Chapter 2

Selective Catalytic Reduction

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2. SELECTIVE CATALYTIC REDUCTION

2.1 Introduction

Selective catalytic reduction (SCR) has been applied to stationary source fossil fuel-fired combustion units for emission control since the early 1970s and is currently being used in Japan, Europe, the United States, and other countries. In the U.S. alone, more than 1,000 SCR systems have been installed on a wide variety of sources in many different industries, including utility and industrial boilers, process heaters, gas turbines, internal combustion engines, chemical plants, and steel mills [1]. Other sources include fluid catalytic cracking units (FCCUs), ethylene cracker furnaces, nitric acid plants, catalyst manufacturing processes, nitrogen fixation processes, and solid/liquid or gas waste incinerators [2, 3]. In the U.S., SCR has been installed on more than 300 coal-fired power plants ranging in size from less than 100 megawatt equivalent (MWe) to 1,400 MWe [1, 4]. Other combustion sources with large numbers of SCR retrofits include more than 50 gas-fired utility boilers ranging in size from 147 MWe to 750 MWe, more than 50 industrial boilers and process heaters (both field-erected and packaged units), and more than 650 combined cycle gas turbines [1]. SCR can be applied as a stand-alone nitrogen oxides (NO_x) control or with other technologies, including selective non-catalytic reduction (SNCR)[1] and combustion controls such as low NO_x burner (LNB) and flue gas recirculation (FGR) [2].

SCR is typically implemented on stationary source combustion units requiring a higher level of NO_x reduction than achievable by selective non-catalytic reduction (SNCR) or combustion controls. Theoretically, SCR systems can be designed for NO_x removal efficiencies up close to 100 percent. In practice, commercial coal-, oil-, and natural gas–fired SCR systems are often designed to meet control targets of over 90 percent. However, the reduction may be less than 90 percent when SCR follows other NO_x controls such as LNB or FGR that achieve relatively low emissions on their own. The outlet concentration from SCR on a utility boiler is rarely less than 0.04 lb/million British thermal units (MMBtu) [1].[2] In comparison, SNCR units typically achieve approximately 25 to 75 percent reduction efficiencies [5].

Either ammonia or urea may be used as the NO_x reduction reagent in SCR systems. Urea is generally converted to ammonia before injection. Results of a survey of electric utilities that operate SCR systems indicated that about 80 percent use ammonia (anhydrous and aqueous), and the remainder use urea [4]. A survey of coal-fired power plants that control NO_x emissions using either SCR or SNCR found anhydrous ammonia use exceeds aqueous ammonia use by a ratio of 3 to 1. Nearly half of these survey respondents also indicated that price is their primary consideration in the choice of reagent; safety is the primary consideration for about 25 percent of the operators [6].

SCR capital costs vary by the type of unit controlled, the fuel type, the inlet NO_x level, the outlet NO_x design level, and reactor arrangement. Capital costs also rose between 2000 and 2010 (at least for utility boiler applications), even after scaling all data to 2011 dollars (2011$).

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[1] A hybrid SNCR/SCR system was demonstrated at the AES Greenidge Power Plant in 2006. However, no hybrid SNCR/SCR systems are currently known to be operating as of February 2016.

[2] Data in the Clean Air Markets Division (CAMD) database also suggest SCR units rarely achieve emissions less than 0.04 lb/MMBtu.

For a small number of early SCR retrofits on utility boilers prior to 2000, the average costs were about \$100/kilowatt (kW), in 2011\$, and there was little scatter in the data. From 2000 to 2007, the SCR costs for 32 utility boilers ranged from about \$100/kW to \$275/kW (2011\$), and a slight economy of scale was evident (i.e., using a regression equation, costs ranged from about \$200/kW for a 200 MW unit to \$160/kW for an 800 MW unit). For 2008 to 2011, the average SCR costs exhibited great variability and again a modest economy of scale was evident (i.e., about \$300/kW for a 200 MW unit to \$250/kW for an 800 MW unit; 2011\$). For eight utility boilers either installed in 2012 or projected to be installed by 2014, the SCR costs ranged from about \$270/kW to \$570/kW (2011\$). The generating capacity for these units ranged from 400 MW to 800 MW [7]. Typical operation and maintenance costs are approximately 0.1 cents per kilowatt-hour (kWh) [8, 9]. Table 2.1a provides capital cost estimates for electric utility boilers, and Table 2.1b presents capital cost estimates for SCR applications of various sizes in several other industry source categories.

The procedures for estimating costs presented in this report are based on cost data for SCR retrofits on existing coal-, oil-, and gas-fired boilers for electric generating units larger than 25 MWe (approximately 250 MMBtu/hr). Thus, this report's procedure estimates costs for typical retrofits of such boilers. The methodology for utility boilers also has been extended to large industrial boilers by modifying the capital cost equations and power consumption (electricity cost) equations to use the heat input capacity of the boiler instead of electric generating capacity.[3] The procedures to estimate capital costs are not directly applicable to sources other than utility and industrial boilers. Procedures to estimate annual costing elements other than power consumption are the same for SCR units in any application. The cost of SCR as part of a new plant often is likely to be less than the cost for retrofitting an SCR at an existing plant. Appropriate factors to estimate the cost of a new plant SCR have been included. In addition, the cost procedures in this report reflect individual SCR applications. Retrofitting multiple boilers with SCR can allow for some economies of scale for installation, thus yielding some reduction in capital costs per SCR application. The cost methodology incorporates certain approximations; consequently, it should be used to develop study-level accuracy (±30 percent) cost estimates of SCR applications. Such accuracy in the cost methodology is consistent with the accuracy of the cost estimates for the other control measures found in this Cost Manual as stated in Section 1.

In the cement industry, pilot tests in the 1970s and 1990s showed that SCR could be a feasible control technology for cement kilns. Building on that experience, SCRs were first installed in Europe in 2001. Today, SCR has been successfully implemented at seven European cement plants in Solnhofer, Germany (operated from 2001 until 2006), Bergamo, Italy (2006), Sarchi, Italy (2007), Mergelstetten, Germany (2010), Rohrdorf, Germany (2011), Mannersdorf, Austria (2012), and Rezatto, Italy (2015) [10, 11, 12]. As of 2015, there is only one cement plant in the U.S. that has installed an SCR. This SCR began operation in 2013 and is installed after an electrostatic precipitator. The control efficiency for the system is reported to be about 80 percent, which is consistent with SCR applications on European kilns. SCRs have not seen widespread use in the U.S. cement industry mainly due to industry concerns regarding potential problems caused by high-dust levels and catalyst deactivation by high sulfur trioxide (SO_3) concentrations

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[3] The term "industrial" boilers as used in the Control Cost Manual includes industrial, commercial, and institutional (or ICI) boilers, unless otherwise noted.

from pyritic sulfur found in the raw materials used by U.S. cement plants. The SO_3 could react with calcium oxide in the flue gas to form calcium sulfate and with ammonia to form ammonium bisulfate. The calcium sulfate could deactivate the catalyst, while the ammonium bisulfate could cause catalyst plugging. There have been concerns expressed about the potential for catalyst poisoning by sodium, potassium, and arsenic trioxide. Finally, other concerns expressed are that dioxins and furans may form in the SCR due to combustion gases remaining at temperatures between 450 degrees Fahrenheit (°F) and 750°F. These and other concerns regarding the implementation of SCR to the cement industry are discussed in detail in “Alternative Control Techniques Document Update – NO_x Emissions from New Cement Kilns” [10]. Due to the small number of SCRs installed at cement plants, information on capital and operating costs for SCRs at cement plants is limited. The installation and operating costs for the SCR installed at the U.S. plant in 2013 are not publicly available at this time. In general, we expect the capital and operating costs would be higher than for low-dust applications due to the need to install catalyst cleaning equipment for SCR systems installed in high-dust configurations and for heating the flue gas in low-dust, tail-end configurations.

Table 2.1a: Summary of SCR Cost Data for Utility Boilers

Source Category	Unit Size	Fuel Type	Capital Cost			$ Year	Comments	Reference
			Min	Avg	Max			
Electric Generating Units	NA[a]	NA	$55/kW		$140/kW	<2000$[b]	Retrofit costs.	[13]
	~300-1,400 MW	NA	~$70/kW		~$120/kW	<2000$[b]	Retrofit costs. Six boilers. No economy of scale.	[13]
	150–1,000 MW	Coal	$80/kWnet[c]		$160/kWnet[c]	2002$	Retrofit costs. Author of referenced document scaled original costs to 2002 dollars. More than 20 boilers. Little to no economy of scale.	[14]
	NA	Coal	$60/kW	$100kW	$200/kW	<2004$[b]	Retrofit costs	[15]
	<300 MW	Coal		$167/kW	$186/kW	<2004$	Costs for 26 boilers.	[16]
	301–600 MW	Coal		$148/kW	$192/kW	<2004$	Costs for 15 boilers.	[16]
	601–900 MW	Coal		$124/kW	$221/kW	<2004$	Costs for 22 boilers.	[16]
	>900 MW	Coal		$118/kW	$195/kW	<2004$	Costs for 9 boilers.	[16]
	100–399 MW	Coal	$70/kW	$123/kW	~$175/kW	<2004$[b]	Costs for 5 boilers.	[17]
	400–599 MW	Coal	$73/kW	$103/kW	~$160/kW	<2004$[b]	Costs for 8 boilers.	[17]
	600–899 MW	Coal	$56/kW	$81/kW	~$100/kW	<2004$[b]	Costs for 9 boilers.	[17]
	>900 MW	Coal	~$80/kW	$117/kW	~$190/kW	<2004$[b]	Costs for 10 boilers.	[17]
	191 MW	Coal		$149/kW		2006$	Retrofit costs.	[18]
	~100 MW-~800MW	NA	~$125/kW	$275/kW	~$440/kW	2008$	Retrofit costs for 15 boilers installed in 2008 to 2010. Most costs between $200/kW and $350/kW. Slight economy of scale—regression average about $340/kW for 100 MW to $250/kW for 800 MW.	[8]
	~400 MW to ~800 MW	NA	~$270/kW	~$420/kW	~560/kW	2011$	Retrofit costs for 8 boilers either installed in 2012 or projected to be installed by 2014.	[7]

[a] Not Available.

[b] Year of reference.

[c] Net kilowatts.

