

**Development and Application of Optimal Design  
Capability for Coal Gasification Systems - Task 1  
(Volumes 1, 2, and 3)**

**Topical Report  
July 1995**

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Work Performed Under Contract No.: DE-AC21-92MC29094

For  
U.S. Department of Energy  
Office of Fossil Energy  
Morgantown Energy Technology Center  
Morgantown, West Virginia

By  
Carnegie Mellon University  
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**Development And Application Of Performance and Cost Models  
For Gas Turbine-Based Selective Catalytic Reduction NOx  
Control - Task 1, Volume 1**

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## **PREFACE**

Increasing environmental awareness, coupled with an array of local, state and federal regulations have placed new requirements on process design for advanced power systems, and increased the need for sophisticated simulation and design tools for examining pollution prevention options. The current research project (DE-AC21-92MC29094) is designed to address these needs. It builds on an earlier project (DE-AC21-88MC24248) sponsored by the U.S. Department of Energy's Morgantown Energy Technology Center (DOE/METC) to develop enhanced modeling capabilities built around the ASPEN process simulator used by METC to analyze the performance of advanced power generation systems. The goals of the current project are to, (1) develop and implement new methodological capabilities in the areas of process optimization and synthesis; (2) develop and implement new performance and cost models of selected processes and process components for advanced power generation; and (3) to demonstrate the application of these new design methods and engineering models in the context of selected advanced power systems of interest to DOE/METC. This Topical Report is one in a series of reports summarizing the results of this project.

## 1.0 INTRODUCTION

Selective catalytic reduction (SCR) is a process for the post-combustion removal of  $\text{NO}_x$  from the flue gas of fossil-fuel-fired power plants. SCR is capable of  $\text{NO}_x$  reduction efficiencies of up to 80 or 90 percent. SCR technology has been applied for treatment of flue gases from a variety of emission sources, including natural gas- and oil-fired gas turbines, process steam boilers in refineries, and coal-fired power plants (Cho, 1994). SCR applications to coal-fired power plants have occurred in Japan and Germany (Behrens et al, 1991a; Chicanowicz, 1988). Full-scale SCR systems have not been applied to coal-fired power plants in the U.S., although there have been small-scale demonstration projects. SCR has become increasingly widely applied in the U.S. to natural-gas fired gas turbine combined cycle systems.

In the remainder of this section, we review the applicability of SCR, as well as the need for post-combustion  $\text{NO}_x$  control, for several power generation systems.

### 1.1 SCR Applied to Gas Turbines

Increasingly strict  $\text{NO}_x$  control requirements are being imposed by various state and local regulatory agencies in the U.S. California has long been the leader in terms of the stringency of its  $\text{NO}_x$  control regulations and, hence, has become a proving ground for new technologies such as SCR.

At the end of 1990, SCR had been applied to 80 natural gas-fired gas turbine sites in the U.S., which represent 110 gas turbine units with a combined capacity of 3,600 MW. Nearly all of the SCR installations were in California, with a few also in New Jersey, Massachusetts, and Rhode Island. At that time, most of the capacity had been on-line less than three years. Thus, the widespread use of SCR technology for gas turbine applications in the U.S. is a relatively recent phenomenon (May et al., 1991). Because many areas of California are non-attainment for ozone and nitrogen oxides with respect to the National Ambient Air Quality Standard (NAAQS), they are required to employ "Lowest Achievable Emission Rate" (LAER) technology, regardless of cost. The South Coast Air Quality Management District (SCAQMD) in California has specified SCR, in combination with steam or water injection into the combustor to minimize thermal  $\text{NO}_x$  formation, as LAER for natural gas-fired gas turbines. Other states are following California's lead. Furthermore, changes in the way in which permitting is done for attainment areas requires emission permit applications to justify why they should not be required to use the most stringent control technologies. This permitting process is referred to as the "top-down"

Best Available Control Technology approach. Unlike LAER, a BACT analysis can take into account economic implications of the technology (Schorr, 1991, 1992).

## **1.2 SCR Applied to Coal-Fired Power Plants**

Increasingly stringent environmental regulations for coal-fired power plants, coupled with the emergence of advanced coal-based technologies which offer the promise of low NO<sub>x</sub> emissions, will lead to requirements for lower NO<sub>x</sub> emissions. The top-down approach to determining BACT is one motivating factor for more stringent emission permitting. Furthermore, implicit in Title IV of the 1990 Clean Air Act Amendment is a national NO<sub>x</sub> emission reduction of 4 million pounds per year. This represents approximately a 25 percent reduction in total NO<sub>x</sub> emissions, of which coal-based power generation contributes approximately one-third.

A number of alternative coal-based power generation technologies are under development. These include integrated gasification combined cycle (IGCC), pressurized fluidized bed combustion (PFBC), and externally-fired combined cycle (EFCC) systems. These types of technologies offer the promise of lower NO<sub>x</sub> emissions than conventional pulverized coal-fired power plants. However, the NO<sub>x</sub> emission rates for some variants of these technologies may be unacceptably high.

## **1.3 NO<sub>x</sub> Control for IGCC Systems**

Nitrogen oxides are formed in combustion systems through several mechanisms, the most common of which are thermal NO<sub>x</sub> and fuel NO<sub>x</sub> (e.g., Flagan and Seinfeld, 1986). Thermal NO<sub>x</sub> results from the high temperature fixation of atmospheric nitrogen. Thermal NO<sub>x</sub> emissions are expected to be quite low for IGCC systems, due to the low heating value of the fuel gas. The presence of thermal diluents in the fuel gas, such as H<sub>2</sub>O, CO<sub>2</sub>, and N<sub>2</sub>, results in relatively low combustor flame temperatures and, hence, reduces the formation rate of NO<sub>x</sub> (Holt et al, 1989). Furthermore, recent trends to develop low-NO<sub>x</sub> combustors that do not require diluent addition also minimize thermal NO<sub>x</sub> formation. For example, premixing of fuel and air prior to combustion yields more uniform fuel/air mixtures and minimizes the peak flame temperatures which contribute to thermal NO<sub>x</sub> formation (Corman, 1994).

However, the fuel NO<sub>x</sub> mechanism may be important for some types of IGCC systems. A conceptual diagram of a generic IGCC system is shown in Figure 1. Fuel-bound nitrogen in coal is converted to ammonia to varying degrees in different types of gasifiers, depending primarily on the gasifier temperature. Lower temperatures, such as

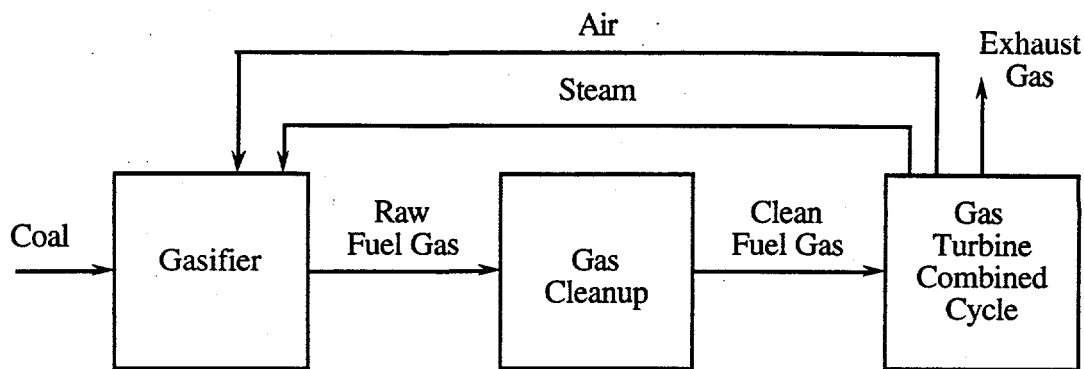


Figure 1. Simplified Conceptual Diagram of an IGCC System.

those characteristic of fixed bed gasifiers, lead to higher ammonia formation than compared to higher temperatures, such as those characteristic of entrained flow gasifiers. If the ammonia is not removed from the coal gas prior to combustion in the gas turbine combined cycle, a substantial fraction of it may be converted to  $\text{NO}_x$ , depending on the ammonia concentration and the type of combustor employed (Bedick et al, 1991; Davis, 1993).

IGCC systems based on "coal gas cleanup" (CGCU) currently hold the advantage with regard to fuel  $\text{NO}_x$  emissions. These technologies, which typically employ entrained-flow gasifiers, are characterized by relatively low fuel-bound nitrogen content in the cleaned coal gas which enters a gas turbine combined cycle, due to the combination of low ammonia yield from the gasifier coupled with wet scrubbing for particulate control and gas cooling, which also removes any ammonia generated in the gasifier. Thus, fuel  $\text{NO}_x$  formation in the gas turbine combustor is prevented. Furthermore, the coal gas can be diluted with nitrogen obtained from an air separation plant, which supplies oxygen to the gasifier, for the purposes of reducing the flame temperature in the combustor and, hence, substantially lowering thermal  $\text{NO}_x$  emissions. The low  $\text{NO}_x$  emissions of such systems have been proven in demonstration plants such as Cool Water (1986).

One promising system concept is the "simplified" IGCC. This concept features the use of a air-blown fixed bed gasifier coupled with "hot gas cleanup" (HGCU). The HGCU system features high temperature fuel gas desulfurization with a mixed metal oxide sorbent (e.g., zinc ferrite or zinc titanate), and high efficiency cyclones and/or ceramic filters for particulate removal. The off-gas from the desulfurization reactor, which contains sulfur compounds, is sent to a sulfuric acid or elemental sulfur plant for byproduct recovery. The advantages of such a system, compared to a base case oxygen-blown system with CGCU (e.g., the Cool Water project), are: (1) it does not require an expensive and energy consuming oxygen plant, (2) it eliminates fuel gas cooling prior to combustion in the gas turbine, thereby improving the plant thermal efficiency; and (3) it

eliminates the need for process condensate wastewater treatment because tars and oils in the fuel gas are not condensed as they would be in a CGCU system. For these reasons, this system concept is "simplified" compared to other alternatives (Corman, 1986).

Due to their relatively low operating temperatures, fixed-bed gasifiers such as the Lurgi technology produce a higher amount of ammonia in the coal gas than do other types of gasifiers. The ammonia, which would be captured in a conventional CGCU system during wet scrubbing, passes through the HGCU system unreacted. In a conventional gas turbine combustor, a large fraction of the fuel-bound ammonia would be converted to  $\text{NO}_x$ . In a previous system study which accounted for uncertainty in estimating the gasifier ammonia yield and the fractional conversion of ammonia to  $\text{NO}_x$  in the combustor, the 90 percent confidence interval for the  $\text{NO}_x$  emission rate was estimated to be between 1.5 and 2.8 lb/10<sup>6</sup> BTU of coal heating value input (Frey and Rubin, 1992). The current federal New Source Performance Standard (NSPS) for conventional power plants is 0.6 lb  $\text{NO}_2$ /10<sup>6</sup> BTU. The future regulatory environment with regard to site-specific environmental permitting is likely to be far more stringent. Thus, the uncontrolled fuel  $\text{NO}_x$  emissions from this system are unacceptably high.

To mitigate  $\text{NO}_x$  emissions, several approaches are possible. One is to develop alternative gas turbine combustor designs. While conventional gas turbine combustors may convert 40 to 80 percent of fuel bound nitrogen to  $\text{NO}_x$  (Bedick et al, 1991), with many design studies assuming 100 percent as a conservative upper limit, more recent combustor designs show promise of lower conversion rates. For example, conversion rates of only 20 to 40 percent were measured during one-half mass flow and pressure laboratory tests of an advanced can-annular combustor concept for the GE MS7001FA gas turbine (Battista et al, 1993). This performance has not yet been confirmed in the field under full-scale conditions. More recent work on a staged rich-quench-lean (RQL) combustor has shown conversion rates of 40 to 55 percent. However, theoretical expectations are for conversion rates as low as 10 percent with RQL designs, and work continues on this concept (Feitelberg, 1994). It thus appears possible that advances in combustor technology may reduce emissions from Lurgi-based systems with HGCU.

Current NSPS are not likely to be the binding standards for future IGCC systems. State and local permitting practices are often more stringent, and the NSPS is likely to undergo more stringent revisions in the coming years. Thus, it may be necessary to consider other approaches for  $\text{NO}_x$  control. SCR is a possible option.





compressor to the temperature needed at the turbine inlet. The flue gas exiting the heat exchanger is further cooled in a heat recovery steam generator to supply energy to a bottoming steam cycle. The exhaust air from the gas turbine, which is at approximately 1,000 °F, is used as inlet air to the combustor. The slagging combustor is potentially a significant source of both thermal and fuel NO<sub>x</sub>. Air staging of the combustion process, using a rich/lean combustor, may be effective for NO<sub>x</sub> control. It is estimated that rich/lean combustion could reduce NO<sub>x</sub> emissions to below 0.1 lb/MMBtu (Vandervort and Seger, 1991). However, it is possible that post-combustion NO<sub>x</sub> control may be required to minimize the risk of high NO<sub>x</sub> emissions.

### **1.5 The Role of SCR**

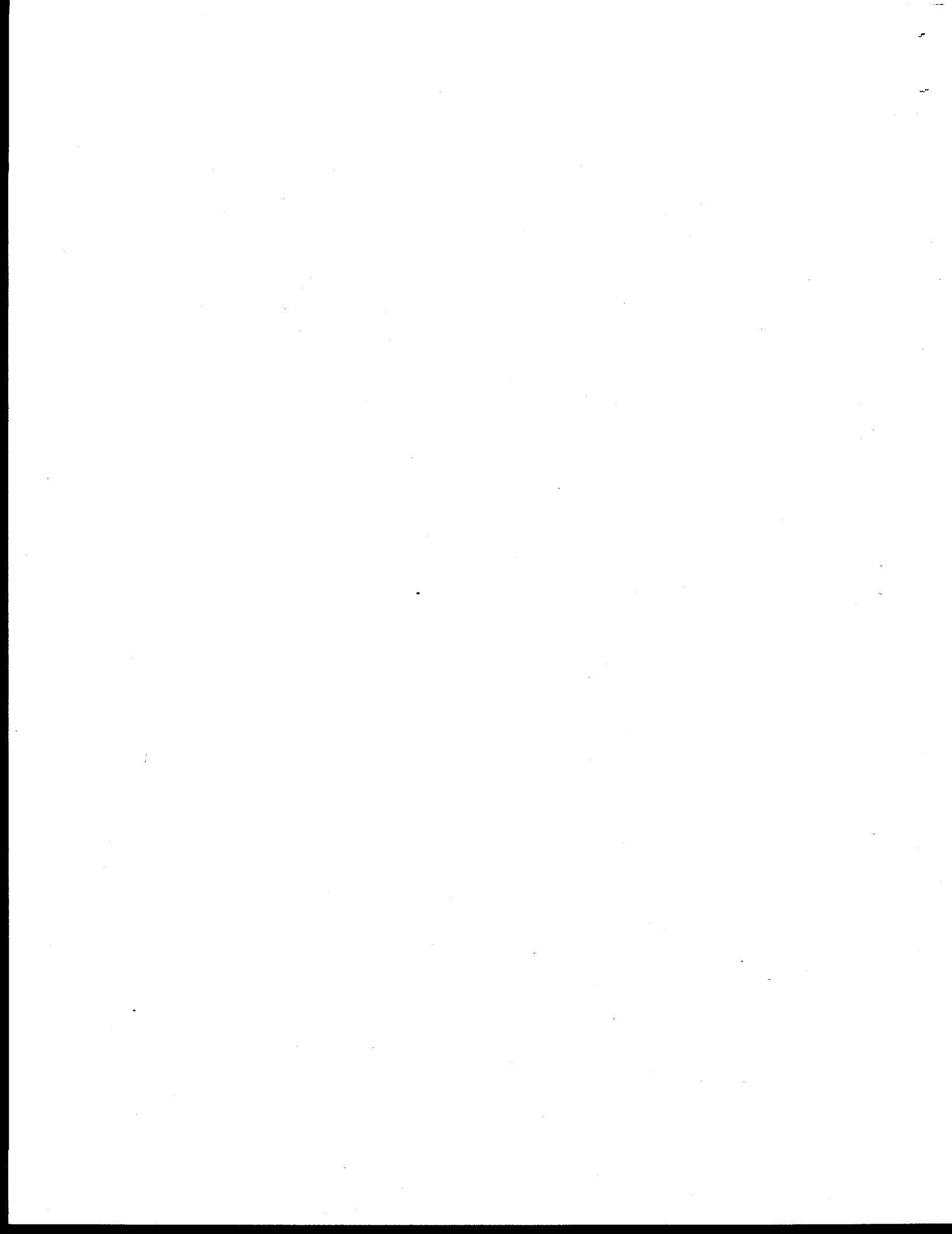
The combination of regulatory and technological factors discussed here may lead to U.S. SCR applications for conventional coal-fired power plants, as well as to coal gasification systems which do not remove fuel-bound nitrogen as part of coal gas cleanup. In the near-term, SCR is likely to be applicable to plants burning low sulfur coals (Robie et al., 1991), whose coal characteristics are most nearly similar to that of coals used at plants with successful SCR experience in Germany and Japan. Due to changing regulations, there may be a strong incentive to adapt SCR technology more generally to U.S. coal-fired power plants with varying coal sulfur contents. However, concern remains over the applicability of SCR technology to U.S. plants burning high sulfur coals or coals with significantly different fly ash characteristics than those burned in Germany and Japan. There is also concern regarding the application of SCR to peaking units due to potential startup and shutdown problems (Lowe et al., 1991).

SCR may serve as a near-term or bridging technology for use with advanced power generation systems such as IGCC, PFBC, and EFCC. For these technologies, research is underway to develop combustion systems that minimize fuel-NO<sub>x</sub> formation. However, if such technologies are not available in time for commercial deployment, SCR may be considered as backup. It is also possible that the stringency of future regulations may motivate the use of SCR, even if the development of advanced combustors is successful.

### **1.6 Scope of this Report**

In this report, the process history of SCR is reviewed. Technical background information is used to provide a basis for the development of a performance models for SCR. Because we consider here the application of SCR to coal-fueled gas turbine-based systems, for which there is no operating experience, information regarding both coal-fired

power plant and gas turbine applications of SCR are reviewed. A cost model is developed for gas turbine-based SCR applications. The performance and cost models are illustrated through several case studies. Future work regarding SCR model development, model implementation as part of other process flowsheet models, and model applications to detailed case studies of  $\text{NO}_x$  control for advanced power generation systems is discussed. One key area for future work is the quantification of uncertainties inherent in making estimates of SCR performance and cost for process environments to which it has not been previously applied. Thus, future case studies will feature a probabilistic approach to technology assessment.



## **2.0 PROCESS HISTORY AND DEVELOPMENT**

SCR was invented and patented in the U.S. in 1959. It was used originally in industrial applications. In the 1970's, SCR was first applied in Japan for control of NO<sub>x</sub> emissions from power plants. Japan was the first country to make widespread use of this technology in response to national emission standards for NO<sub>x</sub>. In Japan, SCR has been applied to gas, oil, and coal-fired power plants. There were over 200 commercial SCR systems operating on all types of sources in Japan in 1985. The Japanese SCR systems tend to run at moderate NO<sub>x</sub> removal efficiencies of 40 to 60 percent (Gouker and Brundrett, 1991). By 1990, a total of 40 systems had been installed on 10,852 MW of coal-fired power plants (Lowe et al., 1991). Coal-based experience has occurred primarily in Japan and Germany. SCR applications to gas turbine-based systems are increasingly prevalent in the U.S.

### **2.1 SCR Experience in Coal-Based Applications**

The Federal Republic of Germany currently imposes more stringent NO<sub>x</sub> emission standards than Japan. To meet the emission requirements, SCR has been adopted and applied to many coal-fired power plants. SCR will be required as a retrofit technology on a total of 37,500 MW of existing capacity. As of 1989, SCR had been applied in 70 pilot plants and 28 full scale retrofit installations, with the latter totaling 7,470 MW of hard coal-fired capacity (Schönbucher, 1989). By 1990, more than 23,000 MW of capacity were fitted with SCR systems (Gouker and Brundrett, 1991). These plants typically burn low sulfur coals (0.8 to 1.5 percent sulfur) with 0.1 to 0.3 percent chlorine. SCR has been retrofitted to power plants with both wet and dry bottom boilers, with variations on the location of the SCR system. As of 1989, 18 installations involve "high dust" placement of the SCR system between the economizer and air preheater, while the remaining involve "low dust" or "tail-end" placement of the SCR downstream of the flue gas desulfurization (FGD) system. Two of the high-dust retrofits involve wet bottom boilers (Schönbucher, 1989). In 1991, 129 systems were reported to have been installed on a total of 30,625 MW of coal-fired capacity (Lowe et al., 1991). The recent German progress in installing retrofit SCR systems is shown graphically in Figure 3.

The process environment for SCR in Germany is typically more demanding than that in Japan, with the requirement for higher NO<sub>x</sub> removal efficiencies in the presence of higher flue gas sulfur and ash loadings (Gouker and Brundrett, 1991). In both Japan and Germany, the SCR systems are not operated during startup or shutdown (Lowe et al., 1991).

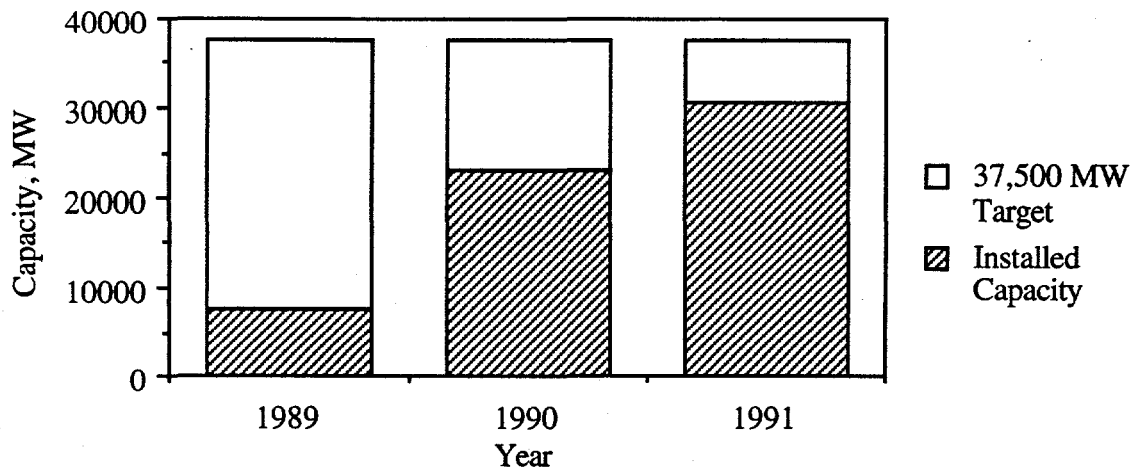


Figure 3. Targeted and Actual Installed Retrofit SCR Capacity in Germany

Recently, a number of U.S. projects for coal-fired applications of SCR technology have been initiated. These include, for example, a U.S. Department of Energy Clean Coal Program funded demonstration of SCR at Gulf Power Company's Plant Crist (DOE, 1992). SCR systems have also been permitted for two coal-fired cogeneration plants to be built in New Jersey (Fickett, 1993).

## 2.2 SCR Experience in Gas Turbine Based Applications

SCR was first used for gas turbine  $\text{NO}_x$  control in 1980 at a Japanese National Railway plant. It was subsequently applied to Tokyo Electric's 2,000 MW Futtsu Power Station, which contains 14 combined cycle units.

SCR is being applied in the U.S. for  $\text{NO}_x$  control of natural gas and oil-fired gas turbine-based power generation systems. In 1990 SCR was installed at a total of 110 units totaling 3,600 MW (May et al, 1991). While the operating environment for these systems is not as demanding as for coal-fired power plant applications, some of these applications do provide experience with systems firing sulfur-bearing fuels that encounter problems analogous to those anticipated in coal-based applications. In particular, ammonium salt formation and downstream effects have been studied (Johnson et al, 1990).

SCR appears to work best in baseloaded combined cycle systems, where the fuel is natural gas. Baseload operation is important to maintain a stable operating temperature for the SCR system. Due to the temperature requirements, SCR must be located within the heat recovery steam generator (HRSG) of a conventional combined cycle system. Natural gas is a clean fuel which lacks impurities that may poison the catalyst (Schorr, 1992).

At this time, successful gas turbine SCR installations have typically been limited to natural gas-fired units. Oil-fired operations with sulfur bearing fuels can lead to downstream affects related to the formation of ammonium salts (Schorr, 1992). A 1990 study reported that on two oil-fired internal combustion engines equipped with SCR only minor problems had been encountered after approximately 3000 hours of operation. However, formation of ammonia salts, and their deposition on heat transfer surfaces, was identified as a potential impact of SCR for oil-fired gas turbine combined cycle systems (Shareef and May, 1990).

### **2.3 SCR Economics**

Since the 1970's, the cost of SCR has dropped substantially. For example, the levelized cost of SCR dropped by a factor of 3 in Japan within a 6 year period, while in recent years costs in Germany have dropped by an additional factor of 2. These improvements are due in part to the international competition among catalyst suppliers. SCR catalysts are available from manufacturers in Japan, Germany, and the U.S. U.S. manufacturers, such as Grace, expect improvements in catalysts to continue, resulting in potential further drops in capital and operating costs. For example, Grace is testing a new catalyst design which is expected to lead to a 50 percent increase in catalyst activity while also increasing catalyst life (Gouker and Brundrett, 1991).

### **2.4 SCR Applicability to U.S. Coal Power Plants**

SCR has not yet been used commercially on conventional coal-fired power plants in the U.S. The experience in Germany, which includes boiler types similar to those in the U.S., provides useful data for predicting SCR performance and cost in the U.S. However, U.S. coals, such as eastern bituminous coals, typically have a higher sulfur content than that of German coals. In addition, fly ash compositions may vary significantly. These differences lead to concerns about maintenance of catalyst activity and potential difficulties downstream of SCR reactors, such as deposition of ammonia salts.

The German experience is particularly useful for U.S. planners because German SCR systems are subject to a more relevant range of flue gas conditions than typical Japanese systems. For example, slagging wet bottom boilers produce different flue gas and flyash characteristics that can significantly affect catalyst performance (Offen et al, 1987).

## **2.5 SCR Applicability to Advanced Coal-Based Systems**

Advanced coal-based power generation concepts, such as IGCC, PFBC, and EFCC, offer substantially different exhaust gas characteristics than conventional coal-fired power plants. For example, in IGCC systems, particulate matter and sulfur species are removed from the fuel gas prior to combustion in the gas turbine. Therefore, the sulfur and particle loadings to a post-combustion cleanup system are expected to be low. SCR catalyst formulations developed for natural gas-fired applications may be adaptable to the needs of advanced coal-fueled gas turbine-based systems.

In contrast, however, EFCC systems may have a high sulfur loading through the heat recovery steam generator (HRSG), in which a conventional gas turbine-based SCR system would be located. Therefore, it may be desirable to employ a tail-end configuration for these systems. In the tail-end approach, the sulfur and particle loading to the SCR system will be low, minimizing the technological risk of the system. The trade-off, however, is for higher costs associated with the tail-end configuration.

### 3.0 PROCESS DESIGN

The general design considerations for the SCR NO<sub>x</sub> control technology for coal-fired power plants are described here. These include the placement of the SCR system in a power plant, and a description of equipment associated with the SCR process area.

#### 3.1 SCR Integration in the Power Plant

Here, we consider the implementation of SCR for two general types of systems: conventional coal combustion-based technologies and advanced coal-fueled gas turbine-based systems.

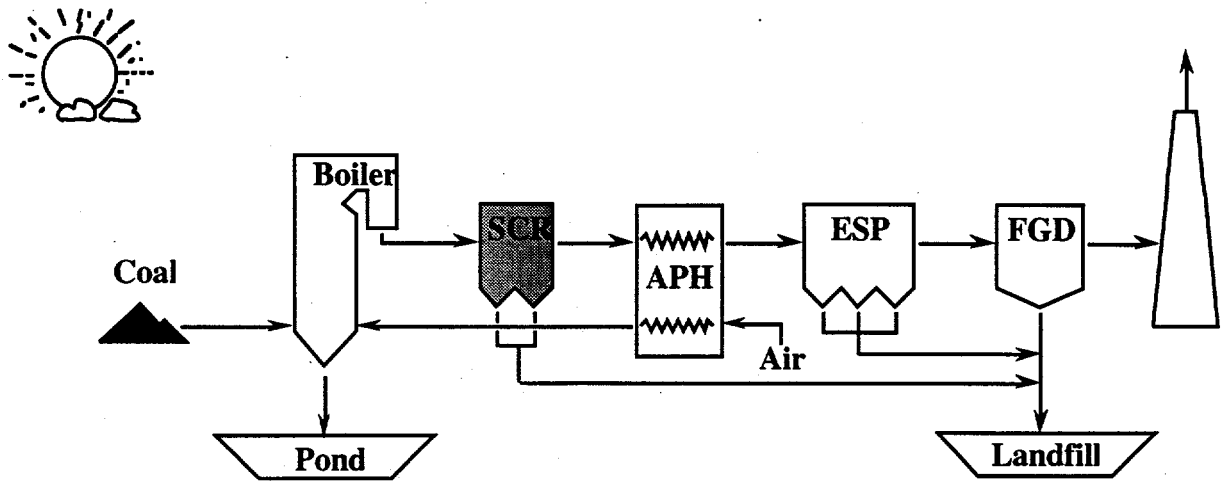
##### 3.1.1 Conventional Coal-Fired Power Plants

The SCR system can be located in several places in the coal-fired power plant flue gas stream (Schönbucher, 1989; Behrens et al, 1991). A key limitation of SCR systems is the operating temperature requirement. The operating temperature window for SCR systems is typically from approximately 550 to 750 °F. Several possible locations are illustrated in Figure 4. These are:

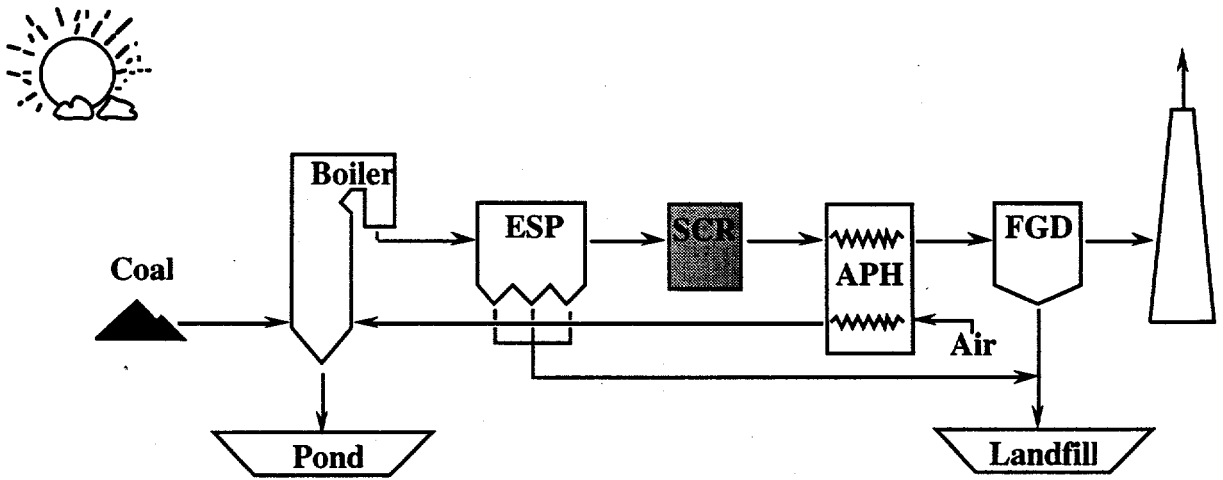
- a) "Hot-side" and "high-dust" SCR, with the reactor located between the economizer and the air preheater. In this configuration, shown in Figure 4(a), the SCR is located upstream of a cold-side electrostatic precipitator (ESP) and, hence, is subject to a high fly ash or "dust" loading. At full-load, the economizer outlet temperature is typically around 700 °F. An economizer bypass is required to supply hot gas to the SCR during part-load operating conditions, in order to maintain the proper reaction temperatures (Lowe et al, 1991).
- b) "Hot-side, low-dust" SCR, which features placement of the SCR system downstream of a hot-side ESP and upstream of the air preheater and FGD systems. This configuration has been employed in some Japanese coal-fired power plants, such as Takehara Power Station Unit 1 in Hiroshima (Behrens et al, 1991). This configuration has the advantage of minimizing the fly ash loading to the SCR catalyst, which leads to degradation in catalyst performance.
- c) "Cold-side" or "Tail-end" placement of the SCR system downstream of the air preheater, particulate collector, and FGD system. This system minimizes the effects that flue gas contaminants have on SCR catalyst design and operation, but requires a gas-gas heat exchanger and duct burners to bring the flue gas up to reaction temperature (Lowe et al, 1991).

The most common configurations envisioned for U.S. power plants are the hot-side high-dust and post-FGD tail-end systems (Robie et al, 1991), with high-dust systems predominating. These are the two most common configurations employed in German coal-fired power plants retrofitted with SCR.

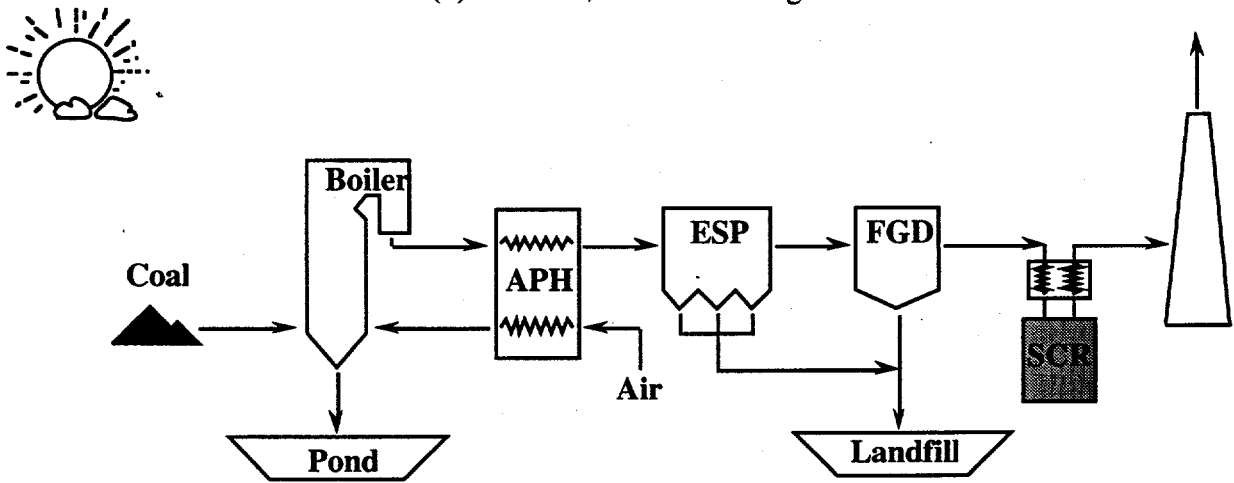




(a) Hot-side, high dust configuration



(b) Hot-side, low-dust configuration



(c) Tail-end configuration

Figure 4. Alternative Placements of SCR Systems in Coal-Fired Power Plants.

Recent German experience indicates increasing acceptance for tail-end systems. Prior to 1987, the number of high-dust installations was twice that of tail-end installation. But since 1987, the number of tail-end installations has been slightly greater. The tail-end systems have tended to be installed on smaller plants, however, and account for about one-third of total installed capacity. They have been preferred for wet bottom boiler applications (Lowe et al, 1991).

In many evaluations, tail-end SCR systems have been charged with the full cost of flue gas reheat, even though a portion of the reheat would be necessary to maintain stack buoyancy for flue gas exiting the FGD system. Similarly, in comparing hot-side and tail-end systems, the significant efficiency penalty for the high-dust system when operating at part load, due to the need for economizer bypass, must be considered (Lowe et al, 1991).

