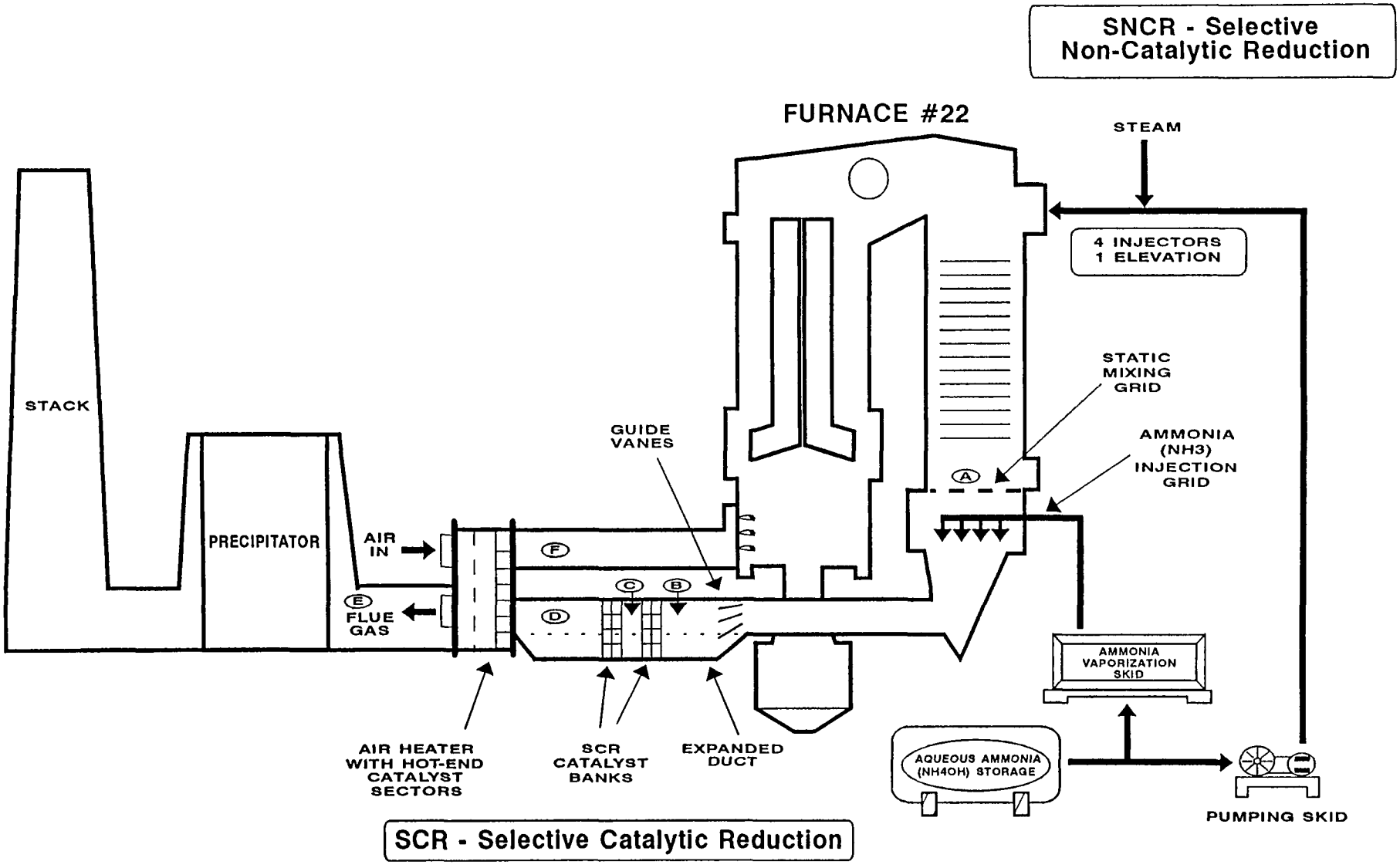


Figure 1 - Side View of Unit No. 2 Reheat Furnace

PSE&G MERCER IN-DUCT SCR/SNCR PROJECT



PSE&G Mercer In-Duct SCR System and SNCR Test Locations

Figure 2

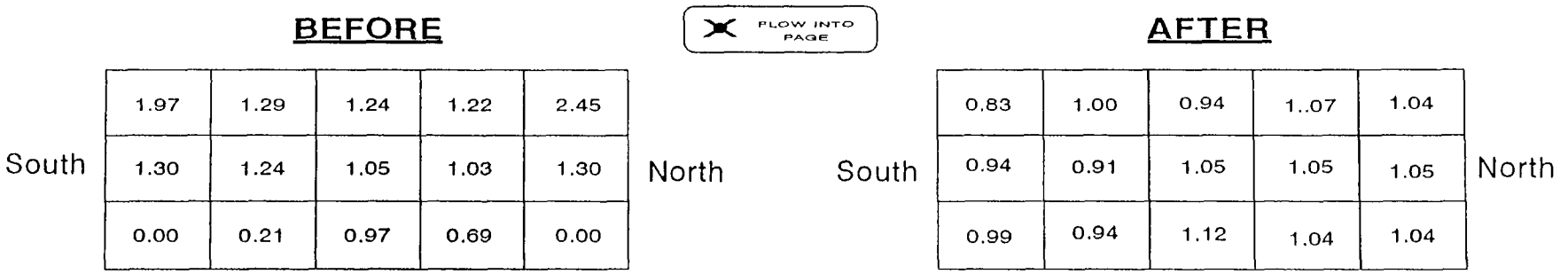


Figure 3 Furnace No. 22 Economizer Exit Cross Section - Normalized AIG Zone Flow Rate Distribution Before and After Installation of the Static Mixer

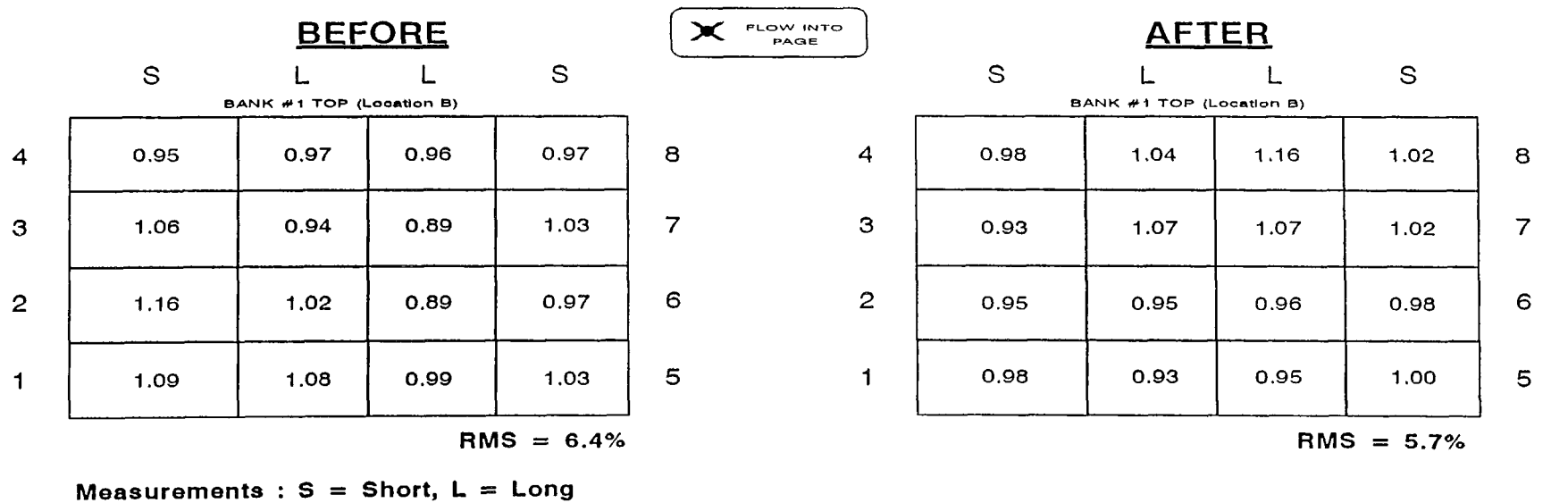
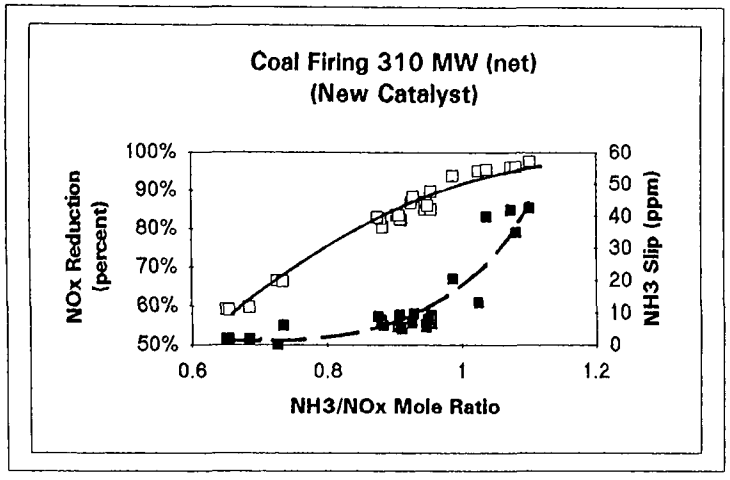


Figure 4 Normalized NH3/NOx Distribution at SCR Bank No. 1 Before and After Installation of the Static Mixer



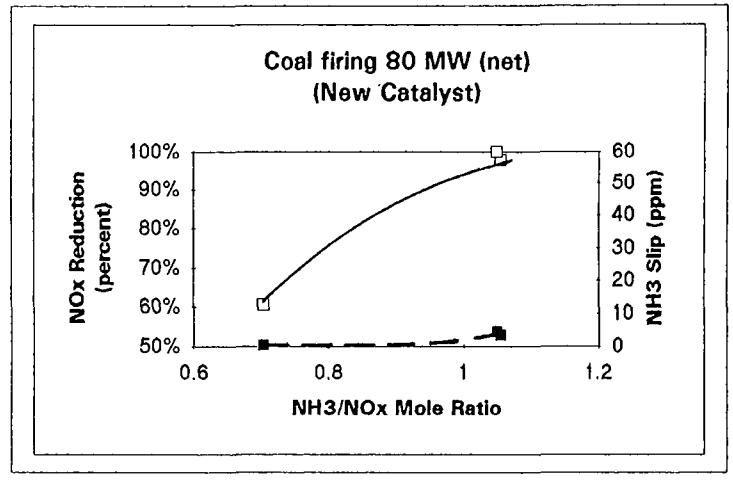
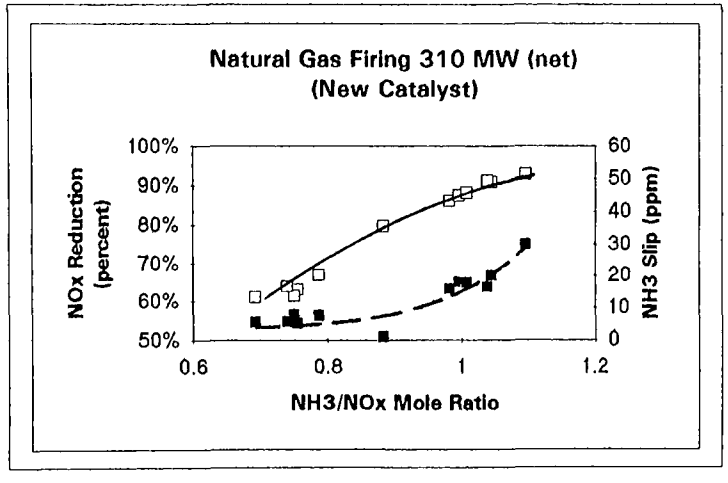


Figure 5 In-duct SCR Perfromance for Coal Firing
NOx Reduction and NH3 Slip as a Function of NH3/NOx Mole Ratio



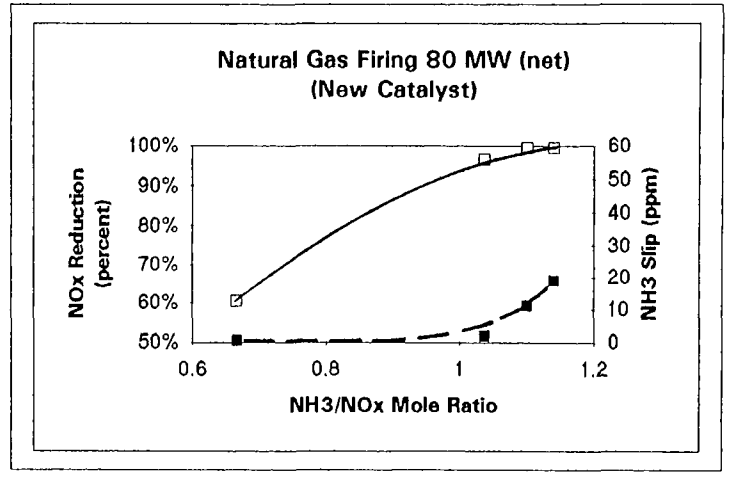


Figure 6 In-duct SCR Perfromance for Natural Gas Firing
NOx Reduction and NH3 Slip as a Function of NH3/NOx Mole Ratio

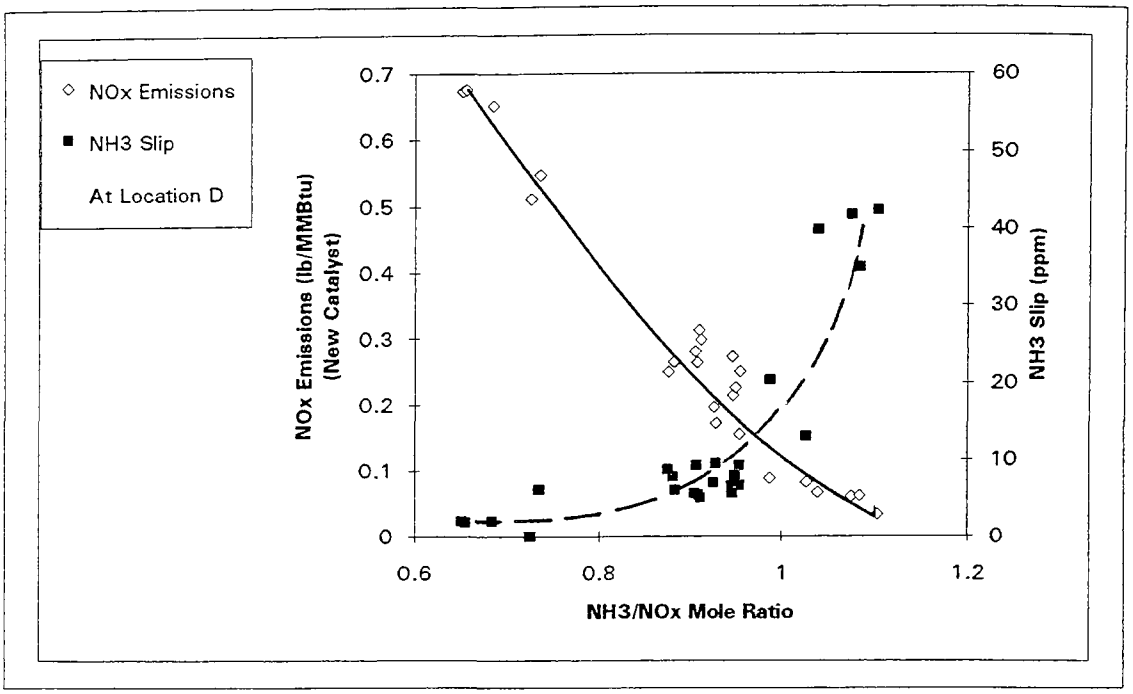


Figure 7 NOx Emissions at In-duct SCR Outlet
Coal Firing 320 MW (net)

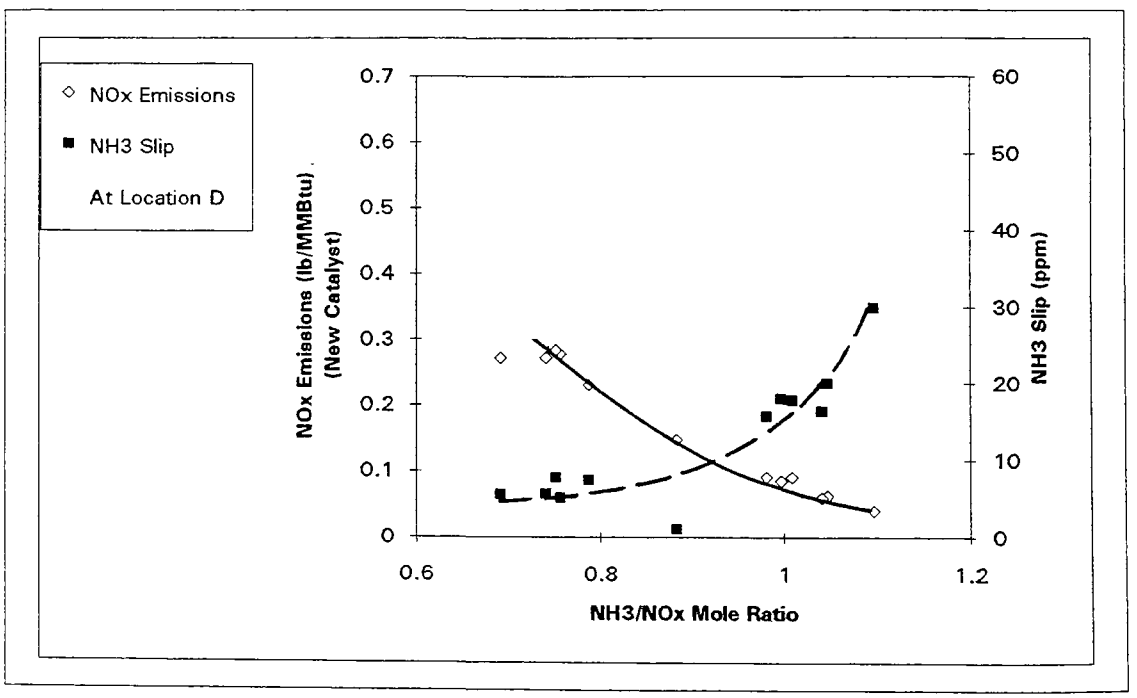


Figure 8 NOx Emissions at In-duct SCR Outlet
Natural Gas Firing 320 MW (net)

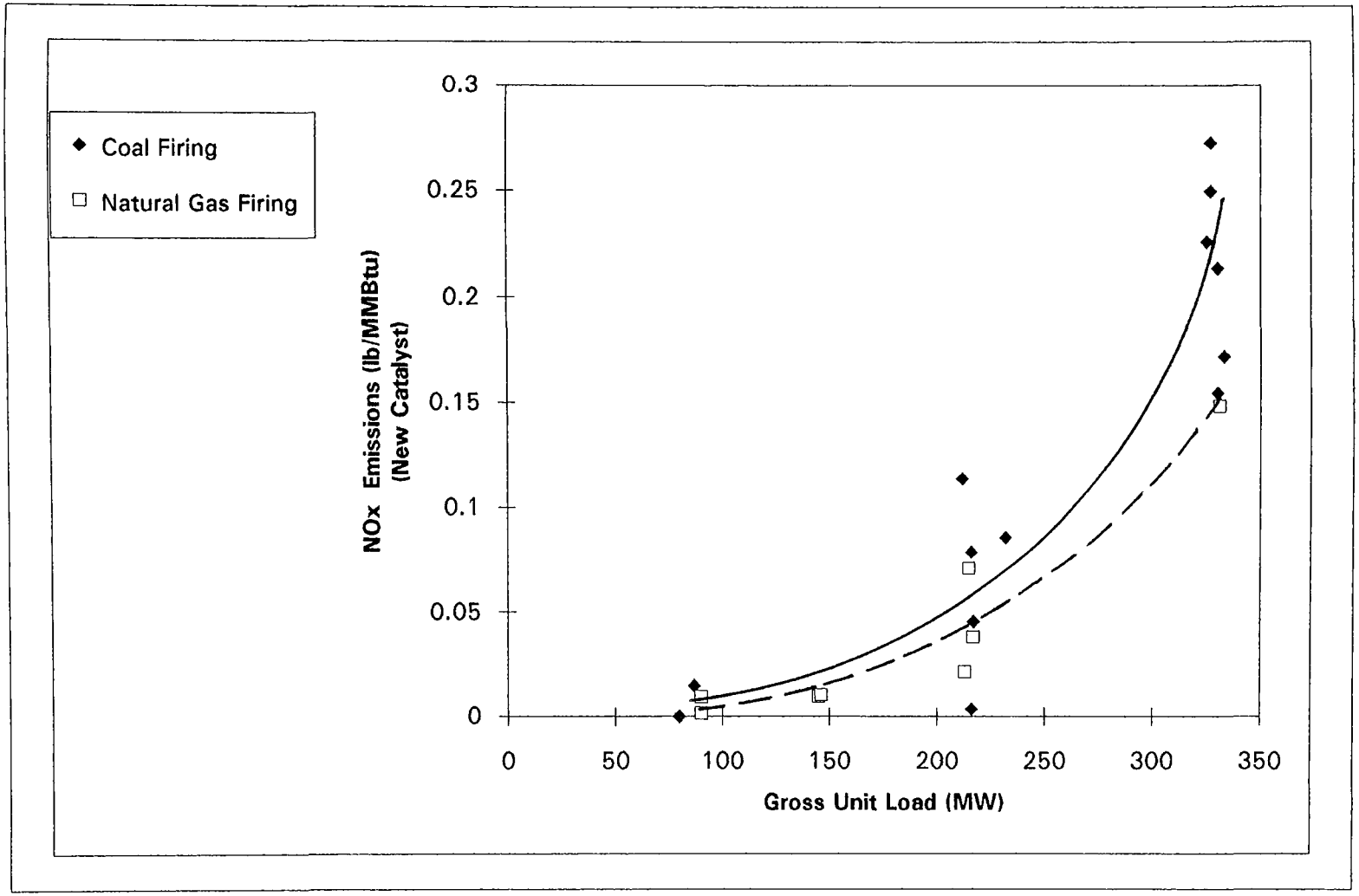


Figure 9 In-duct SCR performance in Most likely Operating Range
NOx Emissions as a Function of Load

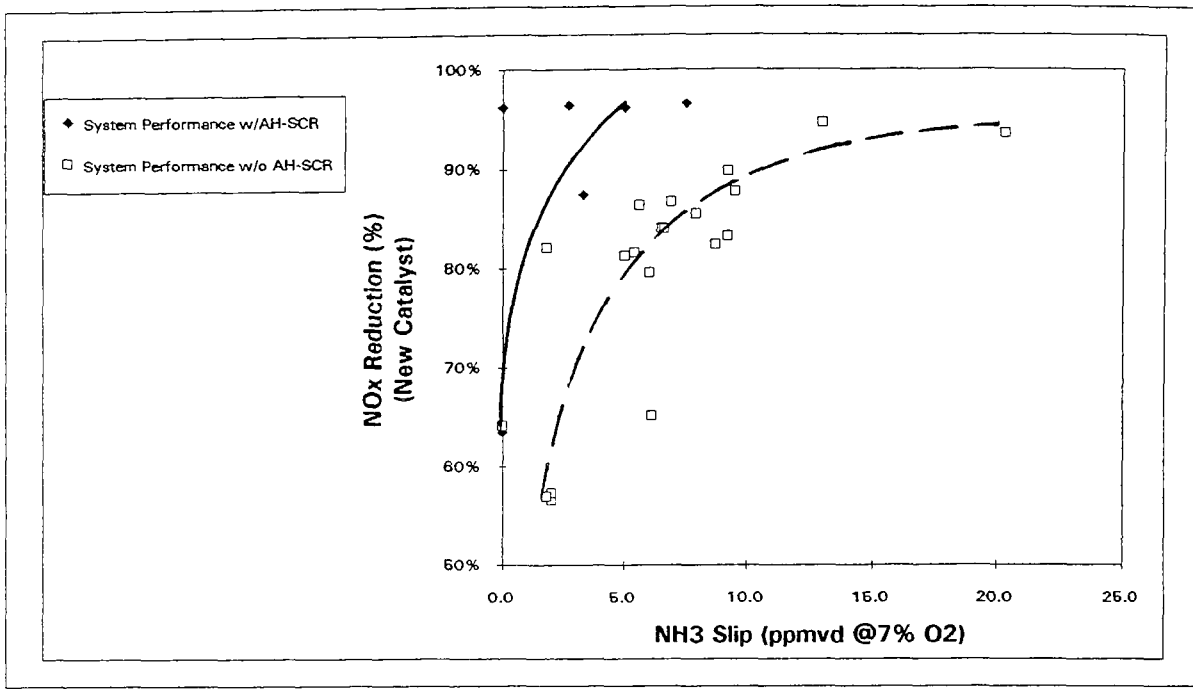


Figure 10 Effect of Catalytic Air Heater on Overall Performance
Coal Firing 310 MW (net)