Table 2.1b: Summary of SCR Cost Data for Miscellaneous Industrial Sources

Source Category	Unit Size	Fuel Type	Capital Cost: average (range)	$ Year	Actual, Vendor Quote, or Estimated?	Comments	Reference
Industrial-Commercial Boilers	350 MMBtu	Coal	NA ($10,000–$15,000/MMBtu/hr)	1999$	Estimated	Retrofit costs. Authors of referenced document estimated the low end of the range assuming a cost of about $100/kW for a 100 MW (1000 MMBtu/hr) utility boiler and assuming that economies of scale would be greater for utility boilers than for industrial boilers (so that the cost for a 350 MMBtu/hr industrial boiler would be comparable to or greater than the cost for a 1000 MMBtu/hr utility boiler on a $/MMBtu basis).	[19]
	100–1,000 MMBtu/hr	Coal	NA ($7,300–$14,600/MMBtu/hr)	1999$	Estimated	Retrofit costs. Generally costs available for one boiler with each type of fuel. Authors of referenced document estimated costs for other sizes assuming ratio of small-to-large $/MMBtu costs are related to ratio of large to small heat inputs raised to the 0.3 power.	[20]
	100–1,000 MMBtu/hr	Oil	NA ($5,550–$11,100/MMBtu/hr)	1999$	Estimated		[20]
	100–1,000 MMBtu/hr	Gas	NA ($4,010–$8,010/MMBtu/hr)	1999$	Estimated		[20]
	100 MMBtu/hr	Gas	NA ($7,500/MMBtu/hr)	1999$[b]	Vendor	Cited source in reference [15] is an unpublished letter from a vendor.	[19]
	350 MMBtu	Oil, Gas, or Wood	NA ($4,000–$6,000/MMBtu/hr)	1999$	Estimated		[21]
	57 MMBtu/hr	Wood	NA (>$560,000 and $9,500/MMBtu/hr)	1999$[c]	Actual/ Estimate	Costs for a new boiler.	[19]
	321 MMBtu/hr	Wood	NA ($1,980/MMBtu/hr)	2006$	Likely Estimated		[22]
Petroleum Refining – Steam Boilers	650 MMBtu/hr	Gas or refinery fuel gas	NA ($3,100–$25,800/MMBtu)	2004$[c]	Estimated	Retrofit costs. Equipment costs based on range of costs found in literature search (references were not provided). Installation costs estimated using factors from the Control Cost Manual for thermal and catalytic incinerators.	[23]
Petroleum Refining – Process Heaters	350 MMBtu/hr	Gas/refinery fuel gas	NA ($3,100–$25,800/MMBtu)	2004$[c]	Estimated	Same comment as above.	[23]
	350 MMBtu/hr	Refinery oil	NA ($3,100–$25,800/MMBtu)	2004$[c]	Estimated	Same comment as above.	[23]

Source Category	Unit Size	Fuel Type	Capital Cost: average (range)	$ Year	Actual, Vendor Quote, or Estimated?	Comments	Reference
	10 MMBtu/hr	Gas or refinery fuel gas/NG combo	$19,200/MMBtu ($12,000–$26,500/MMBtu)	1999[b]	Vendor/ Estimated	Costs are based primarily on quotes from two vendors (and additional discussions). Authors of the referenced report added costs for fan, motor, and ductwork costs based on procedures in the Control Cost Manual.	[24]
	50 MMBtu/hr	Gas or refinery fuel gas/NG combo	$5,140/MMBtu ($4,020–$6,280/MMBtu)	1999[b]	Vendor/ Estimated	Same comment as above.	[24]
	75 MMBtu/hr	Gas or refinery fuel gas/NG combo	$4,190/MMBtu ($3,440–$4,950/MMBtu)	1999[b]	Vendor/ Estimated	Same comment as above.	[24]
	150 MMBtu/hr	Gas or refinery fuel gas/NG combo	$2,730/MMBtu ($2,570–$2,880/MMBtu)	1999[b]	Vendor/ Estimated	Same comment as above.	[24]
	350 MMBtu/hr	Gas or refinery fuel gas/NG combo	$1,550/MMBtu ($1,520–$1,570/MMBtu)	1999[b]	Vendor/ Estimated	Same comment as above.	[24]
	68 MMBtu/hr (Two 32 MMBtu/hr)	Refinery fuel gas	NA ($22,100/MMBtu)	1991	Actual	Retrofit costs.	[19]
Petroleum Refining – FCCU	70,000 barrels/stream day (bbl/stream day)	NA	NA ($9.0 million)	2004$[c]	Vendor	Estimated cost by vendor (for 90 percent reduction).	[3]
	27,000 bbl/stream day	NA	NA ($8-$12 million)	2009	Estimated		[25]

Source Category	Unit Size	Fuel Type	Capital Cost: average (range)	$ Year	Actual, Vendor Quote, or Estimated?	Comments	Reference
	<20,000->100,000 bbl/stream day	NA	NA (order of magnitude range; low end higher than two entries above)	2005 to 2010	Actual	Costs reported by 6 petroleum refining companies for 7 FCCUs in responses to EPA ICR. One new, 6 retrofits.	[26]
	NA	NA	NA ($20 million)	2006	Actual	Approximate average cost for SCR retrofits at several refineries	[27]
Portland Cement (dry kilns)	1.09 million short tpy clinker	NA	NA ($6.9 per short ton clinker)	2006[a]	Estimated	Retrofit cost. Estimate based primarily on SCR procedures for boilers in fifth edition of the Control Cost Manual. Clinker capacity obtained from the second reference.	[28,29]
	1.13 million short tpy clinker	NA	NA ($5.9 per short ton clinker)	2006[a]	Estimated	Same comment as above.	[28,29]
	2.16 million short tpy clinker	NA	NA ($3.9 per short ton clinker)	2006[a]	Estimated	Same comment as above.	[28,29]
	1.4 million short tpy clinker	NA	NA ($5.9 per short ton clinker)	2004	Not clear	Retrofit cost for European kiln. Cost in euros converted to dollars assuming a ratio of $1.3/euro.	[30]
	1.055 million tpy clinker	NA	NA ($4.4 per short ton clinker)	2004	Estimated	Cost for new kiln.	[31]
	1.095 million short tpy clinker	NA	NA ($4.4 per short ton clinker)	2011	Estimated	Cost for new kiln. Cost based on quote for the SCR equipment, and standard installation factors from the Control Cost Manual for other types of control devices.	[32]
Portland Cement (wet kilns)	0.3 million short tpy clinker	NA	NA ($17.5 per short ton clinker)	2006[a]	Estimated	Retrofit costs for 4 kilns. Rated clinker production capacity obtained from the second reference.	[28,33]
	0.320 million short tpy clinker		NA ($15.6-$16.6 per short ton clinker)	2006[a]	Estimated	Retrofit costs for 3 kilns. Rated clinker production capacity obtained from second reference.	[28,29]

Source Category	Unit Size	Fuel Type	Capital Cost: average (range)	$ Year	Actual, Vendor Quote, or Estimated?	Comments	Reference
Gas Turbine, Simple Cycle	NA	Gas	NA ($50-$70/kW)	1999$[a]	Vendor	Retrofit costs.	[19]
	80 MW	Gas	NA ($51/kW)	1999$[a]	Vendor	Retrofit cost, excluding balance of plant costs.	[19]
	2 MW	Gas	NA ($237/kW)	1999$[a]	Vendor	Retrofit cost.	[19]
	12 MW	Gas	NA ($167/kW)	1999$[a]	Vendor	Retrofit cost.	[19]
Internal Combustion Engine	1,800 hp[d]	Diesel (No. 2 fuel oil)	NA ($0.18 million)	1994	Actual	New cost	[19]

[a] Year of reference.

[b] Year analysis was conducted (assumed vendor contacts were made that year).

[c] Commission year of the SCR.

[d] Horsepower.

2.2 Process Description

Like SNCR, the SCR process is based on the chemical reduction of the NO_x molecule. The primary difference between SNCR and SCR is that SCR employs a metal-based catalyst with activated sites to increase the rate of the reduction reaction. The primary components of the SCR include the ammonia storage and delivery system, ammonia injection grid, and the catalyst reactor [2]. A nitrogen-based reducing agent (reagent), such as ammonia or urea-derived ammonia, is injected into the post-combustion flue gas. The reagent reacts selectively with the flue gas NO_x within a specific temperature range and in the presence of the catalyst and oxygen to reduce the NO_x into molecular nitrogen (N_2) and water vapor (H_2O).

The use of a catalyst results in two primary advantages of the SCR process over SNCR. The main advantage is the higher NO_x reduction efficiency. In addition, SCR reactions occur within a lower and broader temperature range. However, the decrease in reaction temperature and increase in efficiency is accompanied by a significant increase in capital and operating costs. The capital cost increase is mainly due to the large volumes of catalyst required for the reduction reaction. Operating costs for SCR consist mostly of replacement catalyst and ammonia reagent costs, and while historically, the catalyst replacement cost has been the largest cost, the reagent cost has become the most substantial portion of operating costs for most SCR [7].[4]

Figure 2.1 shows a simplified process flow schematic for SCR. Reagent is injected into the flue gas downstream of the combustion unit and economizer through an injection grid mounted in the ductwork. The reagent is generally diluted with compressed air or steam to aid in injection. The reagent mixes with the flue gas, and both components enter a reactor chamber containing the catalyst. As the hot flue gas and reagent diffuse through the catalyst and contact activated catalyst sites, NO_x in the flue gas chemically reduces to nitrogen and water. The heat of the flue gas provides energy for the reaction. The nitrogen, water vapor, and any other flue gas constituents then flow out of the SCR reactor. More detail on the SCR process and equipment is provided in the following sections.

There are several different locations downstream of the combustion unit where SCR systems can be installed. Flue gas temperature and constituents vary with the location of the SCR reactor chamber. SCR reactors located upstream of the particulate control device and the air heater (“high-dust” configuration) have higher temperatures and higher levels of particulate matter. An SCR reactor located downstream of the air heater, particulate control devices, and flue gas desulfurization (FGD) system (“low-dust” or “tail-end” configuration) is essentially dust- and sulfur-free but its temperature is generally below the acceptable range. In this case, reheating of the flue gas may be required, which significantly increases the SCR operational costs. Section 2.2.3 discusses the various SCR system configurations.

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[4] Several cost analyses in recent years have shown the largest operating cost is for reagent usage rather than for catalyst costs. For example, for the Navajo Generating Station in Arizona, a 2010 BART analysis report on an 812 MW gross coal-fired unit estimates annual operating costs for ammonia reagent of $1,035,000 (based on $465/ton) and for catalyst replacement of $672,000 (based on $8,000/$m^3$) [34].

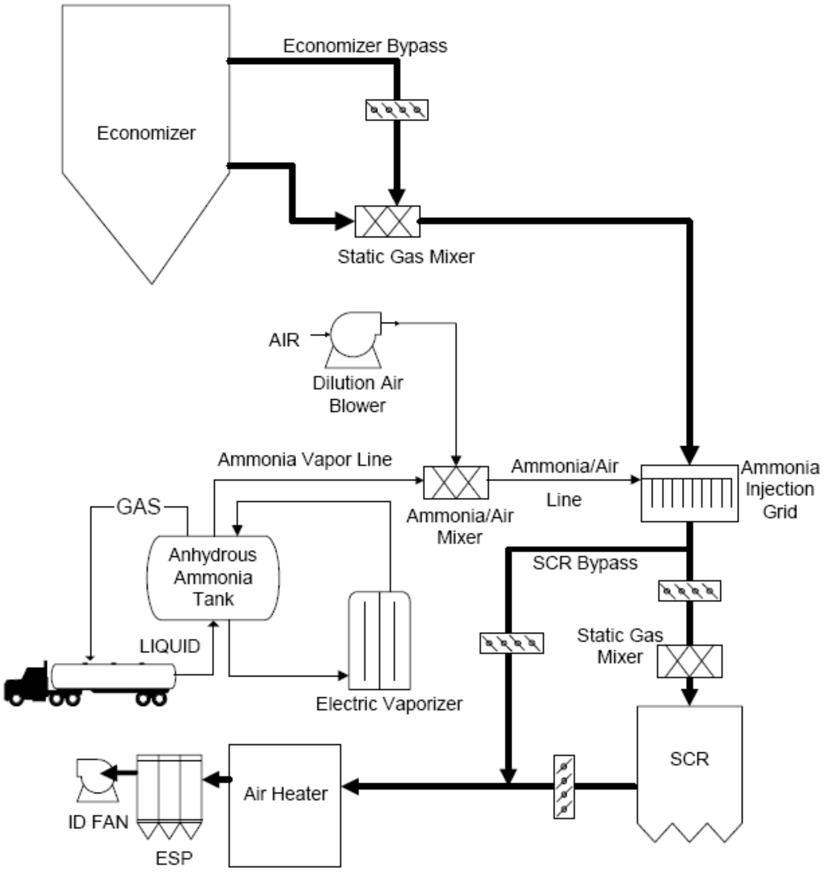

Figure 2.1: SCR Process Flow Diagram [35, 36]