### **3.1.2 Gas Turbine-Based Systems**

Conventional SCR technology requires a temperature window of 550 to 750 °F. Therefore, the location of SCR in gas turbine systems is typically restricted to the heat recovery steam generator (HRSG), as shown in Figure 5. The temperature of the exhaust gas exiting the turbine at full load is typically in the range of 1,000 to 1,100 °F. SCR is installed at a location within the HRSG where the exhaust gas temperature has decreased to within the SCR temperature window. For conventional catalyst formulations, this is typically within the middle of the boiler heat exchanger surfaces. The temperature dependence of SCR places limitations on the load following capabilities of the gas turbine, because the temperature window may vary in location as a function of load (Schorr, 1992). New high-temperature SCR technology is under development. If successful, such technology may enable SCR applications to simple cycle gas turbines (Davis, 1993).

SCR systems employed for natural gas-fired gas turbines are sensitive to fuels containing more than 1,000 ppm of sulfur, such as some light distillate oils. High sulfur levels can result in catalyst poisoning and, therefore, loss of catalyst performance. In addition, ammonia may react with sulfur species, particularly sulfur trioxide, to form ammonia salts such as ammonia bisulfate. Ammonia bisulfate is extremely corrosive, and may condense on downstream surfaces as the exhaust gas cools. Thus, ammonia salt deposition and its impacts are of concern near the discharge of the HRSG. Research is underway to develop catalysts that inhibit the conversion of SO<sub>2</sub> to SO<sub>3</sub> and, thus, mitigate the formation of ammonia salts (Davis, 1993).

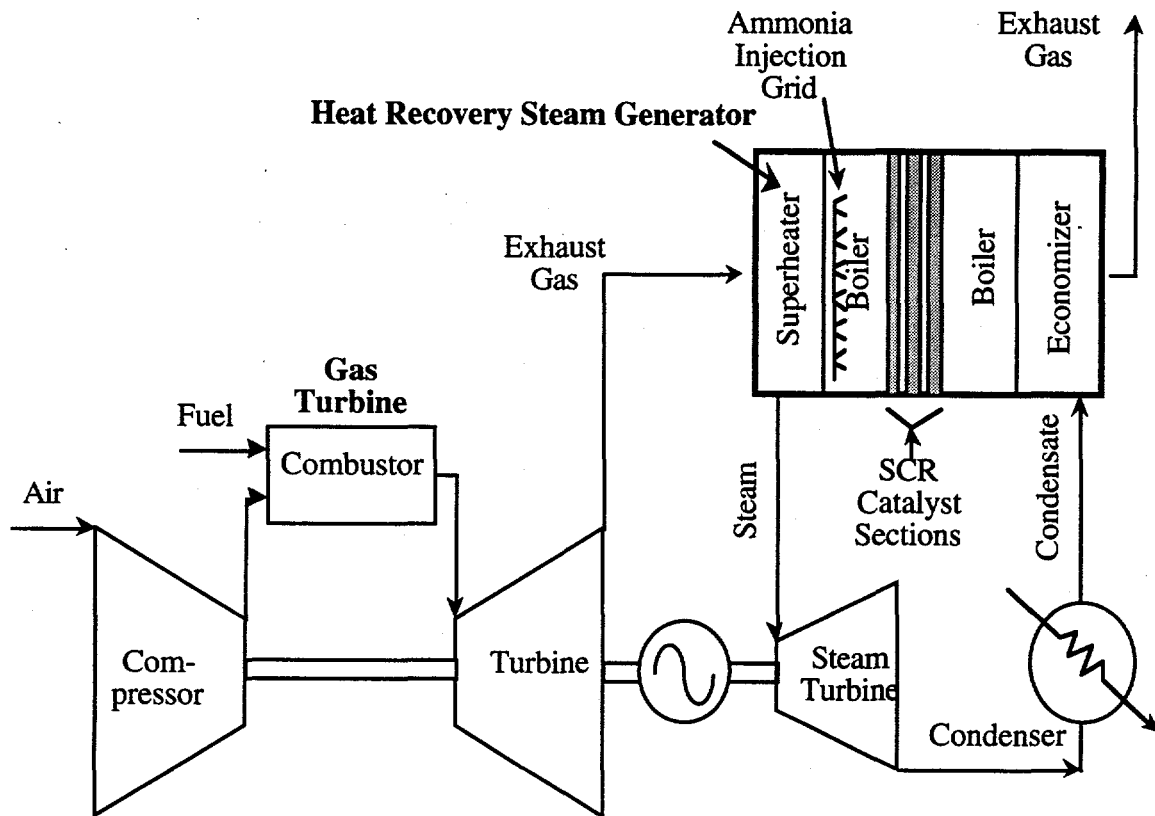


Figure 5. Placement of SCR in a Gas Turbine Combined Cycle System.

### 3.2 The SCR Process Area

A schematic of the SCR system is given in Figure 6. The schematic assumes an SCR location in the hot-side, high-dust configuration for a conventional coal-fired power plant. Ammonia is injected into the flue gas upstream of the SCR reactor vessel. The ammonia/flue gas mixture enters a reactor vessel, containing SCR catalyst. The catalyst promotes the reaction of ammonia and  $\text{NO}_x$  to form nitrogen and water vapor. Products of the SCR reactions may form ammonium sulfate or bisulfate, which can deposit on downstream equipment. In the case of a conventional coal-fired power plant, additional air preheater water washing is expected to be required to remove such deposits. For a gas turbine-based system, water washing of the downstream components of the HRSG may be required.

The SCR system consists primarily of a reactor housing containing catalyst material, an ammonia storage and handling system, an ammonia injection system, and a control system. In addition, air preheater or HRSG wash water pretreatment may be required to remove ammonia from the wastewater.

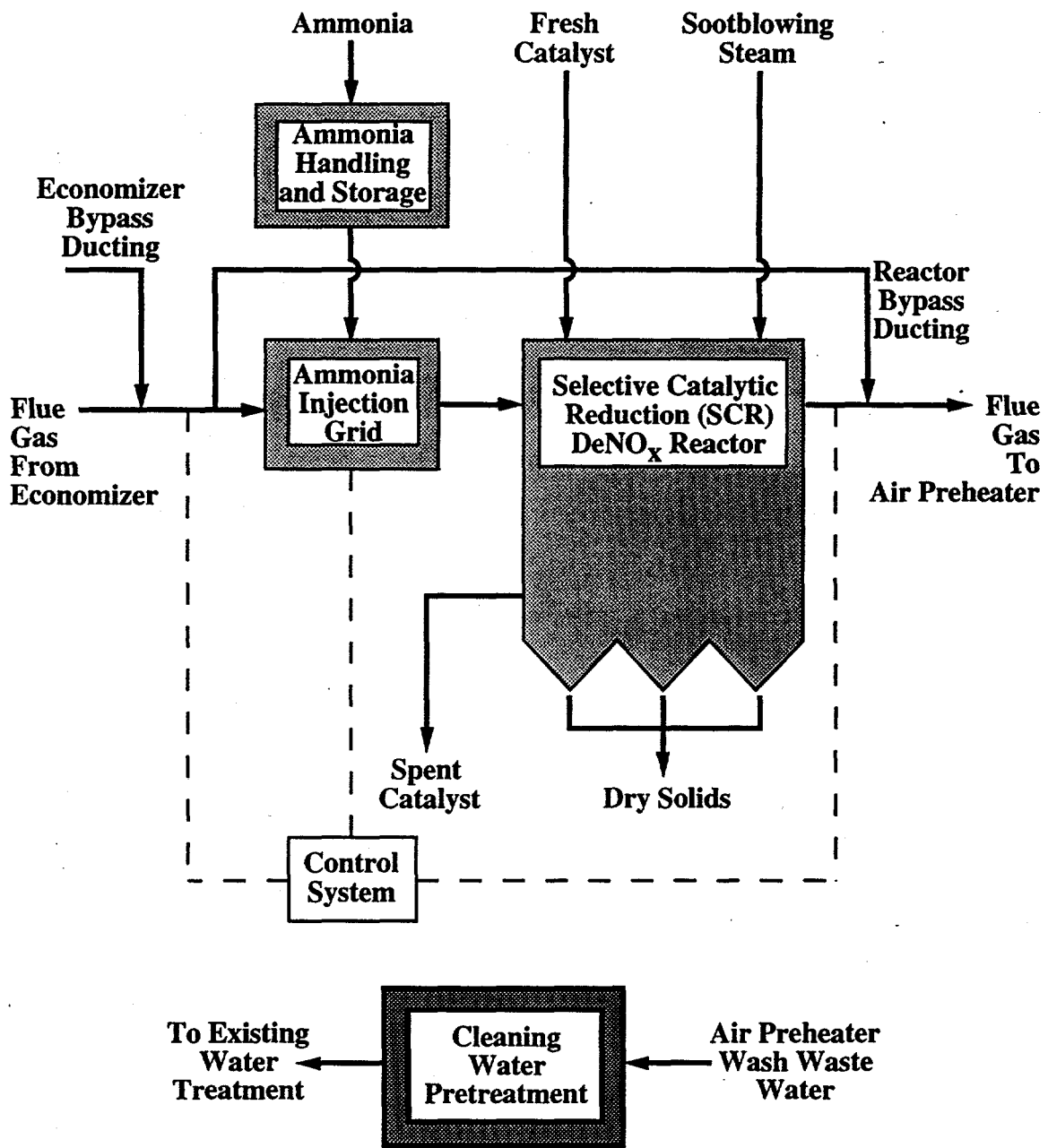


Figure 6. Schematic Diagram of SCR System

A more detailed schematic of the SCR reactor housing for a coal-fired power plant is shown in Figure 7. SCR design philosophy has evolved over the last 20 years due to the accumulation of operating experience in Japan and Germany. Based on Japanese experiences, both vendors and users of SCR technology have identified the following key design considerations for hot-side SCR systems (Lowe et al, 1991):

- Vertical downward flue gas flow to prevent ash accumulation and to allow ash drop out.
- Linear gas velocities of 16-20 ft/sec at maximum continuous rating to prevent ash accumulation. Higher velocities would be expected to increase catalyst erosion.
- Use of grid shaped catalyst with channel spacing (pitch) of 7 to 7.5 mm to allow passage of dust and prevent ash accumulation and erosion.
- Elimination of catalyst seams along the gas flow direction to prevent ash accumulation and erosion.
- Use of a "sacrificial" or dummy initial stage to prevent ash accumulation and erosion in downstream active catalyst layers.
- Removal of deposited ash using intermittent vacuuming or soot-blowing.
- Reliable ammonia feed control, including part load operation.
- Adequate ammonia feed distribution across the cross sectional flue gas flow area.
- Flue gas ducting and guide vane designs that ensure good mixing of the flue gas and ammonia feed.

Most of these features are illustrated in Figure 7.

For a gas turbine-based system, there is no choice regarding flow direction. HRSGs typically have a horizontal gas flow. Even for coal-based gas turbine systems, the ash loading should be very low exiting the gas turbine. Therefore, ash accumulation should not pose a concern as it does for hot-side applications to coal-fired power plants. For this reason, concerns about gas velocity, ash accumulation, catalyst erosion, catalyst pitch, and presence of a sacrificial layer should not be of significant concern for gas turbine based systems.

A tail-end SCR system is illustrated in Figure 8 to show the configuration of the gas-gas heat exchanger and duct burner. The flue gas exiting the FGD system must be heated from approximately 130 °F to a reaction temperature of approximately 625 °F. The flue gas exiting the SCR can be cooled to a stack temperature near 225 °F. Flue gas from the FGD system is preheated with flue gas exiting the SCR system in a heat exchanger. The SCR inlet flue gas is then heated to reaction temperature using a duct burner.

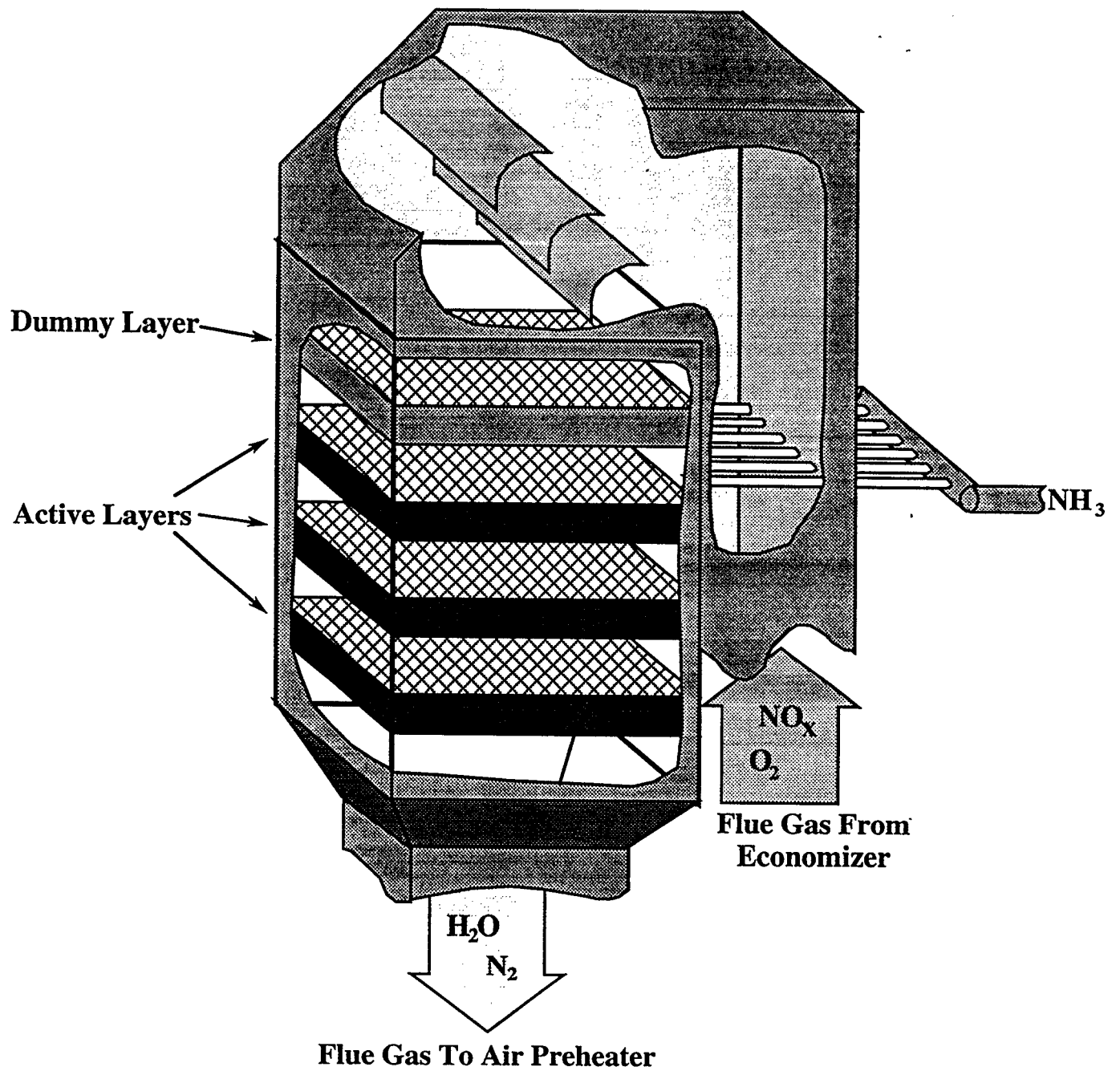


Figure 7. Diagram of SCR Reactor Housing with Downward Flue Gas Flow

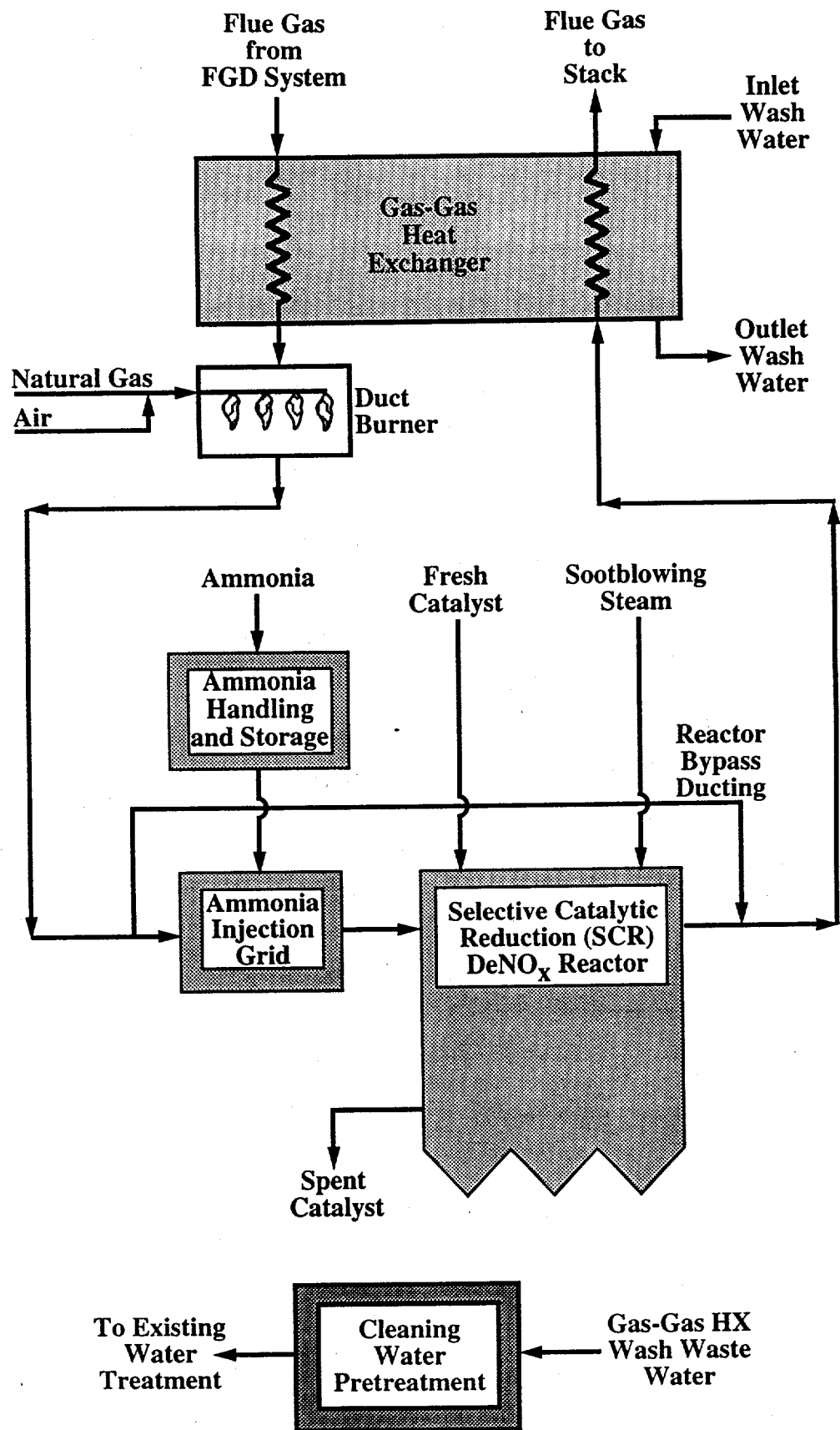


Figure 8. Schematic Diagram of Tail-End SCR System

### 3.2.1 Catalyst

SCR catalysts typically consist of a ceramic honeycomb substrate, a metal "carrier" and active components dispersed by the carrier on the honeycomb surfaces. A typical carrier is titanium dioxide ( $\text{TiO}_2$ ). Vanadium pentoxide ( $\text{V}_2\text{O}_5$ ) and tungsten trioxide ( $\text{WO}_3$ ) are commonly used as active components for hot-side SCR applications (Schönbucher, 1989).  $\text{WO}_3$  provides thermal and mechanical stability to the catalyst (Behrens et al, 1991). Catalysts based on titanium dioxide are best suited for operating temperatures of 280 to 400 °C (536 to 752 °F) (Schönbucher, 1989). At lower temperatures, catalyst activity drops substantially. At higher temperatures, catalyst material phase transition occurs, which causes irreversible activity loss (Bauer and Spendle, 1984). Catalysts using activated carbon may be employed for lower temperature applications near 100 °C (212 °F) (Schönbucher, 1989). The actual catalyst formulations which are offered commercially are closely held propriety information.

A key innovation from Japanese development of SCR technology has been the switch from noble metal oxides to base metal oxides for use as catalyst carrier materials, which has reduced many of the major problems associated with oil- and gas-fired flue gas applications. For coal applications, Japanese catalyst development also focused on improving catalyst geometry. To avoid plugging and erosion, parallel flow honeycomb and plate catalysts were developed. By the early 1980's, ceramic honeycomb and plate configurations have been developed that provide high surface areas while reducing the tendency for flyash plugging. In recent years, research has focused on understanding the deactivation mechanisms of SCR catalyst, particularly due to alkalis and trace metals such as arsenic (Gouker and Brundrett, 1991).

$\text{V}_2\text{O}_5$  is the typical component which controls the reactivity of the catalyst for base metal catalyst formulations. However, it also catalyzes the conversion of  $\text{SO}_2$  to  $\text{SO}_3$  (Behrens et al, 1991; May et al, 1991), which may lead to opacity, ammonium salt deposition, or acid condensation problems downstream. For high-sulfur coal applications, the amount of  $\text{V}_2\text{O}_5$  is minimized by homogeneous distribution throughout the catalyst. To obtain  $\text{NO}_x$  reduction, properly mixed ammonia and  $\text{NO}_x$  must enter micropores in the catalyst, which are the active sites for the reactions which consume  $\text{NO}_x$ .

Other types of catalyst formulations include precious metals and zeolites. Precious metal catalysts are often based on platinum. They are effective at a slightly lower temperature range than base metal catalysts, ranging from 425 to 525 °F. This temperature range can offer advantages in HRSG applications, where the catalyst can be placed more



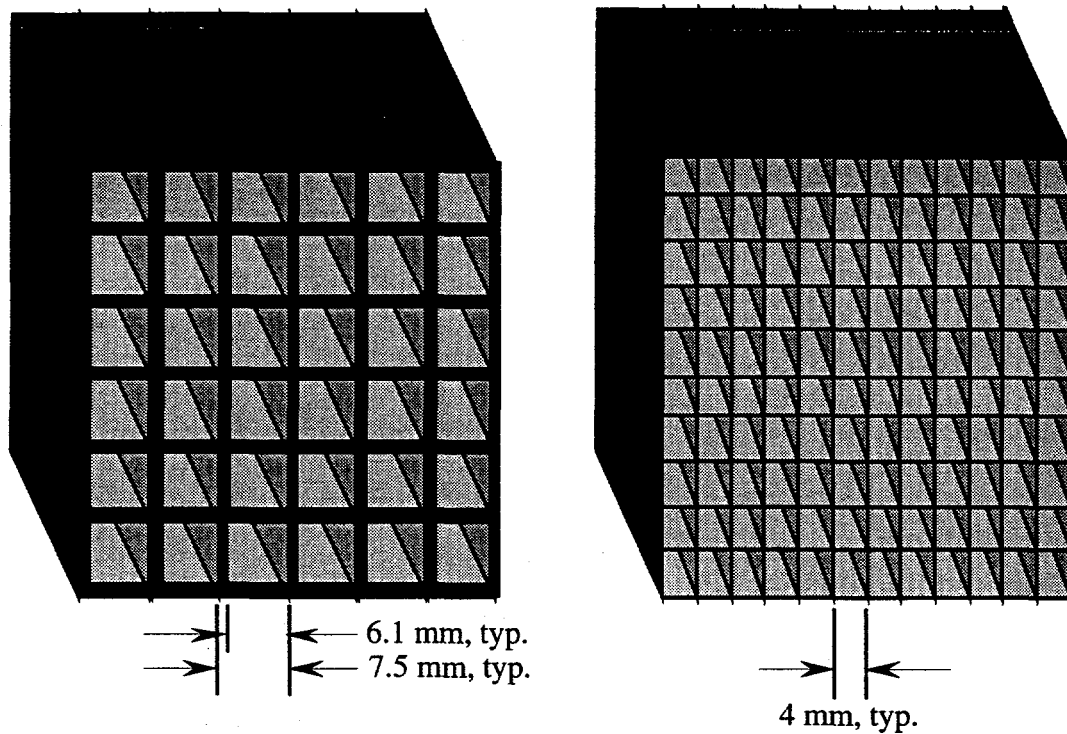
conveniently between the high pressure tubing and the economizer. Precious metal catalysts also offer advantages of recyclability compared to base metal catalysts. However, precious metal catalysts have a high sulfur dioxide oxidation potential and, when operated above their temperature range, catalyze the formation of  $\text{NO}_x$  from ammonia. Base metal catalysts are effective over a wider temperature range, and thus may be somewhat less sensitive to temperature fluctuations than the precious metal catalysts (May et al., 1991; Rosenberg et al., 1992).

Zeolite catalysts are also known as molecular sieve catalysts. They operate at a much higher temperature than either precious metal or base metal catalysts. They are reported to operate at temperatures of approximately 950 °F. This temperature is a reasonable match to the exhaust gas temperature of some simple cycle gas turbines. For this reason, zeolite catalysts may be practical for use with gas turbines that do not have heat recovery systems. There are currently four gas turbines operating with zeolite catalysts (May et al., 1991).

Base metal catalysts are the most widely used of the three major types described here. Therefore, most of the discussions of catalyst performance in this report will be based on base metal catalysts, unless otherwise indicated.

The catalyst is typically installed in a reactor housing in three layers, with provision for a dummy layer for flow straightening and distribution. In some designs, provision is also made for a fourth active catalyst layer. In these cases, the initial catalyst charge consists of three active layers. When catalyst activity drops to the design value, a fourth active layer is added. Then the four layers are changed out periodically to maintain overall catalyst activity. Catalyst modules may be loaded and unloaded from the reactor housing using a fork-lift track assembly and/or rollers (e.g., Behrens et al, 1991)

Ceramic, homogeneous, honeycombed catalyst elements approximately 6 inches square can be extruded to a length of about 39 inches (Behrens et al, 1991). SCR systems subject to high-dust loadings often include a dummy honeycomb or leading edge to control catalyst erosion (Lowe et al, 1991). Catalyst honeycomb design depends on the location of the SCR system in the power plant. For high-dust systems, catalysts with a large pitch (spacing within honeycomb cells) are employed, to allow passage of fly ash. For low dust systems, smaller pitch catalysts can be used (Schönbucher, 1989). These catalyst designs are illustrated in Figure 9.



(a) Monolith for High-Dust Application      (b) Monolith for Tail-End Application

Figure 9. SCR Honeycomb Catalyst Monolith Designs

### 3.2.2 Ammonia Handling

In Germany, strict safety standards have been applied to the shipment and handling of ammonia. Shipments by truck are not permitted if they are larger than 500 liters. Thus, anhydrous ammonia is shipped primarily by rail. A 15 to 30 day supply is typically stored at the plant in two double wall tanks. Double walled piping is also typically employed. The ammonia is diluted to an 8 percent mixture prior to introduction to the flue gas. The ammonia is vaporized in German facilities using warm water (Lowe et al, 1991).

In many U.S. gas turbine installations, electrical heating is used (Lowe et al, 1991). Ammonia can be supplied to the SCR system either as anhydrous grade or aqueous. In many cases, the use of aqueous ammonia with an ammonia content below 28 percent can avoid the need for a permit. For this type of ammonia delivery system, the vapor pressure is only 14.3 psi. Therefore, a pressurized tank is not required. However, compared to anhydrous ammonia storage, the storage capacity of the tanks are reduced by a factor of 2.5 in terms of the weight of ammonia. Steam stripping of aqueous ammonia may be required because of the presence of low concentration salts (e.g., NaCl, KCl, SiO<sub>2</sub>) that are present in the dilution water. These salts can have deleterious impacts on SCR catalyst. The

stripping steam can also be used as a diluent for ammonia injection (Rosenberg et al., 1992).

In order for the SCR system to work properly, good mixing of ammonia and exhaust gas is required. For gas turbine-based systems, an ammonia injection manifold consists of pipes with multiple nozzles designed to provide a uniform ammonia concentration profile over the cross section of the exhaust gas duct (Cobb et al., 1991; Rosenberg et al., 1992).

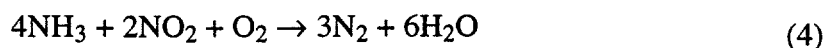
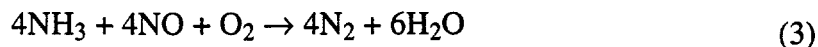
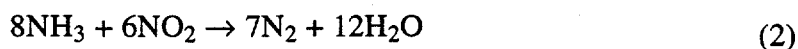
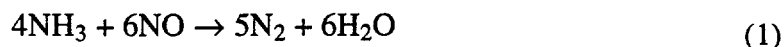
## 4.0 TECHNICAL OVERVIEW

This section presents a detailed technical overview of SCR NO<sub>x</sub> control technology for coal-fired power plants and gas turbine-based systems, with particular focus on the effects of flue gas components on catalyst performance and the effects of the SCR system on the power plant.

### 4.1 Process Chemistry

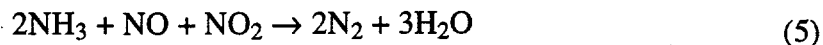
Nitrogen oxides in the flue gas are removed by reduction of NO<sub>x</sub> by ammonia to nitrogen and water. The reduction occurs in the presence of a catalyst. Ammonia is injected in the flue gas upstream of the catalyst, as illustrated previously in Figure 6.

The principle reactions are:



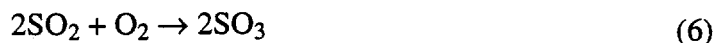
Of these reactions, reaction (3) is usually the most important. There is usually sufficient oxygen in the flue gas to serve as a reactant. In addition, typically 90 to 95 percent of nitrogen oxides in the flue gas are in the form of NO.

The overall reaction that may occur in the SCR unit is (Anderson and Billings, 1991):



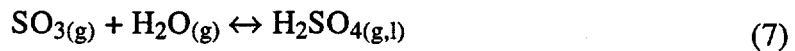
The implication of the above reaction is that a molar ratio of ammonia to NO<sub>x</sub> of 1:1 is sufficient to remove both NO and NO<sub>2</sub> when the NO/NO<sub>2</sub> mixture contains more than 50 percent NO.

Another important reaction occurring in the SCR reactor is the oxidation of sulfur dioxide:



For a typical hot-side coal-fired power plant vanadium-based catalyst, typically, 0.5 to 2 percent of the sulfur dioxide entering the SCR reactor is oxidized to sulfur trioxide (e.g.,

Bauer and Spendle, 1984). Depending on the catalyst formulation, up to about five percent of sulfur dioxide which enters the catalyst may be oxidized to sulfur trioxide (Cobb et al., 1991; May et al., 1991). Some catalyst formulations offer lower SO<sub>2</sub> oxidation potential; however, these catalysts also have a decreased NO<sub>x</sub> reduction activity per unit volume (May et al., 1991). The resulting increased levels of sulfur trioxide at the SCR outlet increases the acid dew point of the flue gas, thus increasing the potential for sulfuric acid condensation on downstream components at temperatures of less than about 350 °F. Sulfur trioxide may react with water vapor to form sulfuric acid (Johnson et al, 1990):



where subscripts (g), (l), and (s) represent gas, liquid, and solid phases, respectively.

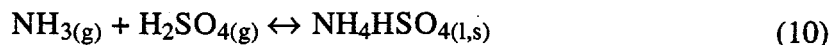
Unreacted ammonia exiting the SCR system ("ammonia slip") can react with sulfur trioxide to form compounds such as ammonium bisulfate and ammonium sulfate that deposit on downstream equipment. These compounds may result in plugging and corrosion. The key reactions for the formation of ammonium sulfate and bisulfate are (Johnson et al, 1990; May et al., 1991):

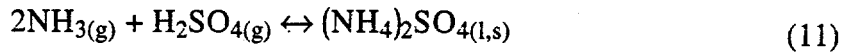


The formation of ammonium bisulfate (Equation (9)) is more sensitive to sulfur trioxide concentration than to ammonia concentration (Lowe et al, 1991).

Ammonium bisulfate is a sticky substance. Deposits of ammonium bisulfate can cause corrosion, plugging, loss of heat exchange efficiency, increased gas flow pressure drop, and shortened equipment life. Ammonium sulfate is a white crystalline compound which can also cause corrosion and plugging problems. These substances typically condense from the flue gas at temperatures below 400 °F, and may be formed even from concentrations of only a few ppm of ammonia and sulfur trioxide (May et al., 1991). Therefore, a water wash capability should be included in the cold end of any HRSG that includes an SCR system. Manual washing of finned tubes may be necessary (Kamali and Tawney, 1989).

As the flue gas cools in downstream equipment, the sulfuric acid may also react with ammonia to form condensate products (Johnson et al, 1990):





The formation of ammonium bisulfate and other ammonia salts can be minimized by reducing the ammonia slip and reducing the formation of sulfur trioxide.

The key design considerations for SCR are the  $\text{NO}_x$  removal efficiency, the ammonia slip, and the  $\text{SO}_2$  oxidation rate. While larger catalyst volumes allow higher  $\text{NO}_x$  removal efficiencies and/or lower ammonia slip, they tend to increase the oxidation of  $\text{SO}_2$ . Furthermore, large catalyst volumes significantly increase the capital and annual costs of SCR.

## 4.2 Catalyst Sizing

A key performance issue associated with SCR systems is the required catalyst volume and the catalyst replacement schedule. One parameter often used to measure catalyst volume is the space velocity, which is given by:

$$\text{SV} = G_{\text{FG}}/V_c \quad (12)$$

An alternative parameter is the area velocity, which is given by:

$$\text{AV} = G_{\text{FG}}/A_c \quad (13)$$

The relationship between the space velocity and area velocity depends on the "geometric ratio" of the catalyst, which is the unit surface area per unit volume of the catalyst.

$$R_g = \frac{A_c}{V_c} \quad (14)$$

For honeycomb monolith catalysts, the important design parameter that determines the geometric ratio is the "pitch," or spacing between the hollow cells through which the flue gas passes. As described in Section 3.2, the catalyst pitch depends primarily on the dust loading in the flue gas. Larger pitches are needed to accommodate high-dust loadings, resulting in a lower geometric ratio. Current catalysts for high-dust application typically have a pitch of around 7.5 mm, with a wall thickness of 1.4 mm and a channel thickness of 6.1 mm. For tail-end applications, pitches of around 4 mm are typical. For gas-fired applications, pitches of 3 mm are reported using ceramic plate-type catalysts. Catalysts using thin metal substrates may have a pitch as low as 0.2 mm for clean flue gas applications (Gouker and Brundrett, 1991).

The space velocity is commonly used to describe the catalyst requirement. However, both the space velocity and area ratio, or the space and area velocities, are

needed in order to specify both the catalyst volume and the catalyst area in contact with the flue gas. For example, a high-dust SCR system typically requires a smaller space velocity (more catalyst) than a low-dust system, due primarily to the lower area ratio of the high-dust catalyst designs.

The space velocity is a function of the desired NO<sub>x</sub> removal efficiency and the required ammonia slip. For example, one EPRI paper suggests a maximum space velocity of 2,500/hr for a NO<sub>x</sub> reduction efficiency of 80 percent and an ammonia slip of 5 ppm for a high-dust system for a coal-fired power plant (Damon et al, 1987). In another example, increasing NO<sub>x</sub> removal efficiency from 80 to 90 percent may entail an increase in catalyst volume of 30 to 40 percent (Kamali and Tawney, 1989).