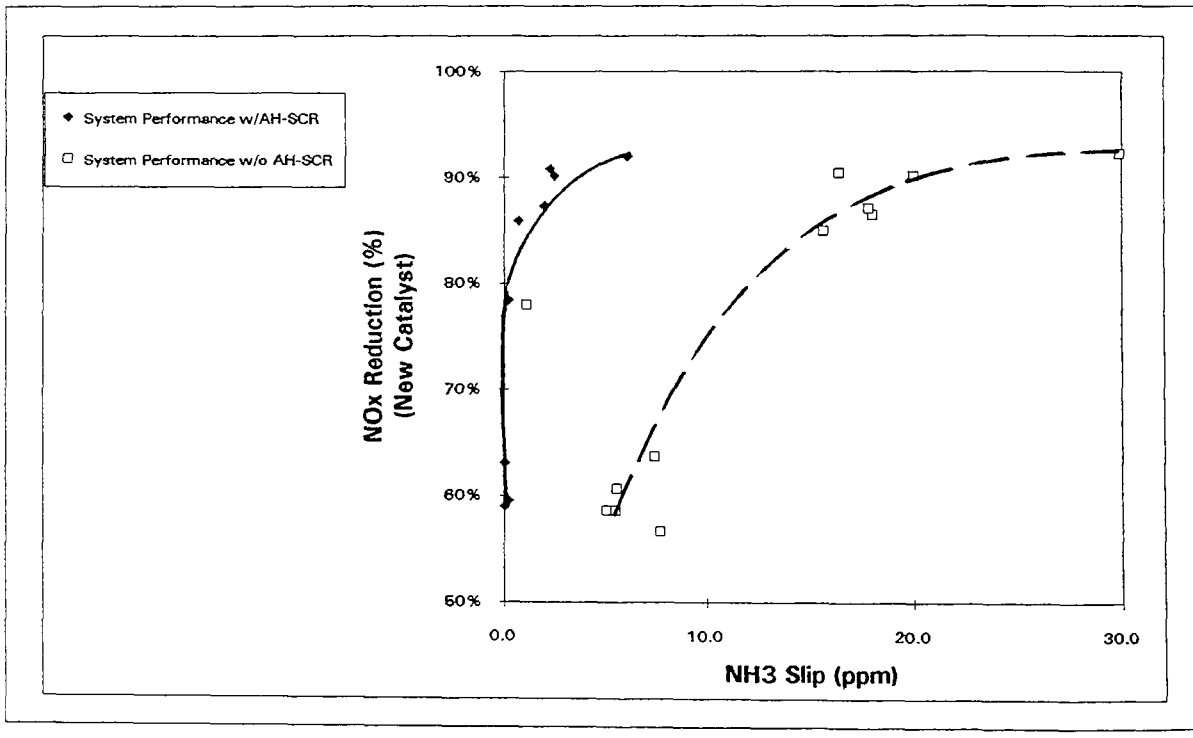


Figure 11 Effect of Catalytic Air Heater on Overall Performance
Natural Gas Firing Firing 310 MW (net)

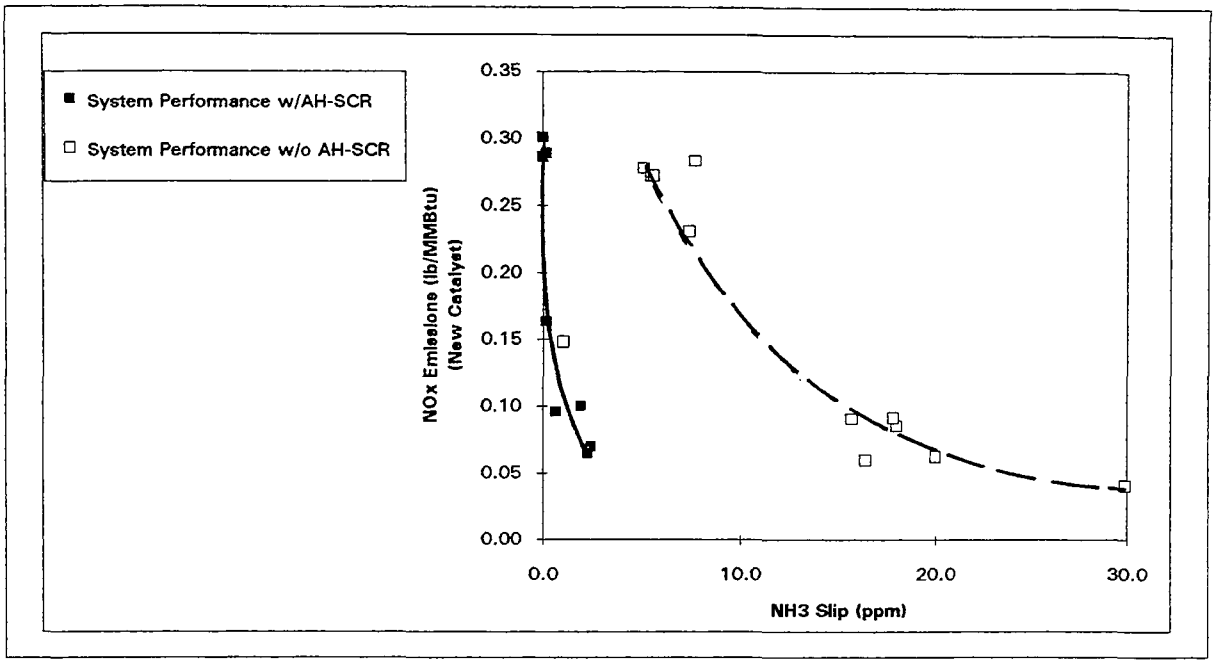


Figure 12 NOx Emissions from In-duct SCR and Hybrid System
Coal Firing 310 MW (net)

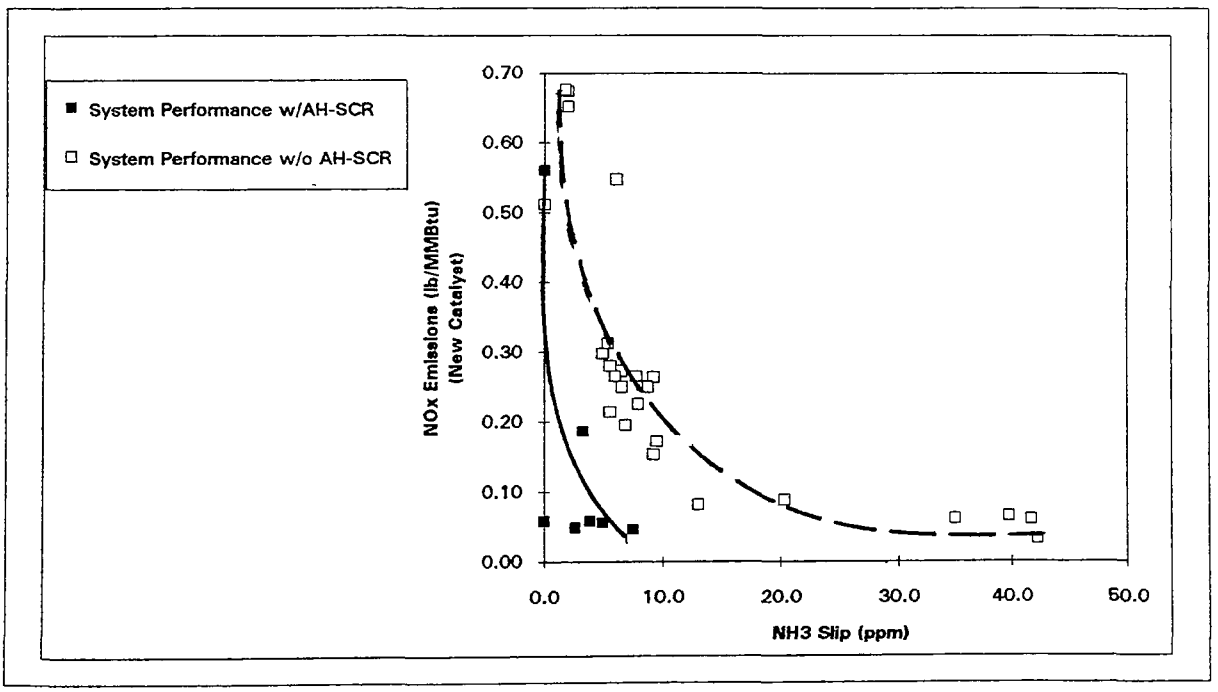


Figure 13 NOx Emissions from In-duct SCR and Hybrid System
Natural Gas Firing 310 MW (net)

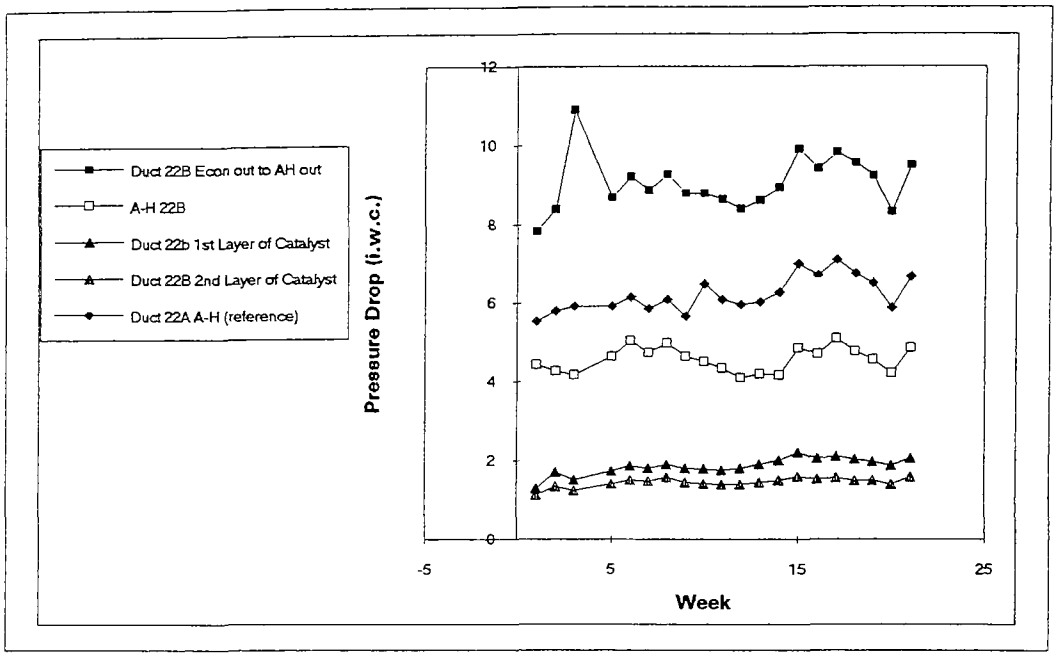


Figure 14 SCR System and Component Pressure Drop vs. Time
Coal Firing 310 MW (net)

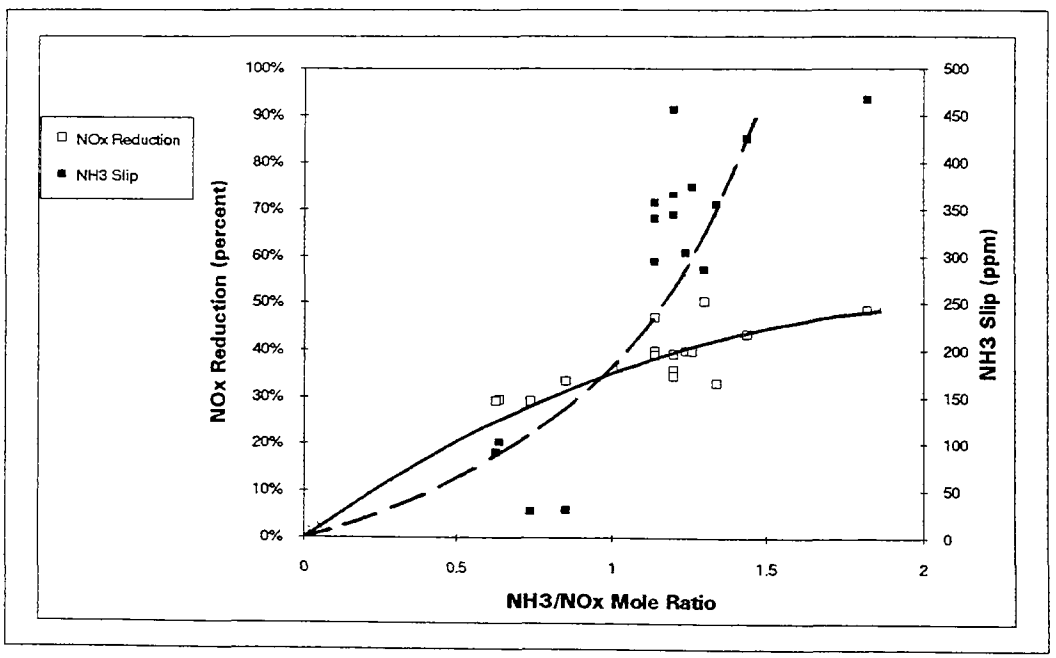


Figure 15 Rear Wall SNCR performance
Coal Firing 310 MW (net)

EVALUATION OF COMBINED SNCR/SCR FOR NO_x ABATEMENT ON A UTILITY BOILER

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Abstract

Public Service Electric and Gas Company has recently completed an evaluation of an SNCR/SCR Hybrid system on a wet bottom, pulverized coal fired utility boiler. This SNCR/SCR Hybrid system utilizes urea based SNCR, provided by Nalco Fuel Tech, and an In-Duct SCR system, supplied by Wahlco Inc., to provide both in-furnace NO_x reduction and/or ammonia to feed the downstream reaction catalyst system. This paper presents the results of the SNCR/SCR Hybrid tests. Performance data for the In-Duct SCR will be presented in a sperate paper.

Introduction

Public Service Electric and Gas (PSE&G) is evaluating the effectiveness of post-combustion NO_x control technologies on a wet-bottomed, coal-fired utility boiler. The technologies under study are conventional urea-based SNCR, In-Duct and air heater SCR, and a combination of SNCR and SCR. While SNCR and, to a limited extent, SCR have been used on coal-fired boilers, these processes have not been demonstrated on a unit with the same configuration as the wet-bottom, continuous slagging, pulverized coal furnaces operated at PSE&G's Mercer Generating Station.

PSE&G's interest in NO_x control for this unit is the result of our overall understanding of the role that NO_x plays in the formation of ground level ozone, and that the Mercer units are high NO_x emitters. When PSE&G assessed the interim state emission inventory it was determined that PSE&G's power plants comprised 27% of the total NO_x emitted in the state of New Jersey. As a result of this emission profile, PSE&G established a public commitment to reduce NO_x from our power plants by 60% by 1995 and 80% by the year 2000. Following the SNCR demonstration in 1993, it was concluded that a greater level of NO_x reduction would be necessary for New Jersey to achieve attainment of the ozone standard. This is because of the severity of upwind transported ozone into New Jersey at levels far in excess of the National Ambient Air Quality Standard. PSE&G subsequently proposed to EPA and the Ozone Transport Commission that a regional NO_x standard of 0.2 lb/mmbtu be adopted on power plants throughout the Ozone Transport Region. The Mercer project demonstrates that emission reductions of the magnitude proposed can be obtained in a cost-effective manner.

In 1993, PSE&G, in partnership with EPRI, conducted a three month demonstration of urea based SNCR on Mercer Unit 2. Results of this demonstration were presented at the 1994 EPRI NO_x Control Workshop, May 11-13 in Scottsdale, Arizona. As a continuing step in the evaluation of post-combustion NO_x control technologies, PSE&G installed a Titanium-Vanadium plate type catalyst in the south (B) duct of furnace 22 between the economizer outlet and air heater inlet, complete with an ammonia injection grid and sootblowers. In addition, the 22 B hot side air heater baskets were replaced with a similar catalyst material. Evaluation of the SCR system began in June and was completed in October, 1994. In order to further evaluate NO_x control technologies, PSE&G set up an additional demonstration program to evaluate a combined SNCR/SCR hybrid system.

The initial goals of the hybrid system evaluation program were:

- Demonstrate the technology for PSE&G and the industry.
- Reduce the load on the catalyst by maximizing the pre-catalytic NO_x reduction
- Identify/quantify NO_x concentrations before and after the catalyst
- Identify/quantify NH_3 slip and effect on the fly ash.
- Minimize the impact on plant operations.

Evaluation of the urea based SNCR/SCR hybrid system was conducted during the last quarter of 1994.

Unit Description

Mercer Generating station has two Foster Wheeler continuous slagging, twin-furnace steam generating units. Units 1&2 are identical with a rated capacity of 321 MW net each. Each unit is designed for 2,060,000 lb/hr of superheated steam flow at 2456 psig and 1050°F. Reheated steam flow is 1,760,000 lb/hr at 445 psig and 1050°F. In this twin furnace design, one furnace generates superheated steam and the other reheated steam. Both units were originally designed with pressurized furnaces and have been converted to balanced draft with some modifications to the radiant superheater to allow for operations at 25%, load without adverse effects on boiler operations. Figure 1 shows a side view of the reheat furnace on Unit No. 2.

The Mercer units burn low sulfur Eastern bituminous coal as the primary fuel, with natural gas available for start-up and as a secondary fuel. Fuel enters each furnace through twelve front wall mounted burners arranged in three levels of four burners per level. There are three water tube division walls located above the top burner elevation and centered between each row. During coal firing, the furnaces are wet-bottomed, continuous slagging by design. The burners are arranged close to the furnace floor to keep the slag in a molten state at all loads. While it is possible to introduce coal and gas through the same burner at one time, general plant practice has been to operate each row of burners on either coal or gas. Fuel distribution between the furnaces is controlled to maintain the appropriate superheat and reheat temperatures.

Furnace gases exit the boiler through a convection superheater in one furnace and a convection reheater in the other. Economizers are installed after both the convection reheater and superheater; however, the economizer in the superheat furnace has more surface area. The flue gas splits into two ducts after each economizer to transfer the gases around the slag tank under each furnace. The flue gases pass through four Ljungstrom horizontal shaft regenerative air heaters (two per furnace) and are ducted to a multi-chamber electrostatic precipitator.

Description of Stand Alone SNCR System

For the 1993 SNCR evaluation, only the reheat furnace of unit 2 (furnace 22) was outfitted with SNCR. Since the basic design of each furnace is the same, the results on one furnace should be the same for the other. Nalco Fuel Tech's (NFT) mobile demonstration unit was used to perform the evaluation of SNCR at four injector levels in the furnace with four injectors per level. There were two SNCR injection control zones setup for this demonstration, each zone had a separate pressure and water/chemical ratio controls. Urea used during the demonstration was NFT's NO_xOUT purchased from a vendor licensed by NFT.

SNCR NO_x reductions were on average of 38% while maintaining ammonia slip of approximately 5 ppm. Based on the results of the SNCR demonstration program, PSE&G has purchased full scale systems to be installed on both Units 1 and 2 for compliance with New Jersey's NO_x RACT regulations operative on May 23, 1995.

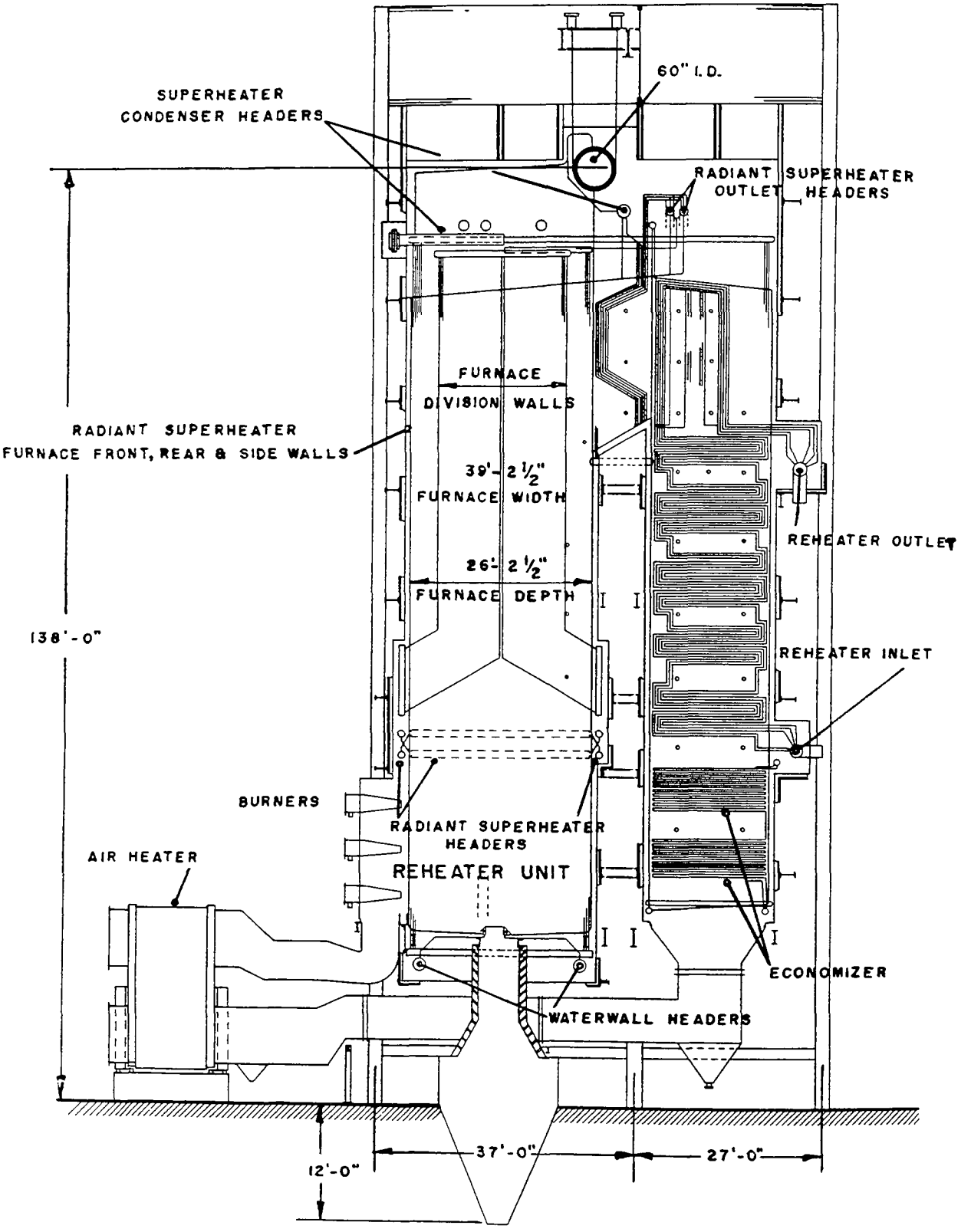


Figure 1 - Side View of Reheat Furnace on Unit 2

In-Duct and Air Heater SCR

As described earlier, the flue gas from each furnace at the economizer exit splits into two separate ducts. For this technology evaluation project, catalyst was installed in only the south duct (22B). This would in theory mean that only one half of the reheat furnace's flue gas passed through the SCR or one quarter of the total unit's flue gas volume (equivalent to $\approx$ 80 MWn). Certain imbalances in the flow characteristics required biasing via a damper in the 22A duct to balance the flows.

The SCR system was jointly designed by PSE&G, Wahlco Inc., and Siemens and consisted of an ammonia (NH_3) injection system, air soot blowers, in-duct catalyst and catalytic air heater baskets. The NH_3 injection system consisted of two NH_3 vaporization skids and an injection grid located in the flue gas duct just downstream of the economizer. Aqueous NH_3, at approximately 27 percent by weight concentration, was vaporized in a hot air stream provided by dilution air fans and electric heaters which heat ambient air. The air/NH_3 mixture was routed to the injection grid through a series of throttle (biasing) valves used to balance the NH_3 distribution to match the flow and NO_x stratification. Downstream of the injection grid, the flue gas made a 90° turn and entered duct 22B. Guide vanes were located downstream of the 90° bend to straighten the flow and to balance the velocity distribution at the catalyst face. This reduces catalyst erosion and pressure drop.

Duct #22B was expanded to approximately three times its original cross sectional area to incorporate the in-duct catalyst banks. Inside are two separate banks (5 feet in depth each) of catalyst elements separated by approximately 5 feet of duct space. **This is the first horizontal In-Duct SCR™ application on a pulverized coal unit in the world.** The catalyst elements were manufactured in Germany where SCR experience has been only applied to vertical down flow configurations. The plate type catalyst was manufactured by Siemens and consisted of a metal substrate covered by catalyst. The active catalytic materials are Titanium Dioxide and Molybdenum/Vanadium Oxide. The catalyst plates were oriented vertically to minimize fly ash deposition. The plates were arranged with metal frames to secure the catalyst in place.