2.2.1 Reduction Chemistry, Reagents, and Catalyst

The reducing agent employed by the majority of SCR systems is gas-phase ammonia (NH_3) because it readily penetrates the catalyst pores. The ammonia, either in anhydrous or aqueous form, is vaporized before injection by a vaporizer. Within the appropriate temperature range, the gas-phase ammonia then decomposes into free radicals, including NH_3 and an amide (NH_2). After a series of reactions, the ammonia radicals come into contact with the NO_x and reduce it to N_2 and H_2O. Since NO_x includes both nitrogen monoxide (NO) and nitrogen dioxide (NO_2), the overall reactions with ammonia are as follows:

$$2NO + 2NH_3 + \frac{1}{2}O_2 \xrightarrow{catalyst} 2N_2 + 3H_2O \tag{2.1a}$$

$$2NO_2 + 4NH_3 + O_2 \xrightarrow{catalyst} 3N_2 + 6H_2O \tag{2.1b}$$

The equations indicate that one mole of NH_3 is required to remove one mole of NO and two moles of NH_3 are required to remove one mole of NO_2. However, Equation 2.1a is the predominant reaction because 90 to 95 percent of NO_x in flue gas from combustion units is NO. Hence, about one mole of NH_3 is required to remove one mole of NO_x. The catalyst lowers the required activation energy for the reduction reaction and increases the reaction rate. In the catalytic reaction, activated sites on the catalyst rapidly adsorb ammonia and gas-phase NO to form an activated complex. The catalytic reaction, represented by Equations 2.1a and 2.1b, results in nitrogen and water, which are then desorbed to the flue gas. The site at which the reaction occurs is then reactivated via oxidation.

The high temperature of the flue gas converts the ammonia to free radicals and provides the activation energy for the reaction. The reaction also requires excess oxygen, typically 2 to 34 percent, to achieve completion. NO_x reduction with ammonia is exothermic, resulting in the release of heat. However, because the NO_x concentration in the flue gas at the inlet of the SCR is typically 0.01 to 0.02 percent by volume, the amount of heat released is correspondingly small. Thermodynamic equilibrium is not a limiting factor in NO_x reduction if the flue gas is within the required temperature range [37].

Reagent

The SCR system can use either aqueous or anhydrous ammonia for the reduction reaction, and some plants use urea-to-ammonia reagent systems where aqueous ammonia is produced onsite (often called onsite urea-derived ammonia production or "ammonia-on-demand"). Anhydrous ammonia is nearly 100 percent pure ammonia. It is a gas at normal atmospheric temperature; therefore, it must be transported and stored under pressure. Anhydrous ammonia is classified as a hazardous material and often requires special permits as well as additional procedures for transportation, handling and storage.

SCR applications using aqueous ammonia generally transport and store it at a concentration of about 29 percent ammonia in water, although some applications use a 19 percent solution [37]. The use of aqueous ammonia reduces transport and storage problems related to safety. In addition, certain locations may not require permits for aqueous ammonia concentrations less than 28 percent. Aqueous ammonia, however, requires more storage capacity than anhydrous ammonia and it also requires shipping costs for the water solvent in the solution. Although the 29 percent aqueous ammonia solution has substantial vapor pressure at normal air temperatures, a vaporizer is generally required to provide sufficient ammonia vapor to the SCR system. Table 2.2 gives the properties of anhydrous ammonia and the properties of a 29.4 percent aqueous ammonia solution (a 21 degree Baumé solution).

The type of reagent used affects both the capital costs and annual costs. Anhydrous ammonia typically has the lowest capital and operating costs, excluding highly site-dependent permitting and risk management planning and implementation costs. Urea systems have the highest capital costs due to the complexity of the processing equipment. Aqueous ammonia systems tend to have the highest operating costs, primarily because of the cost for transportation. Urea systems have the highest energy consumption costs because the energy needed to hydrolyze or decompose urea tends to be higher than the energy needed to vaporize aqueous ammonia. Although the price per ton of anhydrous ammonia is higher than the price per ton of urea, the cost per ton of NO_x removed is higher for urea due to urea's much higher molecular weight. For example, one SCR supplier estimated capital costs for a 130 pounds per hour (lb/hr) ammonia system to be $280,000 for anhydrous ammonia, $402,000 for 19 percent aqueous ammonia, and $750,000 for urea [38]. Another reference reported that the equipment cost for urea is generally twice the equipment cost for anhydrous ammonia [39]. According to one reference, the total SCR system cost is 2 to 5 percent higher when using a urea reagent system instead of an anhydrous ammonia system [14]. Relative to anhydrous ammonia, one reference estimated annual operating costs for 19 percent aqueous ammonia are 50 percent higher, costs for 29 percent aqueous ammonia are 33 percent higher, and costs for urea are 25 percent higher [40]. Another reference stated that as a general rule, operating costs for urea systems are about 50 percent more than the operating costs for anhydrous ammonia [39]. One reference estimated energy costs for an unspecified application to be $167,000 for a urea system, $73,000 to $117,000 for aqueous ammonia systems, and $16,000 for anhydrous ammonia [41].

This presentation is valid for anhydrous or aqueous ammonia; the capital cost procedures are based on the typical mix of systems actually in operation, while the procedures for estimating annual costs apply to any ammonia system (the examples in section 2.5 illustrate the procedures for a system using 29 percent aqueous ammonia as the reagent).

Table 2.2: Ammonia Reagent Properties

Property	Anhydrous Ammonia [42,43]	Aqueous Ammonia
Liquid or gas at normal air temperature	Liquid at high pressure; gas at atmospheric pressure	Liquid
Concentration of reagent normally supplied	99.5 percent (by weight)	29.4 percent (by weight of NH_3)
Molecular weight of reagent	17.03	17.03 (as NH_3)
Ratio of ammonia to solution	99.5 percent (by weight of NH_3)	29.4 percent (by weight of NH_3)
Density of liquid at 60°F	5.1 lb/gal	7.5 lb/gal
Vapor pressure at 80°F	153 psia	14.6 psia [43, p. 3]
Flammability limits in air	16–25 percent NH_3 (by volume)	16 to 25 percent NH_3 (by volume)
Short-term exposure limit	35 ppm	35 ppm
Odor	Pungent odor at 5 ppm or more	Pungent odor at 5 ppm or more
Acceptable materials for storage	Steel tank, rated for at least 250 psig pressure (no copper or copper-based alloys, etc.)	Steel tank, rated for at least 25 psig pressure (no copper or copper-based alloys, etc.)

Catalyst

SCR catalysts are composed of active metals or ceramics with a highly porous structure. Within the pores of the catalyst are activated sites. These sites have an acid group on the end of the compound structure where the reduction reaction occurs. As stated previously, after the reduction reaction occurs, the site reactivates via rehydration or oxidation. Over time, however, the catalyst activity decreases, requiring replacement, washing/cleaning, rejuvenation, or regeneration of the catalyst. Catalyst designs and formulations are generally proprietary. Both the catalyst material and configuration determine the properties of the catalyst.

Originally, SCR catalysts were precious metals such as platinum (Pt). In the late 1970s, Japanese researchers used base metals consisting of vanadium (V), titanium (Ti), and tungsten (W), which significantly reduced catalyst cost. In the 1980s, metal oxides such as titanium oxide (TiO_2), zirconium oxide (ZrO_2), vanadium pentoxide (V_2O_5), and silicon oxide (SiO_2) were employed to broaden the reaction temperature range. Zeolites, crystalline alumina silicates, were also introduced for high temperature (675 to 1,000°F; 360 to 540°C) applications; however, zeolites tended to be cost prohibitive. From 1980 to 2008, the cost of catalyst has dropped from approximately $34,000/m^3 to a range of $5,000 to $6,000/m^3 (costs are in 2011$) [7].[5] This reference also reported that catalyst prices remained in the approximate range of $5,000 to $6,000/m^3 through 2012.

Improvements to the catalyst formulations over time have decreased unwanted side reactions such as sulfur oxide conversions (sulfur dioxide (SO_2) to SO_3) and increased the resistance to flue gas poisons, and newer catalysts can oxidize metallic mercury (Hg) into ionic forms (for easy removal downstream in wet scrubbers and wet electrostatic precipitators [ESPs]) [44]. Improved catalyst designs have also increased catalyst activity, surface area per unit volume, and the temperature range for the reduction reaction. As a consequence, there is a corresponding decrease in the required catalyst volumes and an increase in the catalyst operating life. For coal-fired boiler applications, SCR catalyst vendors typically guarantee the catalyst for an operating life ranging from 8,000 to 24,000 hours [1]. Applications using oil and natural gas have a longer operating life, typically greater than 32,000 hours [45]. In addition, operating experience indicates that actual catalyst deactivation rates are lower than the design specifications [37]. The latest demands on catalyst technology for both higher and lower sulfur coal-fired boilers include design NO_x removal of 90 percent; control of residual NH_3 to 2 parts per million (ppm) (i.e., ammonia slip); guarantees for SO_2 oxidation to less than 1 percent, and in many cases, to less than 0.5 percent; being able to withstand washing/cleaning and regeneration procedures; and guarantees for mercury oxidation [41].

Catalyst formulations include single component, multi-component, or active phase with a support structure. Most catalyst formulations contain additional compounds or supports to give thermal and structural stability or to increase surface area [46]. Catalyst configurations are generally ceramic honeycomb and pleated metal plate (monolith) designs in a fixed-bed reactor, which provide high surface area to volume ratio. Pellet catalysts in fluidized beds are also available. Pellets have greater surface area than honeycombs or pleated plates but are more susceptible to plugging. This limits the use of pellets to clean-burning fuels such as natural gas.

Catalyst elements placed in a frame form a catalyst module. The modules stack together in multiple layers to create a reactor bed of the total required catalyst volume. A typical module is 3.3 ft × 6.6 ft in area (1 m × 2 m) and 3.3 ft (1 m) in height. A crane hoists the large catalyst modules into the reactor from either the interior or exterior of the reactor, depending on the reactor design.

Catalysts greatly accelerate the NO_x reduction reaction rate, but some catalysts have more favorable properties for a given application. Performance requirements that drive the choice of catalyst include reaction temperature range, flue gas flow rate, fuel source, catalyst activity and selectivity, SO_2 oxidation, and catalyst operating life. In addition, the design must consider the cost of the catalyst, including disposal costs [37]. In the past, the initial charge of catalyst costs accounted for 20 percent or more of the capital costs for an SCR system [37], however, as catalyst unit cost has declined over time, this catalyst cost is a smaller percentage of the capital costs [8].

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[5] An earlier reference shows that from 1980 to 2006, the cost of catalyst dropped by 75 percent from approximately $16,000/m^3 to less than $4,000/m^3 [41]. These costs are for the cost year reported and are not adjusted for escalation to current year.