Based on German experience, typical space velocities for high-dust SCR systems with honeycomb catalysts of about 7 mm pitch are 2,000 to 3,000/hr, whereas for tail-end systems employing honeycomb catalysts with pitches of approximately 4 mm space velocities are 4,000 to 6,500/hr. Thus, 50 to 60 percent smaller catalyst volumes per unit flue gas flow can be installed on tail-end systems for equivalent NO<sub>x</sub> removal performance. However, the flue gas volumetric flow rate may be higher for tail-end systems than for high-dust systems, partially offsetting this advantage (Lowe et al, 1991). For natural gas-fired gas turbine-based systems, a typical space velocity is approximately 8,000 hr<sup>-1</sup> to 10,000 hr<sup>-1</sup> for vanadium-based catalysts (Rosenberg et al., 1992; Shareef et al, 1992).

The actual required catalyst volume for a given application depends on a number of site-specific factors. The amount of plugging or catalyst poisoning will determine the effective catalyst activity. The activity will decrease with operating time. Therefore, the initial catalyst volume must be large enough so that at the end of the design life there is sufficient active catalyst to maintain design performance levels.

Most SCR designs have a three layer catalyst. Although most U.S. SCR studies assume that all catalyst layers are changed out simultaneously, most Japanese and German designs are based on periodic replacement of only one catalyst layer at a time. Therefore, at any given time, there may be three layers with differing lengths of service and differing activity levels. In determining the initial catalyst charge, it is necessary to account for the loss of catalyst activity at the design point in the catalyst replacement scheme. For a three layer catalyst, the design point is the fourth catalyst change out.

To determine the catalyst activity at the design point, it is necessary to estimate the loss of catalyst activity as a function of time. Catalyst activity loss is a function of the catalyst formulation and geometry, the operating conditions associated with the flue gas,

including temperature and composition, and the loading and composition of the fly ash. Recent papers, such as those presented at the 1991 Symposium on Stationary NO<sub>x</sub> Control, provide examples of catalyst activity loss curves for specific power plant applications in Germany (e.g., Behrens et al, 1991; Gouker and Brundrett, 1991; Maier and Dahl, 1991) .

### 4.3 Catalyst Fouling and Poisoning

Commercial operating experience in both Germany and Japan have provided insight into the mechanisms for catalyst fouling and poisoning. For coal-fired power plant applications, the primary cause of loss of catalyst activity was attributed to interactions between the catalyst and the flyash. As summarized by Gouker and Brundrett (1991), flyash has several effects on the catalyst, including:

- **Fouling:** Sub-micron ash particles may accumulate on the surfaces of the catalyst, and block the pores of the catalyst. This fouling or masking prevents NO<sub>x</sub> and ammonia from reaching active catalyst sites, thereby reducing the effective catalyst surface area. This leads to a reduction in the performance of the catalyst.
- **Plugging:** Bulk plugging of the catalyst occurs when large accumulations of dust occur. Dust plugging may occur, for example, when large pieces of flyash on upstream equipment "flake" off. Wire screens located upstream of the catalyst help to break up these flakes. Soot blowing may also be required periodically to remove the flakes from the catalyst.
- **Poisoning:** Alkali metals from flyash are a source of catalyst poisoning. Water soluble alkali salts may be leached onto the catalyst due to moisture present on fly ash during startup or shutdown of the SCR unit. Alkali salts have been shown to form inactive complexes with vanadium and tungsten in laboratory studies.
- **Erosion:** Erosion problems may arise due to flue gas flow distribution problems. Flow straightening vanes and dummy "catalyst" layers have been employed in many installations to reduce this type of problem.

Fly ash can also have a beneficial effect, by scouring the catalyst of materials which cause plugging or fouling, and by constantly exposing fresh catalyst material.

#### 4.3.1 Japanese and German Experience

Due to improvements in catalyst technology, catalysts in Germany are not experiencing as much loss of activity as initially predicted based on Japanese high-dust, coal-fired applications (Gouker and Brundrett, 1991).

Behrens et al (1991) report that the typical levels of potential catalyst poisons in the flyash of Ruhr coal have "not appeared to significantly accelerate catalyst deterioration" of the hot-side high-dust SCR unit at the Reuter West power station. These contaminant levels include 4 to 5 percent K<sub>2</sub>O, 6.3 percent of CaO, 1.5 percent of MgO, and 0.6



percent of  $P_2O_5$ . However, data reported by them does indicate that catalyst activity decreased to 65 percent of the level of the fresh catalyst after 30,000 hours.

At the Aichi 40 MW coal-fired boiler burning low sulfur coal, a hot-side high-dust SCR arrangement is employed. The catalyst has experienced low deterioration. After 30,000 hours of operating, the catalyst activity is 75 percent of the fresh catalyst. The primary cause of deactivation is reported to be the deposition of CaSi on the catalyst surface (Behrens et al, 1991).

While arsenic is commonly cited as the most significant catalyst poison, small sticky dust particles can cause more serious deactivation than arsenic. Furthermore, in some cases, SCR fires due to dust accumulations have occurred in Japan (Lowe et al, 1991).

German experience with SCR catalyst indicates that catalyst "lifetimes" of 3 to 4 years are possible and typical for high-dust systems. For tail-end systems, some operators report no measurable catalyst degradation, and expect to achieve up to 80,000 operating hours on a single catalyst charge. Japanese experience on clean flue gases has been similar (Lowe et al, 1991).

In many German wet bottom boilers, fly ash is recirculated to the boiler in order to slag the ash. Arsenic tends to concentrate preferentially in the fly ash during this process by 10 to 100 times compared to cases where no fly ash recirculation is used. A study of 14 wet bottom plants indicates that the actual arsenic concentration obtained in the flue gas is not monotonically proportional to the coal arsenic concentration, but may depend also on the calcium content of the flyash. Calcium oxide in the fly ash tends to getter arsenic, leading to higher arsenic concentrations in the fly ash and lower gaseous arsenic concentrations in the flue gas. This reduces the effect of arsenic as a catalyst poison. However, calcium can also "blind" the catalyst, if it does not react with arsenic (or perhaps other species). Hence, for flue gases where arsenic is not present, Japanese experience has been that fly ash calcium oxide contents of less than 1 percent permit long catalyst lives (e.g., 38,000 hours) while higher calcium contents of 5 to 8 percent result in shorter lives of less than 25,000 hours. For high sulfur coals that yield a gypsum ( $CaSO_4 \cdot 2H_2O$ ) component in the fly ash, the deactivation of catalyst is less pronounced (Lowe et al, 1991).

### 4.3.2 Gas Turbine-Based Experience

A study done of 37 operating SCR units applied to gas turbines ranging from 3.5 to 80 MW included a survey regarding catalyst operating time and catalyst replacement or additions. The total operating time for the SCR systems studied ranged from 1,200 to 40,000 hours. Of the 20 sites containing the 37 units, only three had replaced or added catalyst. One site required six replacements or additions over 40,000 hours (May et al, 1991). For natural gas-fired applications, a nominal catalyst replacement interval of five years is expected (Rosenberg et al., 1992).

A carefully designed and operated SCR system can achieve NO<sub>x</sub> emission reduction efficiencies of up to 90 percent for gas- and oil-fired applications, with ammonia slip levels of 10 ppm<sub>v</sub> or less (Snyder et al., 1992).

The SCR system introduces a new component to conventional gas turbine combined cycle systems. Therefore, the SCR system can be the source of system-wide failures. A study by May et al. (1991) indicates that the SCR system contributed to 20 percent of system-wide failures for a population of 37 units. One-half of the SCR-related failures were due to failure of the continuous emission monitoring (CEM) system, while one-quarter were due to the ammonia system and the remainder were due to the catalyst. Of the CEM failures, almost half were due to the NO<sub>x</sub> analyzer, with other sources of failure being the gas conditioning system, CO analyzer, programming and software, and the O<sub>2</sub> analyzer. For the ammonia system, the ammonia vaporizer and ammonia flow control valve had the highest frequency of failures.

### 4.3.3 Laboratory Studies of Poisons

A laboratory study by Chen et al (1990) examined the effect of several catalyst poisons on a laboratory manufactured sample of 5 percent V<sub>2</sub>O<sub>5</sub> catalyst on a TiO<sub>2</sub> carrier. This work was aimed primarily at understanding the effect of various chemicals from coal combustion on the catalyst. Both catalyst pellets and a ceramic substrate honeycomb catalyst were evaluated. The specific catalyst poisons that were evaluated include five alkali oxides (Li<sub>2</sub>O, Na<sub>2</sub>O, K<sub>2</sub>O, Rb<sub>2</sub>O, and Cs<sub>2</sub>O) and four additional compounds CaO, PbO, P<sub>2</sub>O<sub>5</sub>, and As<sub>2</sub>O<sub>3</sub>. In the experimental work, maximum catalyst activity was observed in the 200 to 300 °C (392 to 572 °F) temperature range. Commercial catalysts also include WO<sub>3</sub> as a component, which allows increased catalyst activity at higher temperatures of 300 to 400 °C (572 to 752 °F).

U.S. coals, and especially Eastern bituminous coals, contain relatively high concentrations of alkali metals. Thus, the effects of alkali and alkaline earth metal oxides on catalyst are important in these applications. Catalyst activity was shown to decrease as the amount of alkali metal dopant was increased in the laboratory tests. The strength ordering of the alkali oxide poisons corresponds to their basicity, with  $\text{Cs}_2\text{O}$  having the most pronounced effect. The deactivation may occur due to acid-base reactions forming alkali-vanadium compounds (e.g.,  $\text{NaVO}_3$ ). The poisoning due to  $\text{CaO}$  is weaker than that of the alkali metals. The basicity of  $\text{CaO}$  is also weaker than the weakest alkali oxide tested,  $\text{Li}_2\text{O}$ . Lead oxide, although a strong poison for automobile catalytic converters, is less important than the top three alkali metal oxides in deactivating the catalyst.

Although arsenic is often cited as the major catalyst poison concern, the experimental results of Chen et al (1990) indicate that  $\text{As}_2\text{O}_3$  is a substantially weaker catalyst poison than the alkali metal oxides  $\text{Na}_2\text{O}$ ,  $\text{K}_2\text{O}$ ,  $\text{Rb}_2\text{O}$ , and  $\text{Cs}_2\text{O}$ .  $\text{P}_2\text{O}_5$  was also found to be a relatively weak poison.

However, the poisoning effect of both  $\text{As}_2\text{O}_3$  and  $\text{P}_2\text{O}_5$  are temperature dependent, with increased catalyst deactivation at lower temperatures.  $\text{P}_2\text{O}_5$  poisoning leads to the formation of phosphate on the catalyst surface, which changes the catalyst surface active properties, and the blockage of surface area and pores. In spite of its relatively low poisoning activity,  $\text{As}_2\text{O}_3$  may be a more notable poison because it is often found in gaseous form, whereas many of the potentially stronger alkali metal oxides are contained in the molten coal ash.

$\text{SO}_2$  entering the SCR reactor is a precursor to the formation of  $\text{SO}_3$ , ammonium sulfates, and sulfuric acid. Under certain conditions, ammonium sulfate may deposit on the catalyst, leading to catalyst deactivation. However,  $\text{SO}_2$  alone has shown a promoting effect on catalyst activity. Formation of surface sulfates on the catalyst may promote the acidity of the surface. The work of Chen indicates that poisoning is associated with increasing basicity of the catalyst due to other contaminants.

Chloride species may have either poisoning or promoting effects on the catalyst. The poisoning effects of chlorides are much weaker than those of the corresponding oxides of the same metals (e.g.,  $\text{NaCl}$  and  $\text{Na}_2\text{O}$ ). In fact, the chlorine atom has a promoting effect, while the alkali metal atom has a poisoning effect, with a net poisoning for  $\text{NaCl}$  and  $\text{KCl}$ .  $\text{KCl}$  is a stronger poison than  $\text{NaCl}$ , analogous to the metal oxide  $\text{K}_2\text{O}$  being stronger than  $\text{Na}_2\text{O}$ .

HCl appears to react with ammonia to form  $\text{NH}_4\text{Cl}$ , which consumes ammonia and reduces ammonia available for  $\text{NO}_x$  conversion. It also deposits on the catalyst at temperatures below  $340\text{ }^\circ\text{C}$  ( $644\text{ }^\circ\text{F}$ ). HCl also appears to react with lower vanadium oxides, which are formed by reduction with ammonia, to form  $\text{VCl}_4$  and  $\text{VCl}_2$ , which are red-brown and green liquids, respectively. Some chlorides, such as  $\text{Cu}_2\text{Cl}$ , act as an SCR catalyst, which may be attributable to its acidity.

In a later study, Chen et al (1991) also added  $\text{WO}_3$  to the catalyst formulation, yielding catalyst samples more representative of commercial offerings. They performed a set of tests with the same poisons as described above.  $\text{WO}_3$  was found to improve catalyst activity and the resistance of catalyst to poisoning. However, similar qualitative results were obtained. Alkali compounds had the most pronounced effect in proportion to their basicity. Lead, arsenic, and phosphorous were also found to be weaker poisons than the strong alkali compounds tested. The addition of  $\text{SO}_2$  decreased the activity of the  $\text{WO}_3$  formulation catalyst, although when doped with alkali activity increased. Similar results were obtained for chloride related effects. In cases where vanadium chlorides form, catalyst activity will decrease.

The results of both studies (Chen et al, 1990; Chen et al, 1991) were not intended to identify the interactive simultaneous effects of multiple catalyst poisons in combination with masking or plugging, such as would occur in an actual flue gas. The purpose was to identify purely chemical mechanisms for catalyst poisoning to provide insight into actual deactivation mechanisms. Deactivation studies on actual flue gas slip streams will be conducted as part of a cooperative pilot plant program between EPRI and selected utilities (Flora et al, 1991).

For many advanced power generation systems, such as IGCC systems, alkali and chloride materials must be removed from the fuel gas prior to combustion, due to their adverse impacts on the hot gas path of gas turbine components. Therefore, the quantity of these materials which would impact an SCR system may tend to be lower than for a conventional coal-fired power plant.

#### **4.4 Catalyst Life**

Catalyst "life" is often reported as the number of operating hours between complete replacement of all catalyst in an SCR reactor. In many papers, catalyst life is described as if it is a property of the catalyst. However, it is actually a design variable. For example, there are trade-off between catalyst life, space velocity, and catalyst replacement

scheduling. In actual installations in Germany and Japan, the catalyst is installed in multiple layers and only one layer is replaced at a time according to a schedule.

In earlier EPRI-sponsored studies (e.g., Bauer and Spendle, 1984), the implicit assumption was that all catalyst would be replaced simultaneously at the end of a specified time interval. This leads to unnecessarily high operating costs. Japanese catalyst vendors and German SCR operators have both reported on the economic benefits of phased catalyst replacement schemes (e.g., Appendix B of Bauer and Spendle, 1984).

In a recent EPRI study, Robie et al (1991) assume that a catalyst life of four years will be realized for U.S. coal-fired high-dust applications. They also assume the same life for tail-end applications even though there are clear differences in operating environments for the two cases. This type of assumption may unnecessarily penalize the tail-end configuration when in fact the major benefit of this configuration is a decrease in catalyst activity loss over time.

Cho (1994) discusses catalyst management strategies, emphasizing that such strategies should consider the addition or replacement of varying quantities of catalyst at different time intervals. The purpose of such strategies are to minimize the lifetime cost of the SCR system.

#### **4.5 Catalyst Disposal**

Catalyst disposal depends on the catalyst formulation. Precious metal catalysts have a large salvage value and, therefore, will typically be recycled. Zeolite and base metal catalysts, however, are more likely to be disposed of as a solid waste. Zeolite catalysts are not considered to be hazardous waste. Therefore, the costs of disposal for zeolite catalysts may be relatively low. In contrast, base metal catalysts containing vanadium may be classified as hazardous in some areas. It is possible that spent catalyst would be classified as a hazardous waste in the U.S. Vanadium pentoxide is on the U.S. EPA's list of Extremely Hazardous Substances. While the quantity of  $V_2O_5$  contained within the catalyst is small, the total volume of catalyst is large. Therefore, the costs of disposal could be substantial if the material is treated as a hazardous waste. At this time, there appears to be no federal requirement that the catalyst be treated as a hazardous waste. However, in the State of California, spent catalyst material has been classified as hazardous (Schorr, 1992; Snyder et al., 1992).

The disposal of base metal catalysts may be done by the SCR plant operator or by the catalyst manufacturer. In the latter case, the catalyst is returned to the vendor. In either case, the catalyst user will bear the costs of disposal.

In Japan and Germany, spent base metal catalyst is returned to the manufacturer. In Japan, catalyst manufacturers have not found it economical to regenerate the catalyst, and the catalyst is often simply disposed of (Lowe et al, 1991). State laws which limit shipment of hazardous waste could make return of catalysts to the vendor prohibitively expensive or illegal (Schorr, 1992).

#### **4.6 Impacts on Other Plant Components**

First, we consider the system implications of SCR for conventional coal-fired power plants. Then we review similar issues for gas turbine based systems.

##### **4.6.1 Coal-Fired Power Plants**

SCR has effects on other components of the power plant, particularly for high-dust designs. According to Robie et al (1991), the main impacts are on the boiler, air heater, and induced draft fan. Other components affected are the FGD process, FGD reheat system, waste disposal system, and water treatment system. These impacts are summarized below.

- **Air Preheater.** Air preheater modifications are required due to the deposition of ammonium sulfates and bisulfates. Heat transfer surfaces must be replaced with heavier gauge metal and, in some cases, modified design surfaces. Additional water wash capability is required for air preheater cleaning. High pressure soot blowers are also required at both the hot and cold ends of the air preheater. Air preheater leakage may increase.
- **Boiler.** Loss of thermal efficiency results from air preheater modifications and, at part load operation, from an economizer bypass, which is required to maintain the reaction temperature in the SCR unit.
- **Induced Draft Fan.** In a new plant, a larger ID fan is required to overcome the pressure drop in the SCR reactor and any other incremental pressure drops associated with downstream effects. This pressure drop may be up to 11 inches of water.
- **Forced Draft Fan.** The forced draft fan for the combustion inlet air to the air preheater will have a higher mass flowrate due to increased air preheater air leakage.
- **Stack.** The increase in the SO<sub>3</sub> concentration of the flue gas could result in increased opacity of the flue gas plume if the SO<sub>3</sub> is not removed in the FGD system. Acid condensation would also be a potential source of concern.
- **ESP.** Because of the lower operating pressure of the ESP due to the pressure drop of the SCR system, higher flue gas temperature, and increased flue gas mass

flow due to air preheater leakage and gases introduced for ammonia injection, the ESP will be required to handle a higher volumetric flow rate. In a new plant, therefore, a larger ESP will be required. The ESP may require additional reinforcement due to the lower, and negative, operating pressure. Although the  $\text{SO}_3$  content of the flue gas will increase, the beneficial effect of this on ESP performance may be offset by the increase in flue gas temperature. Ammonium salt precipitation in the fly ash could improve agglomeration and reduce reentrainment.

- Ash Disposal/Reuse. Ammonia compounds contained in fly ash material decompose and release ammonia at elevated pH. Even at lower pH, ammonia fixation with alkaline species could result in an ammonia odor problem. Flyash containing ammonia compounds may not be suitable for use in cement manufacturing.
- Water Treatment. Water treatment in addition to typical plant waste water treatment is required to convert nitrogen species in the air preheater wash water to free nitrogen.
- FGD/Reheat. Because of the higher flue gas inlet temperature and mass flow rate, there will be an increase in the water evaporation rate for wet limestone systems. In addition, steam would be required for reheat. The FGD liquor recirculation rate may need to be increased to maintain the same  $\text{SO}_2$  removal efficiency. Alternatively, reheat can be accomplished using flue gas duct burners.
- Auxiliary Power Consumption. The net plant output will be decreased by the electricity required to operate SCR process equipment. In addition, during times of soot blowing, the plant efficiency will be decreased slightly due to the use of process steam. Steam is also used for ammonia vaporization, and dilution air for ammonia injection is taken from the discharge of the primary air fans.

The effects of the tail-end SCR system are not as significant as for the high-dust configuration. The tail-end SCR will result in auxiliary power consumption, flue gas pressure drop, water washing of the reheat gas/gas heat exchanger and associated wash water treatment, increased requirement to eliminate mist carryover from the FGD system, and stack effects due to increased  $\text{SO}_3$  concentration and higher stack temperatures. In addition, a duct burner may be required. Sootblowers and ash collection hoppers are not required for the SCR system in the tail-end configuration. In the tail-end configurations, separate dedicated dilution air fans are used for ammonia injection (Robie et al, 1991).

For tail-end systems, the leakage rate of the gas-gas heat exchanger used for reheat has been reported to be as high as 7 percent in German facilities. Such leakage allows untreated flue gas to leak into the treated gas prior to stack discharge, thereby effectively bypassing the SCR system.

#### **4.6.2 Gas Turbine-Based Power Plants**

For gas turbine based systems, the SCR system impacts center on the following issues (Schorr, 1992; Snyder et al., 1992):

- Base-Load Operation. Because SCR systems require a relatively narrow temperature window for effective operation, their placement within the HRSG constrains the acceptable turbine exhaust gas and heat exchanger temperatures that lead to high efficiency NO<sub>x</sub> control. The temperature at the SCR reactor is highly dependent on the electrical load to the combined cycle. Fluctuations in temperature due to load following operation can lead to conditions in which the SCR reactor is no longer within its required temperature window. Therefore, effective use of SCR typically requires base-load operation of the plant.
- Operating Temperature Range. The operating temperature range required for base metal catalysts, which are the predominate catalyst type in use currently, necessitates their location within the HRSG. Therefore, base metal catalyst systems cannot be employed for simple cycle applications. As noted previously, however, recent developments in zeolite catalysts may enable the use of SCR for future simple cycle systems.
- Sulfur-Bearing Fuels. As of 1992, there was no successful operating experience of SCR used in a sulfur-bearing oil-fired gas turbine application. Oil-fired systems that have used SCR successfully tend to have very low sulfur fuels. For example, one facility in Massachusetts used a very low sulfur distillate oil as a backup fuel to natural gas. A gas turbine system in Japan has been operated on kerosene, which has a low sulfur content, with no adverse effect on or due to the SCR system. However, the concern for the use of sulfur-bearing fuels is not adverse effects on the catalyst due to poisoning, but rather it is due to the catalyzed oxidation of sulfur dioxide and the subsequent downstream deposition of ammonia salts.
- Ammonia Salts. As described previously, one of the byproducts of SCR operation on gas turbine systems firing sulfur-bearing fuels is ammonia salt formation. Ammonia bisulfate causes corrosion of boiler tube materials as well as fouling and plugging of the boiler. Deposits increase the exhaust gas pressure drop through the HRSG and, therefore, increase the backpressure on the gas turbine. An increase in backpressure results in a decrease in net power output and plant efficiency. Ammonia compounds also result in emissions of PM<sub>10</sub> (particles less than 10 microns in diameter).
- Effect of Ash. SCR has been used successfully on systems firing sulfur bearing fuels, such as conventional coal or oil-fired boilers. However, in such systems, there is also a significant quantity of fly ash present. The fly ash scours deposits of chemicals off of surfaces and provides a large surface area upon which ammonia salts may deposit, instead of depositing on heat exchanger surfaces. Thus, the formation of ammonia salts in conventional coal-fired power plants is balanced to some extent by the presence of fly ash. Fly ash is typically not present in significant quantities in gas turbine applications.

SCR systems have been applied successfully to natural-gas fired gas turbine combined cycles in baseload service. Preliminary experience suggests that other types of applications, particularly with sulfur-bearing fuels, can lead to adverse impacts. However, catalyst technology is evolving, and catalysts which minimize sulfur oxidation may offer some relief from the problems of downstream ammonia salt deposition.



## 4.7 SO<sub>2</sub> Oxidation

In the U.S., flue gases from coal-fired power plants have typically higher SO<sub>2</sub> and SO<sub>3</sub> concentrations than experienced in Japan and Germany due to the predominance of high sulfur coals in many regions of the country. The oxidation of SO<sub>2</sub> leads to downstream effects such as ammonium sulfate and bisulfate formation and acid condensations as previously described. Sulfur trioxide and sulfuric acid formed downstream of the SCR can also lead to attack of duct liners. The formation of these condensates and deposits depends critically on the presence of ammonia and sulfur trioxide in the flue gas. However, to minimize this type of problem in U.S. applications may require optimization of catalysts for specific U.S. markets (Lowe et al, 1991).

According to data reported by Bauer and Spendle (1984), SO<sub>2</sub> oxidation is primarily a function of catalyst formulation, space velocity, and operating temperature. Catalysts which minimize sulfur oxidation also tend to have a lower NO<sub>x</sub> reduction activity.

## 4.8 Ammonia

A portion of the ammonia injected into the SCR system may pass through the reactor unchanged. Ammonia in the flue gas may react chemically or physically with other constituents of the flue gas, including fly ash. This may lead to maintenance and operational problems. Several key concerns are discussed further.

### 4.8.1 Ammonia Injection

In commercial SCR systems, a critical design issue is the injection of ammonia into the flue gas upstream of the SCR reactor. A key difficulty in ammonia injection is obtaining a uniform mixture of ammonia in the flue gas. Failure to achieve proper ammonia injection and mixing can lead to channeling of ammonia through the SCR system, resulting in high levels of ammonia slip through the SCR reactor and to downstream components in the flue gas path.

Obtaining a uniform distribution of the injected ammonia in the flue gas upstream of the SCR catalyst is often difficult. Flue gas flow modeling and flow straightening devices are often needed to understand and achieve proper flow distribution. Ammonia injection systems typically consist of 30 to 40 injection points per square meter. These injection nozzles are controlled either singly or in groups of several, and the flow of ammonia through them can be optimized to achieve a reasonably uniform ammonia distribution ( $\pm 10$ -30 percent) in the flue gas. However, dust deposits on or around the nozzles can lead to

plugging of some or alteration of the flue gas flow pattern. Thus, the flow patterns may change over time and require periodic checking and adjustment (Lowe et al, 1991).

#### **4.8.2 Ammonia Retention in Catalyst**

Some SCR catalysts may retain ammonia during operation. The ammonia is then released during transients or shut downs. This desorption process may take up to eight hours, based on currently known experience. The ammonia injection rate during low temperature operation should be adjusted to compensate for offgassing of ammonia from the catalyst to maintain ammonia slip within tolerable levels. Because of the absorption/desorption phenomena, changes in NO<sub>x</sub> emissions may lag changes in the ammonia injection rate by 30 minutes (Lowe et al, 1991).

Control problems under load swing conditions, exacerbated by the time-lag phenomena, remain an issue, particularly for potential U.S. high sulfur coal applications (Lowe et al, 1991).

#### **4.8.3 Effects on Downstream Heat Exchangers**

The most common effect of ammonia slip that is discussed in the literature is the deposition of ammonium sulfates on downstream equipment. However, commercial operation has not always substantiated this concern. For example, in the Takehare Power Station Unit 1, featuring a coal-fired power plant with hot-side SCR downstream of a hot side ESP, no additional air preheater washings have been necessary during 34,000 hours of SCR operation. The SCR operates at 80 percent removal efficiency with a NO<sub>x</sub> loading of 300 ppm. The lack of plugging of the air preheater by ammonium salts, even in spite of SCR inlet SO<sub>2</sub> concentrations of 1,000 to 1,500 ppm, is attributed to low NH<sub>3</sub> slip levels. These have been 0.2 ppm or less. SO<sub>2</sub> conversion to SO<sub>3</sub> was typically 0.08 to 0.21 percent (Behrens et al, 1991).

The high-dust hot-side SCR system in the Reuter West power station in Berlin, Germany has operated over 15,000 hours on coals with sulfur contents up to 1.2 percent. The typical NO<sub>x</sub> removal efficiency is 85 percent with an ammonia slip of 1.5 ppm<sub>v,d</sub> and an SO<sub>2</sub> conversion rate to SO<sub>3</sub> of about 0.5 percent (Behrens et al, 1991). No plugging of the air preheater is reported and no washing of the air preheater has been necessary since SCR startup. The SCR catalyst layers receive a weekly sootblowing.

In one German power plant, Neckar (1989) reports that approximately 5 percent of the ammonia leaving the SCR system is deposited as an ammonium salt in the air preheater, with typically about 50 percent of the ammonia absorbed onto fly ash. The ammonium

salts are easily soluble in water, and can be washed. As water washing in the air preheater proceeds, the concentration of ammonia in the exiting water stream decreases. Wash water with a high ammonia concentration must be treated to remove the ammonia prior to entering the regular plant wastewater treatment system. Neckar suggests that the initial wash water with a high ammonia concentration can be pretreated separately from the larger volume of water with a low ammonia concentration, which may be suitable for direct feed to the existing waste water treatment plant.

In pilot plant testing of an SCR system, Shiomoto and Muzio (1986) report that ammonia entering the preheater tends to deposit on air preheater surfaces as solid ammonium compounds or to be absorbed onto fly ash. Furthermore,  $\text{SO}_3$  in the flue gas is consumed in the formation of ammonium sulfate or bisulfate, and also was absorbed onto fly ash. The investigators report that essentially all of the gaseous  $\text{SO}_3$  entering the air preheater during testing was removed from the flue gas.

The effects of deposits include fouling of heat transfer surfaces and increase of pressure drop in flue gas paths. These types of effects may be more pronounced during process upsets (Lowe et al, 1991).

For natural gas-fired gas turbine based systems, problems associated with downstream ammonia deposition are not reported.

#### **4.8.4 Ammonia Absorption by Flyash**

Another concern regarding ammonia slip for coal-fired power plants has emerged in Germany. German experience has been that the typical Japanese criteria of 5 ppm maximum ammonia slip is often not stringent enough to permit commercial use of fly ash as a byproduct. Therefore, in many German installations ammonia slip must be limited to 3 or even 1 ppm (Lowe et al, 1991). Schönbacher (1989) reports that ammonia slip must be limited to 2 ppm to produce a byproduct fly ash acceptable to the cement industry.

Although ammonia does not alter the physical properties of concrete made from fly ash, ammonia captured in the fly ash is released during concrete mixing and may result in a noticeable odor. For ammonia concentrations of less than 60 mg per kg of fly ash, the odor is not noticeable. For the Altbach/Deizisau power station Unit 5 in Germany, it appears that 20 to 80 percent of the ammonia slip is captured in the fly ash, with a mean value near 50 percent (Neckar, 1989).

Experimental studies by Shiomoto and Muzio (1986) indicate that most of the ammonia leaving the SCR system exits as a gas, with very little in the form of solid

compounds or absorbed by fly ash. However, a portion of the gaseous ammonia is absorbed by flyash downstream of the SCR reactor, with higher ammonia partial pressures leading to increased absorption.

#### **4.8.5 Ammonia Slip and Tail-End Systems**

For tail-end systems, ammonia slip is not a significant concern because of the low concentration of sulfur in the flue gas. Therefore, the constraints on catalyst performance are less severe, allowing potentially greater degradation in catalyst activity (and associated increase in ammonia slip) before replacement is required. Ammonia slip constraints for tail-end systems are typically imposed by air emissions regulations, as opposed to downstream process requirements. Ammonia slip as large as 20 to 30 ppm is not expected to lead to operational problems in tail-end systems. Ammonia odor and plumes become noticeable when the ammonia concentration exceeds 50 ppm (Lowe et al, 1991). Ammonium salt deposition is expected to occur in the gas-gas heat exchanger used for flue gas reheat in tail-end SCR systems. Thus, heat exchanger water washing is also required in this case.

#### **4.9 U.S. Outlook**

Because the German and Japanese experiences on coal-fired power plants cannot be directly applied to U.S. applications, EPRI and others are involved in pilot testing of SCR systems on selected slipstreams, analogous to the German testing of over 70 SCR pilot systems. These tests will provide additional data regarding cost and technical feasibility of SCR applied to plants firing domestic medium and high sulfur coals. EPRI will conduct as many as 14 separate tests (Lowe et al, 1991).

In the short term, low-dust tail-end SCR systems hold the most promise of reliable performance for high sulfur coal applications. This type of system would avoid the potentially excessive rate of air preheater fouling and catalyst deactivation that might otherwise be experienced in a high-dust configuration in high sulfur service (Lowe et al, 1991).

There are 105 operating cyclone units in the U.S. totaling over 26,000 MW. These are high NO<sub>x</sub> emission technologies which are not easily amenable to combustion NO<sub>x</sub> control. The typical NO<sub>x</sub> emission rates for these units ranges from 0.8 to 1.8 lb/MMBtu, corresponding to flue gas concentrations of 500 to 1,100 ppm. Many of these units are also located in the midwestern U.S., which is the major source of utility acid rain

emissions. Thus, these boilers would appear to be a prime target for application of tail-end SCR systems (Lowe et al, 1991).

For natural gas-fired gas turbine combined cycle systems in base load service, SCR has been successfully demonstrated. For gas turbine systems firing sulfur bearing fuels, a potential impact is the downstream formation of ammonia salts. Use of liquid fuels containing very low sulfur appears to be successful for gas turbine systems.

Of concern then, is whether gas turbine-based SCR technology can be directly applied to advanced power generation systems, such as IGCC. The fuel specifications for gas turbines in IGCC service tend to be stringent with respect to acceptable fuel gas sulfur, alkali, and particles concentrations. For example, the fuel gas from an IGCC system with hot gas cleanup would typically contain approximately 10 ppm<sub>v</sub> of acid gas (e.g., H<sub>2</sub>S, COS), leading to very low concentrations of SO<sub>2</sub> from the gas turbine combustor exhaust. Equilibrium calculations indicate that for low exhaust gas SO<sub>2</sub> concentrations (e.g., on the order of 2 ppm<sub>v</sub>), ammonium bisulfate formation can be avoided if ammonia slip levels are maintained below 5 ppm<sub>v</sub> (May et al., 1991). Thus, for some types of IGCC systems, the ammonia salt formation problem may be manageable.

For cold gas cleanup-based systems, the sulfur loading in the fuel gas may be as high as several hundred ppm<sub>v</sub>. However, because IGCC systems with cold gas cleanup system successfully remove fuel-bound nitrogen species such as ammonia, they are not as likely to require post-combustion NO<sub>x</sub> control as hot gas cleanup-based systems.

For other power generation systems, such as the EFCC, SCR applications may have to be implemented in a tail-end approach. The EFCC features post-combustion control of sulfur dioxide emissions from a high temperature slagging combustor. The sulfur content of the gas which passes through the HRSG may be unacceptable high for SCR applications. Therefore, it may be necessary to locate the SCR system downstream of the FGD system, which would involve flue gas reheating and the use of a gas-gas heat exchanger for energy recovery.

## **5.0 PERFORMANCE MODELS**

In this section, analytical performance models of SCR systems are presented. These include performance models for both conventional coal-fired power plants and gas turbine-based systems. Gas turbine-based SCR systems may be required for IGCC applications, whereas SCR systems adapted from coal-fired power plants may be required for EFCC applications.

SCR systems for conventional power plants may be divided into high-dust, hot-side and tail-end, low-dust SCR systems. For the hot-side system, downstream effects on the power plant air preheater are modeled. For the tail-end system, a gas-gas heat exchanger and duct burner used for flue gas reheat are modeled.

### **5.1 Catalyst Requirement**

The catalyst requirement is a complex function of the physical and chemical properties of the catalyst, catalyst geometry, catalyst replacement philosophy, reaction temperature, flue gas volumetric flowrate, flue gas characteristics such as  $\text{NO}_x$  concentration, ash concentration, ash composition,  $\text{SO}_2$  concentration, gaseous poisonous species concentration (e.g.,  $\text{As}_2\text{O}_3$ ), the desired  $\text{NO}_x$  removal efficiency, allowable flue gas pressure drop, and the desired ammonia slip. The latter in turn affects downstream precipitation of solids, such as ammonium sulfate and bisulfate.

Although a catalyst requirement model ideally would be sensitive to all of the above factors, insufficient data are available to support the development of such a model. For example, the interactive poisoning effects of multiple flyash constituents is not well-understood. Therefore, the approach taken here is to develop a model of intermediate detail that captures the key functional dependencies between catalyst requirement and process conditions. The model is based on empirical and design assumptions supplied by the user and power plant performance parameters calculated from the power plant performance model.