The success of this catalytic system is dependent upon the operation of a complex soot blowing system. Soot blowers are located ahead of the first and second catalyst banks to periodically clean the catalyst plates of ash build-up. In addition, horizontal soot blowers are located on the duct floor (in front and behind each catalyst bank) to re-entrain any settled fly ash into the flue gas flow. The existing horizontal shaft Ljungstrom air heater was modified by replacing the hot-end baskets with catalytic air heater baskets. The catalyst elements were similar to those used in the in-duct SCR except that they were located in the rotating wheel of the air heater, and therefore, slightly less than 50% of the catalyst was exposed to flue gas at any given time.

During the In-Duct SCR optimization phase of the test program, it was determined that the flue gas velocity profile at the economizer exit was skewed by a factor of about 3:1 across the duct. The NH_3 injection grid did not have sufficient adjustment to accommodate for the high flow bias in the economizer exit. Therefore, a flue gas Turbomix* static mixer was designed/installed by PSE&G/Siemens. The mixer substantially reduced the flow bias which improved the NH_3 distribution and hence, SCR performance. The mixer also reduced the pressure drop across the system apparently by reducing flow separation at the 90° bend in the duct.

Description of SNCR/SCR Hybrid System

The SNCR/SCR Hybrid process consists of utilizing both SNCR and SCR NO_x control technologies simultaneously to achieve a final NO_x reduction target. The SNCR portion of the system was operated in various modes in concert with the In-Duct SCR system at Mercer. The modified SNCR process utilizes more of the SNCR temperature window providing more effective NO_x reduction with increased reagent utilization, while simultaneously producing process ammonia (via SNCR NH_3 slip) to feed the downstream catalyst.

The Nalco Fuel Tech demonstration trailer was installed on Unit 2 with urea injection equipment set up on the same four furnace levels utilized in the 1993 SNCR demonstration (identified as levels 5, 6, 7, and 8). In the hybrid demonstration, injection was on the "B" side (south half) of Furnace 22 and injectors were located in the convective passes on the south side of Furnace 22. Four additional ports were installed on the rear of the Boiler at Level 8 (identified as 8R) using standard flow injectors designed to spray between the support tubes of the down pass. One additional port location was tested on the south sidewall of the boiler utilizing a Multiple Nozzle Lance (MNL level). The MNL was a stationary water cooled injector designed to provide a very fine spray of reagent from multiple nozzles installed along the length of the probe.

Three water tube division walls in the furnace divided gas flow into four straightened flow paths from side to side. The injectors for the SNCR system were aligned to inject between the division walls. By injecting through ports only on the south side of the furnace, the effects of the process were confined primarily to the "B" duct or south side of Furnace #22. The flue gas mixer installed at the economizer outlet in front of the NH_3 injection grid provided mixing of both the NO_x and the ammonia slip.

The additional convection section wall injector (8R) and the MNL ports were designed to take advantage of the greater NO_x reduction capability and reagent utilization available at temperatures lower than would be used for stand alone SNCR (below 1800° F). Together with the existing SNCR injection locations, the SNCR/SCR Hybrid injection strategy was designed to provide stages of chemical injection which yield more in-furnace NO_x reduction with attempts to achieve a uniform ammonia distribution to the catalyst. In some of the tests, reagent feed for catalyst reduction was provided by deliberately utilizing injection at temperatures below the SNCR window limit to produce NH_3.

During the demonstration, the SNCR/SCR Hybrid system was operated in the following configurations:

- Maximum SNCR in-furnace NO_x reduction. Ammonia feed for the catalyst was provided by the SCR 's Ammonia Injection Grid (AIG).
- Maximum SNCR in-furnace NO_x reduction with ammonia feed for the catalyst provided by high NH_3 slip from the SNCR.
- SNCR operated to provide maximum NH_3 feed to the catalyst, with in-furnace NO_x reduction a secondary benefit.
- SNCR Operation to provide NH_3 to the catalyst while maximizing overall reagent utilization.
- Maximum SNCR in-furnace NO_x reduction without concern for ammonia slip.

Description Of Test Programs

In order to evaluate this NO_x reduction technology, a comprehensive test program was developed. Carnot was contracted in 1993 to assist PSE&G in the evaluation of the SNCR process and the SCR and SCR/SNCR Hybrid systems in 1994.

The programs were similar in that baseline NO_x emissions as well as CO, CO_2, O_2, SO_2, SO_3, NH_3, and N_2O data was collected at various loads on coal, gas and combinations of the two fuels. Figure 2 shows the relative sampling locations.

A multi-point probe grid was installed in the ductwork upstream of the catalyst, between catalyst banks and downstream of the in-duct catalyst by PSE&G prior to the testing program. The grid consisted of a series of stainless steel probes that transported the flue gas sample under vacuum to Carnot's CEMS system for determination of NO_x, O_2, CO, CO_2 and N_2O. NH_3 emissions were determined by extracting a metered volume of flue gas through a dilute sulfuric acid solution. The NH_3 concentration of the sample was then determined by spectrophotometry.

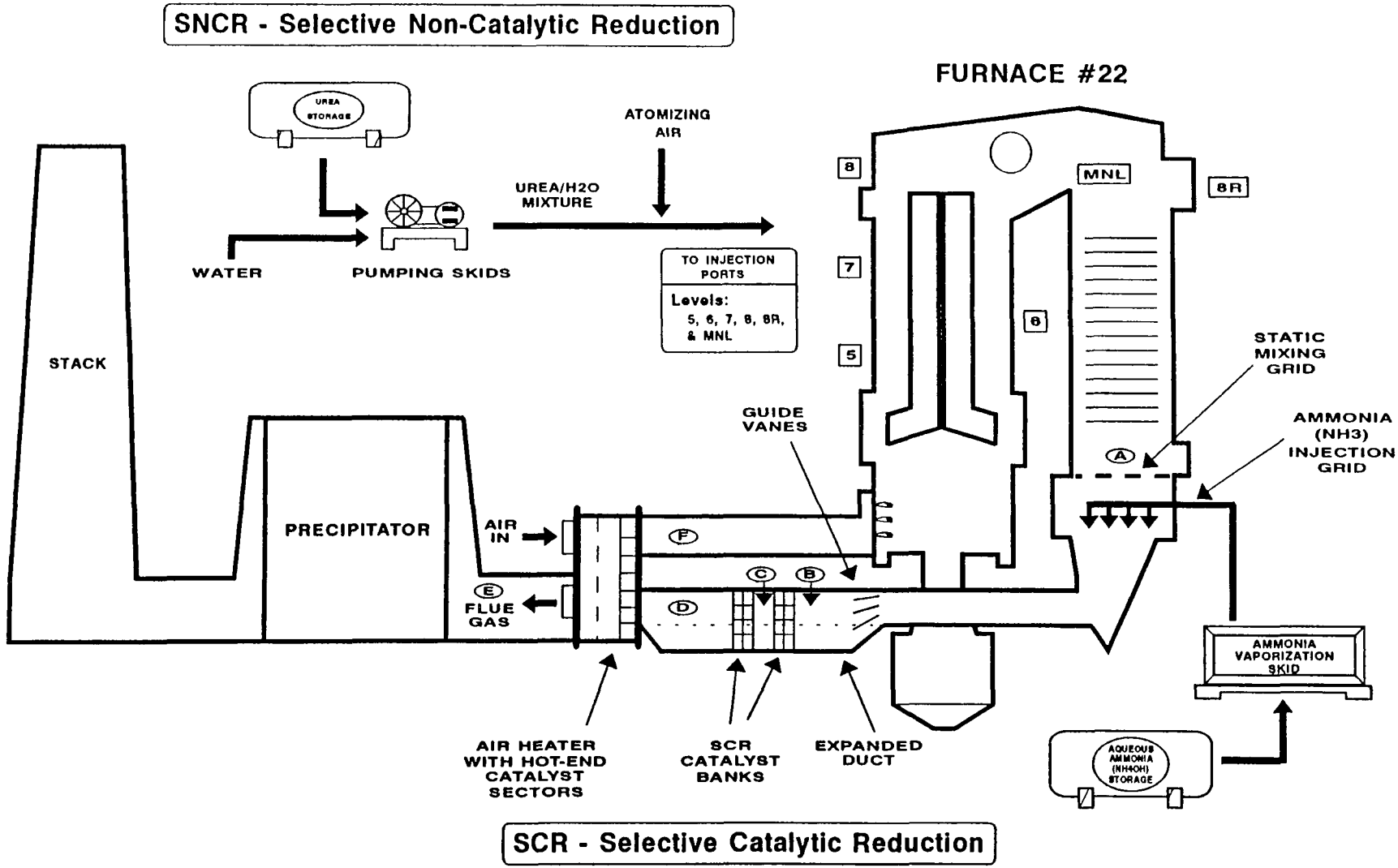
For each of the test conditions composite measurements of NO_x and O_2 were made at Location A and Location D to determine the NO_x reduction performance of the In-Duct SCR. NO_x concentration was determined by measuring NO from a composite sample and taking periodic single point NO_2 measurements. For many of the test points NO_x and O_2 were also measured at Locations C and E. This provided data to evaluate the performance of the first layer of in-duct catalyst and the air heater SCR, respectively.

Urea Based Hybrid SNCR/SCR Performance

The performance of the Hybrid SNCR/SCR system is dependent upon the unit operations. The primary factors which influence the performance of the hybrid system are the load (catalyst space velocity, furnace and duct temperatures, and NO_x concentration), fuel burned (NO_x concentration, flue gas constituents), and the injection configuration (furnace NO_x reduction, ammonia production, NO_x and NH_3 stratification). As with the In-Duct SCR and the catalyst air heater, the greatest demands on the hybrid SNCR/SCR system performance are during full-load, coal-fired operation due to the high space velocity and high inlet NO_x concentration.

During the demonstration, three injection configurations were tested and will be discussed. One configuration employed the in-furnace injectors used during the previous SNCR demonstration. These injectors were operated at increased chemical flow rates in an attempt to provide high furnace NO_x reductions and the ammonia necessary for the SCR process. Total chemical utilization was greater in this operational mode than in traditional SNCR. Another method of operation utilized the in-furnace SNCR injectors with the chemical flow rate set for maximum NO_x reduction and utilization, typically an NSR of 1.1 to 1.3, with the required NH_3 reagent provided by the AIG. The last operating mode involved injecting urea into the back pass just upstream of the economizer. Special ports and injectors were used that would not typically be employed with an SNCR system due to the low temperatures at the location of the chemical release. This configuration resulted in the highest chemical utilization.

PSE&G MERCER IN-DUCT SCR/SNCR PROJECT



PSE&G Mercer In-Duct SCR System and SNCR Test Locations

Figure 2

Figures 3 and 4 show a summary of the SNCR and SCR NO_x reduction as a function of the NSR for the injection configurations tested during full load coal operation. It can be observed in Figure 3 that the NO_x reduction falls onto three curves for the different injection levels. In Figure 4, the injection levels that deliver chemical to the coolest regions of the furnace, MNL and 8R, provide the greatest overall NO_x reduction with the highest chemical utilization when combined with the full scale SCR. Injection through these levels provided chemical utilizations in excess of 75 percent and NO_x reductions in excess of 90% when not considering limits on the NH_3 slip. The overall NO_x reduction provided by levels 8/8R is less for a given NSR than that of the other high ammonia production levels due to the lower utilization of level 8 as a result of the higher temperatures at the point of chemical release.

Figure 3 shows the furnace NO_x reduction possible with the different injection configurations. As observed in this Figure and with Figure 4, the two injection configurations that provided the best combination of high chemical utilization and high overall NO_x reduction (Figure 4) do not correspond to the best furnace NO_x reduction. The MNL/8R configuration provides about 20 percent in-furnace NO_x reduction at an NSR of 1.1. The low furnace NO_x reduction is a result of the low temperature at the point of chemical release. The temperature at the release point is well below the optimum temperature required for the NO_x reduction reactions but is high enough for the efficient production of NH_3. This results in an injection configuration that is very efficient at producing NH_3 to supply the catalyst but is inefficient at reducing NO_x prior to the catalyst.

Levels 5/6/7 as shown in Figure 3 provide in-furnace NO_x reductions of up to 35 percent at an NSR of 1.6. Increasing the injection rate beyond this results in a rapid decrease in the chemical utilization. During the hybrid tests which used the AIG, the urea was injected in the furnace at an NSR of 1.1 to 1.3 (which is similar to the operating conditions determined during the SNCR evaluation performed in June - September 1993) to provide furnace NO_x reductions of approximately 30% and minimize the chemical usage.

Use of levels 5/6/7 and the AIG provided NO_x reductions in excess of 90%, however a greater amount of chemical is required to achieve NO_x reductions similar to MNL/8R. This mode of operation was set for maximum NO_x reduction in furnace with minimal NH_3 slip going to the In-Duct SCR and would best simulate the operation of the Hybrid SNCR/SCR system as a catalyst management tool.

Although high overall NO_x reductions are possible with both the MNL/8R and levels 5/6/7 used with the AIG, the NH_3 slip after the In-Duct SCR is higher for the MNL/8R configuration for a given NO_x reduction. This is a result of the poor NH_3/NO_x distribution at the catalyst face. Since the NO_x distribution was determined to be uniform at the inlet to the catalyst during baseline testing, any stratification resulting from the SNCR or SCR NO_x reduction process may be seen by observing the NO_x stratification at the outlet of the catalyst as long as the injection rate is sufficiently low so that no NH_3 passes through the catalyst unreacted.

Figure 5 compares the NO_x stratification at the outlet of the catalyst for the MNL and 8R injector configurations. The location of the SNCR injector plays an important role in the NH_3 slip distribution across the In-Duct SCR catalyst face. The influence of the injector location can be clearly viewed in the figure, with the MNL providing the highest NO_x reduction at the bottom of the catalyst face (D1,5) and the 8R injector providing the most reduction at the top of the catalyst face

(D4,8). Reference to Figure 2 shows that this corresponds to the placement of the two injection locations in the back pass of the furnace. As shown in Figure 5, use of the MNL and 8R injectors together improves the NO_x reduction profile, however it is difficult to achieve the level of NH_3/NO_x distribution necessary for optimum catalyst efficiency as provided by the AIG. Although it is not shown in this figure, it should be noted that the NO_x reduction associated with the MNL injector is primarily due to ammonia production and reaction in the in-duct catalyst. The NO_x reduction associated with level 8R is more evenly divided between SNCR and SCR reduction.

Levels 5/6/7 were also tested at high injection rates in an effort to provide high furnace reductions while producing sufficient NH_3 to supply the catalyst. Levels 5/6/7 were chosen due to the high levels of NO_x reduction provided during the SNCR demonstration performed in 1993. Reference to Figure 4 indicates that these injectors, in addition to levels 7 and 8 which were also located in the furnace, distribute the chemical in regions that are not conducive to NH_3 slip production. At reduced loads, higher levels of ammonia production were observed, but the chemical utilization was not sufficient to make in-furnace injection a feasible operating mode.

Several hybrid injection configurations are capable of providing high overall NO_x reductions with chemical utilizations in excess of 50%. To more clearly differentiate between the operating modes, the configurations with the highest overall NO_x reduction are shown in Figure 6. The figure shows a comparison of the NO_x reduction and chemical utilization during full load coal firing for the different hybrid configurations. The ideal curve which represents the NO_x reduction that would be achieved with 100% chemical utilization and the conventional In-Duct SCR system performance are included for comparison.

Of the Hybrid SNCR/SCR tests, the MNL/8R injection configuration provided the best performance at chemical flow rates corresponding to an NSR of approximately 0.8; however, above this injection rate, the high ammonia stratification at the inlet to the catalyst limits the overall NO_x reduction performance.

It is also interesting, and somewhat deceiving, to observe the performance of levels 5/6/7 used in conjunction with the AIG. The advantage of this configuration is that furnace NO_x reductions in excess of 30% can be achieved without altering the NH_3 distribution to the catalyst. The chemical utilization for this injection configuration is not fixed, but may approach the In-Duct SCR curve if the SNCR injection rate is decreased. The system could be operated with little or no SNCR injection while the catalyst is new and benefit from high chemical utilizations, and then as the catalyst efficiency decreases overall NO_x reduction could be maintained by increasing the SNCR injection.

Summary

The test programs at Mercer Unit No. 2 demonstrated the ability of SCR and Hybrid SNCR/SCR systems to provide moderate to high NO_x reductions while firing low sulfur coal or natural gas. The variables affecting the performance of each system, and any operational impacts were determined.

The following is a summary of the performance of the systems along with any associated boiler impacts:

- Operation of the hybrid system by injection into cooler regions of the back pass provided overall NO_x reductions of up to 85% with an NH_3 slip of less than 10 ppm prior to the air heater catalyst at full load during coal operation. Overall chemical utilizations of up to 90 percent were demonstrated, approaching that of the In-Duct SCR operated with the AIG.
- Use of injection levels MNL/8R provided SNCR NO_x reductions of 20% during full load coal firing while providing sufficient NH_3 feed for the catalyst with an overall NO_x reduction of 80 percent before the air heater.
- Furnace NO_x reductions of 35% on average and overall reductions in excess of 90 percent were achieved with use of the in furnace injectors and the AIG. The chemical utilization was lower than during injection into the back pass, averaging 50% at high overall NO_x reductions. This mode of operation provided the best SNCR/SCR Hybrid distribution of ammonia across the catalyst face.
- The In-Duct SCR operating with the AIG provided higher levels of NO_x reduction for a given NH_3 slip than the hybrid system utilizing urea. The poor furnace NO_x reduction and NH_3 stratification at the inlet to the catalyst reduced the effectiveness of the hybrid system.
- From the In-Duct SCR system testing, the overall NO_x reduction was increased by approximately 20% while maintaining the NH_3 slip limit by using the catalyst air heater. The NO_x reduction across the air heater was small, averaging 20 ppm, but allowed higher reductions to be achieved by the in-duct catalyst through operation at higher NH_3/NO_x mole ratios and at elevated NH_3 slip.
- Soot blowing performed twice each day maintained the pressure drop across the horizontal in-duct catalyst at acceptable levels over the course of the program.
- SO_2 to SO_3 conversion was moderate, averaging 1.7 percent, or approximately 6 ppm. The impact of the conversion rate was mitigated as the SO_3 formed reacted with the calcium in the fly ash minimizing the detrimental effects of SO_3.

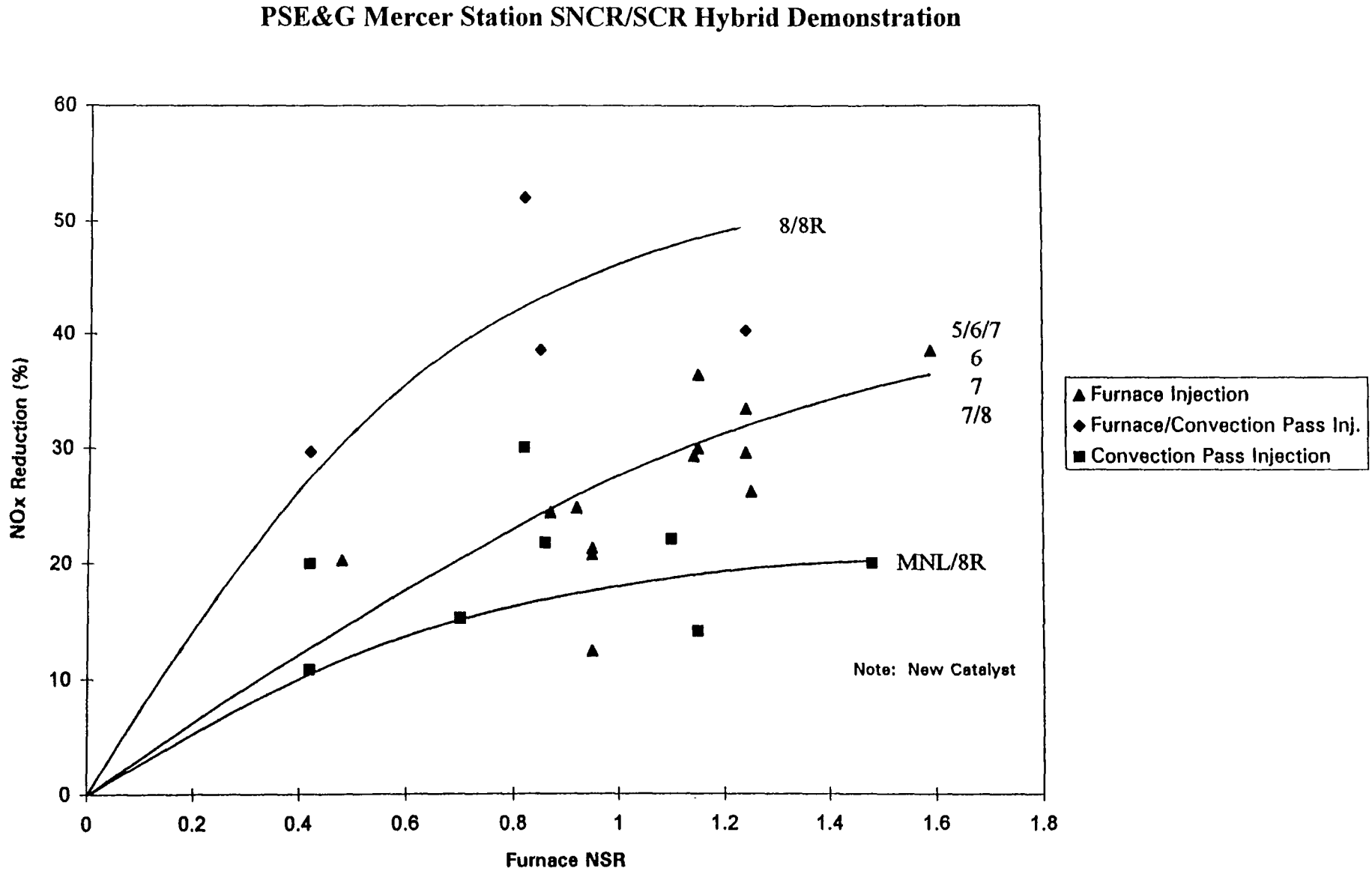


Figure 3. SNCR Reduction Measured at Location B as a function of NSR, Coal Firing 310 MW net

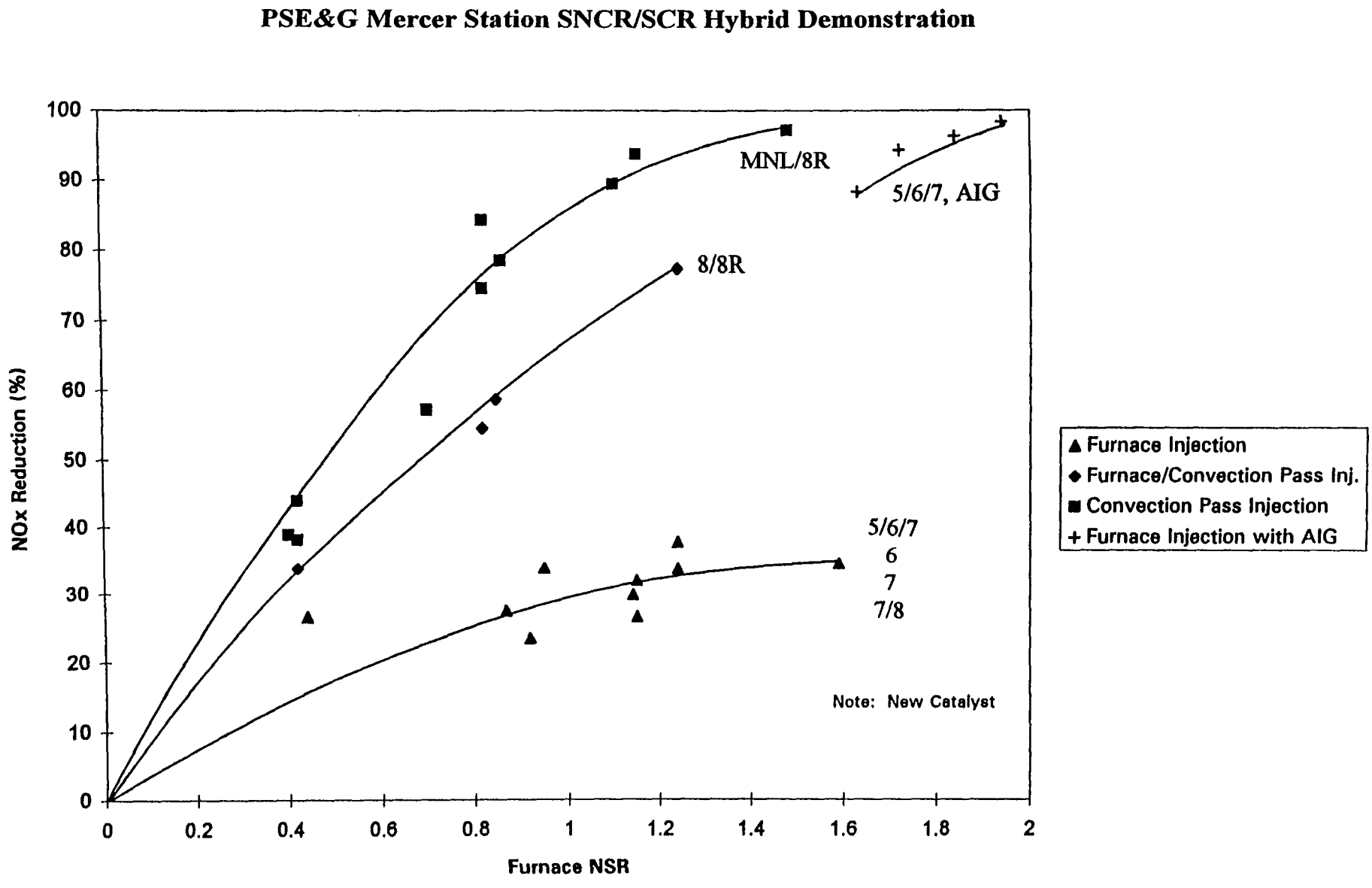


Figure 4. Overall reduction as a function of NSR, Coal Firing 310 MW net.