The catalyst layers may be washed/cleaned, rejuvenated, or regenerated to extend the catalyst life as catalyst activity declines, or they may be replaced. Generally, less than one layer of catalyst is replaced per year for fixed bed designs. Most SCR manufacturers offer a disposal service, in which either the catalyst is reactivated (i.e., rejuvenated or regenerated) for reuse or its components are recycled for other uses [46]. If the catalyst cannot be recycled or reused, the facility operator must dispose of the spent catalyst in an approved landfill. In the United States, most catalyst formulations are not considered hazardous waste [46].

Catalyst cleaning typically means the removal of physical restrictions to the catalyst (i.e., blinding layers and large particle ash [LPA]), rejuvenation means the removal of poisons without replenishing catalytically active compounds in the catalyst, and regeneration typically means the removal of poisons and the restoration of catalytic activity by restoring catalyst active ingredients [47]. These activities may occur online/in-situ or offline, and they may occur onsite or offsite [47]. The catalyst layers may be removed and transported to the cleaning, rejuvenation, or regeneration site. Online catalyst cleaning with soot blowers or sonic horns is conducted on a regular basis to remove ash or particles (soot blowers and sonic horns are discussed below) [48]. Water-based cleaning can also be conducted to remove physical materials that plug or blind the catalyst [49].

Use of rejuvenated and regenerated catalyst has increased since the late 1990s, and for some applications, can be considered equivalent to new catalyst [50]. For rejuvenation processes, the focus is on removal of blinding materials and catalyst poisons [47, 51]. Rejuvenation processes may cause the catalyst to lose structural integrity and mechanical strength [47]. Today, regeneration processes focus on increasing the longevity of the catalyst by maintaining its mechanical strength (both compressive and bonding) and improving its activity or performance [47, 51]. The catalyst layer may be washed in a series of baths that remove ash and particulate, remove poisons, and add chemicals or various metals used in the original catalyst manufacturing [49, 51]. Moisture must also be removed from the catalyst because it reduces the strength of the catalyst [51]. Because regeneration may cause catalysts to lose mechanical strength, recalcination may also be conducted to ensure catalyst impregnation and that catalyst mechanical strength is regained [47, 49, 51]. Improvements to the catalyst activity relative to certain reactions can be made using a regeneration process (e.g., impregnating other chemicals in the catalyst to reduce the conversion of SO_2 to SO_3, or to increase the oxidation of Hg over the conversion rate of the original catalyst while maintaining the same catalyst activity for NO_x) [51]. Damage to the catalyst can occur during operation of the SCR or during transport of the catalyst for rework, so a thorough inspection of the modules is conducted along with replacement or repair of any damaged elements [48].

Benefits for regenerated catalysts include the following: lower regeneration cost per layer as compared to new replacement cost [1]; no catalyst cost [1]; disposal cost savings [1]; full restoration of original catalytic activity [52]; same deactivation rate as a new catalyst in the same SCR installation, i.e., comparable equipment life as new catalyst [52]; lower SO_2 oxidation for some catalysts (i.e., SO_2/SO_3 conversions is no higher than the new catalyst guarantee) [53]; removal of physical restrictions, including fly ash plugging, large particle ash, and blinding layers [47]; and no physical damage to the catalyst and no loss of structural integrity [47].

Use of regenerated catalysts reduces catalyst replacement cost and minimizes the need to dispose of spent catalyst [54]. It is estimated that a typical 500-MW coal-fired power plant will spend approximately \$2,000,000[6] on a single layer of new catalyst [54]. The cost for regenerated catalyst for this same facility would be approximately \$1,000,000 for a single layer of catalyst [54]. Disposal costs when replacing a spent catalyst could be \$50,000 to \$200,000 per layer, and these costs are avoided with regenerated catalysts [48]. Regenerated catalyst typically costs 40 percent less than new catalyst [53, 54. 55].

2.2.2 SCR Performance Parameters

The rate of the reduction reaction determines the amount of NO_x removed from the flue gas. The major design and operational factors that affect the NO_x removal performance of SCR are similar to those presented in Chapter 1, SNCR. The factors discussed previously for SNCR include the following:

- Reaction temperature range;
- Residence time available in the optimum temperature range;
- Degree of mixing between the injected reagent and the combustion gases;
- Molar ratio of injected reagent to inlet NO_x;
- Inlet NO_x concentration level; and
- Ammonia slip.

The majority of the discussion regarding SNCR design and operational factors is valid for the SCR process, except for small variations due to the use of a catalyst and the reaction chamber being separate from the combustion unit. Additional design and operational factors to consider that are specific to the SCR process include the following:

- Catalyst activity;
- Catalyst selectivity;
- Pressure drop across the catalyst;
- Ash management (i.e., mitigating large particle ash (LPA) impacts on the catalyst) and dust loading;
- Catalyst pitch;
- SO_2 and SO_3 concentrations in gas stream;
- Catalyst deactivation; and
- Catalyst management.

The major differences between SNCR and SCR are discussed below.

<u>Temperature</u>

The NO_x reduction reaction is effective only within a given temperature range. The use of a catalyst in the SCR process lowers the temperature range required to maximize the NO_x reduction reaction. At temperatures below the specified range, the reaction kinetics decrease, and ammonia passes through the boiler (ammonia slip), but there is little effect on nitrous oxide

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[6] Cost year not available; data are from 2008 article [Reference 54].

(N_2O) formation. At temperatures above the specified range, N_2O formation increases and catalyst sintering and deactivation occurs, but little ammonia slip occurs.

In an SCR system, the optimum temperature depends on both the type of catalyst used in the process and the flue gas composition. For the majority of commercial catalysts (metal oxides), the operating temperatures for the SCR process range from 480 to 800°F (250 to 430°C) [50]. Figure 2.2 is a graph of the NO_x removal efficiency as a function of temperature for a typical metal oxide catalyst [50]. The figure shows that the rate of NO_x removal increases with temperature up to a maximum between 700 and 750°F (370 to 400°C). As the temperature increases above 750°F (400°C), the reaction rate and resulting NO_x removal efficiency begin to decrease.

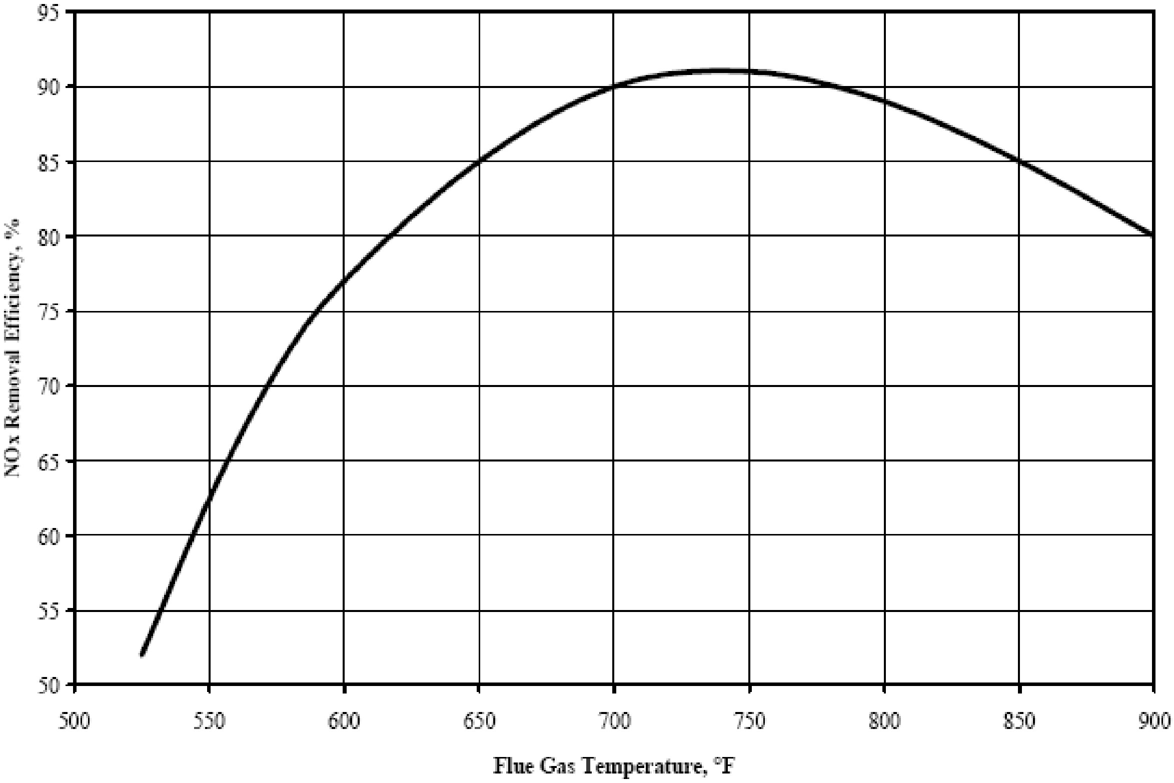

Figure 2.2: NO_x Removal versus Temperature [50]

As flue gas temperature approaches the optimum, the reaction rate increases and less catalyst volume achieves the same NO_x removal efficiency. Figure 2.3 shows the change in the required catalyst volume versus temperature [56]. There is approximately a 40 percent decrease in the required catalyst volume as flue gas temperature increases from 600°F (320°C) to the optimum range, 700 to 750°F (370 to 400°C). This decrease in catalyst volume also results in a significant decrease in capital cost for the SCR system. Less catalyst also results in a decrease in annual operation and maintenance costs. For example, the system pressure drop would be lower, which would reduce the additional electricity needed to run the induced draft (ID) fan. The net effect on catalyst replacement costs is uncertain; although the volume of catalyst replaced would be smaller, deactivation may occur more frequently since the quantity of materials in the emission stream responsible for plugging and poisoning would not be reduced.

The relationships between flue gas temperature, catalyst volume, and NO_x removal are complicated functions of the catalyst formulation and configuration. The physical and chemical properties of each catalyst are optimized for different operating conditions. For a given catalyst formulation, the required catalyst volume or temperature range can even change from one manufacturer of the catalyst to another. Therefore, the selection of the catalyst is critical to the operation and performance of the SCR system.