A number of theoretical models were reviewed as a possible basis for model development. In most cases, these models were not adopted directly here, but were used to identify key functional dependencies that could be modeled based on empirical data.

#### **5.1.1 Factors Affecting Catalyst Requirement**

In a report prepared by Shiomoto and Muzio (1986), there is an appendix containing comments by Kawasaki Heavy Industries, a Japanese manufacturer of SCR

systems. KHI presents the development of a simplified equation for estimating catalyst space velocities based on NO<sub>x</sub> removal efficiency, ammonia slip, linear velocity, reaction temperature, and catalyst activity. The functional form of this equation is a basis for the performance model developed here.

The simplest model of an SCR system for the purpose of determining catalyst requirement is based on a rate model for the chemical reaction of NO with NH<sub>3</sub>, which is the predominate reaction occurring in the SCR reactor. Under the condition of an NH<sub>3</sub>/NO<sub>x</sub> molar ratio of 1.0, pilot plant testing in Japan has shown that a first order reaction occurs:

$$\frac{d[\text{NO}_x]}{dt} = -k[\text{NO}_x] \quad (15)$$

Integration of this equation yields the following relation between flue gas residence time in the catalyst and NO<sub>x</sub> removal efficiency:

$$t = \frac{-\ln(1-\eta_{\text{NO}_x})}{k} \quad (16)$$

The catalyst space velocity is related to flue gas residence time in the catalyst by the following:

$$\text{SV (1/hr)} = \frac{3,600 \text{ sec/hr}}{t \text{ (sec)}} \quad (17)$$

Therefore, the catalyst space velocity is given by:

$$\text{SV} = \frac{3600 k}{-\ln(1-\eta_{\text{NO}_x})} \quad (18)$$

The rate constant,  $k$ , corresponds to the so-called catalyst "activity." The catalyst activity is a complex function of catalyst geometry, chemical formulation, and operating conditions.

In the typical case where the NH<sub>3</sub>/NO<sub>x</sub> molar ratio is less than 1.0, the apparent catalyst activity will be less than the actual catalyst activity. This is because the reaction between NH<sub>3</sub> and NO<sub>x</sub> is 1:1, while for molar ratios of less than 1, the active sites populated with ammonia molecules are fewer than the active sites sought by NO<sub>x</sub> molecules. This is not a limitation of the catalyst, but rather a limitation due to the scarcity of NH<sub>3</sub>.

To adjust for this phenomena, KHI developed an empirical correction factor based on the "end-mole ratio," which is the NH<sub>3</sub>/NO<sub>x</sub> molar ratio at the SCR reactor exit. The

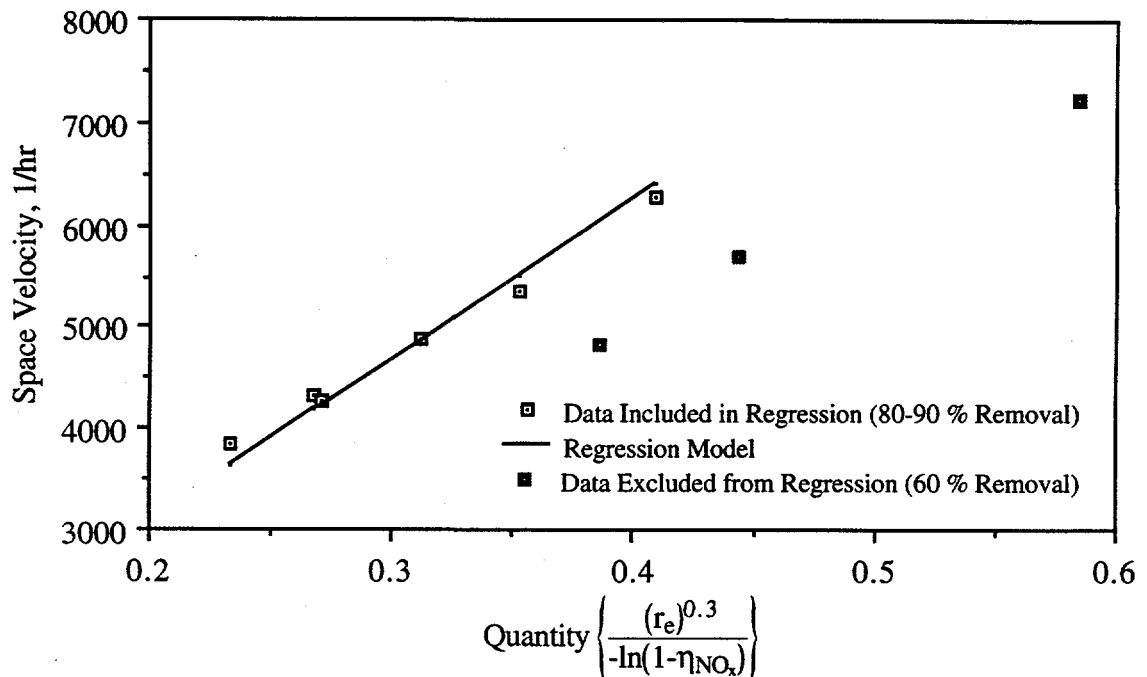


Figure 10. Regression Model of Space Velocity as a Function of NO<sub>x</sub> Removal Efficiency and End-Mole Ratio.

end-mole ratio is the molar ratio of ammonia slip to unreacted NO<sub>x</sub>. Thus, Equation (18) becomes:

$$SV = \frac{3600 C (r_e)^a}{-\ln(1-\eta_{NO_x})} \quad (19)$$

The value of the exponent "a" is obtained empirically from test data obtained for constant conditions except for changes in NO<sub>x</sub> removal efficiency and end-mole ratio. Data plotted by KHI suggest that a typical value of "a" is 0.3.

Equation (19) was used as the basis for a regression model for space velocity based on design data provided by KHI in an earlier SCR design study published by EPRI (Bauer and Spendle, 1984). The purpose of the regression model was to determine the adequacy of the simplified model of Equation (19) for use in an SCR performance model. The results are shown graphically in Figure 10.

The regression analysis indicated that Equation (19) was valid for estimating space velocities for NO<sub>x</sub> removal efficiencies of 80 and 90 percent, with varying end-mole ratios. However, the model could not also be applied simultaneously to the data for 60 percent NO<sub>x</sub> removal. The results here suggest that the catalyst activities differ for the two sets of data. The design details, such as linear velocity, are not reported by Bauer and Spendle (1984); therefore it is possible that design conditions were not the same for the two sets of



data. Thus, it appears that the simple model of Equation (19) is a reasonable basis for correcting space velocity for differences in NO<sub>x</sub> removal efficiency and end-mole ratio when all other factors are held constant.

To compare catalysts under different operating conditions, KHI suggests the following model:

$$\frac{\ln(1-\eta_2)}{\ln(1-\eta_1)} = C_r H_r \beta_r f_r S_r Q_r \left(\frac{m_2}{m_1}\right)^{0.3} \quad (20)$$

From Equation (20), the performance of a specific catalyst may be estimated based on ratios of key performance parameters, assuming a reference catalyst. This formulation suggests that space velocity can be estimated based on a reference catalyst using multiplicative correction factors to adjust for differences in operating conditions.

Other models are possible, such as that reported by Chen et al (1991). The NO<sub>x</sub> removal efficiency is estimated in this model based on detailed design information regarding the SCR catalyst, including catalyst geometry, film mass transfer coefficient, effective diffusivity, and reaction rate constant. However, data to support this detailed model are often not reported in published literature.

### 5.1.2 Model Form

The modeling approach adopted here is to assume a reference catalyst and to apply a series of multiplicative correction factors to adjust space velocity for different design conditions. The general formulation is:

$$SV = SV_{ref} \prod_{i=1}^4 f_i \quad (21)$$

Each correction factor,  $f_i$ , is a ratio that reflects the difference in space velocity from the reference to design conditions due to differences in certain design parameters. If the reference and design conditions are the same, these correction factors have a value of unity. A total of four correction factors have been developed, based on: (1) NO<sub>x</sub> removal efficiency; (2) end-mole ratio; (3) catalyst activity; and (4) reaction temperature.

Default reference conditions are included in the model. The reference parameters required include space velocity, NO<sub>x</sub> removal efficiency, NO<sub>x</sub> inlet concentration, ammonia slip concentration, a catalyst activity curve, a catalyst life, and an operating temperature.

### 5.1.2.1 NO<sub>x</sub> Removal Efficiency

The correction factor for NO<sub>x</sub> removal efficiency is based on the model formulation suggested by KHI. This correction factor is:

$$f_1 = f_{\text{eff}} = \frac{\ln(1-\eta_{\text{ref}})}{\ln(1-\eta)} \quad (22)$$

As the NO<sub>x</sub> removal efficiency increases, the catalyst space velocity decreases, leading to a larger catalyst volume.

For tail-end SCR, a portion of the flue gas exiting the FGD system leaks across the gas-gas heat exchanger used for flue gas reheat. Therefore, a portion of the NO<sub>x</sub> in the flue gas will also pass across the heat exchanger and into the stack gas. Because of this, the NO<sub>x</sub> removal efficiency in the SCR unit must be increased to compensate for the NO<sub>x</sub> that bypasses the SCR system due to leakage. The required NO<sub>x</sub> removal efficiency for the SCR system is calculated based on the overall NO<sub>x</sub> removal efficiency required and the flue gas leakage rate across the gas-gas heat exchanger:

$$\eta = \frac{\eta_{\text{overall}}}{(1 - f_1)} \quad (23)$$

To satisfy this equation, the following constraint must be met:

$$\eta_{\text{overall}} + f_1 \leq 1 \quad (24)$$

This model assumes no additional NO<sub>x</sub> formation in the duct burner.

### 5.1.2.2 End-Mole Ratio

The correction factor for end-mole ratio is based on the design ammonia slip, the design inlet NO<sub>x</sub> concentration, and the design NO<sub>x</sub> removal efficiency. The end-mole ratio is given by:

$$r_e = \frac{[\text{NH}_3]_{\text{out}}}{[\text{NO}_x]_{\text{in}}(1 - \eta)} \quad (25)$$

The correction factor for end-mole ratio is given by:

$$f_2 = f_{r_e} = \left( \frac{r_e}{r_{e,\text{ref}}} \right)^{0.3} \quad (26)$$

### 5.1.2.3 Catalyst Activity

Experimental and commercial SCR operating data indicate that catalyst activity decreases with time, due to physical and chemical changes to the catalyst as previously discussed. The actual rate of catalyst activity deterioration depends on the operating

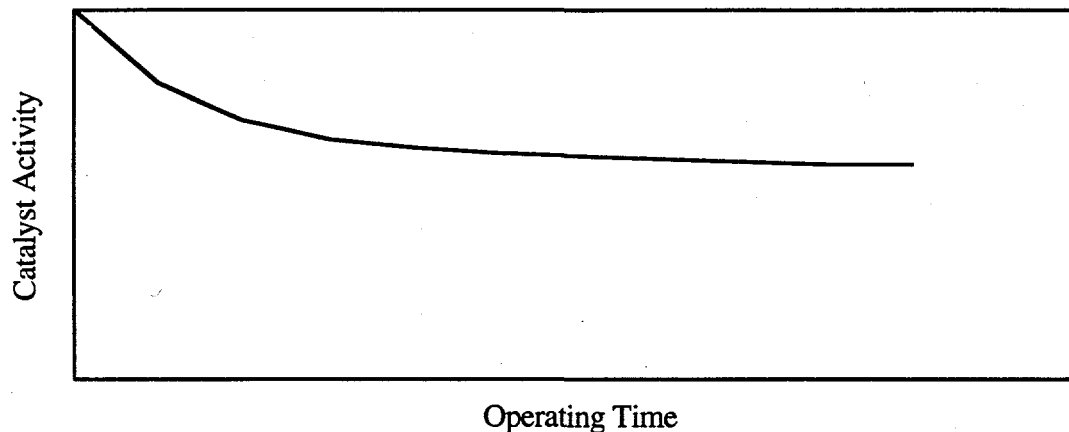


Figure 11. A Typical Catalyst Activity Curve.

conditions for the SCR reactor, and is usually not constant with time. Typically, there is an initial period of relative rapid catalyst deactivation, followed by a period of gradual activity change. A typical catalyst activity curve is shown schematically in Figure 11.

Schönbucher (1989) presents several curves for catalyst activity as a function of time for high-dust and tail-end SCR systems. For wet-bottom boilers with high-dust SCR, catalyst activity loss may range from 25 to 45 percent during the first 2,000 hours of operation, with less rapid changes during subsequent operation. For other applications, such as tail-end systems on wet or dry bottom boilers, catalyst activity loss is slight (e.g., 5 percent) over 10,000 hours of operation and appears to decrease at a relatively constant rate. Similar trends may be true of natural gas-fired gas turbine SCR systems.

To model catalyst activity loss, a simple function is employed to represent the catalyst activity curve. The purpose of this function is to provide a reasonable representation of the qualitative properties of catalyst activity loss for most cases. This function features two components: a minimum activity level and an exponential decay from the initial activity to the minimum activity levels. The initial activity is assumed to have a value of unity, while subsequent activity levels are relative to the initial activity.

$$A(t) = A_{\min} + A_{\text{inc}} \left\{ \exp\left(-\frac{t}{\tau_a}\right) \right\} \quad (27)$$

where:

$$A_{\min} + A_{\text{inc}} = 1 \quad (28)$$

A model user specifies the minimum activity level,  $A_{\min}$ , which determines the incremental initial catalyst activity,  $A_{\text{inc}}$ , that is subject to exponential decay. By also specifying one

data point on the activity curve (an activity level  $A(t_1)$  at time  $t_1$ ), the activity decay time constant,  $\tau_a$ , can be estimated:

$$\tau_a = - \frac{t_1}{\ln\left(\frac{A(t_1) - A_{\min}}{A_{\text{inc}}}\right)} \quad (29)$$

For example, suppose we have a catalyst with a long term activity level of 75 percent of the initial value, and for which the measured activity after 8,000 hours was 85 percent of the initial activity. Then:

$$\begin{aligned} A_{\min} &= 0.75 \\ A_{\text{inc}} &= 1 - A_{\min} = 0.25 \\ \tau_a &= - \frac{t_1}{\ln\left(\frac{A(t_1) - A_{\min}}{A_{\text{inc}}}\right)} = - \frac{8,000}{\ln\left(\frac{0.85 - 0.75}{0.25}\right)} = 8,730 \end{aligned}$$

and:

$$A(t) = 0.75 + 0.25 \left\{ \exp\left(-\frac{t}{8,730}\right) \right\}$$

This example is shown graphically in Figure 12. In addition, a case in which the minimum activity level is assumed to be zero is also shown, to illustrate the flexibility of Equation (27) for representing catalyst activity curves.

In typical SCR systems, multiple catalyst layers are employed. Furthermore, these catalyst layers are generally not replaced simultaneously. The overall catalyst relative activity in the case of multiple layers is the average of the individual catalyst layer relative activities (e.g., Nakabayashi and Abe, 1987). If we have  $N_c$  identical catalyst layers, and if each layer has been on-line for  $t_i$  hours at time  $t$ , then the average catalyst relative activity at any time  $t$  is given by:

$$A_{\text{avg}}(t) = \frac{1}{N_c} \sum_{i=1}^{N_c} A(t_i) \quad (30)$$

However, we are usually interested in the activity at the design point of the catalyst, which corresponds to the activity level at the end of a catalyst layer replacement cycle. If we have  $N_c$  identical catalyst layers that are replaced one-at-a-time every  $t_r$  hours, the catalyst relative activity at the design point is:

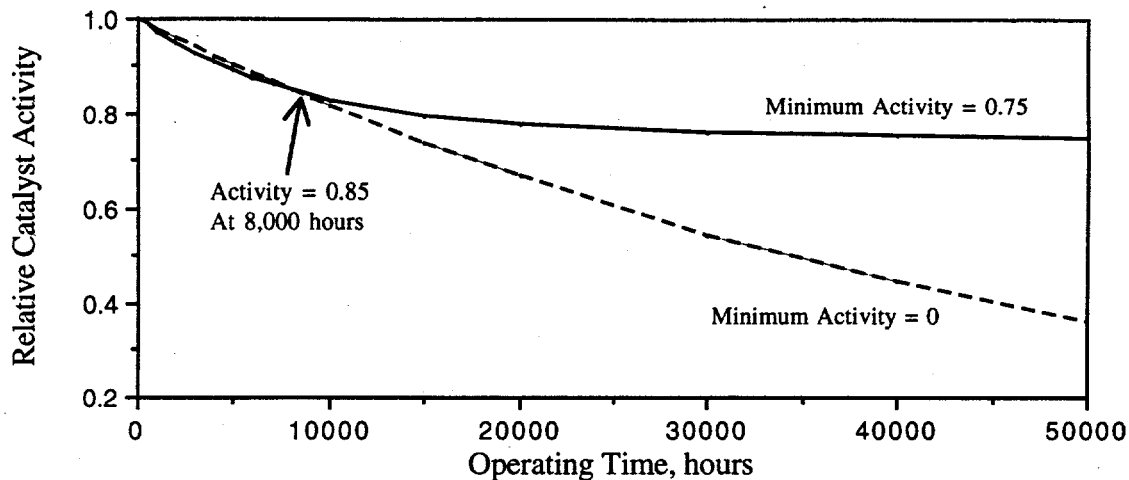


Figure 12. An Example Catalyst Activity Curve.

$$A_{des} = \frac{1}{N_c} \sum_{i=1}^{N_c} A(i \cdot t_r) \quad (31)$$

The implications of Equations (30) and (31) are illustrated in Figure 13. In this figure, the instantaneous average catalyst activity of a three-layer catalyst is illustrated for two cases. The first case, shown in a solid line, assumes that one catalyst layer is replaced every 10,000 hours, with a total time between complete catalyst replacements of 30,000 hours. The second case, shown as a dotted line, assumes that all catalyst layers are replaced simultaneously every 20,000 hours. For this particular illustrative case study, the design activity levels of both schemes are approximately the same. However, it is clear that by replacing individual layers sequentially, rather than simultaneously, the effective catalyst

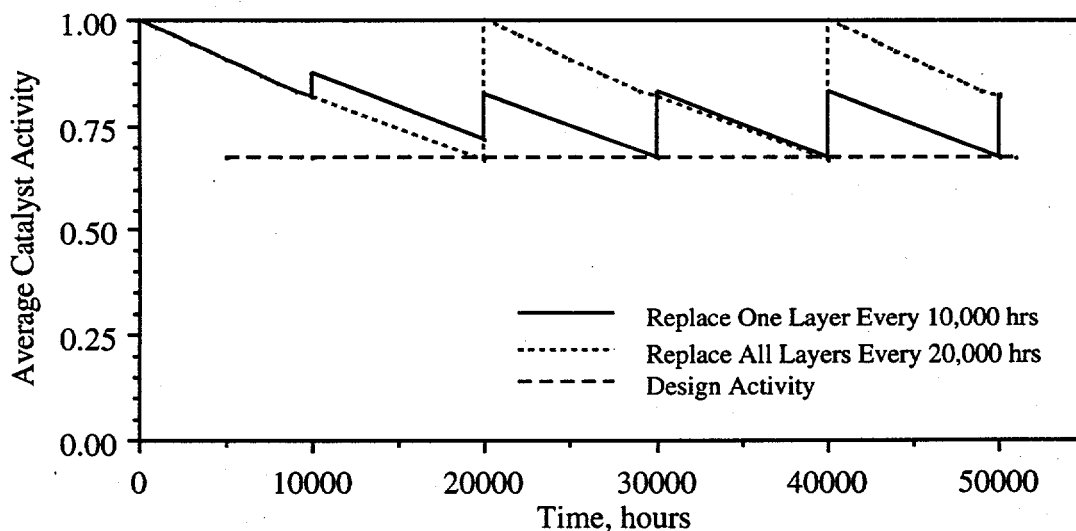


Figure 13. Illustrative Example of Average Catalyst Activity for a Three Layer Catalyst.

"life" is increased for a given volume of catalyst. In the illustrative example, catalyst life is 50 percent greater for the sequential replacement scheme compared to the simultaneous replacement scheme.

In the more general case, sequential replacement schemes may lead to larger initial catalyst volumes in order to achieve the same design activity as a simultaneous replacement scheme. For example, if we simultaneously replace all three catalyst layers in the illustrative problem every 10,000 hours, the design activity level would be approximately 80 percent. To achieve this design activity level with a sequential replacement scheme of one layer every 10,000 hours, the initial catalyst charge would need to be 20 percent larger, which increases capital costs. However, the annual costs associated with catalyst replacement would be reduced by 60 percent, because in the sequential scheme the same amount of catalyst is replaced in 24,000 hours as is required every 10,000 hours in the simultaneous replacement scheme. Thus, selection of a catalyst "life" design value involves trade-offs between capital and annual costs.

Another catalyst charging and replacement scheme involves using an initial charge less than the ultimate design catalyst quantity. For example, three catalyst layers might be used initially, with a fourth layer added at a later time. Then, the four catalyst layers are replaced sequentially similar to the previous case. This example is shown in Figure 14. The example has a design relative activity slightly less than that of the case in Figure 13, which implies that a larger overall catalyst volume is required to achieve the same actual design activity. However, the initial catalyst charge and the periodic catalyst replacement rate are nearly 20 percent less than that for the previous case. This system achieves a better utilization of catalyst. However, a disadvantage of this approach is an increased flue gas pressure drop across the reactor at the design point, due to the requirement for approximately 10 percent additional total catalyst charge compared to the previous case.

For the purpose of estimating a catalyst space velocity based on a reference data point, recall from Equation (18) that space velocity is directly proportional to catalyst activity. Therefore, the correction factor for space velocity due to differences in catalyst activity and catalyst replacement schedules is given by:

$$f_3 = f_A = \frac{A_{des}}{A(t_{ref})} \quad (32)$$

The design activity is calculated using Equation (31) based on the total number of layers to be included in the steady-state catalyst charge (i.e. including layers added to the

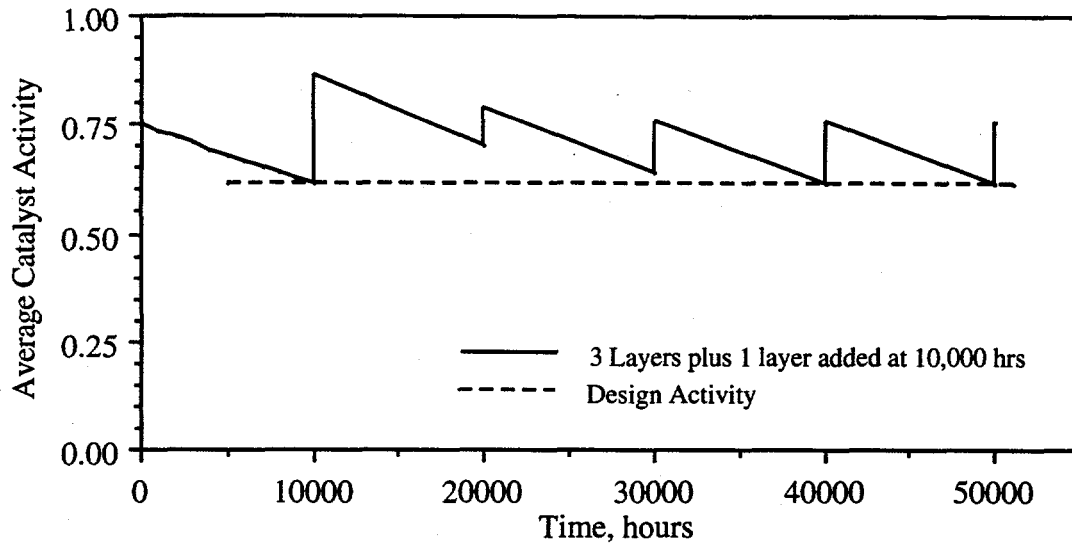


Figure 14. Illustrative Example of A Catalyst Addition and Replacement Scheme

initial charge at a later time). The reference activity level is estimated assuming that the entire catalyst charge is replaced simultaneously at time  $t_{ref}$ .

#### 5.1.2.4 Temperature

A temperature correction is assumed based on the Arrhenius relation for the reaction rate constant, which is given by:

$$k = k_0 \exp\left(\frac{-E_a}{RT}\right) \quad (33)$$

A regression analysis using data from Bauer and Spendle (1984) was done to determine a value for the quantity  $E_a/R$  for use in the performance model. This was accomplished by rewriting Equation (19) as:

$$C = k_0 \exp\left(\frac{-E_a}{RT}\right) = \frac{-\ln(1-\eta_{NO_x})(SV)}{3600(r_e)^a} \quad (34)$$

The quantity  $C$  was calculated using data presented in graphical form for  $NO_x$  removal efficiency as a function of temperature for a given ammonia slip, space velocity, and inlet  $NO_x$  concentration. From the regression analysis, the quantity  $E_a/R$  was estimated to be 7180, with an  $R^2$  of 0.77 for 18 data points. Thus, the correction factor to adjust space velocity for differences in temperature is given by:

$$f_4 = f_T = \frac{\exp\left(\frac{-7180}{T}\right)}{\exp\left(\frac{-7180}{T_{ref}}\right)} \quad (35)$$

### 5.1.2.5 Initial and Annual Catalyst Volume Requirement

The SCR catalyst requirement is calculated based on the space velocity and flue gas volumetric flow rate. In cases where there are no later additions of catalyst layers, the catalyst volume is constant throughout the life of the plant. However, some designs assume that a new catalyst layer is added after some time period. In this case, the space velocity will decrease when the new layer is added. Thus, the total number of active catalyst layers may consist of catalyst layers existing at plant start-up and additional reserve layers added afterwards:

$$N_C = N_I + N_R \quad (36)$$

The initial catalyst volume is given by:

$$V_{c,i} = \frac{G_{FG}(N_I)}{SV(N_C)} \quad (37)$$

The total catalyst volume is similarly given by:

$$V_c = \frac{G_{FG}}{SV} \quad (38)$$

The annual catalyst replacement rate depends in part on the catalyst design. In some studies, it is assumed that all catalyst layers are replaced simultaneously, while others assumed a phased approach to catalyst layer replacement. The number of catalyst layers replaced at the end of each replacement interval is given by:

$$N_{C,r} = 1 \text{ or } N_C \quad (39)$$

If only one layer is replaced at a time, then  $N_{C,r}$  equals one. If all layers are replaced simultaneously, it equals the total number of layers in the catalyst. The number of layers replaced per year is:

$$N_{C,a} = \frac{N_{C,r}(8,760 \text{ hr/yr})c_f}{t_r} \quad (40)$$

Thus, the catalyst volume replaced per year is:

$$V_{C,a} = V_c \left( \frac{N_{C,a}}{N_C} \right) \quad (41)$$

The catalyst "life" can be calculated based on the catalyst layer replacement interval and the number of layers replaced at the end of each interval:

$$L_C = \frac{(N_C/N_{C,r})t_r}{(8,760 \text{ hr/yr})c_f} \quad (42)$$



In the case where all catalyst layers are replaced simultaneously, the catalyst life is the same as the catalyst replacement interval. Of course, the total volume of catalyst associated with each replacement scheme will differ.

## 5.2 Ammonia Requirement

The ammonia requirement is primarily a function of  $\text{NO}_x$  removal efficiency and the ammonia slip. The ammonia slip depends on the catalyst formulation and space velocity. However, it is treated here as a model input because insufficient data are currently available to develop a model of ammonia slip as a function of other variables. For each mole of NO and  $\text{NO}_2$  that reacts in the SCR system, one mole of  $\text{NH}_3$  is required (see Equations (3) and (5)). An excess amount of ammonia is required due to limitations related to diffusion of ammonia and  $\text{NO}_x$  to the catalyst's active sites. Typically, this excess ammonia leaves the SCR system unreacted. Thus, given a specified  $\text{NO}_x$  removal efficiency and ammonia slip, the molar ratio of ammonia to inlet  $\text{NO}_x$  is given by:

$$R_A = \eta + \frac{[\text{NH}_3]_{\text{out}}}{[\text{NO}_x]_{\text{in}}} \quad (43)$$

The ammonia mass flow requirement is then given by:

$$M_{\text{NH}_3, \text{in}} = R_A \cdot M_{\text{NO}_x, \text{in}} \quad (44)$$

Ammonia is stored as a liquid. Many design studies assume that the ammonia is vaporized by mixing it with steam prior to injection into the flue gas. A typical minimum ratio of steam to ammonia is approximately 8, based on the use of medium pressure saturated steam. However, for safety reasons, ammonia dilution to 5 volume percent may be required, leading to a requirement for a steam-to-ammonia ratio of 19. The steam requirement for ammonia injection is given by:

$$M_{\text{steam}} = R_S \cdot M_{\text{NH}_3, \text{in}} \quad (45)$$

## 5.3 $\text{SO}_3$ Oxidation Rate

A portion of the  $\text{SO}_2$  in the flue gas entering the SCR reactor is oxidized to  $\text{SO}_3$ . The percentage of  $\text{SO}_2$  oxidized depends primarily on the reaction temperature, catalyst space velocity, and catalyst formulation. Regression analysis was used to develop two models representative of catalysts formulation for high sulfur and low sulfur operating environments. Data for both of these analyses were obtained from Bauer and Spendle (1984). For the high sulfur catalyst, which would be employed in high-dust hot-side

applications for a conventional coal-fired power plant, the fraction of SO<sub>2</sub> oxidized to SO<sub>3</sub>, based on 31 data points, is given by:

$$f_{\text{ox}} = 2.38 \times 10^{-13} \cdot \text{SV}^{-1.06} \cdot (\text{T}-460)^{5.03} \quad (46)$$

For the low sulfur catalyst, which would be employed in tail-end applications or for natural gas-fired gas turbine systems, the fraction is given by the following regression model based on 26 data points:

$$f_{\text{ox}} = 1.05 \times 10^{-13} \cdot \text{SV}^{-0.996} \cdot (\text{T}-460)^{5.05} \quad (47)$$

The coefficient of determination, R<sup>2</sup>, for both of these regression models exceeds 0.99. SO<sub>2</sub> oxidation increases as space velocity decreases and as temperature increases.

## 5.4 Downstream Effects

Ammonia slip and SO<sub>3</sub> exiting the SCR system can combine to form ammonium sulfate and bisulfate, as previously discussed. Also, in conventional coal-fired power plants, ammonia may be captured by fly ash prior to collection in the ESP. These downstream effects are of concern primarily for the hot-side SCR applications and for gas turbine-based systems firing fuels with a significant sulfur content.

### 5.4.1 Downstream Heat Exchangers

The formation of ammonium salts is treated here empirically. Of primary concern is the amount of ammonia associated with ammonium salts deposited in the downstream heat exchanger surfaces (e.g., air preheater for a coal-fired power plant, HRSG for a gas turbine combined cycle). The fraction of ammonia slip that is deposited as ammonium salts in the air preheater is treated as a parameter in the model, rather than as a calculated variable. The fraction of ammonia that is absorbed onto flyash is also a parameter in the model. The remaining portion of the ammonia slip is assumed to exit the plant with the flue gas leaving the stack. The ammonia partitioning coefficients must satisfy the following condition:

$$f_{\text{NH}_3, \text{dep}} + f_{\text{NH}_3, \text{abs}} + f_{\text{NH}_3, \text{out}} = 1 \quad (48)$$

The molar flow rate of ammonia that is deposited as a solid in the air preheater is given by:

$$M_{\text{NH}_3, \text{dep}} = f_{\text{NH}_3, \text{dep}} \cdot \frac{[\text{NH}_3]_{\text{out}}}{[\text{NO}_x]_{\text{in}}} \cdot M_{\text{NO}_x, \text{in}} \quad (49)$$

Similarly, the molar flows of ammonia absorbed by fly ash and emitted at the stack are given by:

$$M_{\text{NH}_3,\text{abs}} = f_{\text{NH}_3,\text{abs}} \cdot \frac{[\text{NH}_3]_{\text{out}}}{[\text{NO}_x]_{\text{in}}} \cdot M_{\text{NO}_x,\text{in}} \quad (50)$$

$$M_{\text{NH}_3,\text{out}} = f_{\text{NH}_3,\text{out}} \cdot \frac{[\text{NH}_3]_{\text{out}}}{[\text{NO}_x]_{\text{in}}} \cdot M_{\text{NO}_x,\text{in}} \quad (51)$$

The ammonia that deposits in the heat exchanger is removed periodically using water washing. As discussed by Neckar (1989), the concentration of ammonia in the wash water leaving the heat exchanger is initially high, and then gradually decreases. The waste water from the washing may be separated into high and low ammonia concentration streams, with the high ammonia concentration stream requiring denitrification pretreatment before entering the regular plant waste water treatment system. The model includes provision for specifying the average ammonia concentration of the high concentration fraction of the spent waste water, as well as the portion of the deposited ammonia that is removed by this portion of the water. The model also includes a parameter for the ammonia concentration in the "low concentration" wastewater. These parameters are used to estimate the heat exchanger wash water requirement, with the concentrations specified in units of mg/l. The "high concentration" wash water requirement, in gallons/hour, is given by:

$$m_{\text{wash,hc}} = \frac{M_{\text{NH}_3,\text{dep}} \cdot f_{\text{hc}}}{4.90141 \times 10^{-7} \cdot C_{\text{NH}_3,\text{hc}}} \quad (52)$$

and the low concentration wash water requirement is given similarly by:

$$m_{\text{wash,lc}} = \frac{M_{\text{NH}_3,\text{dep}} \cdot (1 - f_{\text{hc}})}{4.90141 \times 10^{-7} \cdot C_{\text{NH}_3,\text{lc}}} \quad (53)$$

#### 5.4.2 Catalyst Sootblowing

Catalyst sootblowing is required to remove ash that may mask or plug the catalyst. A sootblowing design by Bauer and Spendle (1984) is used as a basis to develop a model of the sootblowing steam requirement. The design basis includes steam sootblowing employing multiple sootblower sets. Bauer and Spendle report that the predicted steam requirement is 13,400 lb/hr for a total of approximately one hour per day, or an average of 31 lbmole/hr, for a hot-side application in a coal-fired power plant. There is no indication in the report that the steam requirement is a function of catalyst size, although such a relationship seems plausible. For example, Bauer and Spendle consider catalyst volumes

ranging from 10,000 to 30,000 ft<sup>3</sup> but apparently assume the same sootblowing steam requirement for all cases. It is assumed here that the steam requirement reported by Bauer and Spendle is typical of the base case catalyst design, which for the high-dust configuration had a catalyst volume of 16,146 ft<sup>3</sup>. Furthermore, the steam requirement is assumed to scale with catalyst volume, used here as a measure of catalyst size:

$$M_{\text{soot}} = 31 \left( \frac{SV}{16,146 \text{ ft}^3} \right) \quad (54)$$

This model does not account for any differences in catalyst masking or fouling rates associated with flue gas or fly ash characteristics. For gas turbine-based systems, such sootblowing may not be required as frequently as for conventional coal-fired power plants. Thus, this model will overestimate sootblowing requirements for most gas turbine-based systems.

### 5.5 Pressure Drop

For gas turbine based systems, the pressure drop associated with the SCR system includes pressure drops across: (1) the ammonia injection grid; (2) dummy catalyst layers for flow control (not typically included); (3) active catalyst layers; and (4) HRSG due to the build up of deposits. A typical pressure drop for a gas turbine-based SCR system ranges from 1.9 to 6.1 inches of water (May et al., 1991). We take 4 inches of water to be the nominal pressure drop. The total pressure drop, in inches of water, across the SCR system is:

$$\Delta P_{\text{scr}} = 4 \text{ inches H}_2\text{O} \quad (55)$$

For a hot-side SCR system, the flue gas pressure drop associated with the SCR system includes pressure drops across: (1) ductwork and ammonia injection grid; (2) dummy catalyst layers for erosion control; (3) active catalyst layers; and (4) air preheater due to build up of deposits. Each of these sources of pressure drop are treated as input parameters in the model. The total pressure drop, in inches of water, across the SCR system is:

$$\Delta P_{\text{scr}} = n \cdot \Delta P_{\text{cat}} + n_d \cdot \Delta P_{\text{dum}} + \Delta P_{\text{duct}} + \Delta P_{\text{aph, inc}} \quad (56)$$

An additional consideration for hot-side SCR systems is the effect of the reduction in flue gas side pressure on the air leakage through the air preheater. For hot-side SCR systems, a nominal increase in the leakage rate of 10 percent is assumed. This is a model input parameter.