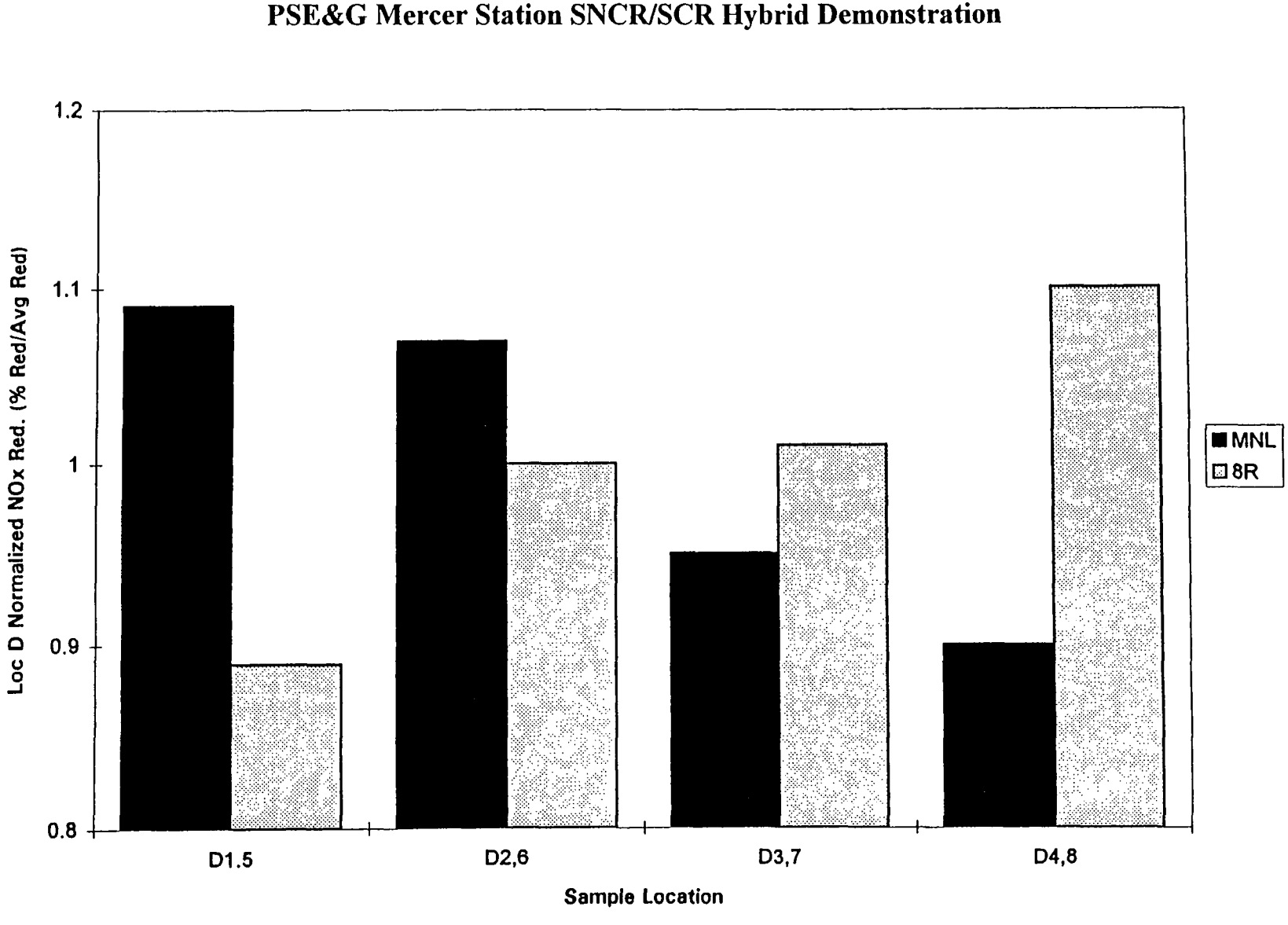


Figure 5. NOx Stratification at Outlet of SCR, Coal Firing 310 MW net

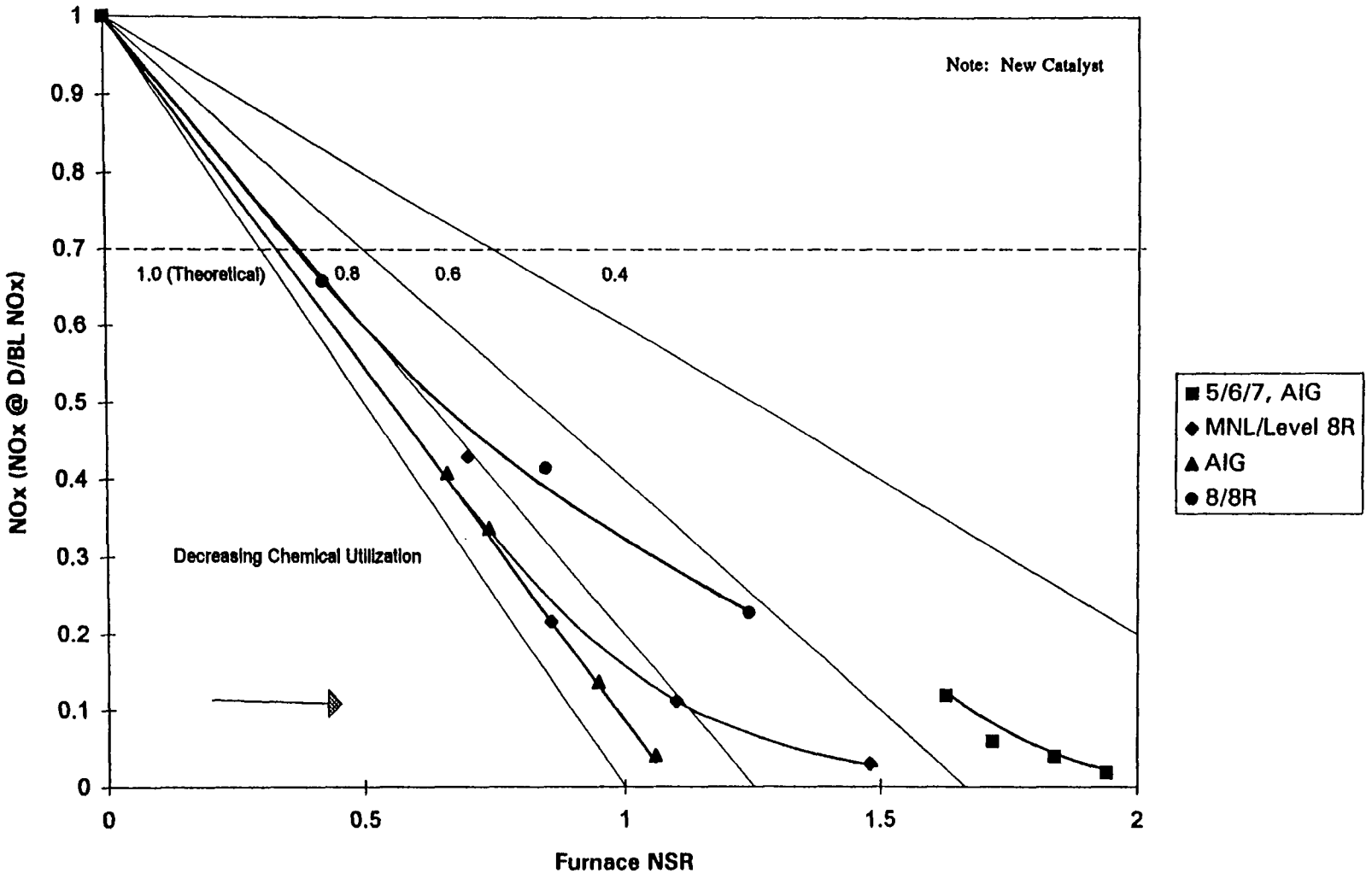


Figure 6. Comparison of NOx Reduction and Chemical Utilization during Full Load Coal Firing

CONCLUSIONS

The SNCR/SCR Hybrid has been successfully demonstrated at PSE&G's Mercer Generating station. Operation of the SNCR in combination with the In-Duct SCR have been determined to be an effective means of reducing NO_x emissions for compliance with New Jersey CAAA compliance regulations. By controlling the amount of ammonia slip prior to the air heater to 5-10 ppm, air heater fouling will not be a major concern.

The SNCR/SCR Hybrid testing has demonstrated the following:

- Deterioration of SCR catalyst performance can be compensated for by use of SNCR without significant plant operational impact. This in effect will allow PSE&G to defer catalyst replacement costs, if necessary.
- When ammonia slip is not a concern, due to the presence of catalyst, the SNCR injection profile can be modified to take advantage of a greater portion of the SNCR temperature window. Thus, higher in-furnace NO_x reductions are possible with higher in-furnace chemical utilization rates than with traditional SNCR applications. The SNCR operations can be configured to provide NH_3 as a feed for a downstream catalyst, while still providing some in-furnace SNCR NO_x reductions. When considering for this mode of operation, the NH_3/NO_x distribution across the catalyst face must be a design criteria in a planning commercial Hybrid SNCR/SCR system.
- Depending on the **specific site**, a Hybrid SNCR/SCR system can be included in the design of an In-Duct SCR system. Use of the SNCR/SCR Hybrid could reduce the overall catalyst size making it fit in existing duckwork or resulting in smaller SCR reactors. Supplemental catalyst ammonia injection (AIG) may or may not be necessary depending on the SNCR injection configuration, and the desired level of overall NO_x Reduction from the combination.

Acknowledgements

(1) In-Duct SCR is a trademark of Public Service Electric and Gas Company.

(2) Turbomix is a registered trademark of Siemens.

EXPERIENCE AND CONSIDERATION OF SNCR-SCR HYBRID SYSTEM

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Kansas City, MO

Abstract

SNCR-SCR Hybrid Systems combine deNOx technologies to overcome the limitations of SNCR-only applications, such as low deNOx efficiency and high concentration of residual reagent, e.g., slip ammonia, while offering, in some cases, lower installation costs than SCR-only applications. Successful application of an SNCR-SCR system must consider:

- Boiler characteristics.
- Required deNOx performance.
- Optimum injection system.
- Combination/interface between SNCR and SCR.

Mitsubishi Heavy Industries (MHI) has applied hybrid systems technology on two full scale utility boilers. A discussion of this experience is presented herein.

Introduction

Mitsubishi Heavy Industries, Ltd. (MHI) has developed several NOx emission control technologies since the 1970's (Table 1). MHI's NOx control developments include low NOx combustion firing systems, NOx reduction by oxidation (wet process) and NOx reduction by NH_3 injection (dry process). MHI has acquired considerable experience commercializing these technologies.

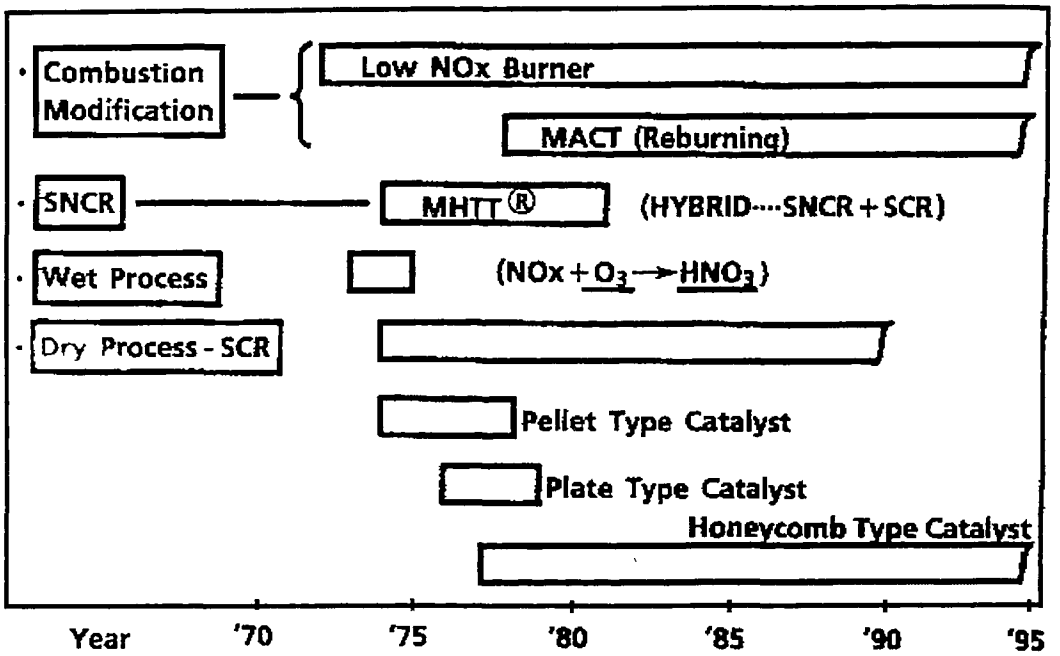

Table 1
History of Mitsubishi deNOx Technology

This paper discusses the combination of two commercially available technologies used to control NOx: Selective Non-Catalytic Reduction (SNCR) and Selective Catalytic Reduction (SCR). Both methods utilize a reagent such as ammonia (NH_3). SNCR requires for the injection of the reagent into the furnace at high temperatures. While this method requires less equipment and thus requires comparatively less capital investment, its performance is generally limited to 20 to 50% NOx reduction and results in high residual NH_3. Further, where SO_3 is present, high NH_3 from SNCR combines with SO_3 to form ammonia salts which may foul downstream equipment. SCR has earned its reputation as a leading NOx control technology by achieving up to 95% reduction with very low NH_3 slip (< 5 ppm) in full scale units. When SCR is installed downstream of an SNCR system, the catalyst reduces not only NOx but also un-reacted NH_3 from the SNCR thus making the hybrid system a viable solution. MHI applied its hybrid system to two units in the 1970's and subsequently patented its technology.

Cormetech, Inc., a licensee of the Mitsubishi SCR technology, engineers, manufactures, and tests catalyst for SCR applications. Cormetech has participated in two tests of hybrid systems: a slip stream demonstration on a coal fired unit and a retrofit on a package boiler. The process and MHI experience are described in this paper.

SNCR Experience

The SNCR Process

The SNCR reaction requires the presence of NH_3, NOx, and oxygen within an appropriate temperature range. The reaction mechanism predicts that NH_3 initially reacts with oxygen at higher flue gas temperature, creating NH_2 radicals which in turn react with NO to form nitrogen and H_2O.

Primary Reactions:

$$4NH_3 + O_2 \rightarrow 4NH_2 + 2H_2O \quad (1)$$

$$NH_2 + NO \rightarrow N_2 + H_2O \quad (2)$$

Net Reaction:

$$4NO + 4NH_3 + O_2 \rightarrow 4N_2 + 6H_2O \quad (3)$$

The main factors influencing the deNOx reduction reaction in SNCR are working gas temperature and residence time. As shown in Figure 1, under laboratory conditions the required reaction will occur at gas temperatures of 800°C to 1,100°C, with the most efficient reactions at 850°C to 1,050°C. Some NH_3 is converted to NO, N_2, and water requiring the higher NH_3:NOx molar ratio (2:1) than the 1:1 implied in equation 3 above.

SNCR Experience

In most operating commercial boilers, deNOx efficiency with this process will be lower than the high efficiency shown in laboratory test data because of insufficient residence time at the appropriate gas temperature. Figure 2 shows the effect of residence time in laboratory tests. The minimum time for the complete reaction is at least 0.1 second, preferably more than 0.4 second. In actual operating conditions, flue gas temperature cools rapidly as gas contacts heat transfer surfaces such as superheaters, reheaters.

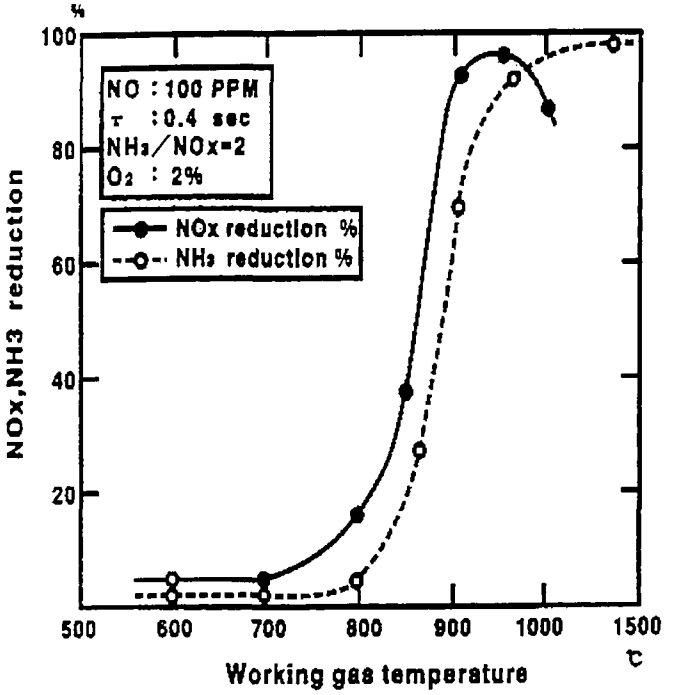

Figure 1
Effect of Working Gas Temperature

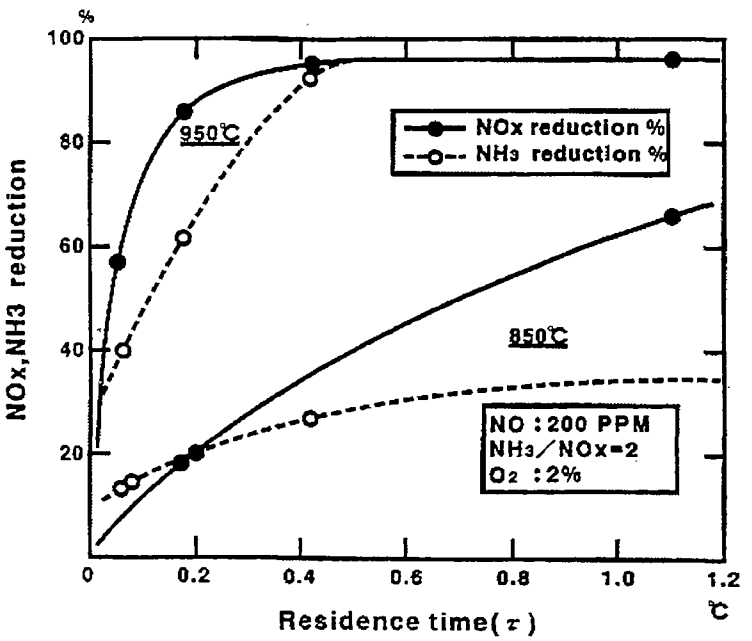

Figure 2
Effect of Residence Time

Additionally, actual operating conditions do not meet the ideal conditions of lab tests; the process of diffusing and mixing flue gas flow with injected NH_3 is not ideal. This diffusion and mixing is critical to deNOx performance. To accomplish sufficient mixing, the system must provide a suitable distribution of NH_3 injection points for the various boiler loads; these injection points may be located either inside the furnace or on the furnace wall. When inside the furnace, in the cross-section of high temperature gas flow, the injection nozzles must be cooled in order to prevent decomposition of the ammonia or overheating of the nozzles. Figure 3 shows an example of the location and the structure of NH_3 injection nozzles. Figure 4 shows deNOx performance results in field tests. It is clear that the location of NH_3 injection point highly influences deNOx efficiency at each boiler load.

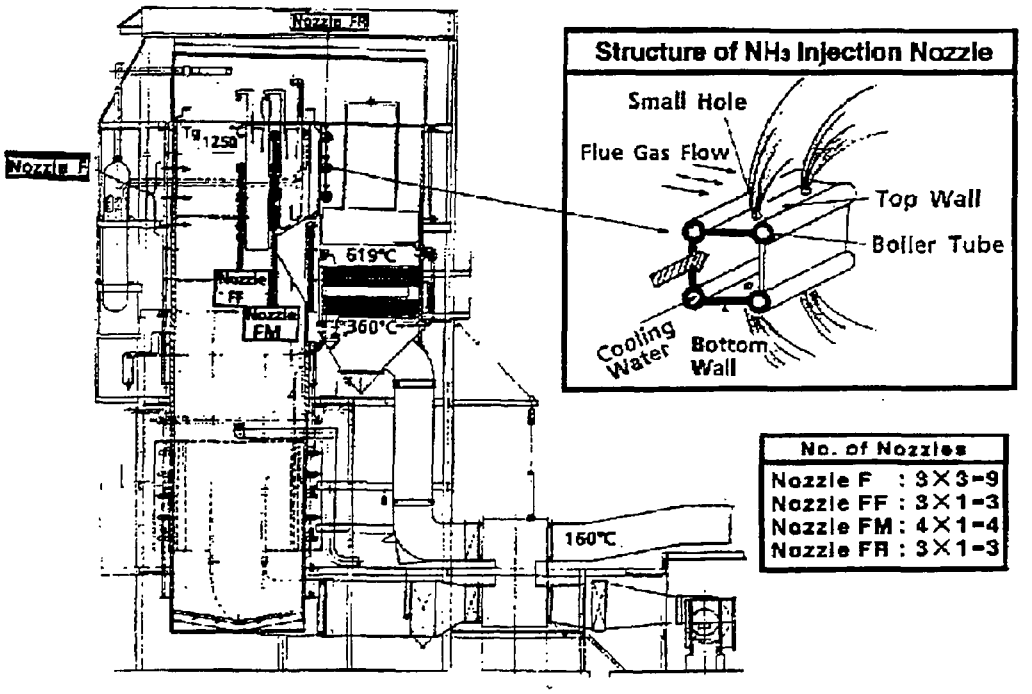

Figure 3
Locations of SNCR Nozzles for Mizushima Unit No. 3

If the ammonia injection nozzle is located on the furnace wall, ammonia can be injected similar to wall sootblower nozzles. However, effective penetration and distribution of ammonia is difficult with such a simple injection system.

Therefore, the performance of the SNCR system is affected by the characteristics of the boiler and injection system. The boiler characteristics, such as its type, performance, structure and heating surface arrangement, operating conditions, etc., impact temperature profile and residence time. Based on these characteristics, each case will present its own unique considerations, such as the location of NH_3 injection nozzles, the nozzle cooling system, NH_3 carrier system, and the gas temperature fluctuation.

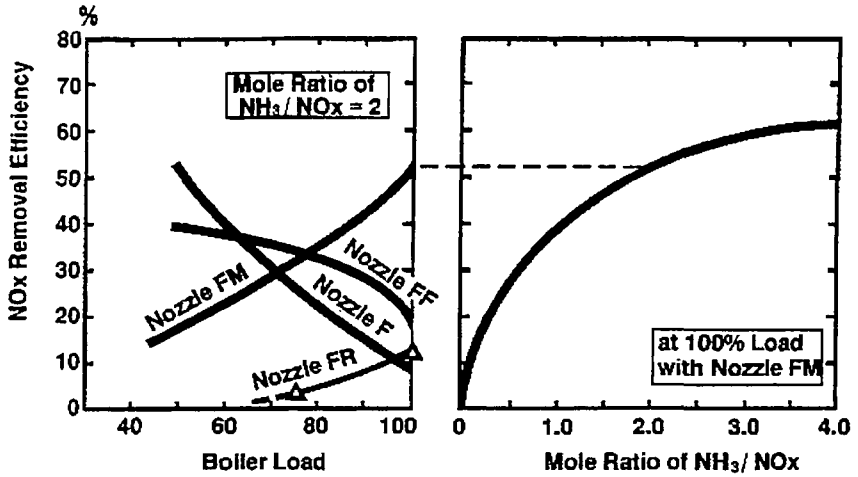

Figure 4
Field Results of Mizushima Unit No. 3

SNCR Discussion

SNCR was applied to several units in Japan in the mid-1970's. However most of these are now out of operation as shown in Table 2. There are two major reasons:

- More stringent environmental requirements, exceeding the capability of SNCR.
- High operating costs due to high consumption of ammonia, blowing medium and auxiliaries.