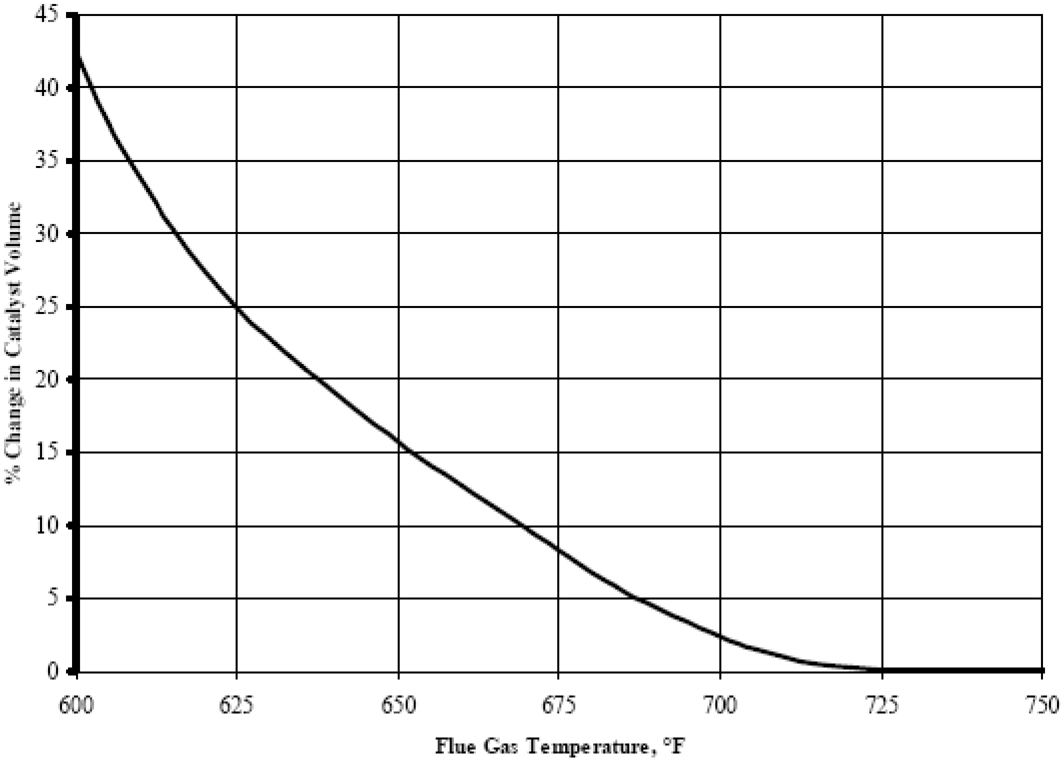

Figure 2.3: Change in Catalyst Volume vs. Temperature [56]

Because the optimum temperature window of the SCR process is lower than that of SNCR, the reagent injection into a reactor chamber occurs downstream of the combustion unit, rather than inside the combustion unit. As discussed previously, there are several options for the location of the SCR reactor. The flue gas temperature at each of these locations is different. Most designs install the reactor downstream of the economizer and prior to the air preheater, where the flue gas is at the appropriate temperature for metal oxide–based catalysts. Reheating of the flue gas may be required for reactors located downstream of the air preheater. Reheating significantly increases SCR operational costs. This continues to be true despite natural gas prices that are relatively low on an historical basis.[7]

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[7] A case study of tail-end SCR indicated that the cost of natural gas for flue gas reheating was about 60 percent of the variable annual operation and maintenance costs (i.e., sum of natural gas, electricity, and reagent costs) when the unit cost of natural gas was \$8/1,000 standard cubic feet (sft^3) [57]. If the natural gas unit cost were \$5/1,000 sft^3, then the natural gas cost for flue gas reheating would have been more than 40 percent of the variable annual operation and maintenance costs.

Boiler operation at reduced loads decreases the gas flow rate. At reduced gas flow rates, the economizer outlet gas temperature decreases because boiler heat transfer surfaces absorb more heat from the flue gas. Typical SCR systems tolerate temperature fluctuations of ±200°F (±93°C) [37]. At low boiler loads, however, the temperature can decrease below the optimum range. For example, a coal-fired utility boiler has an economizer exit flue gas temperature of 690°F (370°C) at 100 percent load, but only 570°F (300°C) at 50 percent load [37]. For low-load operations, an economizer bypass can be used to raise the flue gas temperature. An economizer bypass diverts part of the hot flue gas from within the economizer through a bypass duct and mixes it with the relatively cooler flue gas exiting the economizer. An economizer feedwater bypass also raises the flue gas temperature. The use of an economizer bypass results in less energy transfer to the feedwater for steam generation; consequently, there is a small reduction in boiler efficiency. Lower boiler efficiencies require more fuel to be burned to meet the required boiler steam output.

Residence Time and Space Velocity

Residence time is the time the reactants are within the reactor. Higher residence times generally result in higher NO_x removal rates. Temperature also affects the required residence time. The required residence time decreases as the temperature approaches the optimum temperature for the reduction reaction. Residence time is often expressed as space velocity, the inverse of residence time. The space velocity of a reactor is experimentally determined from the measured flue gas flow rate divided by the superficial volume of the catalytic reactor. The NO_x removal efficiency increases with decreasing space velocity (i.e., increasing catalyst volume) for a given flue gas flow rate.

The optimal residence time for an SCR system is a function of the number of active catalyst sites available for the reduction reaction and the gas flow rates within those active sites (interstitial flow rate). The “area velocity” is a parameter used by SCR vendors that relates the number of sites and the interstitial flow rate to residence time. The area velocity is defined as the space velocity divided by the catalyst pore surface area (specific surface area). For coal-fired boilers, typical specific surface areas range from 90 to 3,800 square feet per cubic feet (ft^2/ft^3) (300 to 1,200 square meters per cubic meters [m^2/m^3]) [37]. Increasing the catalyst specific surface area increases the NO_x removal for a given flue gas flow rate. This can be accomplished by either increasing the catalyst volume, which increases the reactor size, or increasing the pore space of the catalyst, which generally increases the catalyst cost.

Degree of Mixing

The reagent must be dispersed and mixed throughout the flue gas to ensure sufficient contact between the reactants. Mixing is performed by an injection system that injects pressurized gas-phase ammonia into the flue gas. The injection system controls the spray angle, velocity, and direction of the injected reagent. Some systems inject the ammonia with a carrier fluid, such as steam or air, to increase penetration into the flue gas. Injection systems are application specific. Numeric modeling of the flue gas and reagent flow optimizes the design of the injection system (see Section 2.2.6).

Mixing of the flue gas and ammonia occurs before entering the SCR reactor. If mixing is not adequate, the NO_x reduction is inefficient. SCR designs must incorporate adequate duct length between the ammonia injection and the reactor inlet to allow for mixing. Mixing patterns can be improved by

- Installation of static mixers upstream of the reactor;
- Increasing the energy imparted to the injected fluids;
- Increasing the number of injectors and/or injection zones; and
- Modifying the nozzle design to improve the reagent distribution, spray angle, and direction.

Stoichiometric Ratio Factor

The stoichiometric ratio factor (the moles of reagent injected per mole of NO_x removed) defines the quantity of reagent needed to achieve the targeted NO_x reduction. Theoretically, based on reaction equations 2.1(a) and (b), one mole of NO can be removed with one mole of ammonia and one mole of NO_2 can be removed with two moles of ammonia. Since NO_x is mostly comprised of NO (approximately 95 percent), the theoretical stoichiometric ratio for NO_x is close to 1.05 mole of ammonia per mole of NO_x. Hence, SCR systems typically employ a stoichiometric ratio of 1.05 moles of ammonia per mole of NO_x [37]. This assumption of an almost one-to-one linear relationship between the quantity of reagent and the NO_x removed is good up to about 85 percent NO_x reduction [50]. Above 85 percent, the removal efficiency begins to level off and more than the theoretical amount of ammonia is required for additional NO_x removal because of reaction rate limitations. Because capital and operating costs depend on the quantity of reagent consumed, the stoichiometric ratio factor is an important design parameter that is determined by the SCR designer.

Inlet NO_x Concentration

The concentration of the reactants also affects the reaction rate of the NO_x reduction process. In general, higher inlet NO_x inlet concentrations result in higher NO_x removal efficiencies due to reaction kinetics [37]. However, NO_x concentrations higher than approximately 150 ppm generally do not result in increased performance. Low NO_x inlet levels result in decreased NO_x removal efficiencies because the reaction rates are slower, particularly in the last layer of catalyst [37]. The percent removal efficiency achieved is dependent on the inlet NO_x concentration, so that SCR that follow other NO_x controls such as LNB or FGR may achieve an efficiency less than 90 percent. In general, though, SCR achieves greater removal efficiencies than SNCR on sources with low inlet NO_x levels, such as natural gas–fired boilers.

For a given NO_x removal efficiency, higher NO_x levels at the SCR inlet require more catalyst volume. For example, to achieve 90 percent NO_x removal requires 10 percent more catalyst at an inlet NO_x level of 1.7 lb/MMBtu versus an inlet NO_x level of 0.8 lb/MMBtu [58].

Ammonia Slip

Ammonia slip refers to the excess reagent passing through the reactor. Ammonia in the flue gas causes a number of problems, which were discussed in Chapter 1, SNCR, including health effects, visibility of the stack effluent, salability of the fly ash, and formation of ammonium sulfates. Limits on acceptable ammonia slip, imposed by either regulatory limits or design requirements, place constraints on SCR performance.

Ammonia slip does not remain constant as the SCR system operates but increases as the catalyst activity decreases. Properly designed SCR systems, which operate close to the theoretical stoichiometry and supply adequate catalyst volume, maintain low ammonia slip levels, approximately less than 2 ppm [1]. While ammonia slip levels in operating permits are typically in the range of 2 to 10 ppm, in actual practice lower slip levels are achieved, and the slip levels approach permitted levels only when the catalyst is near the end of its service life [1]. Ammonia slip increases with the NSR ratio [59]. Ammonia slip monitoring instruments are commercially available and are in place and operating at a number of coal-fired units. Facilities typically install ammonia slip monitors between the SCR and the air heater and may measure at one or several points. These systems monitor ammonia slip and help the unit maintain slip levels of 2–3 parts per million by volume (ppmv) or less. The capital cost for one ammonia slip monitoring instrument is estimated to be $40,000 for a single measurement point and up to $70,000 in capital cost for three measurement points [60]. Another method for quantifying ammonia slip is to determine the ammonia concentration in collected fly ash [61].

Raw materials at some cement kilns contain constituents that release ammonia to the kiln gas stream when heated to high temperature. Some cement plants have ammonia in the kiln exhaust gas without injecting any ammonia into the gas stream. Therefore, the ammonia slip from unreacted ammonia injected for SCR is difficult to differentiate from the natural fluctuations in ammonia in the stack gas. For this reason, it is important to understand the level of raw material derived ammonia emissions when designing an SCR control system for cement kilns [10].

Catalyst Activity

Catalyst activity is a measure of how much the catalyst accelerates the NO_x reduction reaction rate. A higher catalyst activity results in a faster reaction rate and more NO_x removal. Catalyst activity is a function of many variables, including catalyst composition and structure, diffusion rates, mass transfer rates, gas temperature, and gas composition [62]. As the catalyst activity decreases, the NO_x reduction reaction rate also decreases. This results in lower NO_x removal and higher ammonia slip levels.

The following equation that is in exponential form (e) describes the deactivation of the catalyst activity, K, with time, t [63]:

$$K = K_o e^{(-t/\tau)} \tag{2.2}$$

where K_o is the original catalyst activity and τ is the catalyst operating life time constant. Figure 2.4 shows a typical catalyst deactivation curve based on Equation 2.2. As the catalyst activity decreases, the NO_x removal efficiency is usually kept constant by injecting more ammonia, thereby increasing the ammonia slip. When the ammonia slip reaches the maximum design or permitted level, the catalyst or a catalyst layer must be cleaned, rejuvenated, or regenerated, or new catalyst must be installed.