For tail-end systems, which are downstream of the air preheater, there is no incremental pressure drop associated with solids deposition in the air preheater. However, there is a pressure drop associated with flue gas reheating. Therefore, the pressure drop for the gas-gas heat exchanger used for reheat must be included. This pressure drop must be shared between the SCR system and the FGD system, for which reheat is often required also. Thus, a new parameter is introduced which represents the fraction of the gas-gas heat exchanger pressure drop that is solely attributable to the SCR system.

$$\Delta P_{scr} = n \cdot \Delta P_{cat} + n_d \cdot \Delta P_{dum} + \Delta P_{duct} + \Delta P_{GGH}(1-f_{FGD}) \quad (57)$$

The pressure drop term for ducting includes any pressure drop associated with the duct burner. These values must be specified as model inputs.

## 5.6 Energy Penalties

Here, we separately consider the efficiency penalties for conventional coal-fired power plants versus gas turbine based systems. The former may be more applicable to advanced systems such as the EFCC, whereas the latter is of interest for IGCC systems.

### 5.6.1 Conventional Coal-Fired Power Plants

For conventional coal-fired power plants, the energy penalties for the SCR system include electricity and steam consumption. The largest source of energy use is the incremental electricity required by the induced draft fan to overcome the flue gas pressure drops associated with the SCR system. In addition, electricity is required for the ammonia injection system, primary to compress vaporized ammonia for injection into the flue gas. Steam is consumed for ammonia vaporization and injection and for sootblowing in the SCR reactor. The steam consumption is converted to an equivalent electricity energy penalty based on the difference in enthalpy between the steam and water at standard conditions, the mass flow of steam used, and the steam cycle heat rate.

The energy penalty associated with operation of the induced draft fan to overcome the SCR flue gas pressure drop is given by:

$$E_{scr,fan} = \left( 1.38 \times 10^{-6} \frac{\text{MW}}{\text{cfm-in H}_2\text{O}} \right) G_{scr,in} \Delta P_{scr} \quad (58)$$

The fan equation here assumes a fan efficiency of 85 percent. This equation also represents the energy penalty associated with a forced draft booster fan used in tail-end SCR systems.

The energy penalty associated with ammonia compression is calculated assuming a 100 psi differential compression with an 85 percent compression efficiency:

$$E_{\text{scr,NH}_3} = \left(3.47 \times 10^{-5} \frac{\text{MW}}{\text{lb/hr}}\right) \cdot \left(\frac{17.03 \text{ lb}}{\text{lbmole}}\right) \cdot M_{\text{NH}_3,\text{in}} \quad (59)$$

The energy penalty for steam use in the SCR system is calculated based on the steam mass flows, the enthalpy added to the steam by the steam cycle, and the steam cycle heat rate:

$$E_{\text{scr,steam}} = \frac{(M_{\text{steam}} + M_{\text{soot}}) \left(\frac{18 \text{ lb}}{\text{lbmole}}\right) \left(\frac{1,000 \text{ BTU}}{\text{lb}}\right) \left(\frac{\text{MW}}{1,000 \text{ kW}}\right)}{\text{HR}_s} \quad (60)$$

The total energy penalty for the high-dust SCR system is:

$$E_{\text{scr}} = E_{\text{scr,fan}} + E_{\text{scr,NH}_3} + E_{\text{scr,steam}} \quad (61)$$

For the tail-end SCR system, natural gas used to fire the duct burner represents an additional energy loss. Natural gas is a fuel that could be used to generate electricity. Therefore, the energy penalty is calculated based upon the gross plant heat rate to estimate the equivalent electricity energy penalty associated with natural gas firing:

$$E_{\text{scr,NG}} = \frac{m_{\text{NG}} \text{HHV}_{\text{NG}} \left(\frac{\text{MW}}{1,000 \text{ kW}}\right)}{\text{HR}_g} \quad (62)$$

The total energy penalty for the tail-end SCR system is:

$$E_{\text{scr}} = E_{\text{scr,fan}} + E_{\text{scr,NH}_3} + E_{\text{scr,steam}} + E_{\text{NG}} \quad (63)$$

### 5.6.2 Gas Turbine-Based Power Plants

For gas turbine systems, the energy penalties for the SCR system include electricity consumption for ammonia compression, steam consumption for ammonia vaporization, and the energy penalty due to increased backpressure on the gas turbine. The electricity requirement for ammonia vaporization is given by Equation (59). The energy penalties associated with steam use and gas turbine backpressure are calculated by the ASPEN process flowsheet model.

### 5.7 Flue Gas Reheat for Tail-End SCR

Tail-end SCR may be required for the EFCC. For the tail-end SCR system, the temperature of the flue gas exiting the FGD system must be raised to the reaction temperature required by the SCR system. Furthermore, the temperature of the gas exiting the SCR system can be reduce prior to entering the stack. Therefore, a gas-gas heat exchanger (GGHX) and a duct burner are employed for heating and cooling the flue gas, as shown in Figure 15.

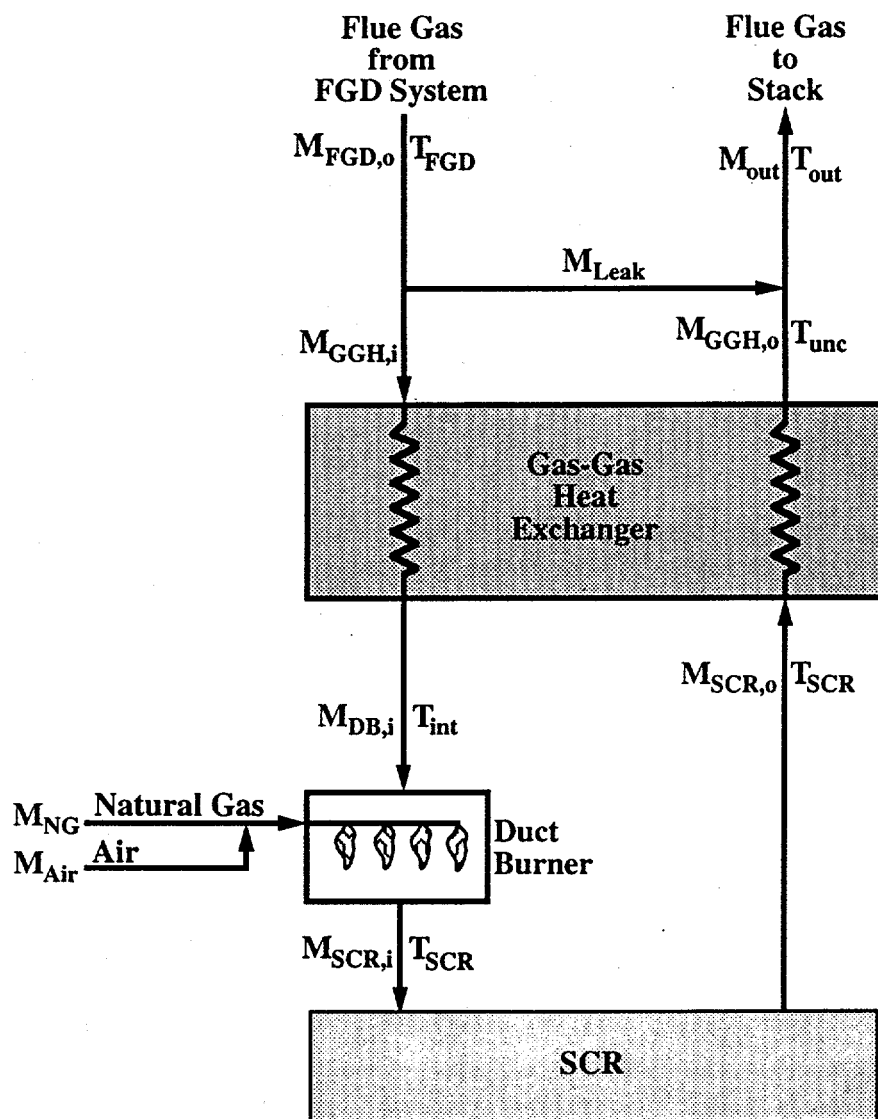


Figure 15. Schematic of Gas-Gas Heat Exchanger Performance Model

Typically, a Ljungstrum type heat exchanger would be used for the flue gas reheat and cooling system. A portion of the higher pressure flue gas entering from the FGD system will leak into the flue gas stream exiting the heat exchanger to the stack. Therefore, an air leakage stream is modeled. For simplicity, it is assumed that the leakage flue gas is at the FGD system exit temperature. The flue gas leakage mass flow is estimated based on a leakage fraction as follows:

$$M_{\text{leak}} = f_l M_{\text{FGD},o} \quad (64)$$

and the flue gas fraction entering the GGHX is given by:

$$M_{\text{GGH},i} = (1-f_l)M_{\text{FGD},o} \quad (65)$$

In the GGHX, the untreated flue gas is heated from the FGD exit temperature to an intermediate temperature by cooling the treated flue gas exiting the SCR system. The heat transferred across the heat exchanger is given by:

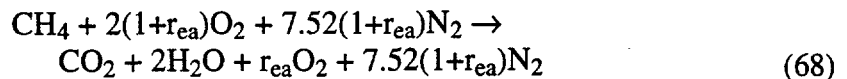
$$Q_{\text{GGH}} = M_{\text{GGH},i}(h_{fg}(T_{\text{int}}) - h_{fg}(T_{\text{FGD}})) \quad (66)$$

In the SCR computer model, the enthalpy of the flue gas is estimated using thermodynamic property data in the ASPEN chemical process simulator (MIT, 1987).

The untreated flue gas must be heated an additional amount to reach the SCR reaction temperature. This additional heating is accomplished by use of a duct burner, which also introduces additional mass streams to the flue gas. The mass balance equation is:

$$m_{\text{SCR},i} = m_{\text{DB},i} + m_{\text{NG}} + m_{\text{air}} \quad (67)$$

Assuming that natural gas consists only of methane, the combustion reaction is:



For each pound of methane consumed, 3.6 pounds of air are consumed at stoichiometric conditions. If excess air is also considered, the air mass flow is then given by:

$$m_{\text{air}} = 3.6(1+r_{\text{ea}})m_{\text{NG}} \quad (69)$$

The energy balance equation for the duct burner may be written as:

$$\begin{aligned} m_{\text{SCR},i}h_{fg}(T_{\text{SCR}}) &= m_{\text{DB},i}h_{fg}(T_{\text{int}}) + m_{\text{NG}}h_{\text{NG}}(T_{\text{NG}}) + \\ &3.6(1+r_{\text{ea}})m_{\text{NG}}h_{\text{air}}(T_{\text{air}}) + m_{\text{NG}}\Delta H_r \end{aligned} \quad (70)$$



The heat of reaction for methane is 345,700 BTU/lbmole. In Equation (71), all variables are specified except for the natural gas mass flow rate. Therefore, this equation can be used to solve for the natural gas requirement:

$$m_{NG} = \frac{m_{SCR,i}h_{fg}(T_{SCR}) - m_{DB,i}h_{fg}(T_{int})}{h_{NG}(T_{NG}) + 3.6(1+r_{ea})h_{air}(T_{air})v + \Delta H_r} \quad (71)$$

In the SCR system, ammonia and steam are added to the flue gas, and the flue gas composition changes due to chemical reactions occurring in the reactor vessel. The SCR exit temperature is assumed to be the same as the inlet temperature. The flue gas exiting the SCR unit is cooled in the GGHX. The uncorrected temperature of the treated flue gas exiting the GGHX is estimated from the following relationship:

$$h_{fg}(T_{unc}) = h_{fg}(T_{SCR}) - \frac{Q_{GGH}}{m_{GGH,o}} \quad (72)$$

This equation is solved using an iterative technique to determine the flue gas temperature.

The treated flue gas from the SCR system is mixed with the flue gas leaking across the GGHX prior to entering the stack. The flue gas temperature entering the stack, corrected for the thermal mixing with the leakage air, is estimated using the following equation:

$$h_{fg}(T_{out}) = \frac{m_{GGH,o}h_{fg}(T_{unc}) + m_{leak}h_{fg}(T_{FGD})}{m_{out}} \quad (73)$$

where:

$$m_{out} = m_{GGH,o} + m_{leak} \quad (74)$$

## 6.0 SCR CAPITAL COST MODELS

In this section, analytical models of direct, indirect, and total capital costs are developed for SCR systems applied to both coal-fired power plants and gas turbine-based systems. The next section focuses on capital cost models for conventional coal-fired power plants, including both high dust and tail end SCR configurations. Section 6.2 presents a capital cost model for SCR applied to gas turbine-based systems. Both types of systems include as major components: (1) reactor housing; (2) catalyst; and (3) ammonia storage and injection.

### 6.1 Hot-Side and Tail-End SCR Systems

Here, the capital cost models for hot-side and tail-end SCR systems for conventional coal-fired power plants are presented. These models may be useful for some types of advanced coal-based power generation systems, such as the EFCC.

#### 6.1.1 Reactor Housing

A recent EPRI study is used here as a basis for developing a reactor cost model (Robie and Ireland, 1991). A total of 14 cost estimates are reported. However, two of these are for oil-fired power plants. All of the remaining 12 data points are for coal-fired power plants. The two oil-fired SCR applications appear to have substantially higher costs than for the coal-fired systems of similar catalyst volume; however, the basis for the difference is not reported. Of the dozen data points for coal-fired systems, one value is a duplicate. Therefore, 11 data points were used to develop a regression model of reactor housing cost versus total catalyst volume. The reactor housing includes flanged gas inlet and outlet, a single vertical downflow reactor, casing, ash hoppers, structural supports for catalyst modules, rectifying plate, baffles, turning vanes, walkways, stairs, monorails, hoists, and sootblowers. The catalyst volume includes both active and spare catalyst layers.

Robie and Ireland (1991) did not report the actual catalyst volumes for all 11 data points used in the regression model. Therefore, the catalyst volume was estimated based on the reported flue gas mass flow and flue gas molecular weight (which were used to calculate flue gas volumetric flow rate), active catalyst space velocity, and the ratio of the number of active plus spare catalyst layers to the number of active layers. Thus, the estimated total catalyst volume is given by:

$$V_{\text{TOT}} = \left( \frac{G_{\text{fg}}^{\circ}}{SV_{\text{act}}^{\circ}} \right) \left( \frac{N_{\text{act}} + N_{\text{sp}}}{N_{\text{act}}} \right) \quad (75)$$

where:

$$G_{fg}^o = \frac{\left(\frac{m_{fg}}{MW_{fg}}\right) R T_{fg}^o}{P_{fg}^o} \quad (76)$$

The space velocity is referenced to a standard temperature and pressure, which in this case is 32 °F and 1 atm. Therefore, the flue gas volumetric flow rate is calculated at standard temperature and pressure. The active catalyst space velocity is adjusted based on the ratio of total to active catalyst layers for the purpose of determining the total catalyst volume for the reactor housing. For the cases where the actual catalyst volumes were reported, these estimates were often in very close agreement and never diverged by more than five percent.

The cost data reported by Robie and Ireland (1991) are based on subcontract costs for the process area to which additional costs have been added. These additional costs are not documented or discussed in the EPRI report, but appear to have been applied consistently for every cost reactor cost estimate. They may reflect installation and integration costs not covered by the subcontract costs. The multiplier for these additional costs is a function of whether the SCR is for a new or retrofit application. The multipliers are substantially larger for retrofit applications, presumably reflecting site access difficulty and congestion impeding equipment installation. All reactor housing costs were normalized to a new plant basis for purposes of the regression analysis. This was done by estimating the multiplier between subcontractor cost and total direct cost for a new plant based on the two case studies for a new installation. The direct costs for the remaining cases were then estimated on a new installation basis by multiplying the subcontract costs with the new installation direct cost factor. The relationship between direct cost and the catalyst volume per reactor housing was evaluated using regression analysis.

The resulting regression model for the direct cost of the reactor housing is:

$$DC_R = 18.65 N_{R,TOT} \left(\frac{V_{TOT}}{N_{R,TOT}}\right)^{0.489} \left(\frac{PCI}{357.3}\right) \quad (77)$$

This model has a coefficient of determination ( $R^2$ ) of 0.94 and a standard error of \$169,000. The regression model is shown graphically in Figure 16. The costs are reported in December 1989 dollars, but may be adjusted to other years using the *Chemical Engineering Plant Cost Index*.

From a statistical perspective, this regression model has significant limitations. The high coefficient of determination is influenced by the wide separation between two groups

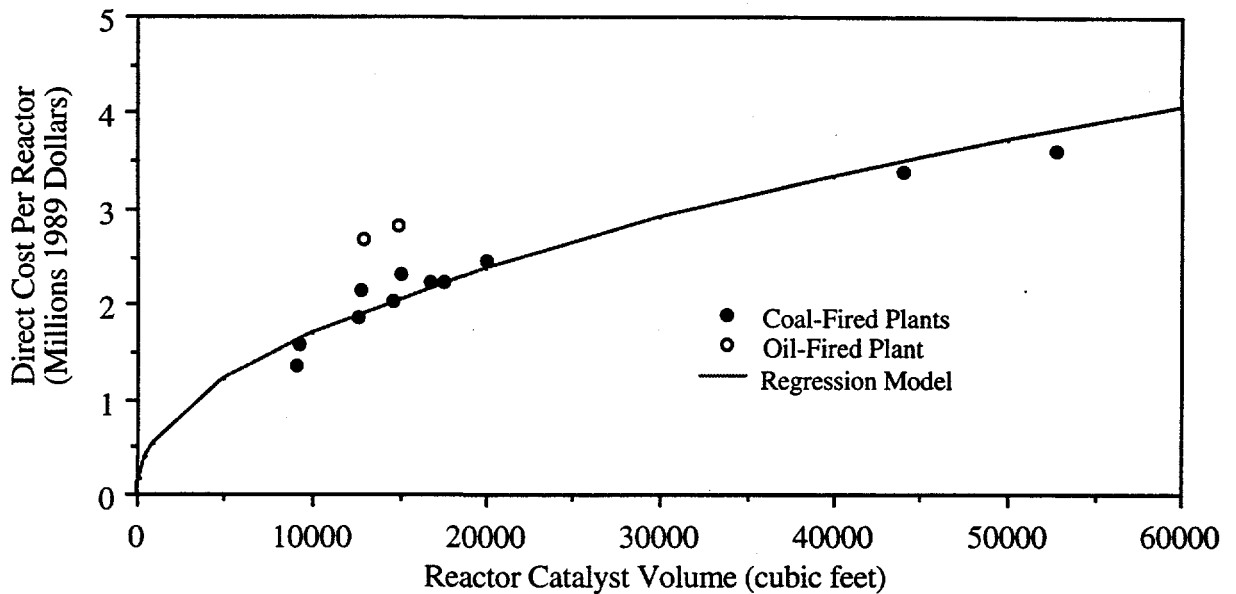


Figure 16. Direct Capital Cost Model for SCR Reactor Housing.

of data points. One set of data points are clustered for space velocities ranging from 10,000 to 20,000 cubic feet, whereas two other data points are close to 50,000 cubic feet. There is considerable scatter among the data points within the first cluster. However, because of the lack of reported detail regarding catalyst volumes and installation costs, it is not possible to reconcile these differences. Furthermore, there is no reported basis for the difference in values between the oil-fired and coal-fired data points. The oil-fired data points were excluded from the regression analysis, but are shown in Figure 16 for comparative purposes.

The regression model is satisfactory in representing the expected trend for SCR reactor housing cost. It is expected that there should be an economy of scale for increasing reactor housing size, and that the key measure of size is the total catalyst volume (including both active and spare catalyst layers).

### 6.1.2 Ammonia Handling and Injection

The direct cost for the ammonia injection process area was estimated based on analysis of eight data points taken from Robie and Ireland (1991). Six duplicate data points contained in their report were excluded from the statistical analysis. The ammonia unloading, storage, and supply system includes a horizontal bullet storage vessel with seven days supply capacity, an ammonia vaporizer, ammonia and dilution air mixer, ammonia injection grid, dilution air ductwork and dampers, and truck unloading station. The latter includes vapor recovery compressors, water deluge system, and transfer piping.

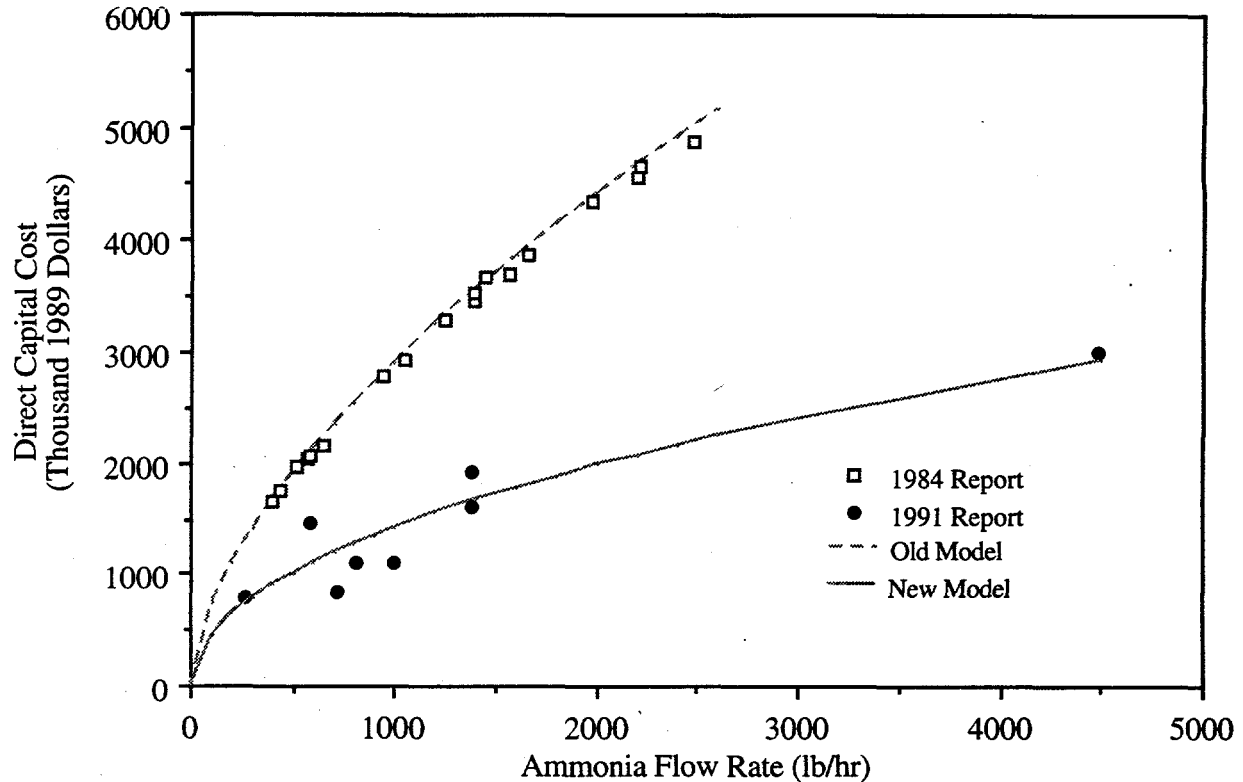


Figure 17. Direct Capital Cost Model for the Ammonia Handling and Injection System

The dilution air requirement is 20 parts air to one part ammonia. The regression model for ammonia process area direct costs is:

$$DC_{NH_3} = 50.8 (m_{NH_3,i})^{0.482} \left( \frac{PCI}{357.3} \right) \quad (78)$$

This model has a coefficient of determination ( $R^2$ ) of 0.87 and a standard error of \$285,000. The regression model is shown graphically in Figure 17. Also shown in the figure is a regression model developed from data reported in an earlier EPRI report (Bauer and Spendle, 1984). The newer model yields costs that are lower than the previous model by a factor of approximately two. The reasons for the difference are not immediately clear. The earlier report used generally more conservative assumptions, reflecting higher levels of uncertainty perceived at that time for this technology.

Like the reactor housing direct cost model, the coefficient of determination for the ammonia injection system direct cost model benefits from the separation between clusters of data points. In this case, however, there is only one data point at the high end of the range of values for the predictive variable, ammonia flow rate. However, the model is satisfactory in reflecting economies of scale for larger sized systems.

### 6.1.3 Ductwork

Ductwork costs are considered both for hot-side and tail-end SCR applications.

#### 6.1.3.1 Hot-Side SCR Applications

For a new hot-side SCR application, the ductwork associated with the SCR process includes economizer bypass ducts, economizer outlet duct, SCR inlet duct, SCR inlet control dampers, SCR outlet duct, SCR air preheater inlet plenum, various expansion joints in the ductwork, and dampers associated with the economizer bypass and air preheater cross-over ducting. Of the six major case studies reported by Robie and Ireland (1991), only two are for a new coal-fired power plant with hot-side SCR. Although these two case studies include five separate performance and cost estimates based on sensitivity analysis of key SCR performance characteristics, they are predicated on just two flue gas flow rates. Therefore, three of these estimates are duplicates. Because only two data points are available from this study to estimate duct costs as a function of flue gas flow, regression modeling was not employed to develop a direct cost model. Instead, a capacity-exponent model of direct cost versus flue gas volume flow rate was assumed as an appropriate function form. This formulation reflects the expected increase in cost that is associated with increases in flue gas volume flow rates. However, it implies that duct runs would be similar for differently sized systems. The parameters of this model, which include a multiplicative constant and an exponential coefficient, were estimated from the two data points. The resulting model is:

$$DC_D = 14.2 (G_{fg})^{0.7} \left( \frac{PCI}{357.3} \right) \quad (79)$$

The two data points used to estimate the parameters of this model are for flue gas flow rates of 3,026 and 2,713 macfm and direct costs of \$4.44 and \$4.10 million, respectively. This is a relative narrow range of values, but the resulting exponential scaling factor of 0.7 is consistent with scaling factors used in a variety of chemical engineering cost models.

#### 6.1.3.2 Tail-End SCR Applications

For an SCR in the cold-side application, there are additional duct costs associated with the gas-gas heat exchanger. There are two data points available for estimating these costs. One is for a gas-gas heat exchanger system which has 5.6 percent gas leakage. The second is for a system with no leakage. In this latter case, there is a higher gas flow rate through the SCR system, thereby leading to increased duct costs. Because of the scarcity of data points, the costs for these two cases are estimated as a multiplier of the costs for ductwork for a hot-side SCR application. The general equation is:

$$DC_{D,CS} = f_{D,CS} 14.2 (G_{fg})^{0.7} \left( \frac{PCI}{357.3} \right) \quad (80)$$

where  $f_{D,CS}$  is 1.90 for a GGH with 5.6 percent air leakage, and 2.15 for a GGH with no air leakage.

#### 6.1.4 Air Preheater Modifications

For hot-side SCR systems, a potentially significant concern is the deposition of ammonia-based compounds on downstream components. Unreacted ammonia exiting the SCR system ("ammonia slip") can react with sulfur trioxide present in the flue gas to form compounds such as ammonium bisulfate and ammonium sulfate that deposit on downstream equipment. These compounds may result in plugging and corrosion. The condensation of these compounds is most likely to occur on the cold and intermediate temperature heat transfer surfaces of the air preheater. In anticipation of deleterious effects, EPRI and others have recommended that air preheaters designed for use with SCR system be constructed with lower gauge (thicker) material, different geometries (e.g., combining intermediate and cold baskets of a conventional Ljungstrom air preheater into a single unit, to minimize seams where corrosion might occur), different materials, and larger motors, structure, and foundation to accommodate the larger weight of these modifications. Furthermore, more stringent provisions are made for soot blowing and water washing of the air preheater to remove the ammonia salts and any associated buildups at regular intervals. To enable an on-line water washing capability, crossover ducts and dampers are required. The changes in air preheater geometry and the effects of fouling can increase gas flow pressure drops. This in turn may increase gas leakage rates between the combustion air and flue gas sides of the heat exchanger.

Therefore, to accommodate the potential impacts of SCR systems on air preheater performance, a number of modifications are included in the design bases. Thicker material is used for the cold and intermediate heat transfer surfaces, in the region of the preheater where ammonia salt deposition is most likely. Furthermore, a corten steel alloy is used instead of carbon steel. A smoother heat transfer surface is used to aid in removing ammonia salts, but at the expense of reduced heat transfer and, hence, large heat transfer surface area. A larger motor is provided for the rotating Ljungstrom heat exchanger. Because of the additional weight of the heat exchanger, additional foundation and structural steel expense is incurred. High pressure steam sootblowers are installed in the cold-end of the heat exchanger. Water wash spray nozzles are also employed for on-line washing.

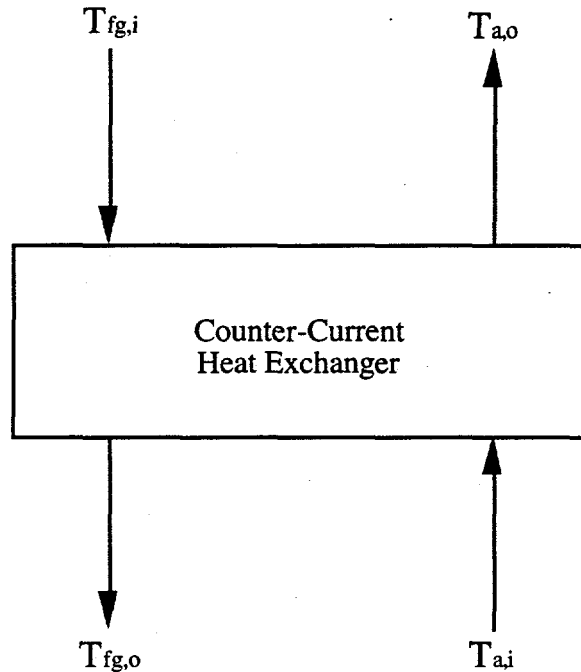


Figure 18. Simplified schematic of a counter-current flow heat exchanger.

Because of the increased heat transfer surface, there is a larger flue gas pressure drop which results, in turn, in a higher air leakage rate across the air preheater.

The costs of the major portions of the air preheater modifications, such as the increase in the heat transfer surface area and the associated increase in costs for special materials and increased structural support, are proportional to the size of the air preheater. Therefore, a cost model was developed for which the key parameter is a measure of the size of the air preheater.

For a counter-current heat exchanger (see Figure 18), the heat transfer is given by:

$$Q_{\text{aph}} = U_{\text{aph}} A_{\text{aph}} \Delta T_{\text{LM,aph}} \quad (81)$$

where

$$\Delta T_{\text{LM,aph}} = \frac{(T_{\text{fg,i}} - T_{\text{a,o}}) - (T_{\text{fg,o}} - T_{\text{a,i}})}{\ln \left( \frac{T_{\text{fg,i}} - T_{\text{a,o}}}{T_{\text{fg,o}} - T_{\text{a,i}}} \right)} \quad (82)$$

is the log-mean temperature difference (LMTD). The product of the universal heat transfer coefficient,  $U$ , and the heat exchanger surface area,  $A$ , is assumed here to be constant for an air preheater before and after modification for use with SCR. Prior to modification, the air preheater has a higher heat transfer coefficient and a lower surface area than after



modification. However, the heat exchanger is designed in either case to accommodate the same inlet and outlet conditions and, hence, the same LMTD. The product UA is calculated based on the known flue gas and air inlet and outlet temperatures, the flue gas molar flow rate, and the average specific heat of flue gas. A typical value for the latter is 7.9 BTU/(lbmole-°R). Thus,

$$(UA)_{\text{aph}} = \frac{q_{\text{aph}}}{\Delta T_{\text{LM,aph}}} \quad (83)$$

$$q_{\text{aph}} = M_{\text{fg}} c_{p,\text{fg}} (T_{\text{fg,o}} - T_{\text{fg,i}}) \quad (84)$$

There are only two data points from the EPRI study which are directly relevant to estimating a reference basis for the product UA. These are the two cases involving installation of a new SCR system involving Ljungstrom air preheaters (Cases 1.0 and 2.0 in the EPRI report). Other cases reported are for modifications to shell and tube heat exchangers or for retrofit modification to Ljungstrom heat exchangers.

For the first case (Case 1.0), the inlet and outlet air temperatures are not given. The amount of heat transfer is estimated based on the flue gas cooling from 725 °F to 270 °F. Assuming an inlet air temperature of 80 °F, and accounting for air leakage across the air preheater, the boiler air outlet temperature is estimated to be approximately 600 °F. Using these values, the LMTD is estimated to be 155 °F and the amount of heat transferred through the heat exchanger is 680 million BTU/hr. Therefore, the UA product in this case is  $4.4 \times 10^6$  BTU/°F. Using a similar approach for Case 2.0 yields a UA product of  $2.3 \times 10^6$ . In this latter case, the reported primary air temperature (air preheater air-side outlet temperature) is inconsistent with the amount of heat transfer obtained from flue gas cooling, assuming an inlet air temperature of 80 °F. Therefore, an independently calculated value of approximately 510 °F was used.

The costs of air preheater modifications for the two cases are \$1.37 million and \$0.81 million, respectively (in December 1989 dollars). Assuming that the cost of the modification is proportional the UA product, then from these two cost estimates the following capacity-exponent cost model is obtained:

$$DC_{\text{aph,mod}} = 1370 N_{T,\text{aph}} \left( \frac{UA_{T,\text{aph}}}{4.4 \times 10^6 N_{T,\text{aph}}} \right)^{0.8} \left( \frac{\text{PCI}}{357.3} \right) \quad (85)$$

While this model is based on only two data points, it nonetheless appears to provide a qualitatively reasonable relationship between air preheater modification costs based on the

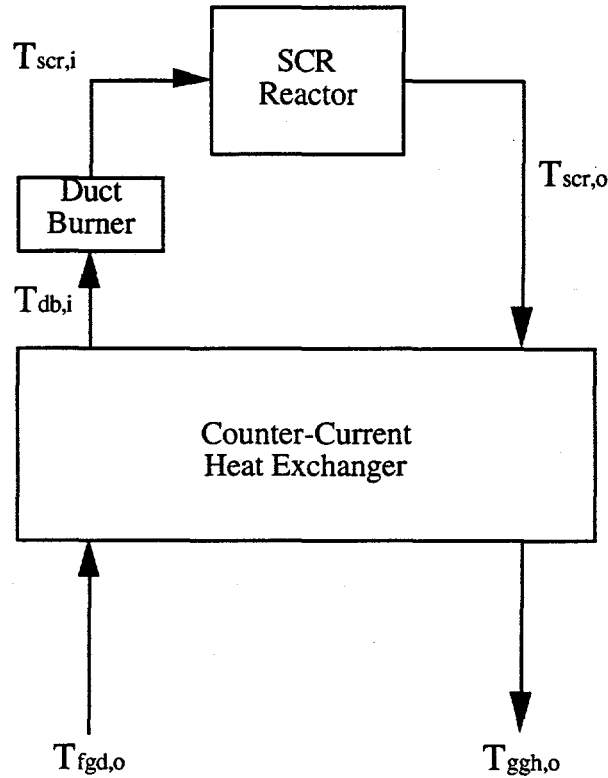


Figure 19. Simplified Schematic of the Gas-Gas Heat Exchanger Employed in Cold-Side SCR Systems.

size of the air preheater. The model suggests a modest economy of scale for modifications to larger air preheaters.

### 6.1.5 Gas-Gas Heat Exchanger

The cost model for the gas-gas heat exchanger (GGH) used for cold-side SCR applications was developed in a manner similar to that for air preheater modifications. The GGH is a Ljungstrom heat exchanger, and typically there is one GGH per SCR reactor. Thus, in a typical 500 MW power plant, there would be two GGHs.