In case of fuels containing sulfur, NH_3-sulfur compounds such as ammonium bisulfate may deposit in lower gas temperature zones, including the air preheater, fouling the heat transfer surface and increasing pressure drop.

Urea can be used as a reagent instead of ammonia, as urea rapidly changes into ammonia at high temperatures in the furnace. However, use of urea may be more expensive than ammonia and may generate additional byproducts including carbon monoxide.

To summarize the advantages and the disadvantages of SNCR systems:

Advantages:

1. Low gas side pressure drop.
2. Simple equipment and small space requirements.
3. Possible combination with other processes to design most flexible system.

Unit Name	Capacity (Flue Gas Flow, Nm3/H)	Fuel	N/R	Start of Operation	Present Operation
Mizushima I-#2, MCI	175,000	L.S. Oil	R	Aug 1974	Out of operation
Mizushima I-#2, MCI	194,000	L.S. Oil	R	Apr 1977	Out of operation
Mizushima I-#3, MCI	540,000	H.S. Oil	R	Jun 1975	Out of operation
Mizushima I-#4, MCI	540,000	H.S. Oil	R	Oct 1974	Out of operation
Yokkaichi, #1, MCI	200,000	H.S. Oil	R	May 1976	Out of operation
Yokkaichi, YD Boiler, MCI	40,000	H.S. Oil	R	Sep 1975	Out of operation
Chita, #2, Chuba Electric	1,000,000 (Unit Capacity 350 MW)	L.S. Oil	R	Feb 1976	Modification to repowering hybrid system
Yokosuka, #4, Tokyo Electric	1,000,000 (Unit Capacity 350 MW)	L.S. Oil	R	Mar 1978	Hybrid system Out of operation

N = New R = Retrofit

Table 2
List of SNCR (MHTT) Applications

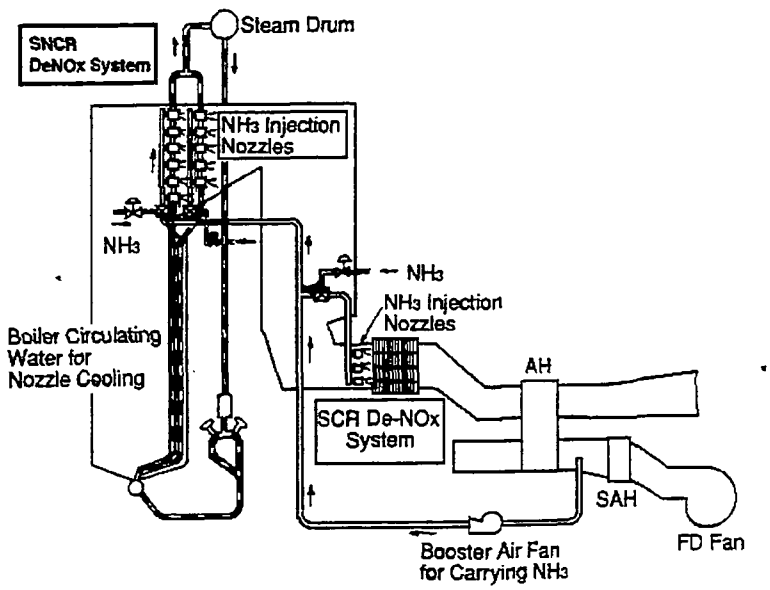

Figure 5
Schematic Diagram of the Hybrid SNCR-SCR System
at Yokosuka Unit No. 4

Disadvantages:

1. NOx removal efficiency is limited.
2. Larger consumption of reagent and blowing medium, such as air, than SCR.
3. High residual NH_3 and resultant ammonium compounds.
4. High operating cost.

Experience with SNCR-SCR Hybrid System

Features and Experience of the Hybrid

The SNCR-SCR hybrid system compensates for the disadvantages of SNCR by combining it with SCR. A combined system can achieve higher deNOx performance and lower residual NH_3 concentration. Figure 5 shows a schematic diagram of a combined SNCR and SCR system applied to an oil fired boiler. An in-duct SCR system is installed downstream the SNCR between the economizer and the air preheater. Figure 6 shows the test results on this plant's total deNOx performance at a rated load. The result achieves the predicted performance and significantly reduces residual NH_3. Based on the successful results, MHI applied for and received a U.S. Patent for its SNCR-SCR hybrid system—U.S. Patent No. 4, 302, 431.

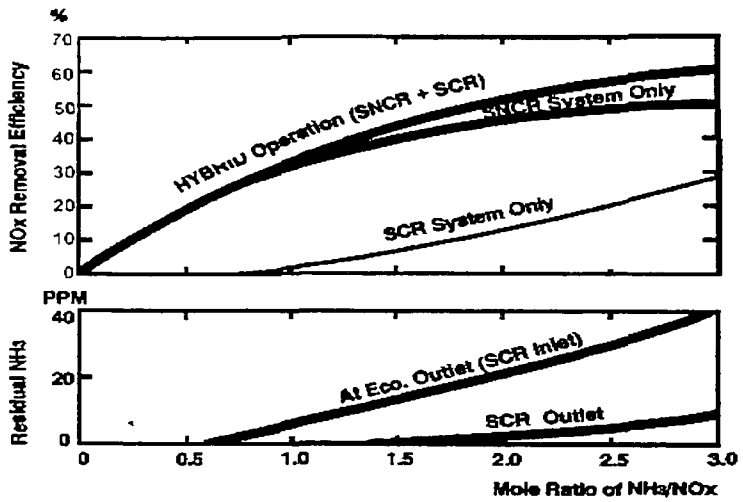

Figure 6
Test Results of Hybrid SNCR-SCR System
at Yokosuka Unit No. 4

Discussion

1. Control of NH_3 Slip: If the fuel contains sulfur, the compounds formed by reaction of the residual NH_3 and SO_3 in the flue gas may plug and damage the air preheater. Therefore, the SCR design must consider reducing residual NH_3 concentration as well as further NOx reduction.

2. Adequate Mixing of SNCR Residual NH_3: In a hybrid system, a larger amount of NH_3 is injected at the SNCR in order to achieve higher SNCR performance. The improved performance associated with higher reagent injection increases the residual NH_3 from the SNCR. If the NH_3 is not well mixed, the higher residual NH_3 may cause plugging of downstream equipment because of higher un-reacted NH_3 through the SCR.

3. System Optimization: The NH_3 consumption rate in an SNCR system is higher than in an SCR system. This is due to the decomposition of some injected NH_3 by its oxidation at high temperature. Therefore, engineers must carefully consider whether the maximum or saturated SNCR performance operation is sufficiently economical and effective.

4. System Characterization: In a hybrid system, a separate NH_3 injection grid can be added at the inlet of the SCR. In this case, the NH_3 injection at the SCR inlet can be equalized to compensate for the amount of residual NH_3 through SNCR. As mentioned in the section discussing SNCR, NOx and NH_3 concentration after the SNCR varies significantly over the load range. During the trial operation, traverse measurement tests must be performed to quantify the operational characteristics of the boiler and SNCR to obtain desirable balance of NOx and NH_3 concentration at the SCR inlet over the entire load range. Figure 7 shows an example of the flue gas and NH_3 injection control curve for SNCR at each boiler load. It is clear that the residual NH_3 is relatively high even with the control of NH_3 injection for SNCR. Therefore, the SCR is required to keep residual NH_3 under control.

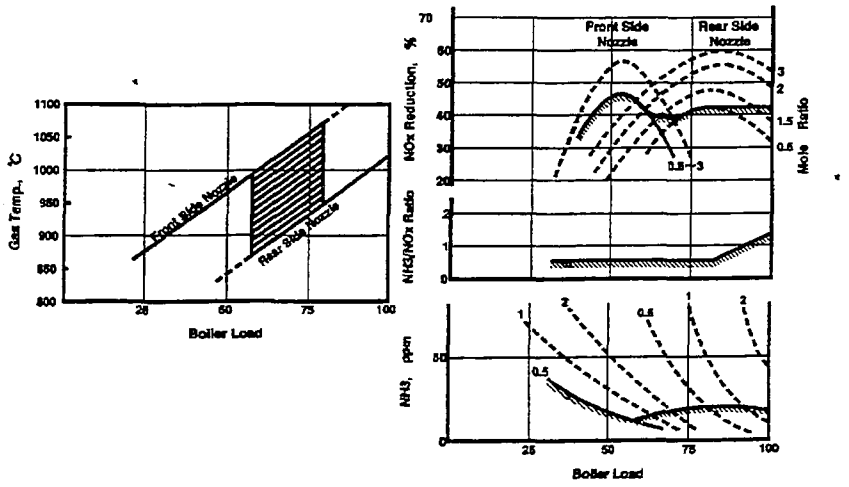

Figure 7
NOx Reduction Performance on 1,225 Ton/Hour Boiler at Various Loads

5. Control of NH_3 Injection: If the system is applied to a coal firing boiler, which produces generally higher NOx concentration than natural gas or oil firing boilers, we may expect higher residual NH_3 from the system. However, due to the presence of sulfur, the residual NH_3 concentration must be kept low to prevent the production of NH_3-sulfur compounds at downstream equipment such as the air preheater. When the imbalance of NOx or NH_3 concentration at SCR inlet is large, there will be concentrated areas of NH_3 at the SCR outlet section, resulting in deposition of NH_3-sulfur compounds as mentioned above. Therefore, operators must inevitably control adjustment of NOx and NH_3 by supplementary injection of NH_3 at the SCR inlet.

Conclusion

For moderate levels of NOx reduction, the SNCR-SCR hybrid system is available and may be feasible under certain conditions. However, as SNCR performance is greatly influenced by boiler characteristics, including boiler load changes, and as SCR performance in hybrid systems is limited by installation space, the performance of hybrid system may carry some elements of uncertainty or complexity.

References

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INDUSTRIAL BOILER RETROFIT FOR NO_X CONTROL: COMBINED SELECTIVE NONCATALYTIC REDUCTION AND SELECTIVE CATALYTIC REDUCTION

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Abstract

A 590 kW (2 MMBtu/hr), oil-fired, three-pass, fire-tube package boiler was retrofit with a combined selective noncatalytic reduction (SNCR) and selective catalytic reduction (SCR) system and demonstrated 85% nitrogen oxide (NO_X) reduction with less than 6 ppm ammonia slip. A urea-based SNCR solution was injected in the first pass, reducing NO_x and providing ammonia reagent for the SCR. A catalyst housing was designed to fit between the second and third passes, where the access doors of the boiler normally attach. The SCR catalyst volume of 0.04 m^3 (1.5 ft^3) provided a space velocity of 10,000 hr^{-1} at a pressure drop of less than 1.5 cm (0.5 in) of water. Numerous runs demonstrated system repeatability and ease of operation.

Introduction

The 1990 Clean Air Act Amendments mandated reduced nitrogen oxide (NO_X) emissions for sources that have the potential to produce more than 22.7 metric tons (25 tons) of NO_X per year. Approximately 54,000 industrial, commercial, and institutional boilers currently in operation in the United States have been identified as sources that produce more than 25 tons of NO_X per year, presenting a need for low cost, high efficiency NO_X removal technologies.[1]

For NO_X control technologies to be applied to existing boilers they must be easy to operate and must not impair the efficiency of the boiler. Some NO_X reduction technologies include selective catalytic reduction (SCR), selective noncatalytic reduction (SNCR), low NO_X burners with advanced over-fired air (OFA), flue gas recirculation, and natural gas reburn which requires OFA in all cases.

This work attempted to develop and demonstrate a combined SNCR/SCR system for retrofit application to an oil-fired package boiler. The objective of the tests was to show that a typical package boiler can be effectively retrofit, optimized, and operated without using excessive manpower and controls. Successful demonstration of this technology will define an option for existing NO_X control sources and will likely provide redesign parameters for new applications. This type of SNCR/SCR system has been demonstrated on gas and coal systems at pilot[2] and full-scale levels. A catalyst housing was designed and attached where the access doors

of the boiler normally attach. A honeycomb-type ceramic catalyst with square cells was installed in the catalyst housing. The SNCR system consisted of a water-cooled, two-fluid reductant injector nozzle that removed a large fraction of the NO_X and supplied reductant to the downstream catalyst.

Experimental

The Boiler and Retrofit

The SNCR/SCR system was retrofitted to a 590 kW (2-MMBtu/hr), three-pass, North American package boiler capable of oil and/or gas firing (see Figure 1). The first pass was the main fire tube, the second pass consisted of 24 6.4 cm (2.5 in) convective tubes, and the third pass consisted of 20 6.4 cm (2.5 in) convective tubes. The burner was a forced air burner located at the front of the boiler that used No. 2 fuel oil at an average rate of 53 L/hr (14 gal/hr) and 173 L/s (367 scfm) of air. Typical baseline operating concentrations were 107 ppm NO_X, 2.8% oxygen (O_2), 12.1 percent carbon dioxide (CO_2), 0 ppm carbon monoxide (CO), 0.6 ppm nitrous oxide (N_2O), and 174 ppm sulfur dioxide (SO_2). Generally, nitric oxide (NO) comprises more than 95% of the total NO_X, and NO was used for the NO_X numbers reported. The average gas flow estimated from the fuel consumption rate and the O_2 and CO_2 concentrations was 177 L/s (377 scfm), but a pitot tube velocity traverse 4 m (13 ft) downstream from the boiler measured 208 L/s (441 scfm); the difference is attributed to measurement error and in-leakage. The back of the boiler has an access plate that has a viewport, a thermocouple port, and a sampling/injection port, the latter centered axially on the 64 cm (25 in) diameter main fire tube. The SNCR reagent was injected countercurrent (toward the burner) through the sampling/injection port. Above and to the either side of the burner are access doors that expose the downstream ends of the second pass (first convective pass) tubes and the upstream ends of the third pass (second convective pass) tubes. The doors were removed, and the SCR catalyst housing was mounted where the doors were. A divider plate was installed to redirect the flow through the catalyst housing (Figure 2).

Sampling System

The sampling system consisted of four sampling lines: a pre-catalyst ammonia (NH_3) (NH_{3RES}) sampling train, a post-catalyst NH_3 (NH_{3SLIP}) sampling train, an SO_2 sample line, and a sample line for the continuous emission monitors (CEMs). The CEM sample system drew a slip-stream from the boiler and then pumped the sample to the CO_2, CO, O_2, and NO on-line analyzers as well as a gas chromatograph (GC) set up for N_2O measurement. N_2O can be a byproduct of the SNCR NO_X reduction system.[3] N_2O is currently not regulated on the Federal level as an air toxic; however, it is a contributor to global warming through the greenhouse effect. The sample gas for the CEMs passed through a Hankison chiller and was pumped through anhydrous calcium sulfate ($CaSO_4$) for removal of water subsequent to the chiller. The NH_3 measurement was aquired by pulling a slip stream of boiler gas through two 1 L impingers in series immersed in an ice bath. The first impinger contained 100 mL of 0.025 N sulfuric acid (H_2SO_4) solution that captured NH_3, and the second impinger was dry and followed by a dry gas meter. The impinger rinse was measured by an ion selective electrode to obtain the NH_3

concentration. The ion selective electrode was calibrated using at least three standards before each set of samples were analyzed, and a spike made from a separate stock was used to check the calibration before and after each sample set was analyzed.

SNCR

The system. The SNCR injector, reagent delivery/dilution system, and the SNCR reagent itself were supplied by Nalco Fuel Tech. The SNCR system consisted of a metering pump and an air atomized reductant injector nozzle. The reagent was similar to NO_XOUT A™, a Nalco Fuel Tech product that consists of 50% urea (NH_2CONH_2), approximately 50% water, and small amounts of anti-scalants and dispersants. The water and urea-based reagent were pumped to a mixing chamber where the total injected liquid amounted to 0.19 L/min (3 gal/hr). The water/SNCR reagent mix was delivered to the injector where it was atomized with air. The normalized stoichiometric ratio (NSR) of reagent nitrogen (N), NH_3, to baseline N, NO_X, was controlled by varying the flowrate of SNCR reagent. SNCR reagent feedrates were measured before and after every data set by pumping SNCR reagent from a burette instead of from the SNCR reagent reservoir barrel. The feedrates before and after the tests were consistent; however, the flow was not monitored during testing.

Optimization. SNCR reagent injection controlled all the post-installation optimization by providing the reductant for the SCR. NH_3, supplied by the breakdown of NH_2CONH_2, reduces NO_X in the presence of the catalyst. The objective when optimizing is to find the injection condition where SNCR removal is most efficient while ensuring that there is sufficient SNCR reagent to supply enough residual NH_3 (NH_{3RES}) for the catalyst to remove the residual, post-SNCR NO_X (NO_{XRES}). The ratio of these two values, NH_{3RES}/NO_{XRES}, defines the pre-catalyst stoichiometric ratio (SR_{RES}). Preliminary tests varied the amount of air, the injector insertion distance into the boiler, the type of nozzle at the tip of the injector, and the total flow of water at a fixed NSR. Although NO_X removal was greater at higher water flowrates, the flow was kept fixed at 0.19 L/min (3 gal/hr) to maintain sufficient boiler efficiency. The temperature profiles acquired before the retrofit indicated that the optimum SNCR injection temperature was close to the end of the main fire tube, but when the SNCR reagent solution and carrier gas were injected they introduced temperature gradients in the main fire tube that resulted in unexpected temperature profile results. The temperature profiles with the SNCR injection indicated that the temperature at the end of the main fire tube was much lower (680 °C) than optimum (about 900°C).[4] This can be attributed to the liquid from the SNCR reagent injection adhering to and then evaporating from the surface of the thermocouple, yielding a gas temperature measurement that is lower than the actual gas temperature. The optimum injector insertion distance for all nozzles, determined by the SNCR NO_X reduction (X_{SNCR}), was between 0 and 20.3 cm (8 in) from the back end of the boiler.

Six different nozzles were tested: three nozzles had round-orifice diameters of 0.1 (0.0388), 0.2 (0.0775), and 0.4 cm (0.155 in) that provided a cone spray pattern; one nozzle had six 0.18 cm (0.07-in) holes evenly distributed 0.28 cm (0.11 in) from the center of the nozzle that provided a cloud spray pattern; one nozzle had an oval orifice that created a fan spray pattern; and one had an oval orifice at a 45° angle to provide an angled fan spray pattern. In the boiler, both fan spray nozzles performed better than the cloud spray, which, in turn, performed better

than the cone spray nozzles. The optimum settings for air pressure and injector distance from the end of the boiler were different for all the nozzles except the two fan spray nozzles. The angled fan spray with the angle pointing up was determined to be the best nozzle to use. The removal was greatest with the injector tip 3.8 cm (1.5 in) from the end of the boiler, and the air at 241 kPa (35 psi). The factors that went into determining the best nozzle were the combined SNCR and SCR total NO_X reduction (X_{TOT}), NH_{3SLIP}, and N_2O emmisions.

SCR

The SCR catalyst was a commercially available titanium and vanadium ceramic catalyst with 7.6 square cells/cm^2 operating at a nominal level of 10,000 h^{-1} space velocity [standard temperature and pressure (STP)]. The catalyst did not appear to degrade over about 240 hours of operation; it also did not collect soot despite overnight shutdown and morning re-start for each test, which could cause temperature changes and increase sooting during startup. Sometimes NH_3 can combine with sulfur trioxide (SO_3) to form ammonium sulfate [$(NH_4)_2SO_4$]. Catalysts can contribute to this formation by converting SO_2 to SO_3. No drop in SO_2 was observed across the catalyst and no NH_3 salts were evident in the boiler. The pressure drop across the catalyst was measured by a manometer and by a Magnahelic pressure gauge to be nominally 1 cm (0.4 in) of water.

Boiler Operation

The boiler was always operated at full load. It was started and run for 3 hours with just water and air injected through the injector to establish temperature, NH_3, and NO_X equilibrium. Then the baseline (no reagent) readings were taken and it was run for 3 more hours with the SNCR reagent flowing. After 3 hours of equilibration time, NH_{3RES}, NH_{3SLIP}, and CEM measurements/samples were recorded. The system was allowed 1 hour to return to equilibrium and post-test baseline readings were taken.

Results and Discussion

The SNCR system typically removed 30 to 40% of the NO_X. Figure 3 shows X_{SNCR} and the resultant SR_{RES} for varying NSR values. The data points are all using the angled fan spray nozzle with the nozzle tip located 1.5 in from the back wall of the boiler. Variation of NSR from 1 to 4 shows little effect on X_{SNCR} and no consistent trend for SR_{RES}.

The SCR NO_X removal performance versus SR_{RES} is shown in Figure 4. The catalyst shows consistent NO_X reduction (X_{SCR}) trends with increasing SR_{RES} until SR_{RES} reaches values above approximately 0.7, where X_{SCR} levels off at about 80%. This is a fairly typical trend for a catalyst performance The points on Figure 4 are a compilation of both fan spray nozzles.

The ratio of the decrease in NO_X to the decrease in NH_3 averaged 1.33/1 across the catalyst. This is usually a 1/1 ratio. NH_3 measurements that are too low (due to NH_3 breaking through the impingers, or due to dry gas meter calibrations returning sample volumes that are higher than actual) could cause the ratio to seem higher than 1/1.

While NSR has little effect on X_{SNCR}, and due to the wide range of testing conditions reported, no obvious relationship with SR_{RES} (Figure 3), the appropriate SR_{RES} (> 0.7, from

Figure 4) is probably obtained at NSR values around 2 or higher. This is more clear from Figure 5, which shows the system's total NO_X removal, X_{TOT}, versus NSR. This shows that under optimal conditions this retrofit established NO_X reduction of 93%. All ammonia slips were less than 6 ppm, so there is clearly enough catalyst to remove the residual NH_3 from the SNCR process.

N_2O formation increases with increasing NSR values (Figure 6). Under fairly typical operating conditions of NSR = 2, measured N_2O emissions are about 10 ppm. These values have not been accounted for in prior NO_X reduction percentages.

The sometimes large degree of scatter in the data are indicative of the varied SNCR chemical injection rates. Practical operation of an SNCR/SCR system would involve finding the optimum conditions for each specific boiler and operating within those conditions. Shakedown, optimization, and long-term operation of this system are ideally suited for a neural network feedback/control or fuzzy logic control system.

Conclusion

An SNCR/SCR hybrid system retrofit to an oil-fired package boiler had an apparent optimum NSR around 2 where 85% NO_X reduction could be achieved without exceeding 6 ppm NH_{3SLIP}. N_2O formation was generally below 15 ppm. At the optimum NSR, it was nominally 10 ppm.

The SNCR/SCR hybrid system was easily retrofit to our existing oil-fired package boiler. Although the design of the SCR catalyst housing was specific to this boiler, similar designs could be developed for other units. Equipping new boilers via minor design changes would probably be even easier and more effective than retrofitting old ones. Once the system is installed, optimization time may be about 1 or 2 days. Changes in temperature profiles may make optimization difficult. Current work is developing a fuzzy logic control system to counter this.

Further testing may be needed to assess long term boiler effects and long term catalyst durability to determine added costs from reagent use and catalyst replacement.

Acknowledgements

The authors acknowledge Nalco Fuel Tech's contribution of the SNCR injector, reagent, and reagent delivery system. Special thanks to John E. Hofmann and M. Linda Lin (Nalco Fuel Tech) for technical advice.

Nomenclature

NH_{3RES}	The residual ammonia in the boiler after the SNCR reaction measured before the catalyst.
NH_{3SLIP}	The residual ammonia in the boiler measured after the catalyst.
NSR	The normalized stoichiometric ratio of input nitrogen (from the SNCR reagent) to the nitrogen in the boiler in the form of NO_X.
NO_{XRES}	The residual NO_X in the boiler after the SNCR reaction measured before the catalyst.
SCR	Selective catalytic reduction.

SNCR	Selective noncatalytic reduction.
SR_{RES}	The stoichiometric ratio of the post-SNCR residual reagent nitrogen, NH_{3RES}, to post-SNCR residual nitrogen measured before the catalyst, NO_{XRES} .
X_{SCR}	The percent reduction of NO_X across the catalyst.
X_{SNCR}	The percent reduction of NO_X from baseline level to the NO_X level measured after the SNCR reaction before the catalyst.
X_{TOT}	The percent reduction of NO_X from baseline to the NO_X level measured after the catalyst.