Catalyst Reaction Selectivity

SCR favors the NO_x reduction reaction over competing reactions if the reactants are at the appropriate temperature and oxygen is present. However, competing reactions still occur, and the catalyst accelerates these reactions as well. Each catalyst has different chemical reaction selectivity properties. In general, catalysts promote the formation of two undesirable compounds, SO_3 and nitrous oxide (N_2O). SO_3 is formed by the oxidation of SO_2 to SO_3, which occurs both during combustion of sulfur-containing fuel and over the catalyst. In cement plant applications, the fuel sulfur is incorporated into the clinker, however, pyritic sulfur contained in some raw materials is oxidized and released into the kiln gas stream [10]. Sulfur oxides (SO_x) are regulated under the 1990 Clean Air Act. SO_3 reacts with ammonia in the flue gas to form ammonia sulfates. Ammonium sulfur salts deposit on the catalyst and on downstream equipment such as the air preheaters. SO_3 and ammonium sulfate formation is primarily a concern for higher sulfur coals. Additional costs for air preheater modifications are included in the cost analysis when the sulfur content of fuel is greater than 2 percent or the SO_2 content of the fuel is greater than or equal to 3 lb/mmBtu. Increasing the number of catalyst layers, while increasing NO_x removal efficiency, can also lead to an increase in the conversion of SO_2 to SO_3 [64]. In addition, SO_3 emissions cause “blue plume” from the unit and have become an emissions concern. Newer catalysts, however, have been developed that limit the formation of SO_3 [41], and these catalysts are now commonly used in SCR installations in the U.S. N_2O is both an ozone depleter and a greenhouse gas. N_2O has a global warming potential (GWP) of 298 as compared to carbon dioxide (CO_2).[8]

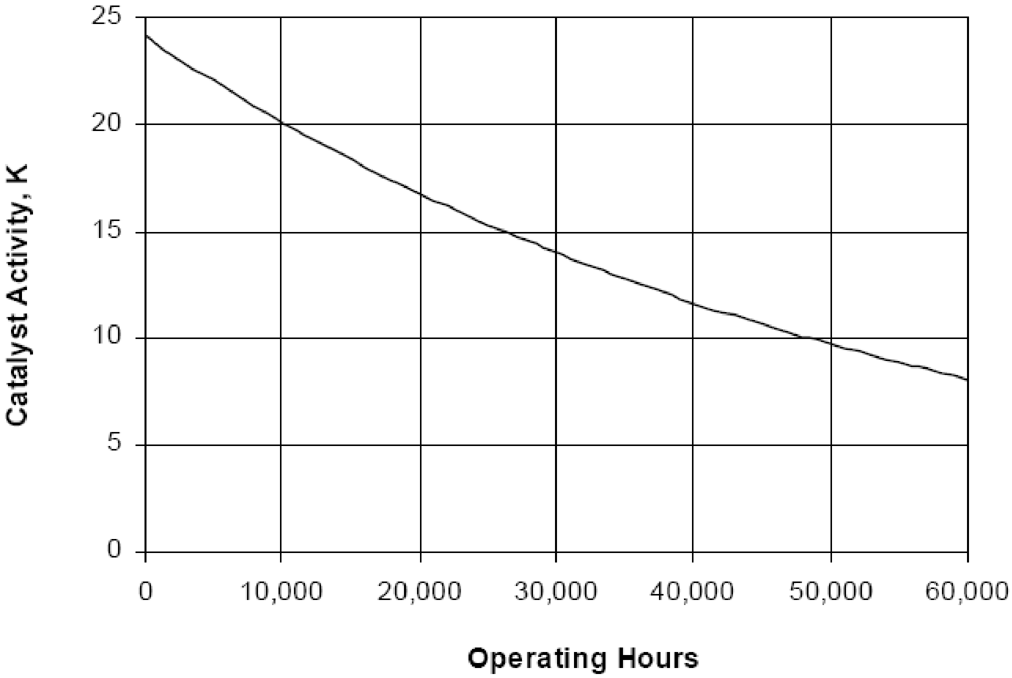

Figure 2.4: Typical Catalyst Deactivation per Equation 2.2 with K_O = 24.12; τ = 55,000

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[8] It should be noted that EPA issued a final rule on November 29, 2013 that changed the GWP of N_2O to 298 (from 310) as part of a notice of data availability concerning the Mandatory Greenhouse Gas Rule. The November 29, 2013 notice can be found in the Federal Register at http://www.gpo.gov/fdsys/pkg/FR-2013-11-29/pdf/2013-27996.pdf .

Pressure Loss

The flue gas pressure decreases as the flue gas flows across the catalyst. The decrease in pressure is a function of the length of the catalyst and the catalyst configuration. Deposition of fly ash and other particulates on the catalyst over time increases this pressure drop across the catalyst. The flue gas pressure can be increased by installing new draft fans or by upgrading existing fans. To minimize the pressure loss across the catalyst, the SCR reactor ductwork can be expanded and flow rectifiers and turning vanes can be installed. Pressure loss is of greater concern in turbine applications, which rely on air flow rather than heat transfer to generate power.

Pressure loss may also be an issue in cement kiln SCR applications as the dust loading in cement kilns can be in excess of 80 grams per cubic nanometer) (g/Nm^3) [10]. In addition, cement dust can be sticky, creating residue buildups that can be difficult to remove, and generally result in higher pressure loss than observed in other industrial operations. As such, it may require additional catalyst cleaning and catalysts with larger pitch [10].

Ash Management and Dust Loading

Ash and dust loading in the SCR catalyst may occur from both fuel combustion (e.g. coal) and raw materials (e.g., cement raw materials) from the process. The amount of ash or dust loading varies depending on fuel used, source type, and raw material feeds to the process. These particulates can mask and block the catalyst surface and inhibit the SCR NO_x reduction efficiency. One source indicates dust loading of 6.7 grams per dry standard cubic foot (gr/dscf) for a coal-fired boiler (Powder River Basin coal (PRB)), dust loading of 4.2 to 15.5 gr/dscf for a typical long wet cement kiln, and 20.1 to 40.2 gr/dscf for a typical PH/PC cement system [65]. One cement industry preheater kiln in Europe has a dust loading of 35.0 gr/dscf (80 g/m^3) [10].

Catalyst Pitch

Catalyst pitch is a term used with honeycomb and metal plate catalyst, and it affects the flue gas velocity in interstitial spaces [37]. As shown in Figure 2.5, pitch (represented as p) is the width of the catalyst cell plus the cell wall thickness, a. For a given flow rate, wider pitch will result in lower interstitial gas velocities. Appropriate catalyst pitch is important to ensure that ash will not deposit and bridge over catalyst cells and pores. Plugging of the catalyst reduces the effective surface area by decreasing the number of active sites available for the NO_x reduction reaction.

For high-dust applications, such as cement kilns and coal-fired boilers, catalysts with larger openings or pitch should be used to reduce the potential for catalyst plugging. However, the wider pitch reduces the surface area per unit volume and may necessitate additional catalyst volume to meet emissions limits for NO_x and ammonia slip.

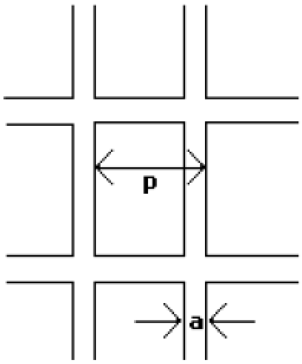

Figure 2.5: Pitch for a Honeycomb Catalyst Configuration

SO_2 and SO_3 Concentrations in Gas Streams

Sulfur in the fuel forms SO_2 during combustion in the boiler, i.e., boiler conversion rate, and some of the SO_2 is further oxidized to SO_3 within the boiler. In a coal-fired boiler, approximately 0.5 to 1.5 percent of SO_2 is converted to SO_3 at typical combustion temperatures, however, as the temperature drops more SO_3 can be formed.[9] Additional SO_3 is formed when SO_2 passes through the SCR catalyst (i.e., SCR conversion rate). SCR catalysts can be designed for ranges of conversion such as 0.2 to 0.8 percent per catalyst layer (for 3 layers of catalyst, this would result in conversion of 0.6 to 2.4 percent). SO_3 will combine with water to form sulfuric acid (H_2SO_4) or sulfuric acid mist. Concentrations of SO_3 and H_2SO_4 of 6 to 10 ppm can cause a visible plume, or a blue plume [66].

Catalyst Deactivation

Catalysts lose their activity over time for various reasons. The primary mechanisms for catalyst deactivation and surface area loss are discussed below.

Poisoning – Certain fuel constituents that are released during combustion act as catalyst poisons. Catalyst poisons include calcium oxide, magnesium oxide, potassium, sodium, arsenic, chlorine, fluorine, and lead. These constituents deactivate the catalyst by diffusing into active pore sites and occupying them. Catalyst poisoning represents the main cause of catalyst deactivation. The water-soluble alkali compounds are known to react with active catalyst sites rendering them inert. The form of the alkali metals is important as water soluble alkali metals have been found to be more detrimental to catalyst activity.

Arsenic is a concern for boilers that reinject fly ash or burn coals that have moderate to high arsenic content with a free lime (CaO) concentration in the ash of less than 2 percent. Some facilities have found it useful to inject limestone into the furnace. The calcium oxide reacts with arsenic to form calcium arsenate, which affectively removes the arsenic from the gas stream thereby preventing it from binding to catalyst and rendering them inactive [67]. The activity of poisoned catalysts may be restored by rejuvenation or regeneration.

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[9] Another source indicates that the amount of SO_3 is approximately 0 to 0.2 percent of the total SOx from the combustion process [51].

In cement kilns, both the raw materials and the fuels can contain compounds that poison catalysts, including inorganic compounds of potassium, sodium, and chlorine. Alkali metal compounds in the raw materials have relatively low melting points and upon reaching the combustion zone of the kiln will readily volatize. Although these compounds are also found in coal-fired boilers, they may be present in higher concentrations in cement kiln exhausts [10]. Catalyst poisoning by phosphorus, chromium, and lead compounds is believed to be a lesser concern for cement kilns than for coal-fired boilers [10]. These compounds are expected to be present in lower concentrations in cement kiln exhaust gases than is typical of coal-fired boilers [10, 28]. Arsenic poisoning is not expected to be an issue for cement plants because the high concentration of CaO should react with any arsenic in the exhaust gas before it reaches the SCR catalyst [65, 67]. However, high CaO levels combined with high sulfur concentrations can result in the formation of calcium sulfate ($CaSO_4$) in cement kilns, which can reduce catalyst activity through masking [10].

Thermal Sintering – High flue gas temperatures within the SCR reactor cause sintering, a permanent loss of catalyst activity due to a change in the pore structure of the catalyst. Thermal sintering can occur at temperatures as low as 450°F (230°C). The amount of thermal sintering depends on the composition and structure of the catalyst. Newer catalyst materials are less susceptible to thermal sintering, which increases their operating life.

Blinding, Plugging, and Fouling – Ammonia-sulfur salts, fly ash, and other particulate matter in the flue gas cause blinding, plugging, or fouling of the catalyst. The particulate matter deposits on the surface and in the active pore sites of the catalyst. This results in a decrease of the number of sites available for NO_x reduction and an increase in flue gas pressure loss across the catalyst.

Erosion – Impingement of particulate matter and high interstitial gas velocities erode the catalyst material. Catalysts with hardened leading edges or increased structural strength are less susceptible to erosion. However, increasing catalyst strength through hardening reduces the number of active pore sites.

Aging – Catalyst aging is a change in the physical and chemical properties of the catalyst pores that occurs over time.

A number of measures can be taken to decrease the rate of deactivation and deterioration of the catalyst. These measures are discussed below.

Catalyst Formulation – Each catalyst formulation has different physical and chemical properties. Catalyst formulations with the following properties will have decreased deactivation:

- Increased activity per unit volume;
- Greater thermal resistance;
- Chemical and physical resistance to poisons;
- Wider thermal operating range;
- Greater structural strength and hardened leading edges; and
- Lower interstitial velocities (i.e., wider catalyst pitch).