A simplified schematic of the GGH is shown in Figure 19. Gas exiting the flue gas desulfurization (FGD) system is heated by counter-current heat exchange with high temperature gas exiting the SCR reactor. The "untreated" gas entering the SCR system is heated by a duct burner.

There is only one case study in the EPRI report by Robie and Ireland which deals with a cold-side SCR. The UA product (see Equation (10)) is assumed here as the key measure of heat exchanger size. Based on reported gas temperatures and flow rates, and correcting for gas leakage through the heat exchanger, the direct cost model is:

$$DC_{GGH} = 9100 N_{T,GGH} \left( \frac{UA_{T,GGH}}{4.4 \times 10^6 N_{T,GGH}} \right)^{0.8} \left( \frac{PCI}{357.3} \right) \quad (86)$$

This model is based on a GGH design with 5.6 percent gas leakage, on a mass basis. For a system with no leakage, multiply the cost given in Equation (86) by a factor of 2.4.

The cost of the duct burner is proportional to the amount of natural gas required to raise the flue gas temperature. The duct burners include combustion air fans, process controls, and a flame safeguard system. The capacity of the duct burners is expressed based on the heating value of the natural gas. The direct cost is given by:

$$DC_{DB} = 264 N_{T,DB} \left( \frac{Q_{NG}}{90 N_{T,DB}} \right)^{0.6} \left( \frac{PCI}{357.3} \right) \quad (87)$$

where

$$Q_{NG} = \frac{m_{NG} HHV_{NG}}{10^6} \quad (88)$$

#### 6.1.6 ID Fan and Booster Fan Costs

For a new SCR installation, the ID fans must be sized to deal with the increased flue gas pressure drop resulting from the additional ductwork and the SCR reactor. Typically, the increase in flue gas pressure drop is approximately 11 inches of water. The size of an ID fan and motor is proportional both to the flue gas flow rate and to the pressure drop. Therefore, the cost of the ID fan modifications is assumed here to be proportional to the difference in flue gas energy requirement necessary to overcome the flue gas pressure drop. This energy requirement is given by:

$$EC_{ID,dif} = \frac{\dot{Q}_{fg} \Delta P_{SCR}}{8,512 \eta_{fan}} \quad (89)$$

The fan efficiency is typically 85 percent. The cost of the ID fan differential is:

$$DC_{ID,dif} = 180 \left( \frac{EC_{ID,dif}}{4,600} \right)^{0.6} \left( \frac{PCI}{357.3} \right) \quad (90)$$

In the case of a cold-side SCR system, a booster fan is required to overcome the flue gas pressure drop throughout the GGH and SCR components. Typically, the pressure drop across the cold-side system is approximately 14.5 inches of water. Thus, using Equation (16) to estimate the energy requirement as a function of the actual flue gas

volumetric flow rate and the pressure drop, the cost of a cold-side SCR system booster fan is given by:

$$DC_{BF} = 810 N_{T,BF} \left( \frac{EC_{BF}}{3,000 N_{T,BF}} \right)^{0.6} \left( \frac{PCI}{357.3} \right) \quad (91)$$

There is typically one booster fan per SCR reactor train.

### 6.1.7 Structural Support

The basis for the SCR cost estimates developed by Robie and Ireland (1991) include a separate cost for structural support. While the definition of this process area is lacking, it appears to be related primarily to the SCR reactor housing, ductwork, and air preheater. In the case of a cold-side SCR system, the structural cost is related to the gas-gas heat exchanger, rather than the air preheater. As part of the data analysis, the structural cost was expressed as a percentage of the direct costs for the reactor housing, ductwork, and air preheater modifications for hot-side applications, and as a percentage of the direct costs for the reactor housing, ductwork, and gas-gas heat exchanger costs for cold-side applications. Of the 14 data points from the EPRI study, 12 of them yielded structural costs as a relatively narrow range of percentages of the appropriate direct costs. The mean value was 18.7 percent, with a range from 16.2 to 20.8 percent and a standard deviation of 1.4 percent. For one retrofit case with a high site congestion, the structural costs were 50 percent of the other selected direct costs. For a new SCR hot-side application, the structural costs are estimated as:

$$DC_s = f_s (DC_R + DC_D + DC_{APH}) \quad (92)$$

and for a cold-side application the structural costs are estimated as:

$$DC_s = f_s (DC_R + DC_D + DC_{GGH}) \quad (93)$$

where  $f_s$  has a mean value of 0.187. For a retrofit application, a value of  $f_s$  of as high as 0.5 may be appropriate.

### 6.1.8 Miscellaneous Other Direct Capital Costs

Other capital costs may be incurred for ash handling addition, water treatment addition, and flow modeling for a hot-side SCR system. For a cold-side system, costs are incurred for water treatment and flow modeling.

Flow modeling costs are similar for all systems and typically represent a flat fee for designing the ammonia injection system to assure proper mixing of ammonia and flue gas. The value used in the EPRI study was \$100,000.

Ash handling addition is required for hot side systems to remove ash deposited in the bottom of the SCR reactor to existing ash piping or to an ash silo. This cost is small compared to the costs of reactors, ammonia injection, heat exchangers, and fans. Here, it is assumed to have a value of approximately \$150,000 for a 550 MW power plant.

For hot-side applications, there is additional waste water burden associated with soot-blowing and water washing in the SCR system. In the cold-side system, additional equipment is required to collect GGH wastewater. For hot-side systems, the additional cost associated with water treatment are less than \$150,000. For cold-side systems, they are approximately \$500,000. These numbers are representative of a typical 550 MW power plant.

The direct cost for miscellaneous expenses of flow modeling, ash handling addition, and water treatment addition for a hot-side SCR system are:

$$DC_{\text{misc}} = \left\{ 100 + 300 \left( \frac{MW_e}{550} \right)^{0.6} \right\} \left( \frac{PCI}{357.3} \right) \quad (94)$$

For a cold-side SCR system, the direct cost is:

$$DC_{\text{misc,CS}} = \left\{ 100 + 650 \left( \frac{MW_e}{550} \right)^{0.6} \right\} \left( \frac{PCI}{357.3} \right) \quad (95)$$

#### 6.1.9 Total Direct Cost

The total direct cost is the sum of all of the direct costs. The cost of the initial catalyst charge is also included here in the total direct cost, because it is such a large and integral part of the SCR system. Thus, the total direct cost for the hot-side SCR system is:

$$TDC_{\text{HS}} = V_{\text{act}} UC_{\text{cat}} + \sum_{i=1}^7 DC_i \quad (96)$$

where there are seven direct cost components (reactor housing, ammonia injection, ductwork, air preheater modifications, ID fan differential, structural, and miscellaneous).

For a cold-side system, the total direct cost is given by:

$$TDC_{\text{CS}} = V_{\text{act}} UC_{\text{cat}} + \sum_{i=1}^8 DC_i \quad (97)$$

where there are eight direct cost components (reactor housing, ammonia injection system, ductwork, gas-gas heat exchanger, duct burner, booster fan, structural, and miscellaneous).

### 6.1.10 Other Capital Costs

Other capital costs include general facilities, various indirect capital costs, preproduction costs associated with startup, and inventory costs associated with providing initial stocks of chemicals and fuels.

General facilities include roads, office buildings, shops, laboratories, and other equipment or facilities that are not part of other process areas but that are associated with the process technology. General facilities costs are typically estimated as a percentage of the direct capital cost of all other process areas. Based on the estimates reported by Robie and Ireland (1991), it appears that general facilities cost is approximately 4 percent of the sum of all other direct costs. Thus, the general facilities cost is given by:

$$C_{GF} = f_{GF} TDC \quad (98)$$

Engineering and home office fees are typically estimated as a percentage of the total direct cost. In this case, a value of 10 percent is assumed as the default.

$$C_{EHO} = f_{EHO} TDC \quad (99)$$

Project contingency costs are also approximately 10 percent, as assumed by Robie and Ireland (1991). Usually, project contingency is assigned as a multiplier of the total direct cost (e.g., EPRI, 1986). For example:

$$C_{ProjC} = f_{ProjC} TDC \quad (100)$$

Process contingency costs are typically evaluated separately for each process area. The total process contingency is given by:

$$C_{ProcC} = \sum_i f_{ProjC} DC_i \quad (101)$$

Typical values of the process contingency are: five percent for reactor, catalyst, structural support, and fans; ten percent for ammonia storage, ductwork, air preheater modifications, and gas-gas heat exchanger; fifteen percent for water treatment addition, and twenty percent for ash handling addition.

The total plant cost, or overnight construction cost, is given by:

$$TPC = TDC + C_{GF} + C_{EHO} + C_{ProjC} + C_{ProcC} \quad (102)$$

An allowance for funds during construction (AFDC) is calculated based on the TPC as a function of the amount of time it would take to construct an SCR system. An 18 month construction period for a new plant is assumed. Methods for computing the AFDC are documented elsewhere (e.g., EPRI, 1986; Frey and Rubin, 1990) and are not repeated here. The total plant investment (TPI) represents the sum of the total plant cost and the AFDC. The procedure for calculating the AFDC is described by EPRI (1986).

$$TPI = TPC + AFDC \quad (103)$$

The final measure of capital cost is the total capital requirement (TCR). The TCR includes the total plant investment plus costs for royalties, startup costs, and initial inventories of feedstocks.

The royalty cost is typically estimated as a multiplier of total direct cost:

$$C_{Roy} = f_{Roy} TDC \quad (104)$$

However, in this case, no costs are assumed for royalties ( $f_{Roy} = 0$ ).

Preproduction costs are intended to cover operating, training, equipment checkout, changes in plant equipment, extra maintenance, and inefficient use of fuel, chemicals, catalysts, and other materials during plant startup. Pre-production costs typically include one month of both fixed and variable operating costs and two percent of total plant investment. In the case of an SCR system, by far the largest portion of the preproduction (startup) costs are represented by the two percent multiplier on TPI.

$$C_{PP} = \frac{N_{m,FOC} FOC + N_{m,VOC} VOC}{12} + f_{PP} TPI \quad (105)$$

where  $N_{m,FOC}$  is the number of months of fixed operating cost to be included in the pre-production cost estimate,  $N_{m,VOC}$  is the number of months of variable operating cost to be included, and  $f_{PP}$  is the pre-production cost factor, typically assigned a value of 0.02.

The inventory capital cost accounts for inventories of consumables required to be on-site and is typically estimated as 0.5 percent of total process capital excluding catalyst. The costs for initial catalysts and chemicals is zero. The SCR catalyst is included in the process capital costs. The inventory capital cost is:

$$C_{InvC} = f_{InvC} TDC \quad (106)$$

Thus, for an SCR system, the total capital requirement is:

$$TCR = TPI + C_{PP} + C_{InvC} + C_{Roy} \quad (107)$$

## 6.2 Gas Turbine-Based Systems

For the gas turbine-based systems, a more simplified cost model was developed. This model is based on data for SCR applied to several gas turbine systems as reported by Rao et al. (1991). Data are reported for five power plant configurations which includes SCR. These include two IGCC systems, two natural gas-fired gas turbine combined cycle (GTCC) systems, and one natural gas-fired humid air turbine (HAT) system. For the IGCC and GTCC systems, two alternatives are considered based on General Electric and ABB gas turbine designs. The reported capital costs include catalyst, reactor housing, ammonia injection system, and all structural supports and auxiliaries associated with the SCR system. Insufficient design detail was reported to allow these costs to be separated. Therefore, the direct cost model for gas turbine-based SCR systems includes all of these components.

The development of the direct capital cost model for SCR applied to a gas turbine-based system involved several steps. Rao et al. (1991) report only the total plant facilities investment (PFI) for a complete power plant with and without SCR for each of five power plant systems. Therefore, the PFI associated with SCR was estimated by difference. The PFI includes contingency costs, which are reported to be ten percent. The PFI less the contingency cost is the base plant cost. For SCR, the base plant cost includes the catalyst and all other SCR components. The cost of the initial catalyst charge was subtracted from the base plant cost in order to develop a cost estimate for all of the installed equipment. Accounting also for indirect construction costs (e.g., engineering and home office fees) which are also part of the base plant estimate, a direct capital cost estimate was developed:

$$DC_{SCR} = N_{T,SCR} 1,767 \left( \frac{V_{CAT,T}}{4,320 \text{ ft}^3 N_{T,SCR}} \right)^{0.637} \left( \frac{PCI}{354.7} \right) \quad (108)$$

This model includes the reactor housing, ammonia injection system, modifications to the HRSG associated with SCR system installation, structural supports, and all auxiliaries associated with SCR. The model is for an SCR system installed within an HRSG. The standard error of the estimate is \$41,000, based on five data points. The model was developed from data which spanned a catalyst volume per train of equipment from 2,800 ft<sup>3</sup> to 6,000 ft<sup>3</sup>.

For gas turbine-based systems, there is no induced draft fan required. Instead, the SCR system increases the backpressure on the turbine exhaust, which reduces the amount of energy that can be recovered by the gas turbine. Therefore, this efficiency penalty is modeled as part of the gas turbine performance model.



The total capital costs for SCR applied to gas turbine systems are estimated using the same method as for conventional power plants. For example, the total direct capital cost of the entire plant is given by:

$$TDC_{GT} = V_{act} UC_{cat} + \sum_{i=1}^n DC_i \quad (109)$$

Other capital costs, including indirects and startup costs, are estimated using Equations (98) through (107).

## 7.0 ANNUAL COSTS

The annual costs for SCR systems include fixed and variable operating costs. Fixed operating costs include operating labor, maintenance labor and materials, and overhead costs associated with administrative and support labor. Variable operating costs include consumables, such as ammonia and catalyst replacement. Costs for steam and electricity consumed from within the plant may also be estimated. First we consider annual costs for conventional coal-fired power plants, followed by annual costs for gas turbine-based systems.

### 7.1 Coal-Fired Power Plants

Annual costs are typically divided into Fixed Operating Costs (FOC) and Variable Operating Costs (VOC).

#### 7.1.1 Fixed Operating Costs

Fixed operating costs include operating labor, maintenance labor and materials, and overhead costs associated with administrative and support labor. The operating labor requirements for SCR are reported to be 4 labor-hours per day per reactor train. Thus, the total cost for operating labor per year is:

$$FOC_{OL} = f_{OL} N_{T,R} 365 UC_L \quad (110)$$

where  $f_{OL} = 4$  labor-hours/train/day.

Annual maintenance costs for new technologies are often estimated as a percentage of the installed capital cost of the facilities. The maintenance costs developed by Robie and Ireland (1991) are approximately two percent of the total direct costs, excluding catalyst:

$$FOC_M = f_M (TDC - V_{act} UC_{cat}) \quad (111)$$

where  $f_M = 0.02$ . Maintenance cost estimates are subdivided into labor and material components, primarily for the purpose of estimating the costs of administrative and support labor. Typically, 40 percent of the total maintenance cost is assumed to be for maintenance labor:

$$FOC_{ML} = f_{ML} FOC_M \quad (112)$$

The maintenance material cost may be estimated by difference:

$$FOC_{MM} = (1 - f_{ML}) FOC_M \quad (113)$$

Administrative and support labor is estimated as 30 percent of operating and maintenance labor costs:

$$FOC_{AS} = f_{AS} (FOC_{OL} + FOC_{ML}) \quad (114)$$

where  $f_{AS} = 0.30$ .

The total fixed operating cost is given by:

$$FOC = FOC_{OL} + FOC_M + FOC_{AS} \quad (115)$$

A typical labor rate would be \$22/hour for a midwest location.

### 7.1.2 Variable Operating Costs

The major component of the variable operating cost is for catalyst replacement. The analytical model for estimating catalyst replacement is given by Equation (41). Similarly, the ammonia mass flow requirement is given by Equation (44), the steam requirement for ammonia injection is given by Equation (45), and the electricity consumption for the SCR systems is given by Equation (63).

The variable operating cost for catalyst replacement is:

$$VOC_{cat} = V_{C,a} UC_{cat} \quad (116)$$

The variable operating cost for ammonia is:

$$VOC_{NH_3} = 8,760 c_f \left[ \left( \frac{M_{NH_3,A,i}}{117.6} \right) UC_{NH_3} \right] \quad (117)$$

where the unit cost of ammonia is in \$/ton and the flowrate is in lbmole/hr.

The SCR system also uses steam and electricity. These costs are accounted for internally within the power plant.

Therefore, the total variable operating cost for the SCR system may be estimated as:

$$VOC = VOC_{cat} + VOC_{NH_3} + 8,760 c_f \left[ \left( \frac{18 M_{steam}}{1,000} \right) UC_{steam} + E_{scr} UC_{elec} \right] \quad (118)$$

The unit cost for catalyst is highly variable. Robie and Ireland (1991) used a value of \$660/ft<sup>3</sup>, but noted that at the time of their study, catalyst costs in Europe were as low as \$330/ft<sup>3</sup>. The unit cost of ammonia is typically \$150/ton. The unit cost of steam is approximately \$3.00/1,000 lb. The cost of power is often assumed to be 5 cents per kWh.

## 7.2 Gas Turbine-Based Systems

### 7.2.1 Fixed Operating Costs

A detailed breakdown of fixed and variable operating costs was not available from Rao et al. (1991) for the purpose of developing a detailed annual cost model. However, the cost model structure of the "Technical Assessment Guide" (EPRI, 1986) may be employed for this purpose.

We assume that no additional plant operating personnel are required for the SCR system. However, additional maintenance personnel and materials are required. The operating labor for the plant is estimated as a function of the number of gas turbine units, with an assumption of four operators required per unit (Frey and Rubin, 1990). The total maintenance cost is estimated as a function of the process area capital costs for the SCR system. A nominal assumption is that the maintenance cost is two percent of the process area capital costs. Process area capital costs include direct costs and a prorated allocation of indirect costs, excluding initial catalyst, allowance for funds during construction, and startup costs. This approach is the same as that used by Frey and Rubin (1990) for cost models of IGCC systems. Thus, the maintenance FOC for SCR is given by:

$$\text{FOC}_{M,SCR} = f_{M,SCR} \left\{ \text{DC}_{SCR} \left( 1 + \frac{\text{TIC} + C_{\text{ProjC}}}{\text{TDC}} \right) + C_{\text{ProcC}} \right\} \quad (119)$$

Here, the term for Total Indirect Costs (TIC) includes taxes, engineering and home office fees, and environmental permitting costs.

The fixed operating cost may be disaggregated into labor and material components using Equations (112) and (113). The total administrative and support labor for the plant is estimated as a function of operating and maintenance labor, as given by Equation (114). The total fixed operating cost is therefore given by Equation (115).

### 7.2.2 Variable Operating Cost

Similar to coal-fired power plants, the variable operating cost for gas turbine based SCR systems includes annual catalyst replacement, ammonia consumption, and electricity consumption for the ammonia compressors. The annual catalyst replacement cost is given by Equation (116). The annual ammonia cost is given by Equation (117). The steam and electricity consumption of the SCR system is accounted for internally within the power plant performance model. These latter factors affect the plant efficiency and, hence, the net plant power output. This, in turn, affects normalized costs (e.g., \$/kW, mills/kWh). Therefore, the variable operating costs for gas turbine based systems are given by:

$$\text{VOC} = \text{VOC}_{\text{cat}} + \text{VOC}_{\text{NH}_3} \quad (120)$$

For gas turbine-based systems, SCR catalyst costs may be as high as \$800/ft<sup>3</sup> (May, 1993). Catalyst cost of approximately \$250/ft<sup>3</sup> were employed in a recent design study (Rao et al., 1991).

### 7.3 Total Annualized Cost Model

The total annualized cost is the levelized annual revenue requirement required to cover all of the capital and operating costs for the economic life of the plant. For electric power plants, the total annualized cost is typically expressed as the cost of electricity. The total capital requirement, fixed operating cost, and variable operating cost are used to calculate the cost of producing electricity that is available for sale from the power plant, based on the net electrical output of the power plant. The net power output is the total power generated from the gas turbines and steam turbines less the total auxiliary power demand:

$$\text{MW}_{\text{net}} = \text{MW}_{\text{GT}} + \text{MW}_{\text{ST}} - 1,000 \sum_i W_{e,i} \quad (121)$$

The cost of electricity in mills (one-thousands of a dollar) per net kWh is given by:

$$C_{\text{elec}} = \frac{[1,000 f_{\text{cr}} \text{TCR} + f_{\text{vcif}} (\text{FOC} + \text{VOC})] \left( \frac{1,000 \text{ mills}}{\text{dollar}} \right)}{\text{MW}_{\text{net}} \times 8,760 \frac{\text{hrs}}{\text{yr}} \times c_f} \quad (122)$$

where,

$f_{\text{cr}}$  = capital recovery factor

$f_{\text{vcif}}$  = variable cost levelization factor

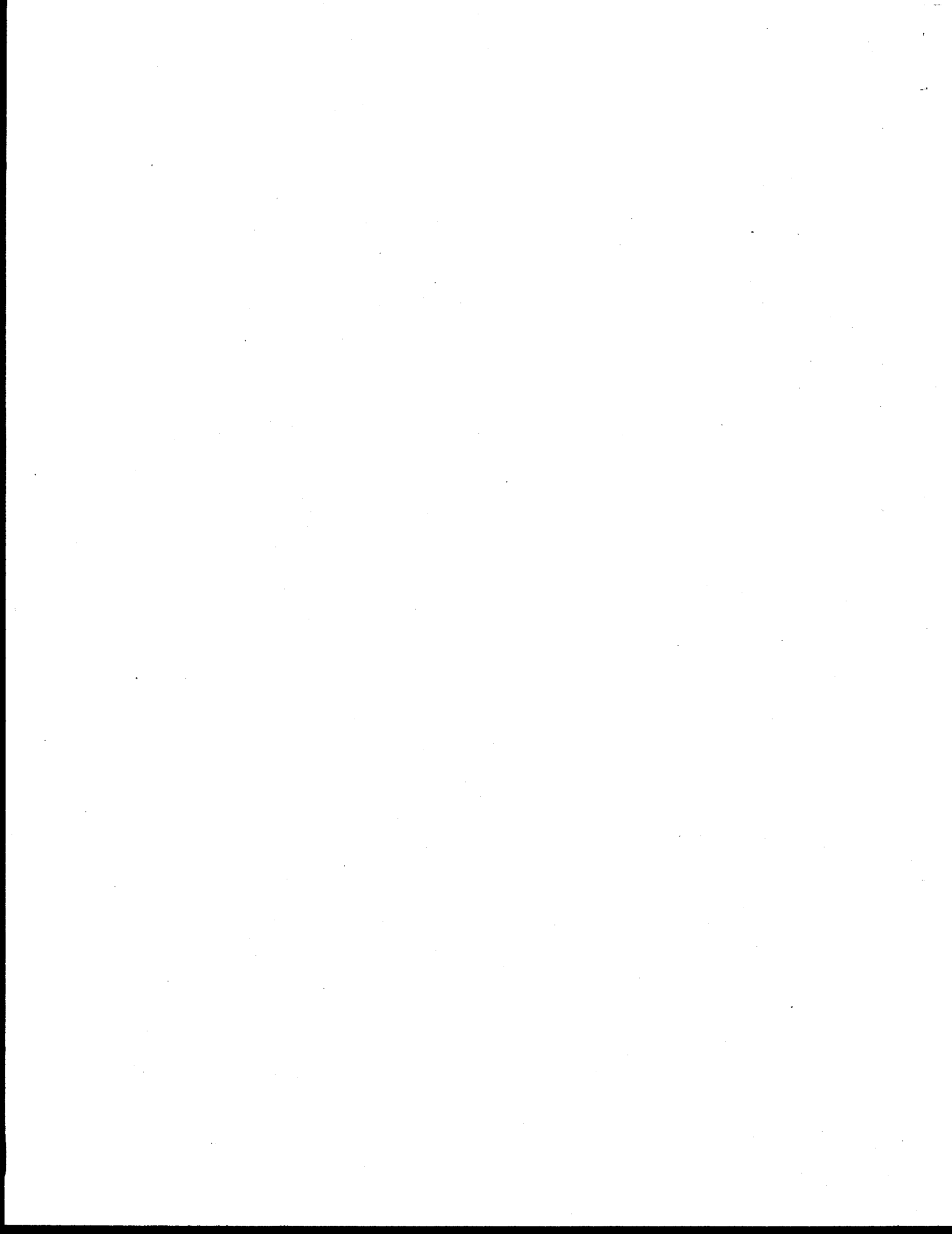
The numerator of Equation (122) is the total annual revenue requirement for the plant, and the denominator is the total net kilowatt-hours of electricity generated in a year. The total capital requirement from Equation (107) is in units of thousands of dollars. Therefore, a factor of 1,000 is used to convert the total capital requirement to units of dollars. The fixed operating cost from Equation (119) and the variable operating cost from Equation (120) are both in units of dollars. The annual revenue requirement shown in square brackets in the numerator is converted from dollars to mills, which is a more convenient unit for reporting the cost of electricity.

The capital recovery factor converts the capital cost into the equivalent levelized annual revenue required to provide a return to equity (stock) and debt (loan) financing

sources, to pay for a portion of the principal, and to pay associated taxes and insurance (EPRI, 1987). The capital recovery factor, therefore, depends on the economic life of the plant, the type of financing used to supply the capital, and the applicable tax laws. A typical capital recovery factor for a 30 year life based on typical assumptions regarding financing and the federal Tax Reform Act of 1986 is 0.1034 (see EPRI, 1986). The variable cost levelization factor converts the annual costs into an equivalent levelized annual cost. If inflation and escalation are assumed to be zero, the variable cost levelization factor has a value of 1.0. The capital recovery and variable cost levelization factors are typically calculated using the standard method described by EPRI (1986).

#### **7.4 Computer Code for the Performance and Cost Model**

The computer code required to implement the performance and cost model of the SCR system for gas turbine-based applications is given in Appendix B. A model application featuring this computer-based model is given in the next section. The SCR performance and cost model has been implemented as part of previously developed performance, emissions, and cost model of IGCC systems. The appendix illustrates the inclusion of SCR in an engineering-economic model of an air-blown Lurgi-gasifier-based IGCC system with hot gas cleanup (Frey and Rubin, 1990; 1991; 1992). The performance models are implemented in ASPEN, while the cost models are written as stand-alone FORTRAN subroutines which are called by the performance model.



## **8.0 MODEL APPLICATIONS**

In this section, we present sample model results to illustrate the application of the SCR performance and cost model. The case studies here focus on two IGCC systems.

### **8.1 Modeling IGCC Systems**

The performance, emissions, and cost of the IGCC systems are modeled using detailed engineering models. Performance models of a Lurgi-based system and a KRW-based system were developed by DOE/METC using the ASPEN chemical process simulator. These performance models have been significantly modified to more completely and accurately represent process performance and emissions (Frey, 1991; Frey and Rubin 1991). For example, gas cleanup and gas turbine performance and emissions are more completely characterized in the modified models. A new performance and cost model of selective catalytic reduction (SCR) NO<sub>x</sub> control has been added to these IGCC models.

The cost models for the two IGCC concepts are based on approximately 30 design studies of IGCC systems (Frey and Rubin, 1990). Direct capital costs are estimated for approximately one dozen major process areas. Typically, several performance and design variables are included in the direct cost models. Indirect and other capital costs are estimated based on approximately 60 cost model parameters. These include process area contingencies, project contingency, indirect construction costs, sales tax, allowance for funds used during construction, environmental permitting costs, spare parts inventory costs, costs for initial inventories of fuels and chemicals, land cost, and startup costs. Fixed and variable operating costs are estimated based on 40 to 50 parameters. Fixed operating costs include maintenance material and labor for each process area, plant operating labor, and administrative and support labor. Variable operating costs include consumables (e.g., water treatment chemicals, zinc ferrite sorbent), ash disposal, fuel, and byproduct credit. Total levelized costs are calculated using the financial assumptions and methodology of EPRI (1986).

### **8.2 Modeling Assumptions**

The case studies developed here are based on system performance and cost assumptions which have been documented elsewhere (Frey and Rubin, 1991; 1992; 1992a; Frey et al, 1994). Here, we focus on assumptions that are unique to the SCR process area. These assumptions are given in Table 1. These assumptions include performance and cost inputs to the SCR models.



Table 1. Modeling Assumptions for SCR Systems

Parameter	Units	Value
<u>SCR Performance Assumptions</u>		
NO <sub>x</sub> Removal Efficiency	%	80
Reference NO <sub>x</sub> Removal Efficiency	%	80
Ammonia Slip	ppmv	10
Reference NO <sub>x</sub> Concentration	ppmv	42
Minimum Catalyst Activity	ratio	0.7
Catalyst Activity at Reference Time	ratio	0.85
Reference Time	hours	8760
Catalyst Layer Replacement Interval	Years	2
Reference Replacement Interval	Years	3
Catalyst Temperature	°F	717
Reference Temperature	°F	717
SO <sub>2</sub> Oxidation	fraction	0.01
Reference Space Velocity	1/hr	10,000
Number of Catalyst Layers		3
Number replaced at the end of each replacement interval		1
Pressure Drop	in. H <sub>2</sub> O	4
<u>SCR Cost Assumptions</u>		
Number of Units		Same as number of gas turbines
Process Contingency	% of process cost	10
Maintenance Cost	% of process cost	2
SCR Catalyst Cost	\$/ft <sup>3</sup>	250
Ammonia Cost	\$/ton	150

### 8.3 Running the Model

The engineering models were exercised to characterize key measures of plant performance, emissions, and cost. The IGCC models were run on a DEC VAXStation 3200 mini-computer using the public version of ASPEN. Model results are reported in Tables 2, 3, and 4. The case studies include an air-blown Lurgi gasifier-based IGCC system with hot gas cleanup and a KRW gasifier-based IGCC system with hot gas cleanup. The model assumptions include a current heavy duty gas turbine with a pressure ratio of 15 and a firing temperature of 2,350 °F. The cases illustrate the sensitivity of SCR costs to the NO<sub>x</sub> loading from the gas turbine combustor.

### 8.4 Modeling Results

Here, we consider a series of case studies that illustrate the insights obtained from the SCR performance and cost model as applied to IGCC systems.

### 8.4.1 SCR and Lurgi Gasifier-Based IGCC Systems

There is uncertainty regarding the future performance of gas turbine combustors with regard to fuel  $\text{NO}_x$  formation. Therefore, we consider two case studies that illustrate the effect of total  $\text{NO}_x$  loading on SCR performance and cost. The results for these two case studies are given in Table 2. In the first case, it is assumed that only 40 percent of fuel-bound nitrogen, in the form of ammonia, is converted to  $\text{NO}_x$ , while in the second case it is assumed that 90 percent of ammonia is converted to  $\text{NO}_x$ . The latter assumption is typical of conventional gas turbine combustors, while the former may be typical of new technology.

For the *Low Fuel  $\text{NO}_x$  Case*, the uncontrolled  $\text{NO}_x$  emission rate is 1.29 lb  $\text{NO}_2/10^6$  BTU of fuel input. This emission rate exceeds emission standards applicable to conventional coal-fired power plants. The use of SCR for post-combustion  $\text{NO}_x$  control can substantially reduce this emission level. With 80 percent control, the  $\text{NO}_x$  emission rate is reduced to 0.26 lb/ $10^6$  BTU, which should be sufficient to comply with current emission standards. A catalyst volume of 28,500  $\text{ft}^3$  is required. This volume is distributed over three gas turbine/HRSG units. This volume is approximately a factor of two greater than required for conventional natural gas-fired gas turbine applications. Such systems work with an inlet  $\text{NO}_x$  concentration of typically 42  $\text{ppm}_v$ , corrected to 15 percent oxygen. In contrast, this case study features an inlet  $\text{NO}_x$  concentration of 450  $\text{ppm}_v$ . While the inlet  $\text{NO}_x$  concentration does not by itself affect the catalyst volume required, the ratio of the outlet  $\text{NO}_x$  to the allowable ammonia slip does (see Equation (25)). To maintain the ammonia slip at a fixed concentration for a fixed removal efficiency as the inlet  $\text{NO}_x$  concentration increases requires a larger catalyst volume. By comparing IGCC systems with and without SCR, it is possible to gain some insight into the performance and cost impacts of SCR. The plant efficiency is reduced by 0.37 percentage points due to gas turbine backpressure, steam consumption for ammonia vaporization, and electricity consumption for ammonia compression. The net plant output is reduced by 6.8 MW. The capital cost for the SCR system is approximately \$50/kW, and the annualized total cost for the SCR system is approximately 1.8 mills/kWh. A sample output for the plant results with SCR is given in Appendix B.

Table 2. Summary of results from deterministic simulations of a 710 MW air-blown Lurgi gasifier-based IGCC system with 2,350 °F gas turbine, hot gas cleanup and SCR.

Parameter <sup>a</sup>	Units <sup>b</sup>	Low Fuel NO <sub>x</sub> Case <sup>c,e,f</sup>			High Fuel NO <sub>x</sub> Case <sup>d,e,f</sup>		
		w/o SCR	w/ SCR	SCR Only	w/o SCR	w/ SCR	SCR Only
<b>Plant Performance</b>							
Thermal Efficiency	%, HHV	39.07	38.70	--	38.99	38.31	--
Coal Consumption	lb/kWh	0.777	0.785	--	0.779	0.793	--
Zinc Ferrite Sorbent Charge	10 <sup>6</sup> lb	7.35	7.35	--	7.37	7.37	--
Sulfuric Acid Production	lb/kWh	0.084	0.085	--	0.084	0.086	--
Process Water Cons.	lb/kWh	1.579	1.664	--	1.582	1.758	--
SCR Space Velocity	1/hr	--	5,204	5,204	--	4,151	4,151
SCR Catalyst Volume	ft <sup>3</sup>	--	28,503	28,503	--	35,732	35,732
<b>Plant Discharges</b>							
SO <sub>2</sub> Emissions	lb/10 <sup>6</sup> Btu	0.041	0.041	--	0.041	0.041	--
NO <sub>x</sub> Emissions	lb/10 <sup>6</sup> Btu	1.29	0.26	--	2.74	0.55	--
CO Emissions	lb/kWh	0.003	0.003	--	0.003	0.003	--
CO <sub>2</sub> Emissions	lb/kWh	1.699	1.715	--	1.702	1.733	--
Solid Waste	lb/kWh	0.082	0.082	--	0.082	0.083	--
<b>Plant Costs</b>							
Total Capital Cost	\$/kW	1,288	1,340	52	1,288	1,357	69
Fixed Operating Costs	\$/kW-yr	40.1	41.0	0.9	40.1	41.4	1.3
Variable Oper. Costs	mills/kWh	17.9	18.7	0.7	18.0	19.2	1.2
Coal	mills/kWh	16.0	16.1	0.1	16.0	16.3	0.3
Sulfuric Acid Sales	mills/kWh	(1.5)	(1.5)	(0)	(1.5)	(1.5)	(0)
Other	mills/kWh	3.5	4.1	0.6	3.5	4.5	1.0
Cost of Electricity	mills/kWh	48.4	50.2	1.8	48.4	51.1	2.7

<sup>a</sup> Coal consumption is on an as-received basis. Water consumption is for process requirements including makeup for steam cycle blowdown, gasifier steam, zinc ferrite steam, and SCR. Solid waste includes gasifier bottom ash and nonrecycled fines from fuel gas cyclones.

<sup>b</sup> HHV = higher heating value.

<sup>c</sup> In the low fuel NO<sub>x</sub> case, 40 percent of ammonia in the coal gas is converted to NO<sub>x</sub>

<sup>d</sup> In the high fuel NO<sub>x</sub> case, 90 percent of ammonia in the coal gas is converted to NO<sub>x</sub>

<sup>e</sup> SCR NO<sub>x</sub> Removal Efficiency is 80 percent and the equivalent catalyst "life" is six years.