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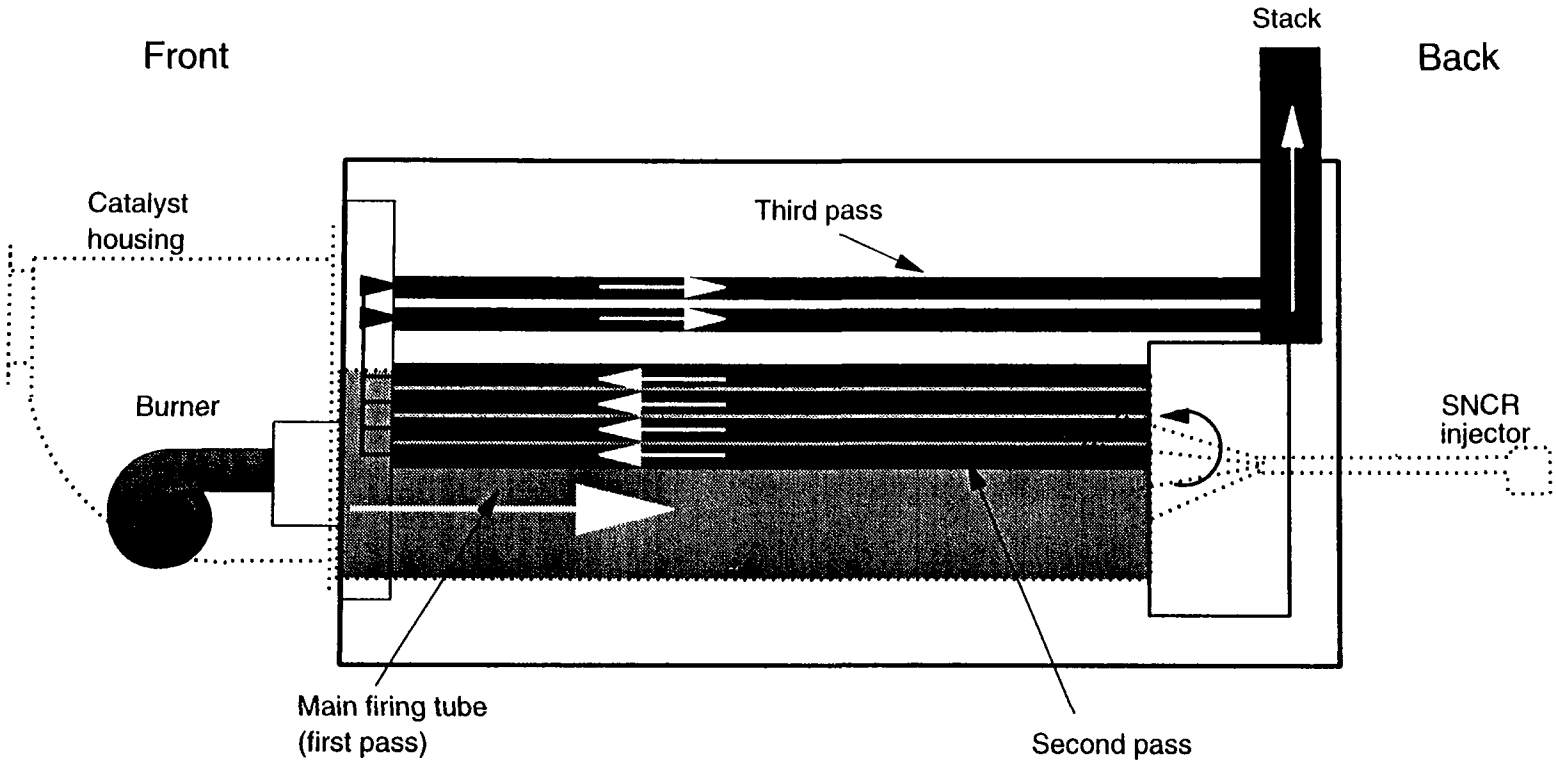

Figure 1. The North American package boiler side view with SNCR/SCR retrofit outlined by dotted lines.

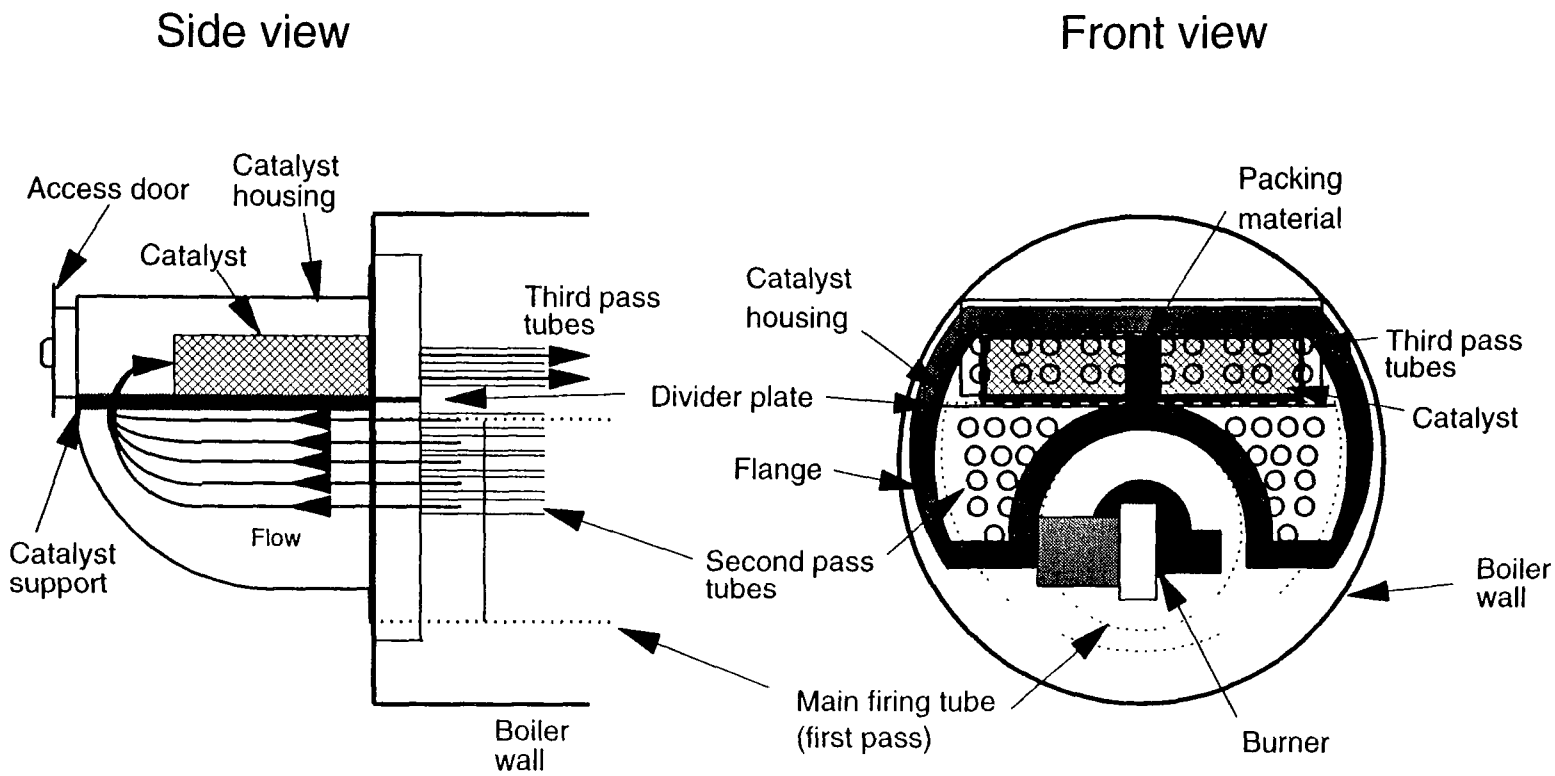

Figure 2. SCR retrofit diagram.

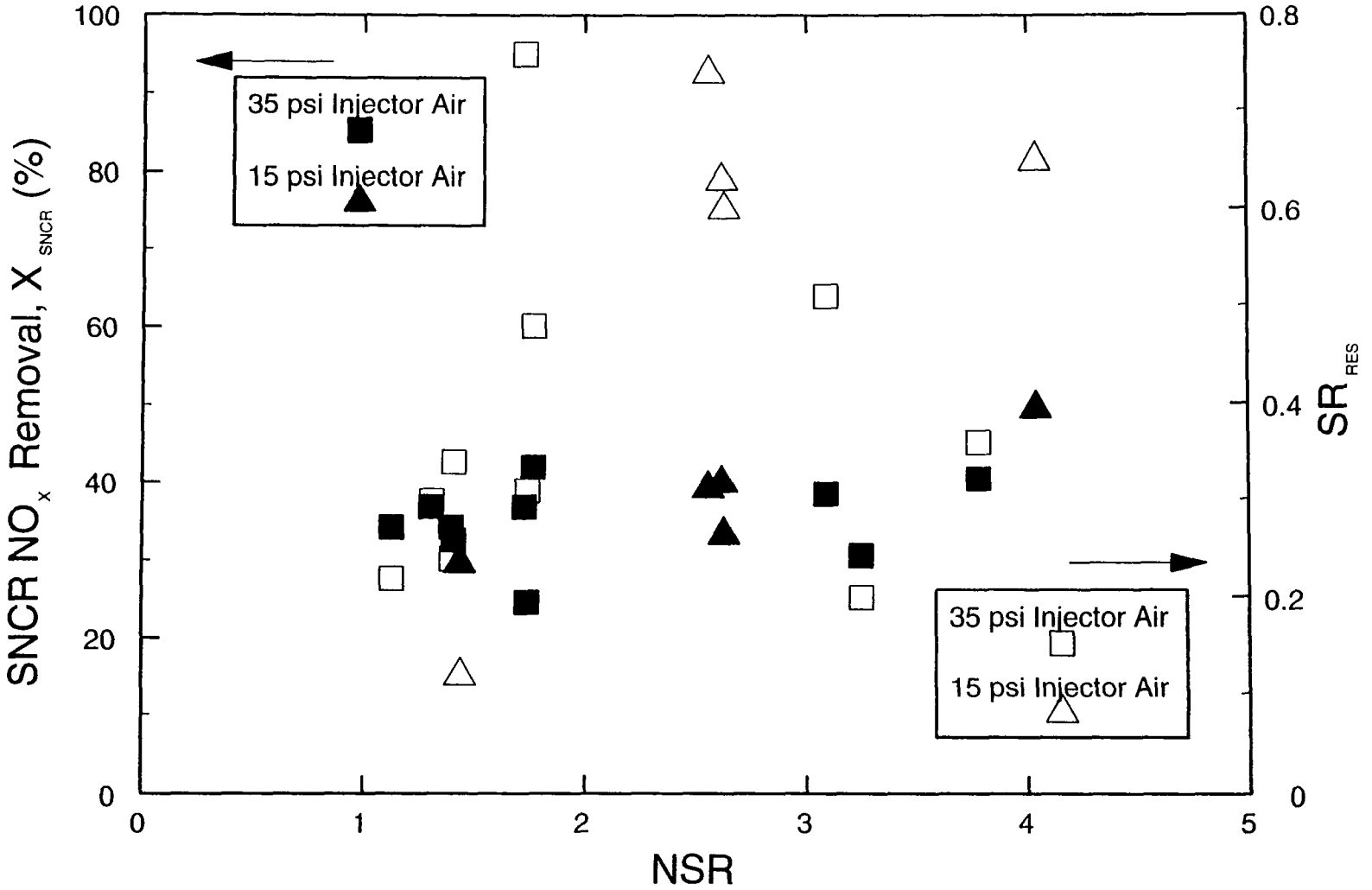

Figure 3. SNCR system performance. Angled fan spray nozzle with injector tip 1.5 in from the back of the boiler.

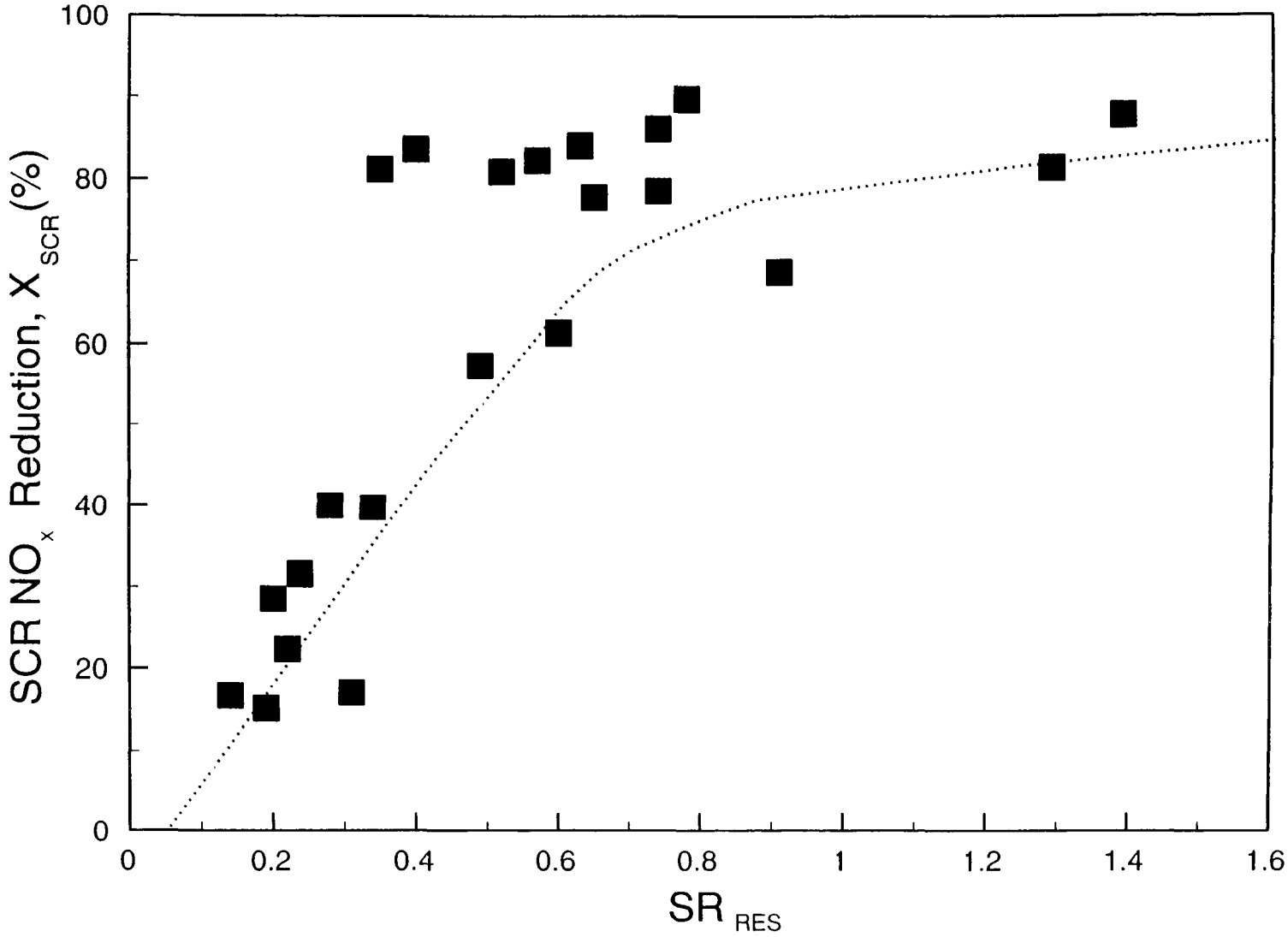

Figure 4. NO_x reduction across catalyst for varying SR_{RES} NH_{3RES} is between 10 and 55 ppm.

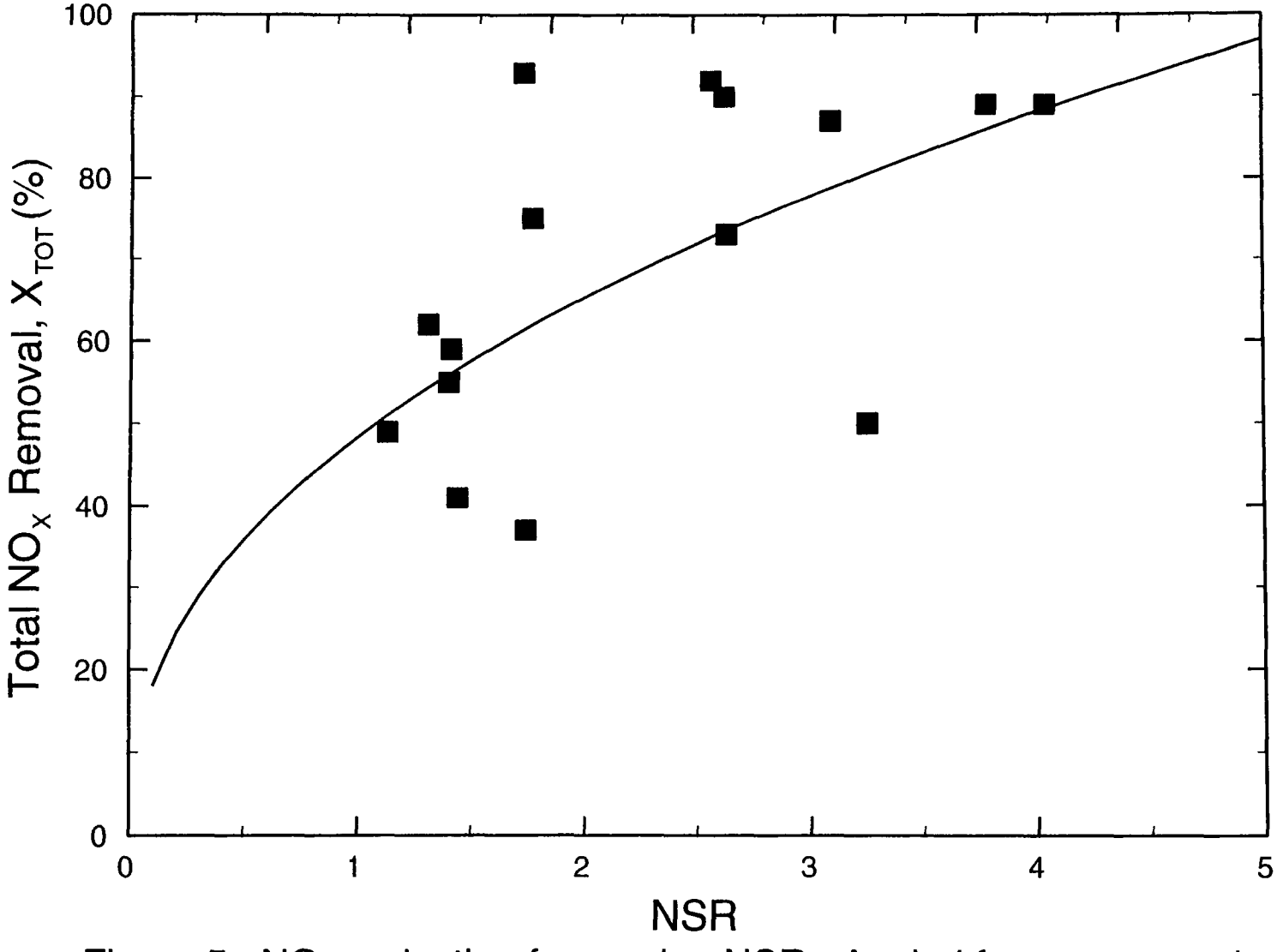

Figure 5. NO_x reduction for varying NSR. Angled fan spray nozzle data only.

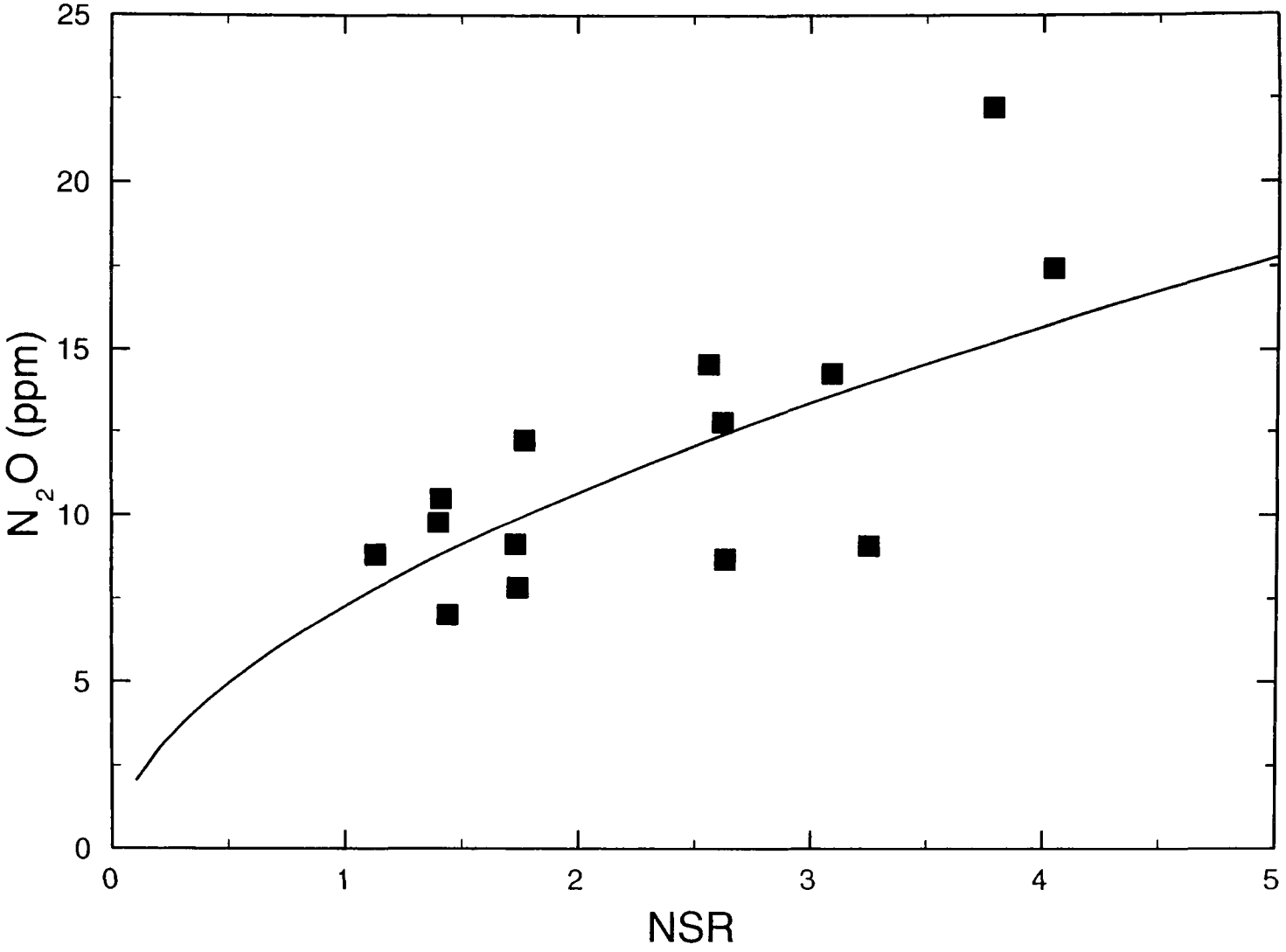

Figure 6. N_2O reduction across catalyst for varying NSR. Angled fan spray nozzle data only.

HYBRID SCR

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Abstract

Hybrid selective catalytic reduction (SCR) systems consist of either a combination of SCR techniques (i.e. in-duct SCR combined with air heater SCR) or selective noncatalytic reduction (SNCR) in combination with SCR. These Hybrid SCR systems can offer substantial benefits in reduced cost and enhanced performance; however, their applicability is very unit specific. This paper presents the results of a study to document the current experience and develop a tool by which utilities can determine the applicability of Hybrid SCR to meet their NO_x reduction goals, a guideline for selecting the best configuration, and a reference for developing the design parameters necessary to implement the technology.

Hybrid SCR systems have been installed and demonstrated on utility boilers. The systems have included in-duct SCR combined with air heater SCR and SNCR combined with SCR as well as systems where SNCR was combined with in-duct and air heater SCR. This document includes a review of the results of these demonstrations as well as comments on the applicability of those results for other utility systems.