To obtain the optimal catalyst formulation and SCR design for an application, the catalyst supplier and SCR vendor should be informed of the fuel and raw material constituents, such as sulfur, chlorine, fluorine, alkali metals, lead, arsenic, and trace metals. The SCR must be designed to operate for the full range of values for the constituent concentrations, therefore, the ranges of these data should be provided to the catalyst supplier. These constituents can be determined by chemical analyses. The associated analytical data can then be used to modify the catalyst composition, determine the appropriate catalyst volume, predict catalyst regeneration and replacement schedules, and design the SCR reactor components.

Soot Blowers and Sonic Horns – Deposits on the surface of the catalyst can be dislodged by soot blowers, which are generally installed between each catalyst layer and operated on a periodic basis, such as once a week. For some higher dust SCR systems, e.g., in the cement industry, more frequent and even continuous catalyst cleaning may be needed. A sonic horn is another option that prevents accumulation of ash deposits on the catalyst surface. A sonic horn may operate at a typical frequency of 10 seconds every 10 minutes [41].

Turning Vanes and Rectifier Grids – Particulate matter can be removed from the flue gas by gas-flow turning vanes and flow rectifier grids near the front of the catalyst layer. Particles impact the surface of the vanes or grid and fall out of the flue gas stream. In addition to removing particles, turning vanes and flow rectifier grids decrease the linear velocity of the flue gas and align its vector with the flow path of the catalyst.

Catalyst Management Plan

Catalyst deactivation is an inherent part of the SCR process. As the catalyst activity decreases with time, the NO_x reduction reaction rate decreases and ammonia slip increases. When the ammonia slip level reaches the design limit, the catalyst must be replaced with regenerated catalyst or new catalyst must be added. The catalyst life is the time the catalyst activity for a given catalyst volume (layer volume) maintains ammonia slip below the design limit. Currently, vendor-guaranteed life for a catalyst layer in coal-fired applications is typically three years [56], and actual catalyst layer lifetimes in such applications are often in the 5 to 7-year range, depending on the condition of untreated flue gas [37]. Gas- and oil-fired applications experience even longer catalyst layer lifetimes.

A catalyst management plan (CMP), as described in a “saw-toothed” graphic shown in Figure 2.6, schedules periodic replacement of catalyst to maintain ammonia slip limits (this CMP has a maximum ammonia slip design value of 2 ppm [68]). In the past, CMP descriptions focused mainly on the catalyst replacement schedules; however, today, a CMP is a comprehensive catalyst management strategy that incorporates both SCR equipment management and catalyst management, along with attention to changes in regulatory requirements. This more comprehensive approach is needed with the move to year-round operation of SCRs for facilities that operate continuously. While operation of SCR for compliance with the NO_x State Implementation Plant (SIP) Call (1998)[10] requirements typically called for ozone season operation only, year round operation is necessary to comply with more

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[10] U.S. EPA. Final Finding of Significant Contribution and Rulemaking for Certain States in the Ozone Transport Assessment Group Region for Purposes of Reducing Regional Transport of Ozone. September 24, 1998. Available at https://archive.epa.gov/ttn/ozone/web/pdf/nxsip.pdf.

recent regulations (e.g., The Clean Air Interstate Rule (CAIR),[11] 1999 Regional Haze Rule,[12] an Ozone Transport Commission (OTC) initiative, and state rules such as the North Carolina (NC) Clean Smokestacks Rule that took effect in 2009[13] and the Texas SIP requirements for the Houston area),[14] to generate NO_x credits, or to comply with settlement agreements with the U.S. EPA and Department of Justice. Continuous, ongoing collection and documentation of data on plant loading and cycling, fuel demands and variation, and ongoing NO_x performance and SO_2 conversion, which can then be compared to catalyst activity data, is conducted to create the plant operating history. Some companies have developed computer software that collects these data and optimizes the costs for catalyst replacement options. In general, an annual SCR system inspection is conducted on the catalyst, the reactor, and the complete NH_3 storage and injection system. Inspection of the catalyst includes a physical inspection along with catalyst sampling and analysis on a bench-scale reactor for activity, SO_2 to SO_3 conversion rate, and pressure drop for each catalyst layer. Annual ammonia injection grid (AIG) tuning and optimization is also conducted to ensure uniform flow rate/velocity and uniform NH_3/NO_x molar distribution. Poor distribution of the NH_3/NO_x decreases the NO_x reduction and increases the NH_3 slip [69]. In situ measurements of the catalyst activity have been developed, where NO analyzers installed before and after the catalyst layer and a small supplemental ammonia controller allow increases in the NH_3/NO_x ratio and measurement of inlet and outlet NO_x samples, contained to a small area of the catalyst. In situ catalyst activity measurements may be important for year-round operation of SCR units [70].

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[11] U.S. EPA. Final Clean Air Interstate Rule. May 10, 2005. Files available at https://archive.epa.gov/airmarkets/programs/cair/web/html/index.html.

[12] U.S. EPA. Final Regional Haze Regulations. July 1, 1999. Available at https://www.govinfo.gov/content/pkg/FR-1999-07-01/pdf/99-13941.pdf.

[13] State of North Carolina. Department of Environmental Quality, Division of Air Quality. Clean Smokestacks Act. Available at https://deq.nc.gov/about/divisions/air-quality/air-quality-outreach/news/clean-air-legislation/clean-smokestacks-act.

[14] Texas Commission of Environmental Quality (TCEQ). Ozone attainment SIPs for Houston-Galveston-Brazoria area. Available at https://www.tceq.texas.gov/airquality/sip/sipplans.html.

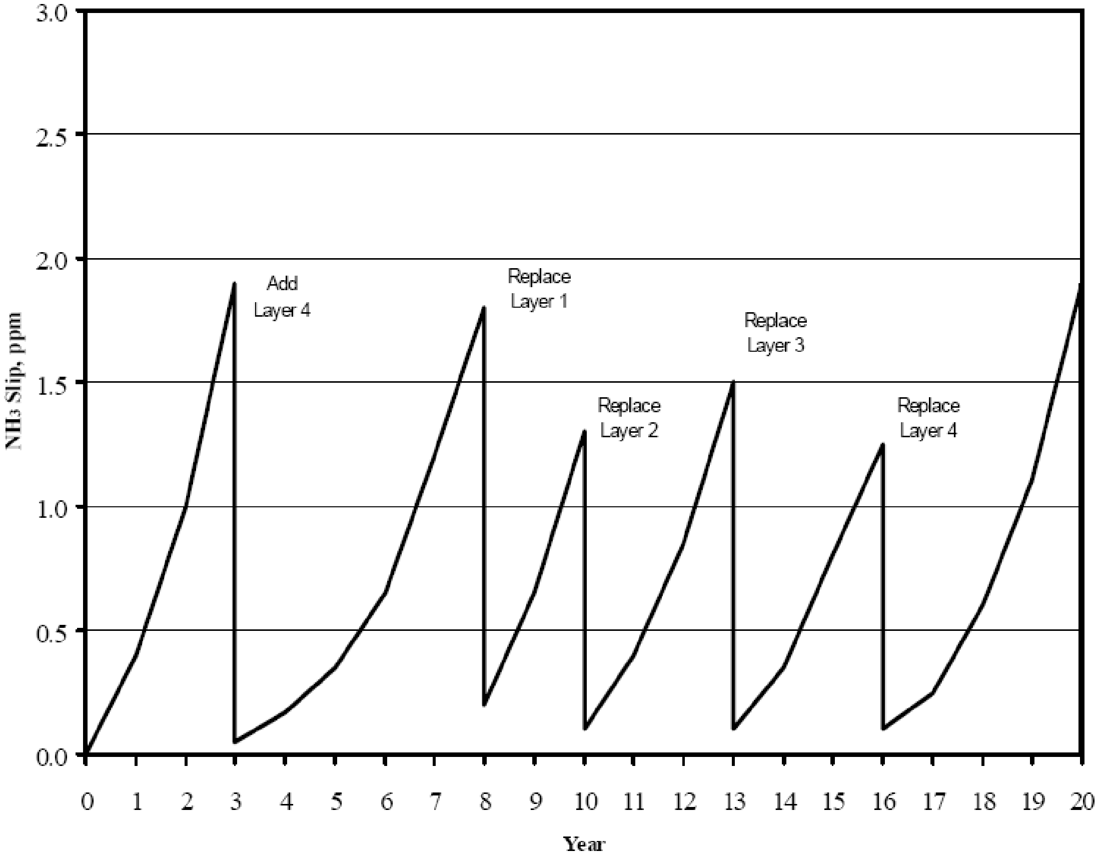

Figure 2.6: Typical Catalyst Management Plan [71]

Most CMPs call for the SCR reactor design to provide two or more layers filled with catalyst and one or more empty or spare catalyst layers (often called “2:1” design). When the initial catalyst layers deactivate to the point where ammonia slip reaches the maximum design value, the facility typically adds catalyst to the empty layer. Catalyst addition is managed so that the total catalyst activity of all the layers (the two or three older catalyst layers plus the new catalyst layers) is sufficient to meet the ammonia slip requirement for a relatively long period of time. As the catalyst continues to deactivate, ammonia slip begins to rise again. When ammonia slip again reaches the maximum design value, one of the older catalyst layers is removed and replaced. The catalyst analysis data identifies which layer should be replaced. With advances in catalyst regeneration, part of a comprehensive CMP is determining whether the catalyst can be regenerated or whether new catalyst must be used. Before a regeneration process is planned, the process should be prequalified on a catalyst sample. If additional catalyst capabilities are needed, review of recent catalyst technology advances for newer catalysts that achieve mercury reductions, lower SO_2 conversion rates, and lower load and temperature operation is advised, although some regeneration processes may offer improvements with these catalyst capabilities as well. Typically, the addition and replacement of catalyst layers is coordinated with plant outage periods if at all possible, and outage frequency should be considered in conjunction with the risk considerations for replacing sooner rather than later [69]. There would likely be additional costs or impacts (e.g., due to lost generation or production) if a facility is unable to coordinate with planned unit outages.

In the past, catalyst cost was a significant portion of the annual cost of operating an SCR system. Under the latest operating approaches that involve using a CMP, only a fraction of the total catalyst inventory, rather than the entire volume, is replaced at any one time. This distributes the catalyst replacement costs more evenly over the lifetime of the system and use of regenerated catalyst may also reduce the overall annual costs [69]. For applications with higher dust loading, such as the dust loading typical for cement kilns, the catalyst management plan may include more frequent catalyst replacement and regeneration schedules than would be typical for low-dust applications.

2.2.3 SCR System Configurations

Electric utility and large industrial boiler applications implement several different SCR system configurations, including high-dust, low-dust, and tail-end arrangements. In a 1997 report, the SCR configurations were reported as 88 percent high-dust SCR, 6 percent low-dust, and 6 percent tail-end [72].[15] More recently for the U.S, it was reported that most SCR configurations are high dust, only one facility has a low-dust SCR, and no tail-end SCR operate in the U.S. [57]. High-dust is generally considered the most economical and straightforward design provided sufficient space is available to construct the SCR close to the economizer and air pre-heater. Boiler units with space constraints must consider low-dust and tail-end SCR designs. SCR configurations for gas turbine applications depend on the type of engine cycle, such as combined-cycle or simple cycle. The various configurations for boilers and gas-fired turbines are discussed below. In addition, there are two different SCR reactor designs, full SCR and in-duct SCR, which are also discussed.