<sup>f</sup> The three columns include the following case studies: (1) an IGCC system without SCR (w/o SCR); (2) an IGCC system with SCR (w/SCR); and (3) the difference between the two (SCR Only)

For the *High Fuel NO<sub>x</sub> Case*, the uncontrolled NO<sub>x</sub> emission rate is 2.74 lb NO<sub>2</sub>/10<sup>6</sup> BTU. The use of SCR for post-combustion NO<sub>x</sub> control can substantially reduce this emission level. With 80 percent control, the NO<sub>x</sub> emission rate is reduced to 0.55 lb/10<sup>6</sup> BTU, which is just below current emission standards. A catalyst volume of 35,700 ft<sup>3</sup> is required. This volume is substantially larger than for the previous case, due to the higher NO<sub>x</sub> emission rate and the constraint that ammonia slip be held to 10 ppm. The plant efficiency is reduced by 0.68 percentage points. This is almost twice as large an energy penalty as for the previous case and is due to the larger ammonia requirement and associated steam and ammonia compression requirements. The net plant output is reduced by 12.6 MW. The capital cost for the SCR system is approximately \$70/kW, and the annualized total cost for the SCR system is approximately 2.7 mills/kWh. These results illustrate the sensitivity of SCR costs to the inlet NO<sub>x</sub> loading.

A second sensitivity analysis is considered for the *Low Fuel NO<sub>x</sub> Case*. In this second analysis, we consider the implications of reduced SCR NO<sub>x</sub> removal efficiency on plant performance and cost. If we assume that the binding emission limit is 0.6 lb/10<sup>6</sup> BTU, then it may be unnecessary to employ an SCR NO<sub>x</sub> removal efficiency of 80 percent. Instead, a removal efficiency of 55 percent may suffice. If this lower removal efficiency is feasible from a regulatory viewpoint, then the catalyst requirement and ammonia consumption will be significantly lower than for the base case. These results are illustrated in Table 3. The results for 80 percent NO<sub>x</sub> removal are repeated in the table for ease of comparison.

The reduced NO<sub>x</sub> removal efficiency leads to a substantial reduction in the catalyst volume requirement, which in turn reduces capital costs. Furthermore, the ammonia-to-NO<sub>x</sub> molar ratio is also reduced, leading to lower annual costs. As seen in the table, the efficiency penalty and costs for the lower removal efficiency case are significantly lower than for the base case.

#### **8.4.2 SCR and KRW Gasifier-Based IGCC Systems**

Two case studies are considered for the KRW-based IGCC system. The results are given in Table 4. These case studies are comparable to the High Fuel NO<sub>x</sub> Case and the Low Fuel NO<sub>x</sub> Case for the Lurgi-based system. In the first case study, we assume that 40 percent of ammonia in the coal gas is converted to NO<sub>x</sub> in the gas turbine combustor. In the second case study, we assume that 90 percent of the ammonia in the coal gas is converted to NO<sub>x</sub>. Because the KRW gasifier operates at a higher temperature

Table 3. Summary of results from deterministic simulations of a 710 MW air-blown Lurgi gasifier-based IGCC system with 2,350 °F gas turbine, hot gas cleanup and SCR: Low Fuel NO<sub>x</sub> Case.

Parameter <sup>a</sup>	Units <sup>b</sup>	w/o SCR	80 % Removal <sup>c,e</sup>		55 % Removal <sup>d,e</sup>	
			w/ SCR	SCR Only	w/ SCR	SCR Only
<b>Plant Performance</b>						
Thermal Efficiency	%, HHV	39.07	38.70	--	38.79	--
Coal Consumption	lb/kWh	0.777	0.785	--	0.783	--
Zinc Ferrite Sorbent Charge	10 <sup>6</sup> lb	7.35	7.35	--	7.35	--
Sulfuric Acid Production	lb/kWh	0.084	0.085	--	0.085	--
Process Water Cons.	lb/kWh	1.579	1.664	--	1.639	--
SCR Space Velocity	1/hr	--	5,204	5,204	8,224	8,224
SCR Catalyst Volume	ft <sup>3</sup>	--	28,503	28,503	18,036	18,036
<b>Plant Discharges</b>						
SO <sub>2</sub> Emissions	lb/10 <sup>6</sup> Btu	0.041	0.041	--	0.041	--
NO <sub>x</sub> Emissions	lb/10 <sup>6</sup> Btu	1.29	0.26	--	0.58	--
CO Emissions	lb/kWh	0.003	0.003	--	0.003	--
CO <sub>2</sub> Emissions	lb/kWh	1.699	1.715	--	1.711	--
Solid Waste	lb/kWh	0.082	0.082	--	0.082	--
<b>Plant Costs</b>						
Total Capital Cost	\$/kW	1,288	1,340	52	1,326	38
Fixed Operating Costs	\$/kW-yr	40.1	41.0	0.9	40.8	0.7
Variable Oper. Costs	mills/kWh	17.9	18.7	0.7	18.5	0.6
Coal	mills/kWh	16.0	16.1	0.1	16.1	0.1
Sulfuric Acid Sales	mills/kWh	(1.5)	(1.5)	(0)	(1.5)	(0)
Other	mills/kWh	3.5	4.1	0.6	3.9	0.4
Cost of Electricity	mills/kWh	48.4	50.2	1.8	49.7	1.3

<sup>a</sup> Coal consumption is on an as-received basis. Water consumption is for process requirements including makeup for steam cycle blowdown, gasifier steam, zinc ferrite steam, and SCR. Solid waste includes gasifier bottom ash and nonrecycled fines from fuel gas cyclones.

<sup>b</sup> HHV = higher heating value.

<sup>c</sup> SCR removal efficiency is 80 percent.

<sup>d</sup> SCR removal efficiency is 55 percent

<sup>e</sup> SCR NO<sub>x</sub> Removal Efficiency is 80 percent and the equivalent catalyst "life" is six years.

Table 4. Summary of results from deterministic simulations of a 830 MW air-blown KRW gasifier-based IGCC system with 2,350 °F gas turbine, hot gas cleanup and SCR.

Parameter <sup>a</sup>	Units <sup>b</sup>	Low Fuel NO <sub>x</sub> Case <sup>c,e,f</sup>			High Fuel NO <sub>x</sub> Case <sup>d,e,f</sup>		
		w/o SCR	w/ SCR	SCR Only	w/o SCR	w/ SCR	SCR Only
<b>Plant Performance</b>							
Thermal Efficiency	%, HHV	41.84	41.68	--	41.83	41.59	--
Coal Consumption	lb/kWh	0.726	0.728	--	0.726	0.730	--
Zinc Ferrite Sorbent Charge	10 <sup>6</sup> lb	4.88	4.88	--	4.81	4.81	--
Sulfuric Acid Production	lb/kWh	0	0	--	0	0	--
Process Water Cons.	lb/kWh	0.714	0.737	--	0.714	0.754	--
SCR Space Velocity	1/hr	--	6,770	6,770	--	5,640	5,640
SCR Catalyst Volume	ft <sup>3</sup>	--	21,800	21,800	--	26,200	26,200
<b>Plant Discharges</b>							
SO <sub>2</sub> Emissions	lb/10 <sup>6</sup> Btu	0.013	0.013	--	0.013	0.013	--
NO <sub>x</sub> Emissions	lb/10 <sup>6</sup> Btu	0.390	0.083	--	0.711	0.148	--
CO Emissions	lb/kWh	0.005	0.005	--	0.005	0.005	--
CO <sub>2</sub> Emissions	lb/kWh	1.684	1.690	--	1.684	1.694	--
Solid Waste	lb/kWh	0.227	0.228	--	0.227	0.226	--
<b>Plant Costs</b>							
Total Capital Cost	\$/kW	1,351	1,385	34	1,405	1,447	42
Fixed Operating Costs	\$/kW-yr	45.2	45.8	0.6	47.8	48.5	0.7
Variable Oper. Costs	mills/kWh	19.1	19.5	0.4	19.1	19.6	0.5
Coal	mills/kWh	14.9	15.0	0.1	14.9	15.0	0.1
Sulfuric Acid Sales	mills/kWh	0	0	0	0	0	0
Other	mills/kWh	4.2	4.5	0.3	4.2	4.6	0.4
Cost of Electricity	mills/kWh	51.6	52.7	1.1	53.0	54.4	1.4

<sup>a</sup> Coal consumption is on an as-received basis. Water consumption is for process requirements including makeup for steam cycle blowdown, gasifier steam, zinc ferrite steam, and SCR. Solid waste includes gasifier bottom ash and sulfated sorbent.

<sup>b</sup> HHV = higher heating value.

<sup>c</sup> In the low fuel NO<sub>x</sub> case, 40 percent of ammonia in the coal gas is converted to NO<sub>x</sub>

<sup>d</sup> In the high fuel NO<sub>x</sub> case, 90 percent of ammonia in the coal gas is converted to NO<sub>x</sub>

<sup>e</sup> SCR NO<sub>x</sub> Removal Efficiency is 80 percent and the equivalent catalyst "life" is six years.

<sup>f</sup> The three columns include the following case studies: (1) an IGCC system without SCR (w/o SCR); (2) an IGCC system with SCR (w/SCR); and (3) the difference between the two (SCR Only)

than the Lurgi gasifier, its ammonia yield is substantially lower. Therefore, the fuel NO<sub>x</sub> emissions and the total NO<sub>x</sub> loading to the SCR system are lower than for the comparable Lurgi-based system case studies.

For the Low Fuel NO<sub>x</sub> Case, the uncontrolled NO<sub>x</sub> emissions are approximately 0.4 lb/10<sup>6</sup> BTU. With 80 percent post-combustion NO<sub>x</sub> control, this emission rate is reduced to approximately 0.08 lb/10<sup>6</sup> BTU. There is a 0.16 percentage point reduction in the plant thermal efficiency, an additional capital cost of \$34/kW, and an additional annualized total cost of 1.1 mills/kWh.

For the High Fuel NO<sub>x</sub> Case, the uncontrolled NO<sub>x</sub> emissions are approximately 0.7 lb/10<sup>6</sup> BTU. With 80 percent post-combustion NO<sub>x</sub> control, this emission rate is reduced to approximately 0.15 lb/10<sup>6</sup> BTU. There is a 0.24 percentage point reduction in the plant thermal efficiency, an additional capital cost of \$42/kW, and an additional annualized total cost of 1.4 mills/kWh. Thus, the higher NO<sub>x</sub> loading results in higher costs for the SCR system.

The High Fuel NO<sub>x</sub> Cases exhibits significantly higher costs than for the Low Fuel NO<sub>x</sub> Cases. This is because the coal flow rate for the High Fuel NO<sub>x</sub> Case is just large enough that an additional gasifier train is required. The additional gasifier increases the direct capital cost by approximately \$20 million. However, the differences between the two cases for comparisons with and without SCR are not sensitive to the difference in the gasification island costs, because SCR capital costs depend primarily on the exhaust gas volumetric flow rate and factors that determine space velocity.

When compared to the results for the Lurgi-based IGCC system, it is apparent that the costs of SCR for the KRW-based system are substantially lower. For the High Fuel NO<sub>x</sub> Cases, the levelized cost for SCR for the KRW-based system is almost one-half that of the Lurgi-based system. These results illustrate the variability of SCR costs for different types of IGCC systems.

Figure 20 illustrates the sensitivity of catalyst requirements to design assumptions, based on modeling of a KRW-based system. The initial catalyst volume is smallest for the most frequent catalyst replacement intervals. This is because the catalyst activity at the design point is relatively high. In contrast, for long catalyst layer replacement intervals, the catalyst activity will decline and result in the need for additional initial catalyst to compensate for the activity loss at the design point. Thus, while the annual volume of catalyst that must be replaced decreases as the replacement interval increases, the initial

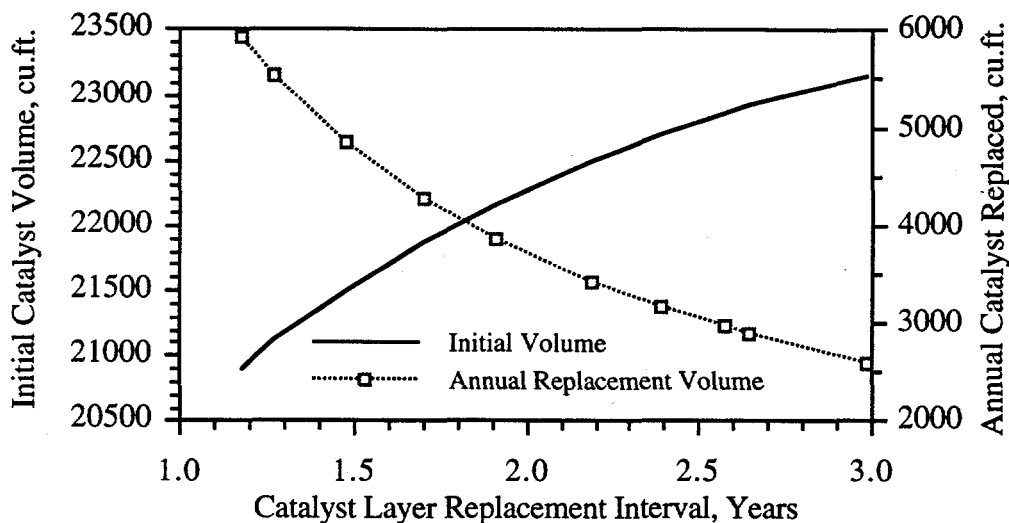


Figure 20. Sensitivity of Initial Catalyst Volume and Annual Catalyst Replacement Volume to Catalyst Layer Replacement Interval.

catalyst requirement increases. Furthermore, there are diminishing returns for annual replacement reductions as the replacement interval increases.

These sensitivity analysis results emphasize that catalyst "life" is a design issue subject to optimization.

### 8.5 Summary

Detailed performance and cost models of SCR systems have been developed. These models have been implemented as computer code to augment previously developed performance, emissions, and cost models of selected IGCC systems. The SCR performance and cost models will be deployed as part of other advanced power generation system models, such as for EFCC and PFBC systems.

The SCR models allow the implications of alternative  $\text{NO}_x$  control strategies to be explored, and also permit sensitivity and optimization case studies to minimize the total cost of environmental control for advanced power generation systems. Future work will include the application of new process optimization techniques to flowsheets which include SCR (Diwekar et al., 1994).

Because SCR systems have not yet been applied to many of the advanced power generation systems, it is critically important to consider the role of uncertainties in making predictions of SCR performance and cost. Because many of these technologies are of interest to research and development (R&D) planners and to electric utilities, it is necessary



to fully understand their environmental characteristics and the costs of complying with current and future regulations. Thus, the use of probabilistic modeling techniques, as described by Frey and Rubin (1991), to evaluate SCR will be a key area for future work.

The SCR performance and cost models developed here will need to be revised as new data and technologies become available. For example, while model development has focused on base metal catalysts, additional models could be developed for precious metal and zeolite catalyst systems. As experience is gained regarding long-term catalyst durability, model input assumptions may be revised. As better understanding is gained regarding the factors that affect catalyst poison, perhaps more detailed mechanistic models can be developed to characterize catalyst durability. Therefore, the models developed here should be viewed as a first step in an iterative model development and adaptation process, as new data become available and as the need emerges to adapt these models for applications to new types of systems not currently envisioned.

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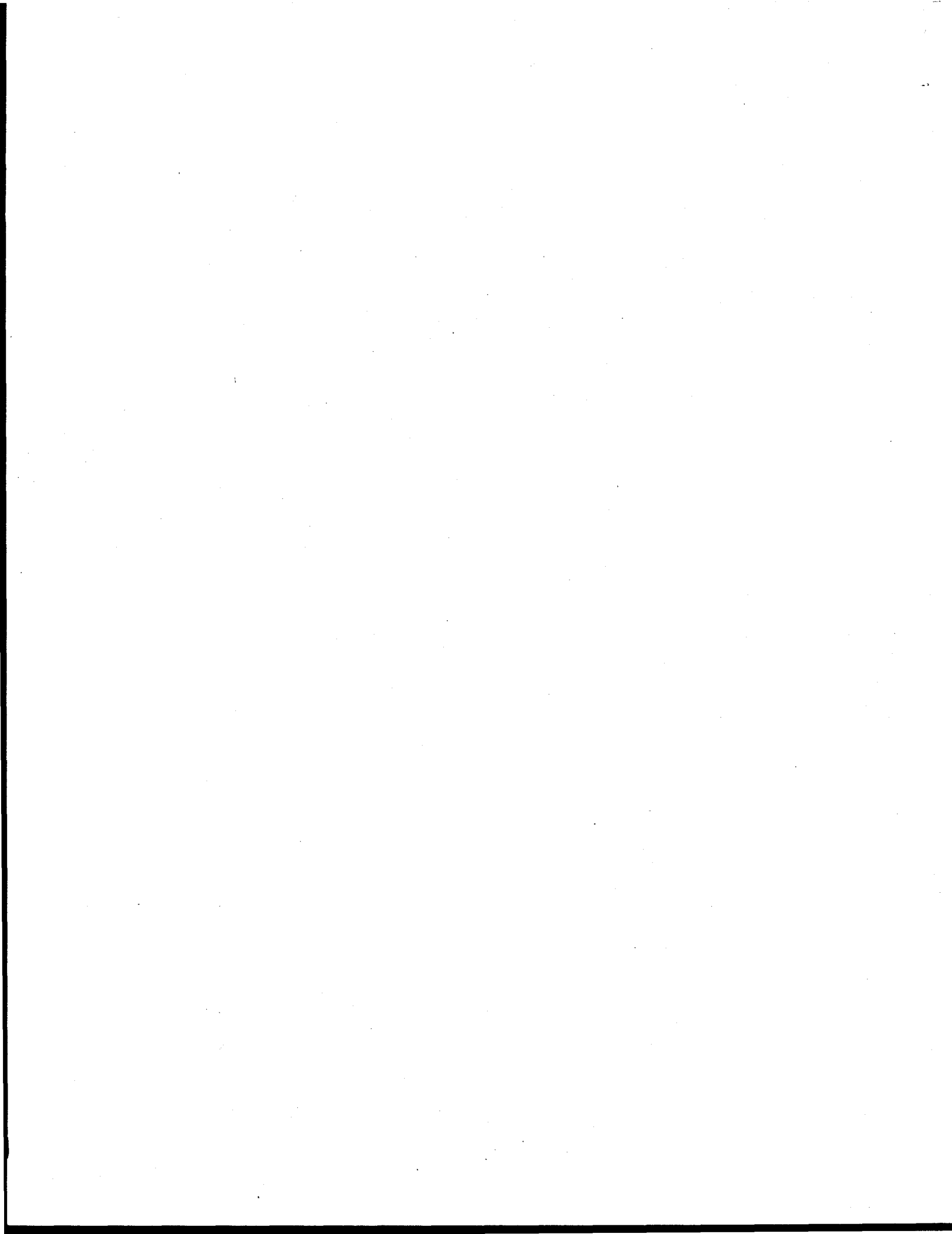
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## APPENDIX A: NOMENCLATURE

### A.1 English Letter Symbols

- a = Constant
- $A_c$  = Catalyst area, ft<sup>2</sup>
- $A_{min}$  = Minimum catalyst activity level
- $A_{inc}$  = Incremental catalyst activity subject to exponential decay ( $1-A_{min}$ )
- AFDC = Allowance for funds used during construction (\$)
- AV = Area velocity, ft/hr
- $c_f$  = Annual plant capacity factor, fraction of year
- C = Constant representing catalyst activity for a given operating condition (e.g., temperature, inlet NO<sub>x</sub> concentration) with NH<sub>3</sub>/NO<sub>x</sub> ratio equal to 1.0
- $\vec{C}_i$  = Vector of concentrations of trace species in stream i, ppmw
- $C_{NH_3,hc}$  = Concentration of ammonia in "high concentration" wash water, mg/l
- $C_{NH_3,lc}$  = Concentration of ammonia in "low concentration" wash water, mg/l
- $C_{ProcC}$  = Cost of project contingencies, \$1,000
- $C_r$  = Ratio of catalyst activities for two catalysts.
- $DC_{aph,mod}$  = Direct capital cost of air preheater modifications, \$1,000
- $DC_{BF}$  = Direct capital cost of the booster fan for cold-side SCR, \$1,000.
- $DC_D$  = Direct capital cost of ductwork, \$1,000
- $DC_{D,CS}$  = Direct capital cost of ductwork for a cold-side SCR system, \$1,000
- $DC_{DB}$  = Direct cost of duct burners for a cold-side SCR system, \$1,000.
- $DC_{GGH}$  = Direct capital cost of the gas-gas heat exchanger for cold-side SCR, \$1,000.
- $DC_{misc}$  = Direct capital cost of miscellaneous items, \$1,000.
- $DC_{NH_3}$  = Direct capital cost of the ammonia injection systems, \$1,000
- $DC_R$  = Direct capital cost of reactors, \$1,000.
- $E_a$  = Activation energy
- $E_i$  = Energy penalty for system i, MW
- $E_{scr}$  = Electricity requirement for SCR, kW
- $EC_{ID,dif}$  = Electricity consumption associated with SCR flue gas pressure drop, kWh.
- $f_{AS}$  = Factor for administrative and support costs (fraction of total labor)
- $f_{D,CS}$  = Factor for cold-side SCR duct costs.
- $f_{FGD}$  = Fraction of gas-gas heat exchanger pressure drop that is normally associated with FGD flue gas reheat.
- $f_{hc}$  = Fraction of ammonia deposited on air preheater surfaces removed in high concentration wash water.
- $f_i$  = Generic correction factor
- $f_{InvC}$  = Factor for inventory capital costs (fraction of non-catalyst direct costs)
- $f_l$  = Fraction of flue gas exiting FGD system that leaks across gas-gas heat exchanger in tail-end SCR system
- $f_M$  = Factor for total maintenance cost (% non-catalyst total direct cost)
- $f_{NH_3}$  = Fraction of ammonia slip partitioned to fate i (e.g., "dep"=deposition on air preheater surfaces, "abs"=absorbed by flyash, "out"=emitted with flue gas)
- $f_{OL}$  = Factor for operating labor requirement (labor hours/train/day)
- $f_{ox}$  = Fraction of SO<sub>2</sub> oxidized to SO<sub>3</sub> in the SCR reactor.



- $f_{pp}$  = Factor for preproduction costs (fraction of TPI)  
 $f_{ProjC,i}$  = Project contingency factors for each process area (fraction of direct cost).  
 $f_r$  = Ratio of correction factors for linear velocity and temperature  
 $f_{Roy}$  = Factor for royalty costs (fraction of TPI).  
 FOC = Fixed operating cost (\$/yr)  
 $G_{fg}^o$  = Flue gas volumetric flow rate, referenced to a standard temperature (32 °F) and pressure (1 atm), ft<sup>3</sup>/hr.
- $G_{FG}$  = Flue gas volumetric flow rate, ft<sup>3</sup>/hr at standard conditions  
 $G_{scr}$  = Flue gas volumetric flow rate in the scr system, ft<sup>3</sup>/min  
 $h_i(T_j)$  = Enthalpy of gas i at temperature j, BTU/lbmole  
 $HHV_{NG}$  = Higher heating value of natural gas, BTU/lb  
 $H_r$  = Ratio of catalyst layer heights  
 $HR_g$  = Gross plant heat rate, BTU/kWh  
 $HR_s$  = Steam cycle heat rate, BTU/kWh  
 $k$  = Reaction rate constant. Also interpreted as catalyst "activity."  
 $LC$  = Catalyst life, years  
 $m$  = End-mole ratio  
 $m_i$  = Mass flow rate of species i, lb/hr  
 $M_i$  = Molar flow rate of species i, lbmole/hr  
 $m_{fg}$  = Mass flow rate of flue gas, lb/hr.  
 $m_{NG}$  = Mass flow rate of natural gas, lb/hr  
 $m_{NH_3,i}$  = Mass flow rate of ammonia injected into the flue gas, lb/hr  
 $M_{NH_3,A,i}$  = Molar flow rate of ammonia injected into the SCR system, lbmole/hr  
 $M_{steam}$  = Molar flow rate of steam required for ammonia injection, lbmole/hr  
 $MW_{fg}$  = Equivalent molecular weight of flue gas, lb/lbmole.  
 $n_d$  = Number of dummy catalyst layers  
 $N_{act}$  = Number of active catalyst layers at plant startup.  
 $N_C$  = Number of active catalyst layers  
 $N_{C,a}$  = Number of active catalyst layers replaced each year, yr<sup>-1</sup>  
 $N_{C,r}$  = Number of active catalyst layers replaced at each replacement interval  
 $N_I$  = Number of active catalyst layers installed initially  
 $N_{m,FOC}$  = Number of months of fixed operating costs required during startup  
 $N_{m,VOC}$  = Number of months of variable operating costs required during startup  
 $N_{sp}$  = Number of spare catalyst layers at plant startup.  
 $N_R$  = Number of reserve catalyst layers installed at end of first catalyst replacement interval  
 $N_{R,TOT}$  = Total number of reactor housings.  
 $N_{T,aph}$  = Total number of air preheaters  
 $N_{T,DB}$  = Total number of duct burners  
 $N_{T,GGH}$  = Total number of Gas-Gas Heat Exchangers.  
 $P_{fg}^o$  = Reference pressure of the flue gas at standard conditions, 1 atm.  
 PCI = *Chemical Engineering Plant Cost Index*. For Dec. 1989, PCI = 357.3.  
 $\dot{Q}_{fg}$  = Flue gas volumetric flow rate, actual ft<sup>3</sup>/min.  
 $Q_{NG}$  = Heating value of natural gas used in duct burners, BTU/hr.  
 $Q_r$  = Ratio of flue gas flow rates  
 $p_i$  = Partition factor within a partition factor matrix for trace species i

- $P_{i \rightarrow j}$  = Partition factor matrix for partitioning of trace species from stream  $i$  to stream  $j$ .  
 $r_e$  = Exit molar ratio of  $\text{NH}_3$  to  $\text{NO}_x$  at SCR reactor outlet.  
 $r_{ea}$  = Ratio of air to stoichiometric requirement (excess air ratio).  
 $R$  = Universal gas constant  
 $R_A$  = Ratio of ammonia to nitrogen oxides, molar basis  
 $R_g$  = Geometric ratio,  $\text{ft}^2/\text{ft}^3$   
 $R_S$  = Ratio of steam-to-ammonia, molar basis  
 $R_r$  = Ratio of reactor cross-section areas  
 $SV$  = Catalyst space velocity, 1/hr  
 $SV_{act}^o$  = Active catalyst space velocity, referenced to a flue gas temperature of 32 °F, 1/hr.  
 $SV_{ref}$  = Reference catalyst space velocity, 1/hr  
 $t$  = Time, seconds  
 $t_r$  = Catalyst replacement interval, hours  
 $T$  = Temperature, °R.  
 $T_{a,i}$  = Temperature of air entering a heat exchanger, °R.  
 $T_{a,o}$  = Temperature of air exiting a heat exchanger, °R.  
 $T_{fg}^o$  = Reference temperature of the flue gas at standard conditions, 32 °F.  
 $T_{fg,i}$  = Temperature of flue gas entering a heat exchanger, °R.  
 $T_{fg,o}$  = Temperature of flue gas exiting a heat exchanger, °R.  
 $TDC$  = Total direct cost (\$)   
 $TPC$  = Total plant cost (\$)   
 $TPI$  = Total plant investment (\$)   
 $TCR$  = Total capital requirement (\$)   
 $UA$  = Product of universal heat transfer coefficient and heat exchanger surface area,  $\text{BTU}/^\circ\text{R}$   
 $UC_{cat}$  = Unit cost of catalyst,  $\$/\text{ft}^3$ .  
 $UC_{elec}$  = Unit cost of electricity,  $\$/\text{kWh}$   
 $UC_L$  = Unit cost of labor,  $\$/\text{hour}$   
 $UC_{\text{NH}_3}$  = Unit cost of ammonia,  $\$/\text{ton}$   
 $UC_{steam}$  = Unit cost of steam,  $\$/1,000 \text{ lb}$   
 $V_c$  = Total catalyst volume,  $\text{ft}^3$   
 $V_{c,i}$  = Initially installed catalyst volume,  $\text{ft}^3$   
 $V_{c,a}$  = Volume of catalyst replaced each year,  $\text{ft}^3/\text{yr}$   
 $V_{c,r}$  = Volume of catalyst replaced at each replacement interval,  $\text{ft}^3/\text{interval}$   
 $V_{TOT}$  = Total volume of the catalyst, including active and spare layers,  $\text{ft}^3$ .  
 $VOC$  = Variable operating cost ( $\$/\text{yr}$ )  
 $[\text{NH}_3]_{out}$  = Ammonia slip, ppm  
 $[x]$  = Concentration of species  $x$ , molar basis

## A.2 Greek Letter Symbols

- $\beta_r$  = Ratio of fractions of reactor plugging  
 $\Delta H_r$  = Heat of reaction,  $\text{BTU}/\text{lbmole}$  reactant  
 $\Delta P_i$  = Pressure drop for component  $i$   
 $\eta_{\text{NO}_x}$  =  $\text{NO}_x$  removal efficiency, fraction

$\tau_a$  = Time constant for catalyst activity decay.

### A.3 Subscripts

cat = Catalyst  
DB = Duct burner  
duct = Ducting  
dum = Dummy catalyst  
FG = Flue gas  
FGD = Flue gas desulfurization  
GGH = Gas-gas heat exchanger  
i = in  
NG = Natural gas  
o = out  
org = organics  
scr = Selective catalytic reduction  
stm = steam  
tr = trace species  
WW = Wash water

## B.0 COMPUTER CODE FOR SCR PERFORMANCE AND COST MODEL

This appendix provides a sample of the computer code required to implement the performance and cost model of SCR in an ASPEN input file for calculating overall plant performance and in a FORTRAN cost model subroutine for calculating plant economics. Please see Frey and Rubin (1990) for detailed documentation of cost models for selected IGCC systems.

### B.1 Performance Model

The following computer code must be added to that of an ASPEN input file for a process technology. The example here is taken from modifications to the input file for the air-blown Lurgi-based IGCC system with hot gas cleanup.

#### B.1.1 Initializing Model Variables

This code is included as part of a FORTRAN block to initialize flowsheet variables:

```
F      COMMON/USRSCR/ SCRRE,  SCRAE,  XNH3S,  XNOXC,  AMIN,  TAU,
F      &                REPHRS, TEMPR,  TEMPA,  STMNH3, RNH3,  FNO,
F      &                REFHRS, SCRSO2, DPSCR,  SCRsvr,
F      &                NCATI,  NCATT,  NREP
;
;
;  DEFINE PH1 BVAR  BLOCK=HRSG1  SENTENCE=PARAM VAR=PRES
;  DEFINE PHSMIX BVAR  BLOCK=SCRMIX SENTENCE=PARAM VAR=PRES
;  DEFINE PHS BVAR  BLOCK=SCRRCT  SENTENCE=PARAM VAR=PRES
;  DEFINE PH2 BVAR  BLOCK=HRSG2  SENTENCE=PARAM VAR=PRES
;  DEFINE PH3 BVAR  BLOCK=HRSG3  SENTENCE=PARAM VAR=PRES
;  DEFINE PSG STREAM-VAR STREAM=STACKGAS SUBS=MIXED VAR=PRES
;  DEFINE POG STREAM-VAR STREAM=OUTGAS  SUBS=MIXED VAR=PRES
;
;  SCR Design Assumptions
;
;  SCRRE = SCR reference NOx removal efficiency, fraction
;  SCRAE = SCR actual removal efficiency, fraction
;  XNH3S = SCR NH3 Slip, ppm
;  XNOXC = Design NOx concentration, ppm (actual)
;  AMIN = Minimum relative catalyst activity
;  TAU = Catalyst activity curve time constant
;  NCATI = Number of initial catalyst layers
;  NCATT = Number of total catalyst layers
;  NREP = Number of catalyst layers replaced at a time
;  REPHRS = Replacement interval, hours
;  TEMPR = Reference temperature
;  TEMPA = Actual temperature, F
;  STMNH3 = Steam to ammonia ratio, volume
;  SCRSO2 = SCR catalyst SO2 oxidation rate
;
F      SCRRE = 0.80
F      SCRAE = 0.80
F      IF(SCRAE.GT.0.9) SCRAE = 0.9
F      XNH3S = 10
```

```

F      XNOXC = 42.0
F      AMIN  = 0.70
F      TIME1 = 8760.
F      ACT1  = 0.85
F      TAU   = 8000.
F      IF(ACT1.GT.AMIN) TAU = -TIME1/log((ACT1-AMIN)/(1.-AMIN))
F      REPHRS = 2.*0.65*8760.
F      IF(REPHRS.GT.20000.) REPHRS = 20000.
F      IF(REPHRS.LT.1000.) REPHRS = 1000.
F      REFHRS = 3.*0.65*8760.
F      TEMPR  = 717.
F      TEMPA  = 717.
F      STMNH3 = 19.
F      SCRSO2 = 0.01
F      SCRSVR = 10000.
F      NCATI  = 3
F      NCATT  = 3
F      NREP   = 1
;

```

The code here initializes the pressure drops in the HRSG, including that associated with the SCR system.

```

F      DPHRSG1 = 1./3. * 20./27.68
F      DPHRSG2 = 1./3. * 20./27.68
F      DPHRSG3 = 1./3. * 20./27.68
F      DPSCR   = 0.
F      IF(SCRAE.GT.0.) DPSCR = 4.0/27.68
;
F      POG     = APRES
F      PSG     = POG
F      PH3     = PSG
F      PH2     = PH3 + DPHRSG3
F      PHS     = PH2 + DPHRSG2
F      PHSMIX  = PHS
F      PH1     = PHS + DPSCR
F      GTTPO   = PH1 + DPHRSG1
;

```

### B.1.2 Modifications to the Gas Turbine Power Generation Section

The gas turbine flowsheet must be modified to include a stoichiometric reactor for the SCR system. New stream variables must be defined for ammonia and steam. The steam required for ammonia vaporization is taken from the plant steam cycle.

A FORTRAN block, SETSCR, is added to determine the flowrates of ammonia and steam, and to set chemical reaction conversion rates in the stoichiometric reactor, SCRRECT.