Finally this document provides a reference for the development of design parameters for the implementation of Hybrid SCR. There are a number of technical and commercial considerations which must be resolved prior to designing or procuring a Hybrid SCR system. The boiler operating, temperature and emissions data necessary for the final design are presented along with the process design variables which must be specified. Procurement suggestions are included to assist the user in addressing some of the more pertinent commercial issues.

Introduction and Background

Promulgation of increasingly stringent regulations limiting NO_x emissions from utility boilers has created a need for technologies capable of achieving high levels of NO_x reduction. Selective catalatic reduction (SCR) is capable of achieving the desired reductions; however, high costs and/or technical limitations caused by unique boiler configurations often make stand-alone SCR a less than optimum solution. Hybrid SCR systems offer the high levels of NO_x reduction necessary for compliance and can often overcome the limitations caused by unique boiler configurations.

Description of Hybrid SCR Options

There are four different configurations which fall under the general heading "Hybrid SCR." Each configuration has its own individual benefits and applicability. The first application would be the application of selective noncatalytic reduction (SNCR) with in-duct SCR. This application utilizes the NH_3 slip from the SNCR process as part or all of the NH_3 for the SCR reaction. This Hybrid SCR configuration can be further categorized according to the type of reagent used, urea or NH_3, and by whether or not supplemental NH_3 is injected upstream of the SCR. The benefits of each of these options will be discussed below.

The second option is to combine an in-duct SCR with an air heater SCR. Typically a single NH_3 injection grid is installed upstream of the in-duct SCR. NH_3 slip from the in-duct SCR provides the inlet NH_3 for the air heater SCR.

The third option is the combination of SNCR with an air heater SCR. As with the first option this Hybrid SCR configuration can utilize either an NH_3 or urea based reagent for the SNCR and supplemental NH_3 injection at the SCR inlet. The utilization of an air heater SCR differentiates this option from option 1 due to the unique characteristics of the air heater SCR. The fourth option is the combination of SNCR with both in-duct SCR and air heater SCR.

Benefits of Hybrid SCR

The application of Hybrid SCR must offer direct benefit over alternative technologies (typically stand-alone SCR) to justify consideration by the end user. The benefits offered by Hybrid SCR vary according to the configuration selected as well as many boiler specific parameters. Below we present a qualitative assessment of several of the benefits. A quantitative assessment would require a boiler specific evaluation including capital and operating cost estimates, which are addressed in detail in the EPRI technical report.

Higher Overall NO_x Reduction Possible without Extensive Unit Modifications. Any of the four Hybrid SCR configurations discussed above will offer a higher level of NO_x reduction compared to a stand-alone SCR, if the quantity of catalyst is limited by space available or by allowable pressure drop. This limitation applies to some degree for most retrofit applications. If the regulatory limits can be achieved within the limitations of catalyst volume, there may not be a direct benefit to Hybrid SCR. However, in many cases the regulatory limits will not be achievable without extensive modifications to the unit. In these

cases, the benefit of Hybrid SCR will be avoiding the cost of unit modifications such as air heater relocation, duct work, and structural modifications necessary for installing an external reactor or fan modifications to overcome pressure drop.

Lower System Pressure Drop. For a given quantity of NO_x reduction, a Hybrid SCR system will typically have a lower pressure drop than a stand-alone SCR. If the configuration utilizes SNCR the lower NO_x concentration at the SCR inlet will decrease the volume of catalyst necessary and therefore the pressure drop. Air heater SCR achieves NO_x reduction with little or no increase in pressure drop compared to the original air heater. Therefore, if the Hybrid SCR utilizes an air heater SCR to reduce the volume of in-duct catalyst required, the total system pressure drop will be less than if the entire reduction was achieved with in-duct SCR. Lower pressure drop will reduce operating costs and may avoid the need for fan upgrades or replacements on fan limited units. In an extreme case on a forced draft unit which is pressure limited the pressure drop of an SCR could require the installation of induced draft (ID) fans. Avoiding the cost associated with this extensive modification would be a substantial benefit if Hybrid SCR could achieve the same reductions within the pressure limitations of the boiler.

Safer and Less Costly Chemical Storage. This benefit applies only to Hybrid SCR systems which utilize urea based SNCR systems and do not use supplemental NH_3 injection at the SCR inlet. NH_3 is a hazardous chemical whether stored in the anhydrous or aqueous form. SNCR systems which utilize a urea based chemical as their reagent would avoid the need for storage of NH_3 on-site. While providing a safer work environment, this would also save the cost of preparing and implementing risk management prevention plans.

Lower NH_3 Slip. With a Hybrid SCR incorporating air heater SCR, it is possible to achieve the same emissions reductions as an equivalent stand-alone SCR at lower NH_3 slip. This may be critical on systems where the slip must be limited to ultra low values to avoid air heater fouling and or fly ash contamination. It may be possible to achieve the same reduction in NH_3 slip by installing additional catalyst in the duct; however, the amount necessary may exceed the volume limitations and the catalyst may not be as effective in the duct as on the air heater baskets.

The rotation of the air heater may offer two benefits, particularly when ultra low NH_3 slip is desired. The first is that the rotation will reduce the impact of stratification of NH_3 and NO_x in the flue gas. Frequently when trying to obtain ultra low NH_3 slip, the limiting factor is the ability to adequately mix the NH_3 and NO_x such that a consistent NH_3 to NO_x mole ratio is achieved at the SCR inlet. Even a slight variation will make it difficult to achieve high or ultra high NO_x reduction at an ultra low NH_3 slip. As the air heater rotates, unreacted NH_3 can be adsorbed within the catalyst in one location and react with NO_x in another location. This benefit is limited as the rotation will only impact stratification in one of the two axis within the duct.

The second benefit of the air heater rotation is the desorption of the NH_3 into the combustion air. As the air heater rotates from the flue gas into the combustion air stream, a portion of the unreacted NH_3 will desorb into the air stream. The NH_3 in the combustion air stream

will enter the boiler through the burner and the vast majority of it will be oxidized to NO_x. On the surface this may seem like a disadvantage as it will increase the baseline NO_x emissions and require greater reductions to achieve compliance with a given NO_x emission limit. However, the increase is small (typically 5 to 10 ppm) compared to the overall baseline and the benefit of being able to operate the in-duct SCR at a higher slip will usually outweigh the increase in baseline NO_x.

Operational Flexibility. The Hybrid SCR offers the benefit of operational flexibility allowing the utility to optimize operation for specific, and frequently changing, operating conditions. Hybrid SCR which includes SNCR can also be operated as a stand-alone SCR by shutting down the SNCR injection and injecting NH_3 only through the grid upstream of the catalyst. This will reduce overall system performance; however, under certain operating conditions this will be acceptable and the operating cost savings can be substantial.

The first scenario where this approach would be applicable is to comply with seasonal NO_x limitations. Future regulations are being considered for Phase II NO_x emission reductions in the ozone transport region which include lower emission limits during the "ozone season" (May 1 through September 30) than during the remainder of the year. This regulatory approach creates an opportunity for Hybrid SCR. If a Hybrid system consisting of SNCR and SCR is designed to meet the lower NO_x emission limit, it is likely that the SCR alone would meet the higher limit. Compared to a stand-alone SCR, which would have to be designed to meet the lower limits, the Hybrid SCR would have substantially less catalyst, reducing the initial capital cost as well as catalyst replacement costs. The fact that the SNCR system would be used only during the five-month ozone season will minimize the drawback of the higher reagent consumption.

The second scenario where the Hybrid SCR would add operational flexibility is for load following units with moderate or low capacity factors. Unless the emissions averaging period is long (i.e., 30 days), most utilities will design the NO_x reduction system to achieve the regulatory limit at the most stringent design point, typically full load. This is to assure compliance even if the unit must run at full load for an extended period of time. The majority of the time, however, the unit will be operating at moderate or low loads, which typically have less stringent NO_x reduction requirements. In this scenario the Hybrid SCR consisting of SNCR and in-duct SCR would be designed to provide the NO_x reduction requirements at full load. At lower loads the SNCR injection rate would be decreased and below a specified set point would be shut down completely. The load at which the SNCR system could be shut down would be dependent on the NO_x reduction capability of the SCR which would improve at lower loads due to the lower inlet NO_x and lower space velocity. As in the first scenario the Hybrid SCR approach would reduce the catalyst volume required, lowering the initial capital costs as well as the catalyst replacement costs. The high operating cost associated with higher reagent consumption would only be applicable at higher unit loads.

Drawbacks to Hybrid SCR

Although the above benefits may appear to make Hybrid SCR a very attractive technology, there are drawbacks which at times will outweigh the benefits. As with the benefits, the

drawbacks vary according to the selected Hybrid SCR configuration as well as unit specific parameters. Below we present a qualitative assessment of the potential drawbacks.

System Complexity. Hybrid SCR systems are inherently more complex than a stand-alone SCR system. This increase in complexity will increase the design costs as well as the operating and maintenance costs and may make the system more difficult to trouble shoot.

Higher Chemical Costs. The costs of chemicals for a Hybrid system utilizing SNCR will be greater than the chemical costs for a stand-alone SCR achieving the same emission reductions. There are two factors that affect the cost of chemicals. The first factor is the higher chemical consumption required for SNCR systems and applies whether the reagent is NH_3 or urea. A Hybrid system utilizing SNCR may require a normalized mole ratio of 1.2 to 1.5, whereas a stand-alone SCR would only require a 0.9 to 1.0 mole ratio. The second cost factor applies only to urea based SNCR systems and reflects the differential cost for urea over NH_3. The cost, normalized to moles of nitrogen in the reagent is typically 90 percent higher for urea vs. aqueous NH_3 and 150 percent vs. anhydrous NH_3 (based on average reagent costs from vendor survey conducted as part of EPRI Report TR-103885, "SNCR Feasibility and Economic Evaluation Guidelines for Fossil-Fired Utility Boilers").

Potentially Higher Capital Costs Associated with Air Heater SCR. The capital cost for an air heater SCR to achieve a given amount of NO_x reduction will typically be higher than a stand-alone SCR. One reason for this higher cost is that to obtain the same space velocity the air heater will require more than twice the volume of catalyst. This is due to the fact that slightly more than 50 percent of the air heater wheel is out of the flue gas path at any given time. Another factor in the capital cost is the higher cost to manufacturer the catalytic air heater baskets compared to typically stationary SCR catalyst. This cost comparison is based on the cost of installing catalyst in an existing duct. In addition to the higher catalyst costs, the air heater SCR may require modifications to the air heater wheel or the intermediate and cold end baskets to accommodate catalytic baskets with a greater depth than the original hot end baskets. If it is necessary to install a external reactor or to make significant duct work or support modifications, these cost may outweigh the higher costs for the air heater SCR.

Utility Experience with Hybrid SCR

Three utility Hybrid SCR systems have been installed and tested in the United States as demonstration projects: San Diego Gas and Electric's Encina Power Plant Unit 2, Southern California Edison Company's Mandalay Generating Station Unit 2, and Public Service Electric and Gas's Mercer Generating Station Unit 2. Additional testing has been conducted on pilot scale reactors and preliminary designs were completed for Houston Lighting and Power's Webster Station when that utility solicited turnkey proposals for a Hybrid SCR system. Presented below is a brief system description for the three utility demonstration projects, as well as performance testing results. A summary of the Webster solicitation is included presenting average values for predicted performance and for costs. Finally an overview of some of the pilot scale testing is presented.

San Diego Gas and Electric's Encina Power Plant Unit 2

Encina Unit 2 is a 110 MW Babcock and Wilcox front wall-fired balanced draft boiler designed to burn natural gas and oil. Prior to the installation of post-combustion NO_x controls the unit maintained compliance with the San Diego Air Pollution District's (SDAPCD) NO_x emission limits of 125 ppm on natural gas and 225 ppm on oil utilizing burners out of service (BOOS) and hopper flue gas recirculation.

SNCR System. The SNCR system is a urea based system installed and tested extensively prior to the Hybrid SCR demonstration. The system injects urea solution through one of two elevations. The first elevation has four injectors located on the front wall. The second elevation has six injectors, four of which are located on the front wall and two of which are located on the side walls. Urea is received and stored as a 48 percent aqueous solution and is diluted as necessary to maintain a constant injection flow of 6.7 gallons per minute regardless of mole ratio or unit load.

In-duct SCR System. At the economizer exit flue gas is divided into two ducts (designated north and south) which route the flue gas to two L'jungstrom regenerative air heaters. Approximately 186 ft^3 of Catalyst was installed in each of the two ducts. The catalyst, which was manufactured by Grace, is a vanadia/titania formulation in a conventional honeycomb configuration with a 7.4 mm pitch. The in-duct SCR is designed to operate at a space velocity of approximately 35,800 hr^{-1} and a pressure drop of 3 in. wc.

An NH_3 injection grid is installed upstream of the in-duct SCR to supplement the NH_3 slip from the SNCR process. The grid initially consisted of eight pairs of vertically stacked pipes with orifices drilled on both sides to inject NH_3 perpendicular to the flue gas flow. The injection grid was subsequently modified by increasing the number of orifices and was relocated further upstream to provide greater residence time for improved mixing.

Air Heater SCR. The existing air heater is a vertical shaft L'jungstrom regenerative air heater manufactured by Air Preheater Company (APCO). The hot end baskets were replaced with catalytic baskets. No additional modifications were made to the air heater intermediate or cold end baskets. The installation of catalytic baskets did not cause any loss in thermal efficiency or any increase in pressure drop when compared to the original air heater. The air heater SCR has a space velocity of approximately 22,800 hr^{-1}.

Performance Results. The demonstration project was separated into three different phases. The first phase of testing was to develop an understanding of the overall design of the Hybrid system, its compatibility with boiler operations, and the impact of key design parameters on performance. The second phase was designed to apply the knowledge gained in phase one to maximize the NO_x reduction with only minor hardware changes. The objectives of the third phase were to maximize performance after significant hardware changes were made (redesign of the NH_3 injection grid and installation of an improved air heater SCR). Table 1 presents the results of the phase three testing.

Table 1
SDG&E Encina Unit 2
Hybrid SCR - Individual and Combined Performance
Phase III Results

Unit Load (MW)	SNCR Performance (NSR=3.0)		In-duct SCR Performance (NH_3/NO_x=1.0)		Air Heater SCR Performance (NH_3/NO_x=1.0)		Hybrid SCR Performance	
	% NO_x Red	Outlet NO_x (ppm)	% NO_x Red	Outlet NO_x (ppm)	% NO_x Red	Outlet NO_x (ppm)	% NO_x Red	Outlet NO_x (ppm)
22	38	56	66	19	57	8	91	8
70	50	45	48	23	26	17	81	17
108	41	53	36	34	26	25	72	25

Note: Baseline NO_x emissions are approximately 90 ppm for all loads.

Southern California Edison's Mandalay Generating Station Unit 2

Mandalay Generating Station Unit 2 is a 215 MW front wall-fired, forced draft Babcock and Wilcox boiler capable of burning natural gas or fuel oil. The normal NO_x emissions on natural gas at full load are 160 ppm without any post combustion NO_x control; lower emissions (approximately 115 ppm) can be achieved by operating the unit with additional burners out of service. The unit was retrofitted with SNCR and air heater SCR under two separate programs during 1991. Both systems were tested independently and in combination (Hybrid).

Air Heater SCR. Mandalay Unit 2 is equipped with two vertical shaft L'jungstrom regenerative air heaters originally supplied by APCO. The air heater SCR was designed to provide the same thermal efficiency as the original air heater and to increase pressure drop by no more than 20 percent. The installation of air heater SCR on one wheel only caused a flue gas flow imbalance which affected the assessment of system performance. This makes an extrapolation to a complete balanced flow system difficult.

An anhydrous NH_3 injection system was installed with a grid upstream of the air heater SCR inlet. The grid consists of seven horizontal pipes with a total of 165 injection nozzles. The injection grid was designed and used for testing of the air heater SCR as a stand-alone technology. The injection grid was not used during the Hybrid SCR testing.

SNCR System. A low-energy, urea-based SNCR system was installed on Mandalay Unit 2. The system consists of three levels of injectors located on the front wall of the furnace and one level on the rear wall with four injectors on each level. Urea solution was injected through only the upper elevation two injectors in the north half of the furnace during demonstration of the hybrid system because the air heater SCR was on the north duct. The injection rate was controlled by an automatic control system based on the unit load and set to ensure an ammonia slip of less than 10 ppm at the outlet of the air heater.

Performance Results. The component systems, urea based SNCR and air heater SCR, which comprise the hybrid system at SCE Mandalay Station Unit 2 were each designed as stand alone NO_x reduction technologies. The air heater SCR system optimized and tested extensively as an independent system prior to the testing as a hybrid system. Table 2 presents the results of full load performance tests for the component systems both as stand alone technologies and as part of the hybrid system. This table indicates the synergy achieved with hybrid systems as the performance of the individual components increases when operated in a hybrid configuration.

Figure 1 presents the Hybrid SCR system performance, as well as the performance of the individual component systems, as a function of load. As can be seen from the figure, NO_x emission reductions of 82 percent were achieved at full load and exceeded 90 percent for loads of 50 percent or less.

The test results presented in Table 2 are based on 8 BOOS; however, when the boiler was operated with 6 BOOS (normal operation) the baseline NO_x emissions are higher and the

Table 2
SCE Mandalay Station Unit 2
Independent SNCR/SCR versus Hybrid Performance
Full Load Natural Gas Firing

	NO_x Reduction (%)[Note 1]	
Technology	Independent Performance	Hybrid Performance
Urea Based SNCR	16	30
Air Heater	70	74
Total System	75[Note 2]	82

Note 1: NO_x reductions are from a baseline of approximately 115 ppm (8 BOOS) and in all cases NH_3 emissions were less than 10 ppm.
Note 2: Theoretical system performance based on combined independent performance.

benefit of the Hybrid system is greater. With 6 BOOS at 200 MW, Hybrid SCR with urea injection only increases performance from 65 percent to 81 percent when compared to air heater SCR only performance.

Public Service Electric and Gas's Mercer Station Unit 2

Mercer Generating Station Unit 2 is a 321 MW Foster Wheeler continuous slagging twin furnace boiler capable of firing coal and natural gas. Originally designed as a pressured furnace Unit 2 has been modified to balanced draft. In the twin furnace design one furnace generates superheated steam and the other generates reheated steam. SNCR was installed and tested on Unit 2 in 1993. In June through October 1994 an in-duct SCR and an air heater SCR were installed on one of four ducts for unit 2 and were tested without SNCR. The final step in the series of demonstrations was to test the combined performance of SNCR with the SCR. Two separate test programs were conducted in late 1994 and early 1995, one with a urea based reagent and one with aqueous NH_3.

SNCR System. The system was initially designed with three levels of injection. Each level has four lances located on the rear wall for the lowest of the three levels and on the front wall for the upper two levels. During the SNCR only testing a fourth level of injection was added below the original three with two lances on the front wall and one on each of the side walls. During the Hybrid SCR testing, additional ports were added to inject reagent at cooler temperatures.

The reagent used during the SNCR only testing was NO_xOUT, a proprietary urea based reagent licensed by Nalco Fuel Tech. Two independent tests were performed during the Hybrid SCR testing; the first was with NO_xOUT and the second was with aqueous NH_3. During the Hybrid testing with NO_xOUT no supplemental NH_3 was injected at the injection grid. During the Hybrid SCR testing with aqueous NH_3 as the SNCR reagent, supplemental NH_3 was injected at the NH_3 injection grid.

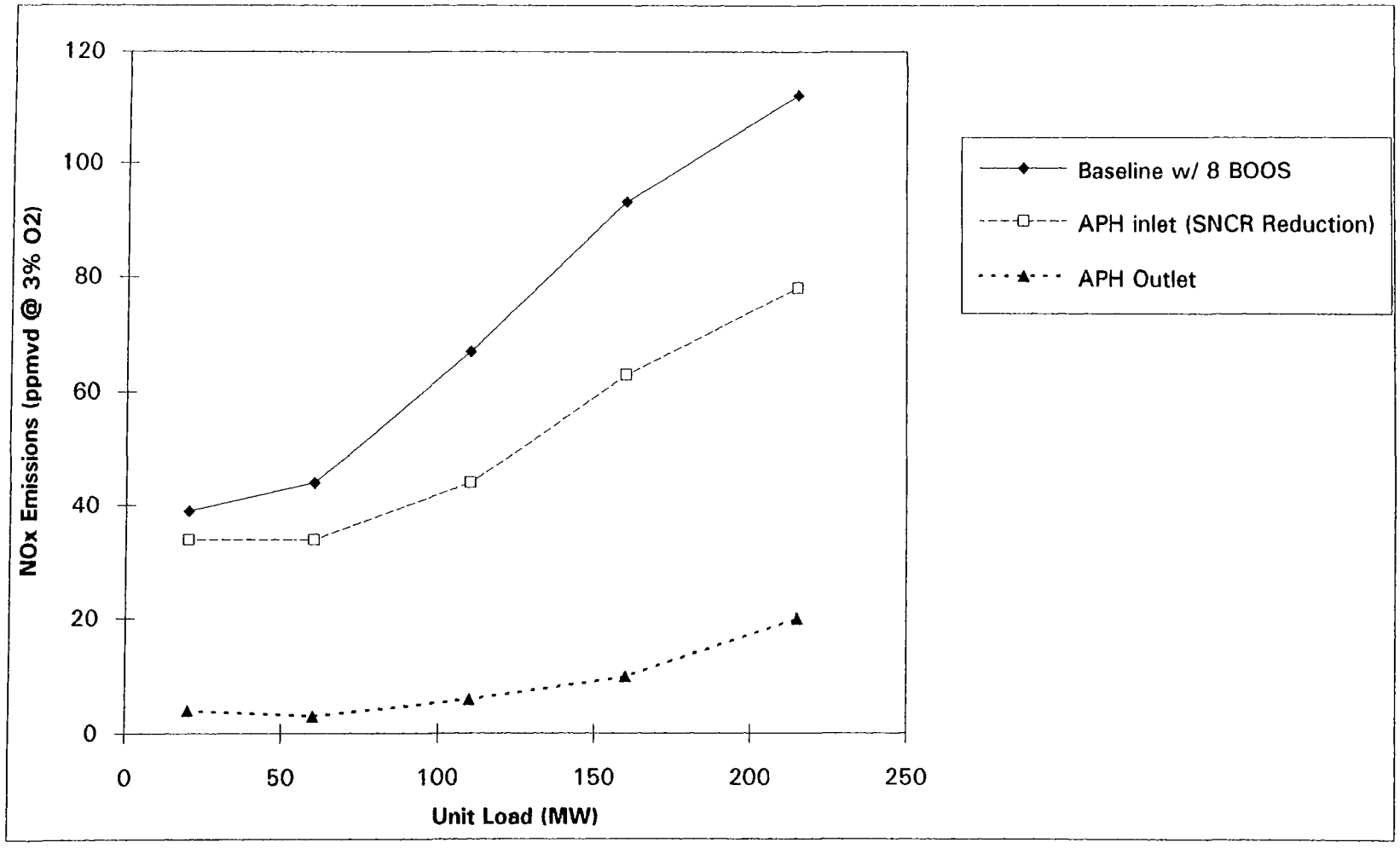

Figure 1
SCE Mandalay Unit 2
Hybrid SCR System Performance as a Function of Load

In-duct SCR System. As stated previously the unit consists of two furnaces each with a separate flue gas path. At the exit of the economizer of each furnace the flue gas is divided into two separate ducts. The in-duct SCR is installed on one of these ducts and therefore treats approximately 25 percent of the flue gas from Unit 2. The in-duct SCR consists of two banks of plate catalyst manufactured by Siemens. The catalyst is a titanium dioxide and molybdenum/vanadium oxide formulation designed for low SO_2 to SO_3 conversion. The flue gas flow through the SCR is horizontal, the first time this had been done on a coal-fired boiler. The catalyst plates are oriented vertically to minimize the build-up of fly ash.

Aqueous NH_3 at a concentration of approximately 27 percent by weight is vaporized in a hot air stream provided by dilution air fans and electric heaters. The air/NH_3 mixture is routed to the injection grid which consisted of 15 horizontal pipes. During the optimization phase of the SCR testing it was determined that the flue gas flow stratification was greater than could be accommodated by the NH_3 flow biasing valves. Therefore, a flue gas static mixer was installed upstream of the NH_3 injection grid.

Air Heater SCR. One of the four existing L'jungstrom air heaters was retrofit by replacing the hot end baskets with catalyst coated baskets. The catalyst elements were manufactured by Siemens using the same catalyst formulation as the in-duct catalyst. The rotation of the horizontal shaft air heater creates cyclic stress on the catalytic baskets and increases the concern for loss of catalyst material. Air heater baskets must be designed to avoid differential expansion between the catalyst material and the metal substrate. However, tests baskets installed prior to the testing program did not show adverse impact.

Performance Results. As stated previously several testing programs have been conducted at Mercer Generating Station Unit 2. The first test program was SNCR only and therefore does not apply to this Hybrid SCR paper. Three additional programs were conducted which qualify as Hybrid SCR and are therefore discussed below.

The second program was to test the in-duct SCR and the air heater SCR without SNCR. This six month program optimized the performance and then tested the two stage configuration of the Hybrid SCR system under a variety of conditions. It is not practical to present all of the results in this report; however, we have included two figures which summarize the results. Figure 2 presents the full load NO_x reduction as a function of inlet NH_3 to NO_x mole ratio for coal firing. Figure 3 presents full load NO_x reduction as a function of NH_3 slip for coal firing. The later figure graphically depicts the benefits of the air heater SCR when installed in conjunction with an in-duct SCR. A typical SCR will be designed and operated to maintain a specified NH_3 slip limit (either a regulatory limit or a process limitation). As can be seen from these two figures, the performance for the in-duct and air heater Hybrid SCR configuration is much greater than for the in-duct alone for a given slip limit.

The third and fourth test programs both tested SNCR in combination with the installed SCR system. The third program utilized NO_xOUT as a reagent and the majority of the testing was conducted without supplemental NH_3 injection through the injection grid. The results of this testing are summarized in Table 3. The fourth program utilized aqueous NH_3 as the SNCR

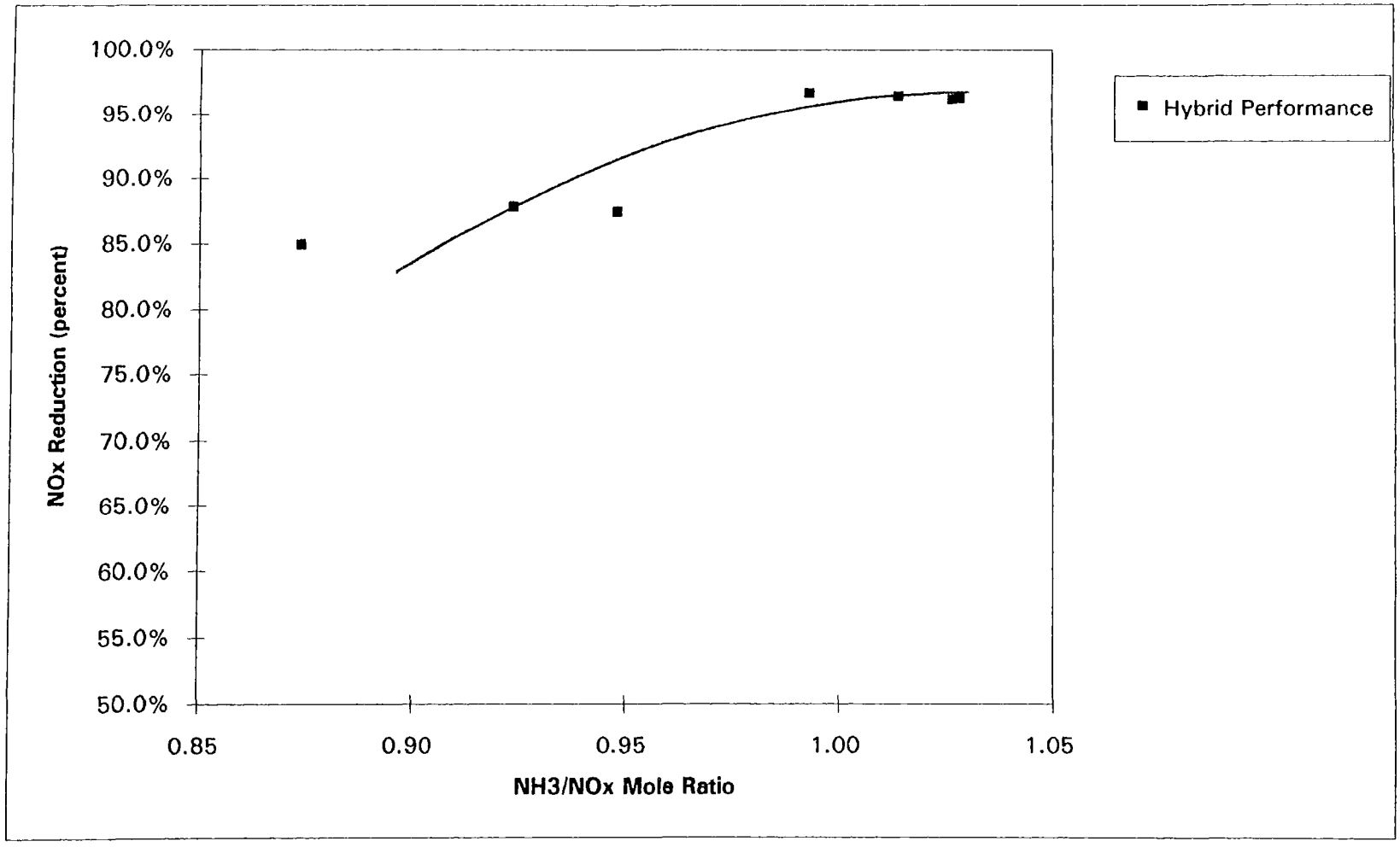

Figure 2
PSE&G Mercer Station Unit 2
Hybrid System Performance Coal Firing 320 MW

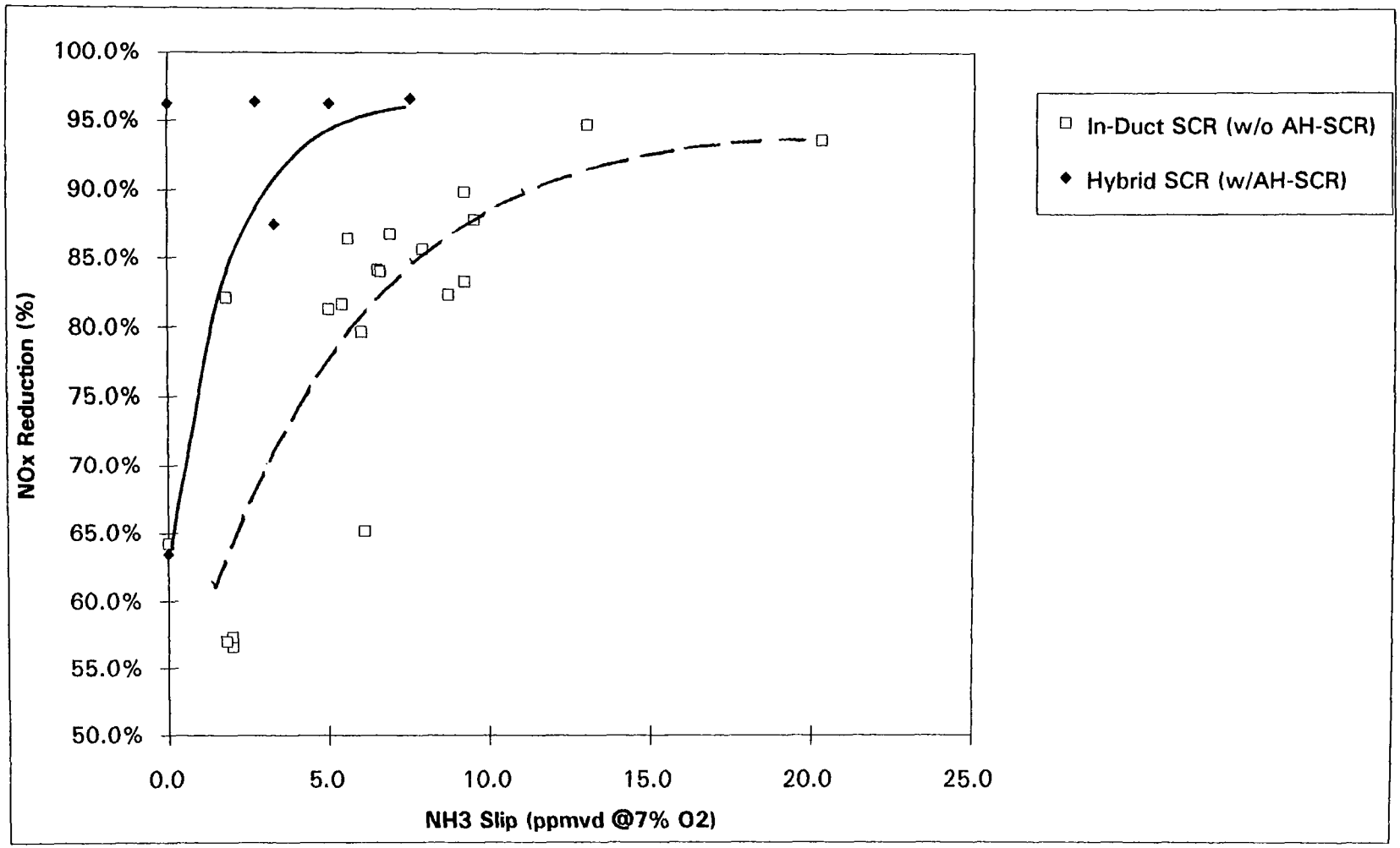

Figure 3
PSE&G Mercer Unit 2
Performance of In-duct SCR and Hybrid SCR as a Function of NH_3 Slip
Coal Firing 320 MW

Table 3
PSE&G Mercer Hybrid Urea SNCR/In-Duct SCR Evaluation
November 1994

Load (MWn)	Fuel	Furnace NSR	Baseline NO_x (ppm @ 7% O_2)	SNCR NO_x Red. (%)	Inlet NO_x @ In-duct SCR (ppm @ 7% O_2)	In-duct SCR NO_x Red. (%)	Outlet NO_x (ppm @ 7% O_2)	Hybrid System NO_x Red. (%)	NH_3 Slip In-duct @ SCR Outlet (ppm @ 7% O_2)
310	Coal	0.82	840.5	30	588.7	77.5	132.7	84.2	9.9
310	Coal	1.15	686.1	14.2	589	92.6	43.5	93.7	29.2
310	Coal	1.84	918.5	31.2	631.5	94.5	34.7	96.2	11.7
80	Coal	1.2	357.9	14.1	307.3	88.2	36.3	89.9	1.3
310	Gas	1.63	659.7	39.2	401	83.3	66.9	89.9	10.7
135	Gas	1.14	226.5	12.5	198.3	92.9	14	93.8	4.2
80	Gas	1.24	148	11	131.7	93.9	8	94.6	3.3

Table 4
PSE&G Mercer Hybrid NH_3 SNCR/In-Duct SCR Evaluation
December 1994 - January 1995

Load (MWn)	Fuel	Furnace NSR	Baseline NO_x (ppm @ 7% O_2)	SNCR NO_x Red. (%)	Inlet NO_x @ In-duct SCR (ppm @ 7% O_2)	In-duct SCR NO_x Red. (%)	Outlet NO_x (ppm @ 7% O_2)	Hybrid System NO_x Red. (%)	NH_3 Slip In-duct @ SCR Outlet (ppm @ 7% O_2)
310	Coal	1.44	1050	41	621.9	85.2	92.3	91	17
310	Coal	1.35	1080	26	799	87.8	97.2	91	12

reagent and tested with supplemental NH_3 injection. Table 4 presents the results of this testing.

Houston Lighting and Power's Webster Generating Station

In 1994 Houston Lighting and Power (HL&P) in conjunction with EPRI solicited turnkey proposals for a Hybrid SCR system at Webster Station. Proposals were received, reviewed and a vendor was selected; however, due to pending regulatory changes the project has been indefinitely postponed. The system was to include SNCR, in-duct SCR and air heater SCR for the 380 MW gas-fired boiler. The proposal effort provided a good assessment of Hybrid SCR performance and costs for the individual component systems as well as the total system. Due to the proprietary nature of the proposals and need to maintain confidentially we cannot publish the individual performance projections or costs; however, we have complied a summary, Table 5, presenting average NO_x reduction and cost (in $/kW) by technology.

Table 5
HL&P Webster Generating Station
Hybrid SCR Proposal Summary

	Average Cost ($/kW)	Average Performance (% NO_x Reduction)
SNCR	5.0	30
In-duct	15.0	60
AH-SCR	8.0	54
Complete Hybrid System	28	87

Pilot Scale Hybrid SCR Studies

A number of pilot scale studies have been performed to test and demonstrate the feasibility of Hybrid SCR. These studies have shown very promising results, with SNCR reduction exceeding 70 percent and total system reductions exceeding 95 percent at NH_3 slip values of less than 5 ppm. These results, however, are under optimum temperature, mixing, and residence time conditions and are not directly applicable to full scale commercial applications. The studies did indicate the potential of Hybrid SCR and provided data to characterize the impact of such variables as inlet NO_x and SNCR reductions on overall system performance.

Hybrid SCR Applicability

As stated previously the applicability of Hybrid SCR is very unit specific. A number of different factors regarding boiler design, fuel quality, operating modes, and regulatory limitations all impact the performance and cost of Hybrid SCR. This in turn affects the

benefits and drawbacks of Hybrid SCR when compared to alternative technologies such as stand-alone SCR. Presented below is a brief description of key variables affecting applicability. Further details as well as an applicability decision tree are included in the EPRI report.

NO_x Reduction Required

The first issue to be addressed in determining the applicability of Hybrid SCR is to determine the level of NO_x reduction required. If the emission reductions required are low, Hybrid SCR may not be applicable as other technologies, such as combustion modifications and SNCR, may achieve compliance at lower costs.

If the NO_x reduction requirements are greater than can be achieved with these lower cost technologies, either a stand-alone SCR or a Hybrid SCR will usually be required. The selection between stand-alone SCR and Hybrid SCR will depend on the issues addressed below.

High Baseline NO_x Emissions. The performance of an SCR is dependent on the inlet NO_x concentration. For a given space velocity and NH_3 slip value, the NO_x emission reduction performance for an SCR is inversely proportional to the inlet NO_x (i.e. for a given volume of catalyst the higher the inlet NO_x the lower the reduction possible). Along the same lines, to achieve the same reduction a unit with higher NO_x will require a lower space velocity and therefore a greater volume of catalyst. Therefore units with high baseline NO_x emissions are candidates for Hybrid SCR, which can reduce the inlet NO_x upstream of the SCR.

Limited Allowable Pressure Drop. The installation of catalyst in an in-duct or conventional SCR will increase the system pressure drop. This pressure drop, which is a function of the cross sectional area and depth of the catalyst bed, can be as high as 6 to 8 in. w.c. The allowable increase in pressure drop is often restricted for existing boilers, due to fan limitations or furnace pressure limitations. If the SCR pressure drop cannot be accommodated within these limitations, unit load is frequently curtailed.

Hybrid SCR can address pressure drop concerns in two ways. The first is through the use of SNCR which reduces the inlet NO_x to the SCR and thus the required catalyst volume. This in turn reduces the system pressure drop. The second way that Hybrid SCR can reduce system pressure is that air heater SCR typically adds little or no additional pressure drop beyond that of the existing air heater.

Space Limitations. The length of duct work between the economizer and the air heater on existing units is often very limited. This was done to reduce original construction costs as well as to minimize the footprint of the unit. In addition to the short duct length, the plant layout often further restricts the extent of duct modifications which can be made. The cost of relocating the air heater and or installing an external reactor can often be a substantial portion of the total cost of the SCR system. Hybrid SCR can avoid the need to relocate the air heater or install an external reactor by decreasing the volume of catalyst installed in the

duct. The catalyst volume is reduced by the lower inlet NO_x due to SNCR and/or because some of the catalyst is located within the air heater.

Structural Limitations on Boiler Steel. The installation of catalyst either in an existing duct or in an external reactor will increase the loading on existing boiler steel and foundations. In many cases the support steel and or foundations will have to be modified or a new independent structure will be required. In all cases the existing structure will have to be analyzed to determine the impact of the additional loading. The complexity of the analysis as well as the probability of the retrofit requiring substantial modifications to the boiler steel are dependent on the following:

- Weight of catalyst
- Weight of reactor or duct modifications
- Height of catalyst above grade
- Seismic zone
- Design margin of original structure

Hybrid SCR can reduce these concerns because the Hybrid SCR will require less catalyst and therefore less weight. Secondly, the Hybrid design will eliminate the need for an external reactor, allowing the catalyst to be installed in the existing or modified duct. Frequently the duct is lower in elevation than an external reactor would be. This will reduce the height of the new load reducing the modifications necessary to the structural steel and the foundations. The savings realized from lowering the load are more significant in high seismic activity areas such as California.

NH_3 Slip Limit Less than 5 ppm. There are two sources of NH_3 slip limits. The first is regulatory, and these limits have typically been 10 ppm for SCR systems and 20 ppm or higher for SNCR. The second limit is a process limit to avoid problems downstream of the SCR. The potential down stream problems include formation of ammonia salts which can cause pluggage in the air heater and fly ash contamination. For gas-fired boilers there are typically no process limits. For coal- and oil-fired boilers the process limit is dependent on the sulfur content of the fuel; and for coal, whether or not the fly ash is sold. Frequently the process limit for coal- or oil-fired boilers is 5 ppm or less.

The performance of an SCR is directly proportional to the allowable NH_3 slip. The difference between an allowable NH_3 slip of 10 ppm and 5 ppm can substantially change the performance of an SCR. As an example for the in-duct SCR installed as part of the PSE&G Mercer demonstration the full-load NO_x removal during coal firing increased from approximately 80 percent to 88 percent when the NH_3 slip changed from 5 to 10 ppm. The same NO_x reduction can be achieved with a lower slip by increasing the volume of catalyst.

A Hybrid SCR configuration with an air heater SCR can offer greater benefit than simply adding catalyst in the duct. As discussed in Section 2.0 the air heater SCR will allow operation at lower NH_3 slip values in two different ways. First, the air heater improves mixing and thus reduces stratification problems which make achieving low NH_3 slip values difficult. Secondly the air heater will transfer some of the NH_3 from the flue gas stream to the combustion air stream. Although this desorption process will increase the baseline NO_x

slightly (typically less than 10 ppm) the overall benefit for system performance will more than outweigh the increase.

Existing SNCR System. The application of Hybrid SCR to a unit which already has an SNCR will often provide the most cost effective method of achieving compliance with a new stricter regulation. If the regulation is established with two or more increments of compliance, a staged implementation may offer substantial savings. For many units which do not have the unique design parameters discussed above, a stand-alone SCR will be the most cost effective solution for a single large reduction in NO_x emissions. However, if the required reductions are incremental, the optimum solution may be Hybrid SCR implemented by installing an SNCR system first and delaying the installation of SCR until the final regulatory limits are in force. This will reduce the overall present value cost by delaying the large expenditure of capital necessary for SCR.

Selecting the Optimum Hybrid SCR Configuration

There are several steps in selecting the optimum configuration for Hybrid SCR. These steps include establishing the unit baseline data and design parameters, performing technology assessments to project the performance of each technology, and performing an economic analysis. Each of these steps is addressed below.

Establishing Unit Baseline Data

The first task in selecting the optimum Hybrid SCR configuration is to establish the baseline data for the unit to be modified. This data will be used to project the performance of the individual systems and ultimately the combined performance of the Hybrid system. The more comprehensive the data collection and the more accurate the data the more reliable the projections will be. The data required includes the following:

- Uncontrolled Emissions
- Temperature Profiles
- Residence Time
- Fan Capabilities
- Space Available for Catalyst
- Air Heater Configuration and Performance
- Unit Operating Mode and Capacity Factor

The EPRI technical report on Hybrid SCR includes detailed description of the baseline data requirements, the impact of key variables and suggest methods to obtain the data. This level of detail was beyond the scope of this paper.

Technology Assessment

Once the utility engineer has compiled the above data it can be used to perform a technical assessment of the individual technologies as well as the overall Hybrid SCR system. The EPRI technical report includes a guideline for technical assessment which will determine the

general sizing of the system and major components in sufficient detail to estimate the capital and operating costs, and will provide performance projections. This, along with the economic assessment included in the report, will allow the utility to select the optimum configuration for the Hybrid SCR.

Establishing Design Parameters

Once the determination has been made that Hybrid SCR is applicable to a given unit and the configuration has been selected, the utility will need to establish the design parameters. The design parameters will be used in the preparation of a performance specification if the utility chooses to purchase a system from a vendor or will be used to complete the detailed design if the utility chooses to design and manufacturer the system in-house. For the purposes of this study it will be assumed that the Hybrid system will be purchased from a vendor specializing in the design of Hybrid SCR systems. The work completed under the Technical Assessment will provide the basis for the majority of the design parameters. Below we outline the design parameters which should be provided to the system designer as part of the performance specification.

System Performance

The synergy of the individual systems within the Hybrid SCR can provide performance greater than the sum of the individual systems. As an example the performance of an SNCR system will be greater if the NH_3 slip is increased to provide the inlet feed for an in-duct or air heater SCR. When specifying the performance the utility has the option of specifying individual performance or specifying the overall Hybrid system performance. The advantage of specifying the overall system performance is that this allows the vendor to optimize the system to provide the greatest reduction in the most cost effective manner for each unit's design and operational constraints. Different vendors may have technical skills and/or intellectual properties that allow them to design systems with better performance in one area. Except in some unique situations (see below) specifying individual performance would establish an artificial and possibly unnecessary constraint on the system designs. Specifying system performance only will provide the greatest flexibility and maximize the options available for the design. The design criteria which are applicable to the entire system are as follows:

- Baseline Emissions
- NO_x Reduction Requirement
- Allowable NH_3 Slip
- Boiler Performance Data
- Fuel Composition
- Unit Operational Parameters
- Bid Evaluation Criteria

SNCR Design Parameters

The following design parameters are specific to the design of the SNCR process:

- Boiler Dimensions
- Time and Temperature Profiles

In-duct SCR

The following design parameters are specific to the in-duct SCR process:

- Flue Gas Flow Rates
- Temperature
- Duct Dimensions
- Allowable Pressure Drop
- Catalyst Life

Air Heater SCR

The following parameters are specific to the air heater SCR process:

- Air Heater Configuration and Performance
- Catalytic Basket Life

Specifying Individual Performance

The discussion of design parameters and specifications presented above is based on procuring a system from a single vendor and specifying the overall performance. As discussed above there is substantial benefit to this approach which allows the vendors to individually optimize the performance of the components. However, there are unique cases where the utility may wish to specify the performance of the individual systems. These cases would be where for one of several reasons the utility wishes to utilize only one of the systems under certain operating conditions.

One example where the utility may wish to specify individual performance would be seasonal NO_x limits that are lower in the summer (ozone season) than in the winter. During the summer the utility may need the entire Hybrid system to achieve compliance. During the winter, when the higher limits are applicable the system could operate with just the SCR portion of the Hybrid system. This would eliminate the operating cost for the SNCR system.

Another example would be for a utility which burned two different fuels such as coal and natural gas. Burning coal would necessitate higher NO_x reduction and therefore the entire Hybrid system. Firing natural gas would only require the SCR portion of the system, again saving the operating cost of the SNCR system.

A third example would be specifying individual performance for load following or peaking units which only operate at high loads a small percentage of the time. Due to higher

emissions and higher flow rates the critical design point for most units is full load. It is likely that a unit designed to achieve a specified limit at full load would be over designed at most lower loads. The Hybrid system would be operated at high loads and at lower loads the SNCR would be shut down to reduce operating costs.

In each of these cases it may be prudent to specify the individual performance; however, it would still be advisable to work with the vendor to optimize the overall hybrid system performance.

Commercial Considerations

There are a few commercial considerations to be addressed when purchasing a Hybrid SCR. The first is licensing of technologies and patent issues. It is not possible within the scope of this report to fully address the broad array of patent and licensing issues regarding SNCR, SCR, and Hybrid SCR. The user should be aware that there are patents for each technology, and combination thereof and that different vendors have rights to different aspects of different technologies. To provide protection from possible future patent or license infringement law suits it is advisable that the utility obtain indemnification from the vendor before purchasing a system. The vendor should also be required to prove his right to offer any technology and consideration should be given to the ability of the vendor to defend the utility against any patent or license infringement law suits.

Whether to choose a single vendor for the entire Hybrid SCR system or to purchase the systems individually will depend on the implementation schedule. If the entire Hybrid system is to be installed at once a single vendor would be best as they will provide the most comprehensive guarantee. If the systems are to be installed at different times to comply with increments of compliance, it may be advantageous to purchase them separately; however; this approach may lead to a situation where the different vendors blame each other for shortcomings in the overall system performance.

The warrantee should stipulate system performance at the end of the catalyst guarantee life. This will include NO_x reduction, reagent utilization, NH_3 slip, and pressure drop. Given the expected catalyst degradation it would be advisable to obtain a guaranteed performance for each technology at the start of life as well as the end of life. This will allow the utility to determine if the performance is acceptable when the system is installed without waiting until the system drops below the final guaranteed performance before taking corrective action.

SNCR/SCR Hybrid Capabilities and Limitations

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