The SCR reactor is inserted in the middle of the HRSG at a point where the exhaust gas temperature is within the SCR temperature window.

```

;
; Gas-side HRSG
;
HRSG1      IN=GTPOC                OUT=HRSGGAS1 QHRSG1
SCRMIX     IN=HRSGGAS1 NH3SCR STMSCR OUT=SCRIN
SCRRCT     IN=SCRIN                OUT=SCROUT
HRSG2      IN=SCROUT              OUT=HRSGGAS2 QHRSG2
HRSG3      IN=HRSGGAS2            OUT=STACKGAS QHRSG3
STACK      IN=STACKGAS            OUT=OUTGAS
;
STREAM NH3SCR
; Ammonia injected upstream of SCR Unit. Ammonia flow
; set by SETSCR
SUBSTREAM MIXED T=59 PRES=14.95
MOLEFLOW NH3 1.0
;
STREAM STMSCR ; SCR NOX CONTROL STEAM
SUBSTREAM MIXED T=460 P=465
MASSFLOW H2O 1. ; SET BY FORTRAN BLOCK SETSCR

BLOCK SCRMIX MIXER
PARAM NPK=1 KPH=1 MAXIT=100
;
BLOCK SCRRCT RSTOIC
;
; Selective Catalytic Reduction Reactor
; Added by H.C. Frey, 2/19/92
; Reaction conversions set by fortran block SETSCR
;
PARAM PRES=14.95 T=717.
STOI 1 MIXED NH3 -2 / NO -1 / NO2 -1 / N2 2 / H2O 3 /
      2 MIXED NH3 -4 / NO -4 / O2 -1 / N2 4 / H2O 6 /
      3 MIXED SO2 -2 / O2 -1 / SO3 2
CONV 1 MIXED NO2 0.8 /
      2 MIXED NO 0.79121 /
      3 MIXED SO2 0.0
;
SEQUENCE GT-SEQ3 &
HRSG1 SETSCR SCRMIX SCRRCT HRSG2 HRSG3 &
GT-SPEC4 GT-AIRH1 GT-BOOST (RETURN GT-SPEC4) (SEQUENCE SC-SEQ)
;
FORTRAN SETSCR
;
; H.C. Frey, NC State Univ., 10/17/94
; Set performance parameters for SCR system:
; Ammonia injection ratio
; Ammonia injection mass flow
; Steam mass flow for ammonia injection
; NO and NO2 removal efficiency
; SO2 Oxidation Rate
;
F COMMON/USRSCR/ SCRRE, SCRAE, XNH3S, XNOXC, AMIN, TAU,
F & REPHRS, TEMPR, TEMPA, STMNH3, RNH3, FNO,
F & REFHRS, SCRSO2, DPSCR, SCRSVR,
F & NCATI, NCATT, NREP
DESC 'SET FLOW OF NOX CONTROL AMMONIA AND STEAM TO SCR'

```

```

IN-UNITS ENG
DEFINE STMCR SUBSTREAM-VAR STREAM=STMSCR SUBS=MIXED VAR=MOLEFLOW
DEFINE XNH3CR SUBSTREAM-VAR STREAM=NH3SCR SUBS=MIXED VAR=MOLEFLOW
DEFINE GASFL STREAM-VAR STREAM=HRSGGAS1 VAR=MOLEFLOW
DEFINE FNOIN MOLE-FLOW STREAM=HRSGGAS1 SUBS=MIXED COMP=NO
DEFINE FNO2IN MOLE-FLOW STREAM=HRSGGAS1 SUBS=MIXED COMP=NO2
DEFINE CNO2 BVAR BLOCK=SCRRCT SENTENCE=CONV &
      VARIABLE=CONV ID1=1
DEFINE CNO BVAR BLOCK=SCRRCT SENTENCE=CONV &
      VARIABLE=CONV ID1=2
DEFINE CSO2 BVAR BLOCK=SCRRCT SENTENCE=CONV &
      VARIABLE=CONV ID1=3
F CNOXIN = (FNOIN+FNO2IN)/GASFL*1.D6
F IF(SCRAE.GT.0.) THEN
F RNH3 = SCRAE + XNH3S/CNOXIN
F CSO2 = SCRSO2
F ELSE
F RNH3 = 0.
F CSO2 = 0.
F ENDIF
F XNH3CR = RNH3 * (FNOIN + FNO2IN)
F STMCR = STMNH3 * XNH3CR
F CNO2 = SCRAE
F CNO = (SCRAE*FNO-SCRAE*(1-FNO))/FNO
C F CNO = SCRAE * (2 * FNO - 1)/((1 + SCRAE) * FNO - SCRAE )
F WRITE(NRPT,871) CNOXIN, RNH3, XNH3CR, STMCR, CNO2, CNO, CSO2,
F & SCRAE, XNH3S, FNOIN, FNO2IN, FNO, SCRSO2
F 871 FORMAT('Inlet NOx concentration: ',F12.4,/
F & 'NH3/NOx Ratio: ',F12.4,/
F & 'Ammonia Molar Flow Rate ',F12.4,/
F & 'Steam Injection Flow ',F12.4,/
F & 'NO2 Removal Efficiency ',F12.4,/
F & 'NO Removal Efficiency ',F12.4,/
F & 'SO2 Oxidation Rate ',F12.4,/
F & 'NOx Removal Efficiency ',F12.4,/
F & 'Ammonia Slip ',F12.4,/
F & 'NO Flowrate in ',F12.4,/
F & 'NO2 Flowrate in ',F12.4,/
F & 'Fraction NOx as NO ',F12.4,/
F & 'SO2 oxidation from commo ',F12.4)
;

```

### B.1.3 FORTRAN Block to Call Cost Model

A FORTRAN block at the end of the input file is used to initialize cost model variables, call the cost model, and obtain cost model results. Such FORTRAN blocks have been previously documented by Frey and Rubin (1990). Shown here are the modifications required to accommodate the SCR performance and cost models.

```

FORTRAN ALHCOST
;
; Author: H.C. Frey, Carnegie-Mellon University
; Date: September 15, 1990
; Modified: H.C. Frey, NC State Univ., October 17, 1994
; Description: This block accesses the key variables required to

```

```

; estimate capital and annual costs, and then calls subroutine
; USRALHC (USer Air-blown Lurgi-based igcc with Hot gas cleanup
; Cost model).
;
; All key parameters in the cost model are assigned values
; in the main program Fortran block. Therefore, it is not necessary
; to make any changes to the cost model subroutine when redefining
; case studies.
;
F COMMON /USRSCR/ SCRRE, SCRAE, XNH3S, XNOXC, AMIN, TAU,
F & REPHRS, TEMPR, TEMPA, STMNH3, RNH3, FNO,
F & REFHRS, SCRSO2, DPSCR, SCRSCR,
F & NCATI, NCATT, NREP
;
; access flowsheet variables for SCR
;
DEFINE SCRGAS STREAM-VAR STREAM=HRSGGAS1 VAR=MOLEFLOW
DEFINE SCRTEMP SUBSTREAM-VAR STREAM=HRSGGAS1 SUBS=MIXED VAR=TEMP
DEFINE SCRPRS SUBSTREAM-VAR STREAM=HRSGGAS1 SUBS=MIXED VAR=PRES
DEFINE SCRNO MOLE-FLOW STREAM=HRSGGAS1 SUBS=MIXED COMP=NO
DEFINE SCRNO2 MOLE-FLOW STREAM=HRSGGAS1 SUBS=MIXED COMP=NO2
DEFINE SCRNH3 STREAM-VAR STREAM=NH3SCR VAR=MOLEFLOW
DEFINE SCRSTM STREAM-VAR STREAM=STMSCR VAR=MOLEFLOW
;
; ----- PROCESS CONTINGENCY FACTORS -----
; The following are process contingency factors for each process
; area. (Only the SCR variables are shown here)
;
F FPCCR = 0.100
;
; ----- MAINTENANCE COST FACTORS -----
; Maintenance cost factors, based on plant section capital cost
; (Only the SCR variables are shown here)
;
F FMCCR = 0.02
;
; -----UNIT COSTS AND PRICES -----
; (Only the SCR variables are shown here)
;
F BCSCRC = 250.00
F BCNH3 = 150.00

```

## B.2 Cost Model

Shown here are those parts of the cost model which are affected by the addition of SCR. For complete cost model documentation, see Frey and Rubin (1990).

```

SUBROUTINE USRALHC15
IMPLICIT DOUBLE PRECISION (A-H,O-Z)

```

```

C Revision 1.51 H.C. Frey Date: Oct. 17, 1994
C
C Short Title: Cost Model for Air-blown Lurgi-IGCC With Hot Gas
C Cleanup
C Description: Capital and annual cost model of Lurgi-based

```



C integrated gasification combined cycle system with hot gas  
 C cleanup. Based on  
 C ASPEN input file developed by Klara, Rastogi, Regenhardt, Stone,  
 C and Craig over time.  
 C Cost model documented in Frey and Rubin (1990).

(Only the SCR variables are shown here)

C  
 COMMON /USRSCR/ SCRRE, SCRAE, XNH3S, XNOXC, AMIN, TAU,  
 & REPHRS, TEMPR, TEMPA, STMNH3, RNH3, FNO,  
 & REFHRS, SCRSO2, DPSCR, SCRSVR,  
 & NCATI, NCATT, NREP

C  
 C ----- UNIT COSTS FOR FUEL, CHEMICALS, AND CATALYSTS -----  
 C The following are unit costs for fuels, chemicals, and catalysts.  
 C The costs are adjusted to  
 C other years using the Chemical Engineering magazine industrial  
 C chemicals producers prices index.

C  
 UCSCRC = BCSCRC \*(CICPPI/411.25)  
 UCNH3 = BCNH3 \*(CICPPI/411.25)

C  
 C ----- DIRECT CAPITAL COSTS -----

C  
 C Direct capital cost of Selective Catalytic Reduction (SCR)

C  
 NOCR = 0  
 NTCR = 0  
 FEFF = 0  
 ACTAVE = 0.  
 VACTI = 0.  
 VCATT = 0.  
 VREF = 1.  
 XNCANN = 0.  
 VCATAN = 0.  
 DCCR = 0.  
 IF (SCRAE.GT.0.) THEN  
 NOCR = NOHR  
 NTCR = NTHR  
 FEFF = Log(1-SCRRE)/Log(1-SCRAE)  
 ENDIF  
 XNOXCA = (SCRNO+SCRNO2)\*1.D6/SCRGAS  
 SCREMR = XNH3S/(XNOXCA\*(1-SCRAE))  
 REFEMR = XNH3S/(XNOXC\*(1-SCRRE))  
 FRE = (SCREMR/REFEMR)\*\*0.3

C  
 C Calculate catalyst activity:  
 C 1. Calculate average long term steady state activity  
 C 2. The number of layers installed initially may be less  
 C than the number of layers installed finally. Therefore,  
 C calculate activity at end of first interval before  
 C an additional layer is installed (NCATT = NCATI + 1)  
 C 3. If the catalyst activity at the end of the first  
 C interval is less than the long term average, then  
 C delayed addition is not feasible and all layers must  
 C be installed initially. Therefore, set NCATI = NCATT

C

```

IF (NCATI.LT.(NCATT-1)) NCATI = NCATT - 1
IF (NCATT.LT.1) NCATT = 1
IF (NCATI.LT.1) NCATI = 1
ACT2 = 0.0
ACT3 = 0.0
ACT4 = 0.0
ACT1 = AMIN + (1-AMIN)*EXP(- 1.* REPHRS/TAU)
IF (NCATT.GE.2) ACT2 = AMIN + (1-AMIN)*EXP(- 2.* REPHRS/TAU)
IF (NCATT.GE.3) ACT3 = AMIN + (1-AMIN)*EXP(- 3.* REPHRS/TAU)
IF (NCATT.GE.4) ACT4 = AMIN + (1-AMIN)*EXP(- 4.* REPHRS/TAU)
ACT1I = ACT1
ACTREF = AMIN + (1-AMIN)*EXP(-REFHRS/TAU)
IF (NCATI.GT.0) ACTAVI = ACT1I
IF (NCATT.GT.0) ACTAVE = (ACT1 + ACT2 + ACT3 + ACT4)/NCATT
IF (NCATI.LT.NCATT) THEN
  IF (ACTAVI.LT.ACTAVE) NCATI = NCATT
ENDIF
FACT = ACTAVE/ACTREF
FTEMP = EXP(-7180./(TEMPR+460.))/EXP(-7180./(SCRTMP+460.))
SCRSV = SCRSVR * FEFF * FRE * FACT * FTEMP
GSCR = SCRGAS * 10.73 * 520 / 14.7
IF (SCRSV.GT.0.) THEN
  VCATI = (GSCR/SCRSV) * (NCATI/NCATT)
  VCATT = (GSCR/SCRSV)
  VREF = (GSCR/SCRSVR)
ENDIF
IF (NREP.GT.1) NREP = 1
IF (NREP.LT.1) NREP = 1
IF (REPHRS.GT.0.) XNCANN = NREP * 8760. * CF / REPHRS
IF (NCATT.GT.0) VCATAN = VCATT * (XNCANN/NCATT)
XLCAT = (NCATT/NREP)*REPHRS/(8760.*CF)
IF (SCRAE.GT.0.) DCCR = NICR * (1767.0 *
& (VCATT/(NOCR*4320))**0.637 + ERRCR)* (CI/354.7)
XNH3IN = SCRNH3 * 17.03 / 2000. * 8760. * CF

```

C

C

C

C

C

C

C

C

C

C

C

----- CAPITAL COSTS: TOTAL PLANT INVESTMENT -----

The total plant investment is estimated in this section. The total capital requirement, which includes items dependent on operating costs, is not calculated until after the operating costs are estimated.

Total Direct Cost (TDC)

```

TDC= DCCH+DCL +DCOF+DCG +DCSS+DCCY+DCZF+DCSA+DCS +DCBF
& +DCGT+DCHR+DCCR+DCST+DCGF

```

C

C

C

Total Indirect Cost (TIC)

```

CICC=FICC*TDC
CTAX=RTAX*(0.8*TDC+0.1*CICC)
CEHO=FEHO*(TDC+CICC+CTAX)
TIC=CICC+CTAX+CEHO+CEP

```

C

C

C

Process Contingency Costs

```

CPCCR = FPCCR*DCCR*(1.+TIC/TDC)

```

C  
CPC=CPCCH+CPCL+CPCOF+CPCSS+CPCCY+CPCG+CPCS+CPCZF+CPCSA  
& +CPCBF+CPCGT+CPCHR+CPCCR+CPCST+CPCGF

C  
C Project contingency cost

C  
C CPJ=FPJ\*(TDC+TIC+CPC)

C  
C Total Plant Cost (TPC)

C  
C TPC=TDC+TIC+CPC+CPJ

C  
C Total plant investment (TPI)

C  
Z=(1+INTRST)/(1+EREAL)  
IF(Z.LE.1) THEN  
AF = 1  
ELSE  
AF=(Z\*\*NOYEARS-1)/(NOYEARS\*(Z-1))  
ENDIF  
TPI=AF\*TPC  
AFDC=TPI-TPC

C  
C ----- NET PLANT ELECTRICAL OUTPUT -----

C  
C The net electrical output must be estimated in order to estimate  
C a number of plant costs. The gross power plant electrical output  
C is reduced by plant auxiliary loads, including coal handling, the  
C oxygen plant, the gasification section, limestone handling,  
C sulfation, general facilities, and power consumption in the  
C steam cycle. The pump and compressor power consumption  
C is estimated directly from the ASPEN simulation. The other areas  
C are estimated from regression relationships taken from Frey and  
C Rubin (1990).

C  
C Gas Turbine Output: WGTE                  Steam Turbine Output: WSTE

C  
C The auxiliary loads have been converted to MW from Frey and Rubin  
C (1990)

C  
WSCR = 3.47D-5 \* XNH3IN \* 2000./(8760. \* CF)  
WAUXE = WCHE+WLE+WOFE+WGE+WSE+WSAE+WBF+WGFE+WSCR  
WNETE = WGTE+WSTE-WAUXE

C  
C ----- INITIAL CATALYSTS AND CHEMICALS -----

C  
C Total cost of initial catalyst and chemicals  
C (These costs are in dollars, and must be divided by 1,000 for  
C conversion to thousands of dollars)

C  
TCICC=(UCFUEL\*FUELI+CCHICW+UCZFSO\*TSCZF+  
& UCCOKE\*CICOKE+UCBNTN\*CIBENT+UCSCRC\*VCATI)/1000.

C  
C ----- FIXED OPERATING COSTS -----  
C

C Fixed Operating Costs: Operating Labor  
C

C NOPERS=15+2\*NOG+4\*NOGT  
C OCL=ALABOR\*2080\*NSHIFTS\*NOPERS

C Fixed Operating Costs: Maintenance  
C

OCMCR = FMCCR \* (DCCR \* (1+(TIC+CPJ)/TDC) + CPCCR)

C  
C OCM = (OCMOF+OCMCH+OCML+OCMG+OCMS+OCMZf+OCMBF+  
C & OCMSS+OCMCY+OCMSA+OCMCC+OCMCR+OCMGF)\*1000  
C OCMM=0.60\*OCM  
C OCML=0.40\*OCM

C Fixed Operating Costs: Administration and Supervision  
C

OCAS=0.30\*(OCL+OCML)

C Total fixed operating costs  
C

FOC=OCL+OCM+OCAS

C  
C ----- VARIABLE OPERATING COSTS -----  
C

C Estimate of consumable operating costs  
C

OCSCRC = UCSCRC \* VCATAN  
OCNH3 = UCNH3 \* XNH3IN

C  
C OCCONS = OCSAI + OCNAOH + OCNA2H + OCHYDR + OCMORP + OCLIME +  
C & OCSODA + OCCORI + OCSURF + OCCHLR + OCBIOC + OCZFSO +  
C & OCLMST + OCFOIL + OCPLTA + OCRAWW + OCSCRC + OCNH3 +  
C & CWW + OCFLRL + OCSACT + OCCOKE + OCBENT

C TOTAL VARIABLE OPERATING COST (No byproduct sales)  
C

VOC = OCFUEL + OCCONS + OCASH - OCBYP

C  
C ----- PREPRODUCTION CAPITAL COSTS -----  
C

PPFC = FOC/(12000)  
PPOC = (0.083/CF) \* (OCCONS+OCASH)/1000  
PPFUEL=(0.021/CF) \* OCFUEL/1000  
PPC=PPFC+PPOC+PPFUEL+0.02\*TPI

C  
C ----- Inventory Capital Costs -----  
C

IC=(0.164/CF) \* (OCFUEL+OCCONS)/1000

C  
C  
C

----- Total Capital Requirement (TCR) -----

ALAND=-93+0.065\*QCOAL/1000000  
TCLAND=ALAND\*UCLAND/1000  
TCR=1.01\*TPI+PPC+IC+TCICC+TCLAND

## C.0 SAMPLE MODEL OUTPUT

Shown here is a sample cost model output for an Air-Blown Fixed-Bed Gasifier IGCC with SCR. This is the same output used to estimate the High Fuel NO<sub>x</sub> Case results with SCR, as described in Section 8 of the report.

Air-Blown Fixed-Bed Gasifier IGCC w/SCR: Deterministic Analysis  
SYSTEM SUMMARY

### \*\*\* GASIFIER CONDITIONS \*\*\*

COAL FLOW RATE: 0.560774E+06 LB/HR  
SORBENT FLOW RATE: 0.376997E+00 LB/HR  
AIR FLOW RATE: 0.137681E+07 LB/HR  
REQUIRED STEAM: 0.688406E+06 LB/HR  
SUPPLIED STEAM: 0.998983E-03 LB/HR  
SO<sub>2</sub> FLOW RATE: 0.999983E-06 LB/HR  
ASH FLOW RATE: 49093.911185 LB/HR  
GASIFIER PRESSURE: 307.5 PSIA  
GASIFIER TEMPERATURE: 1100.0 F

ZINC FERRITE SULFUR REMOVAL PROCESS  
RESULTS GENERATED ON OCT 17, 1994 AT 15:18

### \*\* SO<sub>2</sub> BALANCE (LB/HR) \*\*

H<sub>2</sub>S IN RAWGAS 19669.24  
H<sub>2</sub>S IN FUELGAS 41.31  
SO<sub>2</sub> IN SO<sub>2</sub>GAS 36896.25  
PPMV OF H<sub>2</sub>S IN FUELGAS 11.

### \*\* FLOW RATES (LB/HR) \*\*

RAWGAS 0.26D+07 MW = 23.11  
FUELGAS 0.26D+07 MW = 23.08  
SO<sub>2</sub>GAS 547120.81 MW = 20.35  
STEAM REQUIRED TO MAINTAIN  
CO/CO+CO<sub>2</sub> RATIO (SHFT-STM) 1.00  
DILUENT REQUIRED TO MAINTAIN  
REGENERATOR TEMPERATURE (GEN-STM) 407845.64  
AIR FLOW TO REGENERATOR (REGENAIR) 133094.60

### \*\* GENERAL \*\*

CO/(CO+CO<sub>2</sub>) RATIO: 0.281  
SULFUR CAPTURE EFFICIENCY: 0.998  
TEMPERATURE OF THE REGENERATED GAS: 1450.2  
VOLUME PERCENT AIR IN REGENERATION GAS: 16.766

### \*\*\* MS7000F GAS TURBINE CONDITIONS \*\*\*

FUEL FLOW RATE: 0.256371E+07 LB/HR  
AIR FLOW RATE: 0.988548E+07 LB/HR  
AIR DIVERTED TO GASIFIER: 0.137681E+07 LB/HR

AIR DIVERTED TO ZINC.FER: 0.133092E+06 LB/HR  
 FUEL LHV (WET BASIS): 1729.4 BTU/LB, 105.5 BTU/SCF  
 FUEL HHV (WET BASIS): 1938.9 BTU/LB, 117.9 BTU/SCF  
 FIRING TEMPERATURE: 2350.0 F  
 COMBUSTOR EXIT TEMPERATURE: 2436.8 F  
 TURBINE EXHAUST TEMPERATURE: 1134.7 F  
 THERMAL EFFICIENCY (LHV): 0.4315  
 GENERATOR EFFICIENCY: 0.9850  
 GASIFIER AIR INTERCOOLER DUTY: 0.000000E+00 BTU/HR

Ratio of Inlet Air to Exhaust Gas: 0.902  
 Ratio of Boost Air to Inlet Air: 0.153  
 Ratio of Gasif Air to Inlet Air: 0.139  
 Ratio of Regen Air to Inlet Air: 0.013  
 Ratio of Fuel Gas to Inlet Air: 0.259  
 CO2 Emissions: 194.3 LB/MMBTU  
 CO Emissions: 0.32280 LB/MMBTU  
 CO Emissions: 186.08 ppmv  
 SO2 Emissions: 0.01217 LB/MMBTU  
 NOx Emissions: 0.25813 LB/MMBTU  
 NOx Emissions: 90.57 ppmv  
 Oxygen in Exhaust: 9.9045 percent  
 Vapor in Exhaust: 15.7150 percent  
 CO2 in Exhaust: 7.1294 percent

\*\*\* STEAM TURBINE CONDITIONS \*\*\*

STEAM DIVERTED TO GASIFIER: 0.688406E+06 LB/HR  
 STEAM DIVERTED TO ZINC FER: 0.407839E+06 LB/HR  
 STEAM TURBINE FLOW RATE: 0.128315E+07 LB/HR  
 SUPERHEATED STEAM TEMPERATURE: 1000.0 F  
 EXPANDED STEAM QUALITY: 0.9000  
 ISENTROPIC EFFICIENCY: 0.9227  
 GENERATOR EFFICIENCY: 0.9850

\*\*\* SUMMARY OF SAMPLED ASPEN FLOWSHEET PERFORMANCE PARAMETERS \*\*\*

PLANT SECTION	FLOWSHEET PERFORMANCE PARAMETERS	VALUE	UNITS
Coal Feed	Mass flow of coal to gasifier	560783.36	lb/hr
Limestone Feed	Mass flow of limestone (mixed)	0.00	lb/hr
	Mass flow of limestone (cisolid)	0.38	lb/hr
Oxidant Feed	Work to boost air compressor	0.138477E+05	Watts
	First precooler inlet air temp.	750.33	F
	First precooler outlet air temp.	679.51	F
	First precooler inlet BFW temp.	505.54	F
	First precooler outlet BFW temp.	605.29	F
	Heat transfer in first precooler	0.276352E+08	BTU/hr
	Heat trans. to BFW from HRSG	0.572365E+09	BTU/hr
	Heat trans. to BFW from Regen.	0.181136E+09	BTU/hr
Gasification	Heat leaving from economizer	0.609957E+09	BTU/hr
	Gasifier output syngas pres.	307.50	psia
	Gasifier output syngas temp.	1100.00	F
	Gasifier output syngas density	0.183078E-01	lbmole/ft3
Zinc Ferrite	Gasifier recycle fines flow rate	18646.05	lb/hr
	Zinc Ferrite inlet syngas flow	111094.47	lbmole/hr
	Off-gas to sulfuric acid plant	26885.10	lbmole/hr

	SO2 to sulfuric acid plant	575.94	lbmole/hr
	Off-gas temp. to sul. acid plant	1233.45	lbmole/hr
	Zinc ferrite inlet H2S flow	577.15	lbmole/hr
	Zinc ferrite inlet COS flow	0.00	lbmole/hr
Sulfation	Gasifier ash removal	49094.74	lb/hr
	Gasifier ash removal	8794.46	lb/hr
	Gasifier fines removal	245.34	lb/hr
	Gasifier fines removal	736.03	lb/hr
BFW Treating	Raw water to power plant	1189633.05	lb/hr
	Steam turbine condensate	793299.96	lb/hr
Gas Turbine	Gas turbine inlet air flow	9885645.28	lb/hr
Steam Cycle	HRSG outlet HP steam pres.	1465.00	psia
	HRSG outlet HP steam flow	1283176.10	lb/hr
	Gas turbine net shaft work	-.570841E+09	Watts
	Steam turbine net shaft work	-.181810E+09	Watts
	Steam cycle auxiliary power	0.353504E+07	Watts
	Acid gas auxiliary power	0.000000E+00	Watts
Miscellaneous	Moisture in coal feed	0.12	wt %
	Ash in coal	0.10	wt %
	Temp. of ambient air	59.00	F
	Heating value of coal	12774.00	BTU/lb
SCR	SCR Inlet Gas Flow	390819.35	lbmole/hr
	SCR Inlet Gas Temperature	717.00	lbmole/hr
	SCR Inlet Gas Pressure	15.33	lbmole/hr
	SCR Inlet NO loading	168.14	lbmole/hr
	SCR Inlet NO2 loading	8.85	lbmole/hr
	SCR Ammonia Requirement	145.50	lbmole/hr
	SCR Dilution Steam Requirement	2764.54	lbmole/hr
Environmental	High Pressure Blowdown	26187.27	lbmole/hr
	Low Pressure Blowdown	26187.27	lbmole/hr
	CO2 from gas turbine	27863.06	lbmole/hr
	CO from gas turbine	72.72	lbmole/hr
	SO2 from gas turbine	1.20	lbmole/hr
	COS from gas turbine	0.00	lbmole/hr
	CH4 from gas turbine	0.00	lbmole/hr
	H2S from gas turbine	0.00	lbmole/hr
	NH3 to gas turbine	396.37	lbmole/hr
	NO from gas turbine	33.63	lbmole/hr
	NO2 from gas turbine	1.77	lbmole/hr
	Actual coal heating value	11248.73	BTU/lb
	COST VAR WARNING ---- Variable MCFGI value of	560783.362	
	in DCSS above the upper limit of	433000.000	
	COST VAR WARNING ---- Variable VSNZFI value of	9194.144	
	in DCCY above the upper limit of	6000.000	
	COST VAR WARNING ---- Variable MRW value of	1189633.049	
	in DCBF above the upper limit of	614000.000	

#### COST SUMMARY

Air Blown Lurgi-Based IGCC System with Hot Gas Cleanup

#### A. KEY INPUT ASSUMPTIONS -----

##### Performance Assumptions:

Gasifier Availability:	0.87	Sorb. Sulfur Loading:	0.168
Max. Desulf. Vessel Diameter:	12.50	Superficial Velocity:	2.00
L/D Ratio:	3.00	Absorption Cycle Time:	30.0



Maximum Space Velocity (1/hr): 2000. Sorbent Bulk Density: 82.0  
 Sorb. Attrition Rate/80 cycles: 0.800 Generator Efficiency: 0.9850  
 Gasifier Coal Throughput: 1.01

Economic Assumptions:

Cost Year: January 1989	Inflation Rate: 0.000
Plant Cost Index: 351.5	Real Escalation Rate: 0.000
Chemicals Cost Index: 411.3	Plant Booklife: 30
Plant Capacity Factor: 0.65	Sales Tax Rate: 0.05
General Facilities Factor: 0.20	Real Return on Debt: 0.046
Indirect Construction Factor: 0.20	Real Ret. on Pref.: 0.052
Engr & Home Office Fees: 0.10	Real Ret. on Equity: 0.087
Project Contingency Factor: 0.17	Debt Ratio: 0.500
Byproduct Marketing Factor: 0.10	Pref. Stock Ratio: 0.150
Average Operating Labor Rate: 19.70	Fed. & State Taxes: 0.380
Number of Shifts: 4.25	Investment Tax Credit: 0.000
Construction Interest Rate: 0.10	Prop. Taxes & Insur.: 0.020
Years of Construction: 4	

Process Contingency and Maintenance Cost Factors:

Plant Section	Process Contingency	Maintenance Cost Factor
Coal Handling	0.050	0.030
Limestone Handling	0.000	0.030
Oxidant Feed	0.100	0.020
Gasification	0.200	0.030
Coke, Ash, & Bent. Subsystems	0.050	0.020
High Temp. Cyclones	0.050	0.030
Zinc Ferrite	0.400	0.030
Sulfuric Acid Plant	0.100	0.020
Sulfation	0.600	0.045
Boiler Feedwater Treatment	0.000	0.015
Gas Turbine	0.250	0.020
Heat Recovery Steam Generator	0.025	0.015
Selective Catalytic Reduction	0.100	0.025
Steam Turbine	0.015	0.050
General Facilities	0.015	

B. CALCULATED DIRECT CAPITAL AND PROCESS CONTINGENCY COSTS (\$1,000) --

Plant Section	Number of Units Operating	Total Capital	Direct Cost	Process Contingency
Coal Handling	1	1	22640.	1548.
Limestone Handling	0	0	0.	0.
Oxidant Feed	3	3	4854.	664.
Gasification	11	13	86034.	23531.
Coke, Ash, & Bent. Subsystems	11	13	15624.	1068.
High Temp. Cyclones	22	26	6414.	439.
Zinc Ferrite	11	26	12542.	6861.
Sulfuric Acid Plant	1	1	25083.	3430.
Sulfation	0	0	0.	0.
Boiler Feedwater Treatment	1	1	5832.	0.
Gas Turbine	3	3	96000.	32821.
Heat Recovery Steam Generator	3	3	28573.	977.
Selective Catalytic Reduction	3	3	8679.	1187.
Steam Turbine	1	1	28421.	972.

General Facilities                      N/A    N/A                      68139.                      4659.

C. CALCULATED TOTAL CAPITAL REQUIREMENT (\$1,000) -----

Description	Capital Cost
Total Direct Cost	408834.
Indirect Construction Cost	81767.
Sales Tax	16762.
Engineering and Home Office Fees	50736.
Environmental Permitting	1000.
Total Indirect Costs	150265.
Total Process Contingencies	78156.
Project Contingency	111520.
Total Plant Cost	748774.
AFDC	119991.
Total Plant Investment	868766.
Preproduction (Startup) Costs	24057.
Inventory Capital	20327.
Initial Catalysts and Chemicals	33680.
Land	2605.
TOTAL CAPITAL REQUIREMENT (\$1,000) ----->	958123.

D. CALCULATED FIXED OPERATING COSTS (\$/year) -----

Description	Annual Cost
Operating Labor	8533252.
Maintenance Costs	16285642.
Administration and Supervision	4514253.
TOTAL FIXED OPERATING COST (\$/year) ----->	29333147.

E. CALCULATED VARIABLE OPERATING COSTS -----

1. CONSUMABLES (\$/year)

Description	Assumed Unit Cost	Calc. Material Requirement	Calc. Annual Operating Cost
Sulfuric Acid:	110.00 \$/ton	2344.2 ton/yr	257859.
NaOH:	220.00 \$/ton	488.4 ton/yr	107439.
Na2 HPO4:	0.70 \$/lb	2866.2 lb/yr	2006.
Hydrazine:	3.20 \$/lb	13798.6 lb/yr	44156.
Morpholine:	1.30 \$/lb	12877.2 lb/yr	16740.
Lime:	80.00 \$/ton	580.6 ton/yr	46452.
Soda Ash:	160.00 \$/ton	643.5 ton/yr	102961.
Corrosion Inh.:	1.90 \$/lb	115612.5 lb/yr	219664.
Surfactant:	1.25 \$/lb	115612.5 lb/yr	144516.
Chlorine:	250.00 \$/ton	18.3 ton/yr	4573.
Biocide:	3.60 \$/lb	20336.0 lb/yr	73209.
SCR Catalyst:	250.00 \$/ft^3	4750.6 ft^3/yr	1187645.
Ammonia:	150.00 \$/ton	7054.6 ton/yr	1058187.
Zinc Fer Sorb:	3.00 \$/lb	2952898.6 lb/yr	8858696.
Limestone:	18.00 \$/ton	0.0 ton/yr	0.
Sul Acid. Cat.:	1.90 \$/l	29573.6 l/yr	56190.
Coke:	150.00 \$/ton	1574.6 ton/yr	236192.
Bentonite:	0.03 \$/lb	4565335.7 lb/yr	132395.
Fuel Oil:	42.00 \$/bbl	40550.4 bbl/yr	1703115.
Plant Air Ads.:	2.80 \$/lb	3041.3 lb/yr	8516.
Raw Water:	0.73 \$/Kgal	811984.0 Kgal/yr	592748.
Waste Water:	840.00 \$/gpm ww	0.0 lb/hr	0.
LPG - Flare:	11.70 \$/bbl	3548.2 bbl/yr	41513.

TOTAL CONSUMABLES (\$/year) -----> 14894772.

2. FUEL, ASH DISPOSAL, AND BYPRODUCT CREDIT (\$/year)

Coal:	1.61 \$/MMBtu	560783.4 lb/hr	65669751.
Ash Disposal:	10.00 \$/ton	706.4 ton/day	1676045.
Sulfuric Acid:	40.00 \$/ton	172784.8 ton/yr (	6220253.)

TOTAL VARIABLE OPERATING COST (\$/year) -----> 76020315.

F. CALCULATED COST OF ELECTRICITY -----

Power Summary (MWe)		Auxiliary Loads (MWe)			
Gas Turbine Output	562.28	Coal Handling	1.06	Sulfation	0.00
Steam Turbine Output	179.08	Limestne Hdlg	0.00	Acid Rem.	1.70
Total Auxiliary Loads	26.53	Oxidant Feed	13.85	Steam Cycle	3.54
Net Electricity	714.83	Gasification	2.80	SCR	0.09
		Zinc Ferrite	0.00	General Fac	3.50

Capital Cost:	1340.35 \$/kW
Fixed Operating Cost:	41.04 \$/(kW-yr)
Incremental Variable Costs:	4.07 mills/kWh
Byproduct Credit:	1.53 mills/kWh
Fuel Cost:	16.13 mills/kWh

Variable Operating Cost: 18.68 mills/kWh

COST OF ELECTRICITY -----> 50.22 mills/kWh

Fixed Charge Factor: 0.1034 Variable Cost Levelization Factor: 1.0000

The plant heatrate (HHV) is: 8825. BTU/kWh. Efficiency: 0.3870

G. ENVIRONMENTAL SUMMARY -----

INPUTS:	Coal	0.7845 lb/kWh
	Water	1.6642 lb/kWh
OUTPUTS:	Blowdown	0.0733 lb/kWh
	Ash	0.0824 lb/kWh
	Sulfuric Acid	0.0849 lb/kWh
	CO2 emissions	1.7151 lb/kWh
	CO emissions	0.0028 lb/kWh
	SO2 emissions	0.0414 lb/MMBtu
	COS emissions	0.0000 lb/MMBtu
	CH4 emissions	0.0000 lb/kWh
	H2S emissions	0.0000 lb/MMBtu
	NOx emissions	0.2581 lb/MMBtu

NOTE: NOx emissions are based on fuel bound ammonia

I. SCR SUMMARY -----

Reference Removal Efficiency	0.80 fraction
Actual Removal Efficiency	0.80 fraction
Ammonia Slip	10.00 ppmv
Reference NOx Concentration	42.00 ppmv
Minimum Relative Cat. Activ.	839.92 ratio
Activity Time Constant	12638.01
Replacement Interval	11388.00 hrs
Reference Temperature	717.00 F
Actual Temperature	717.00 F
Dilution Steam/NH3 Ratio	19.00 lbmole/lbmole
NH3/NOx Ratio	0.82 lbmole/lbmole

Fraction NOx as NO	0.95
Reference Replacement Interv.	17082.00 hrs
SO2 Oxidation Rate	0.01 fraction
SCR Pressure Drop	0.14 psi
Reference Space Velocity	10000.00 1/hr
Calculated Space Velocity	5204.32 1/hr
No. Initial Catalyst Layers	3
No. Total Catalyst Layers	3
No. Layers Replaced at a time	1
Factor, efficiency	1.00
Factor, end mole ratio	0.4900
Factor, activity	1.0621
Factor, temperature	1.0000
Initial Catalyst Volume	28503.48 ft <sup>3</sup>
Total Catalyst Volume	28503.48 ft <sup>3</sup>
Reference Cat. Volume	14834.12 ft <sup>3</sup>
Annual Layers Replaced	0.50
Annual Volume Replaced	4750.58 ft <sup>3</sup> /yr
Catalyst Life, yrs	6.00 yrs
Direct Capital Cost	8678.86
Ammonia Flow, tons/yr	145.50