High-Dust SCR

Figure 2.7 shows a high-dust SCR system for coal-fired boiler applications. The SCR reactor location is downstream of the economizer and upstream of the air heater and particulate control devices. The flue gas temperature in this location is usually within the optimal temperature window for NO_x reduction reactions using metal oxide catalysts. In this configuration, however, the flue gas contains particulates when it enters the SCR reactor.

Coal-fired boilers generally use a vertical SCR reactor, where the flue gas flows downward through the catalyst. The reactor generally contains multiple layers of catalyst. The volume of catalyst required varies with each installation, as discussed previously. Soot blowers or sonic horns are installed to remove particulates from the catalyst surfaces. For designs that use a honeycomb catalyst, the catalyst pitch is typically about 7 to 9 millimeters (mm) (compared with 3 to 4 mm for gas-fired boilers) to allow easy passage of ash particles without deposition and ease of cleaning with soot blowers or sonic horns. To obtain uniform gas flow and remove particulates, high-dust SCR designs usually include turning vanes and a flow-rectifying grid in the ductwork prior to the reactor. High-dust SCR typically require 3 or 4 layers of catalyst [57].

A hopper at the bottom of the SCR reactor collects ash and particulates separated from the flue gas stream. The hopper outlet connects to the plant fly ash handling system for periodic removal of the accumulated ash. Flue gas exits the reactor via an opening at the top of the hopper

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[15] In a 2006 report, one utility/vendor reported that of their 24 SCRs, 71 percent were high-dust, 4 percent were low-dust, and 25 percent were tail-end [73]. These data are from a single vendor; the data above in the text represent multiple vendors.

and is directed to the air heater inlet. Some designs eliminate the need for hoppers by keeping flue gas velocities high enough in these areas that fly ash remains entrained in the flue gas.

Natural gas–and distillate oil–fired boilers generate flue gas that is relatively free of dust and SO_2 (for low-sulfur oil). Consequently, SCR systems for these boilers place the reactor upstream of the air heater, in the high-dust SCR configuration.

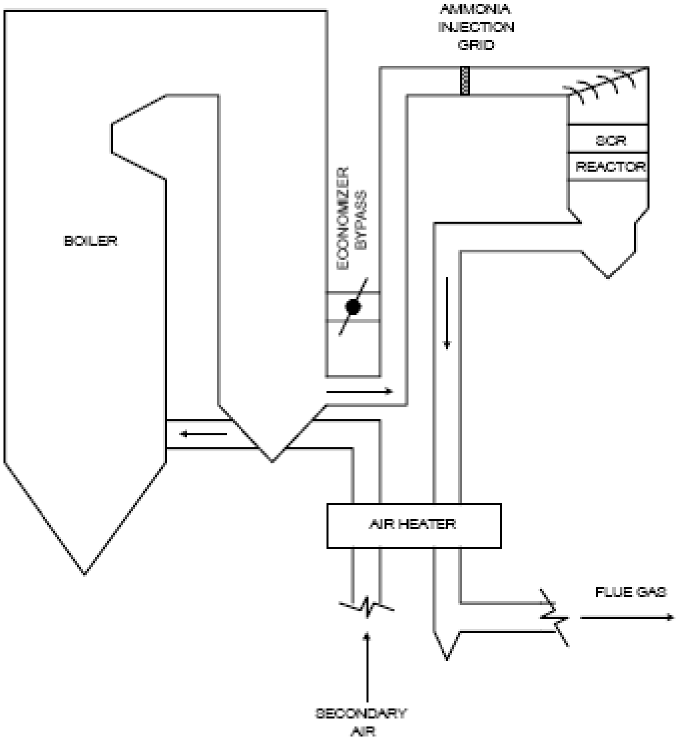

Figure 2.7: High-Dust SCR Arrangement [46]

Low-Dust SCR

Coal-fired units with an ESP located upstream of the air heater (hot-side ESP) typically use a low-dust SCR configuration. Figure 2.8 shows a low dust configuration, which locates the SCR reactor downstream of the ESP. In this location, the flue gas is relatively dust free. The ash removed by the ESP typically contains arsenic, alkali metals, and other constituents that are detrimental to catalyst performance and life.

A low-dust SCR system increases catalyst life by mitigating concentrations of particulates and catalyst poisons in the SCR reactor. In addition, low-dust SCR configurations do not need ash hoppers. For designs employing honeycomb catalyst, the catalyst pitch can be reduced to approximately 4 to 7 mm, resulting in lower catalyst volume. Low-dust SCR typically requires only 2 layers of catalyst [57]. Longer catalyst life, lower catalyst volume, and the elimination of the ash hopper mean lower costs for low-dust SCR compared to high-dust configurations. The only disadvantage of low-dust SCR is the temperature drop of the flue gas as it flows through the ESP. Flue gas temperatures generally do not decrease to the point where reheating is required. However, an increase in the size of the existing economizer bypass duct may be required to maintain the flue gas temperature within the optimal range.

In the low-dust SCR installed at a U.S. cement kiln in 2013, the gas stream passes through a hot electrostatic precipitator to remove the majority of the dust prior to entering the SCR. The gas stream exiting the SCR may then pass through a second, more efficient particulate control device to remove the remaining particulate to acceptable emissions rates.

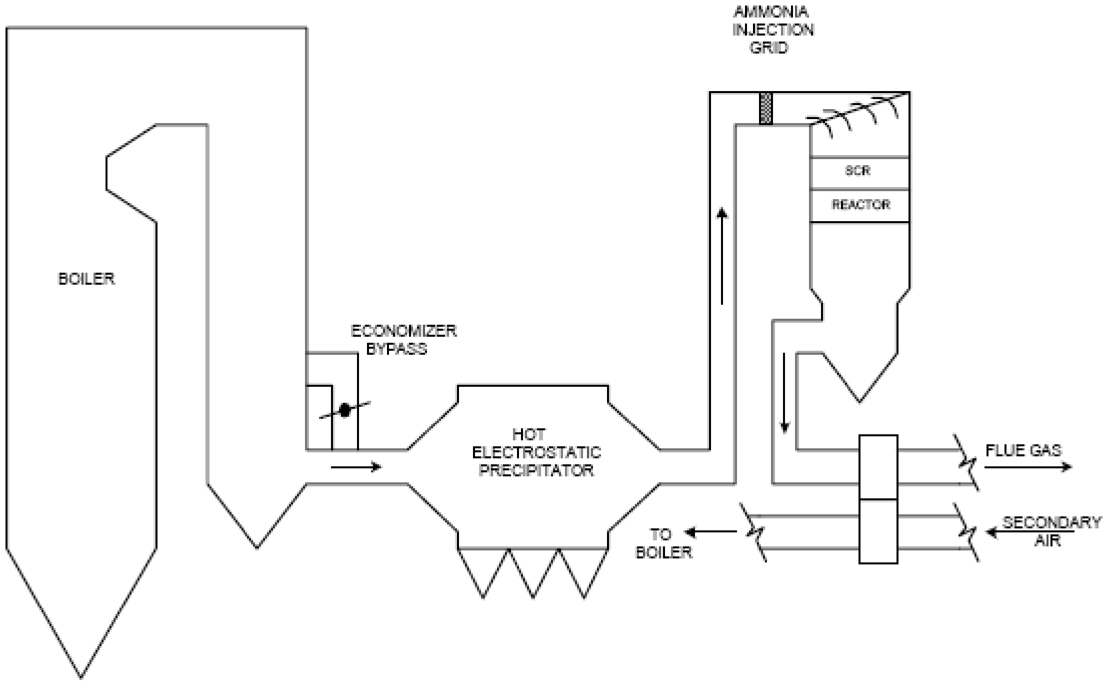

Figure 2.8: Low-Dust SCR Arrangement [46]

Tail-End SCR

The tail-end SCR configuration places the SCR reactor downstream of all air pollution control equipment installed on a unit. Figure 2.9 depicts a tail-end system for a plant with a particulate control device and a wet FGD system. The air pollution control equipment removes most flue gas constituents detrimental to SCR catalysts before the flue gas enters the SCR reactor. The tail-end SCR configuration is often a technically feasible alternative for situations where the high-dust SCR configuration is impractical [74].

Because the flue gas temperature at the tail end is below the range required for the NH_3/NO_x reaction, the flue gas must be reheated. Tail-end SCR systems use oil- or natural gas–fired duct burners or steam coil gas heaters for reheating. Some of the energy used to reheat the gas is recovered in a recuperating gas-to-gas heater. Some use catalysts specially designed for temperatures between 300 to 550^{o}F and low pressure drops, which lowers the costs flue gas reheating [75, 76, 77].

A tail-end system may have higher capital and operating costs than the other SCR systems because of the additional equipment and operational costs associated with flue gas reheating and heat recovery. However, these costs are in part offset by reductions in catalyst costs. Tail-end units require less catalyst because they can use catalysts with smaller pitch and higher surface area per unit volume. Tail-end SCR typically require only 2 layers of catalyst [57], although some use four half-layers of catalyst to allow for greater flexibility for catalyst replacement [78]. In addition, because there is less fly ash, catalyst poisons, and SO_2 in the flue gas for tail-end units, the catalyst lifetime is significantly increased, and less expensive catalyst may be used [57]. Some sources have reported catalyst lifetimes for tail-end SCRs to be over 100,000 hours [57, 74, 78]. The tail-end SCRs may also have longer lifetimes due to the lower operating temperatures and lower levels of dust and SO_3.

Tail-end SCRs have been used since the late 1980s and were initially used on coal-fired power plants. They are currently used at a variety of different applications in Europe, Japan, and the U.S., including power plants, incinerators, refinery crackers, cement plants, and ethylene crackers [74, 78]. They have been installed on units burning a wide range of fuels, including fuels of variable composition, such as biomass (including wood waste and chicken litter), hazardous waste, municipal waste, and wastewater sludge [79, 80]. They are often easier and less complex to install than the high-dust and low-dust SCR configurations for retrofit situations and can be installed with less disruption to production. The tail-end SCR configuration has been used in many retrofits of existing power plants in Europe. In some situations, particularly where combustion units have space constraints, the capital cost for retrofitting high-dust SCRs may be higher than for tail-end SCR [57]. Modular tail-end SCR systems are also available that are designed to be installed with minimal plant disruption [81].

One other major advantage of the tail-end SCR configuration is that its preheater enables the SCR to operate independently of the combustion unit. This arrangement enables greater operating flexibility, allowing the combustion unit to operate in a wider range of operating loads and fuel types [74]. Because tail-end units follow the ESP and wet scrubber, the flue gas has cooled and must be reheated to an appropriate temperature for the NO_x reaction to occur in the SCR. For tail-end units, the flue gas is typically sent through a gas-gas heat exchanger and then to either a natural gas-fired duct burner or steam coil to heat to the appropriate SCR operating temperature. Most tail-end SCR in Europe use steam coil reheating, which has advantages over a duct burner such as lower operating cost, no increase in flue gas flow rate from combustion byproducts, and no moisture condensation on the SCR catalyst.[16]

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[16] A case study for a tail-end SCR achieving 84 percent NO_x removal efficiency on a 600 MW boiler burning bituminous coal indicated annual reheating cost for steam coil of $2.5 million/yr and for natural gas burner of $12 million/yr (2008$) (assuming approximately $4/1000 lb steam and $8/1000 sft^3 natural gas) [57]. For comparison, the annual reheating cost for natural gas burner would be $7.8 million/yr (assuming approximately $5/1000 sft^3 natural gas).

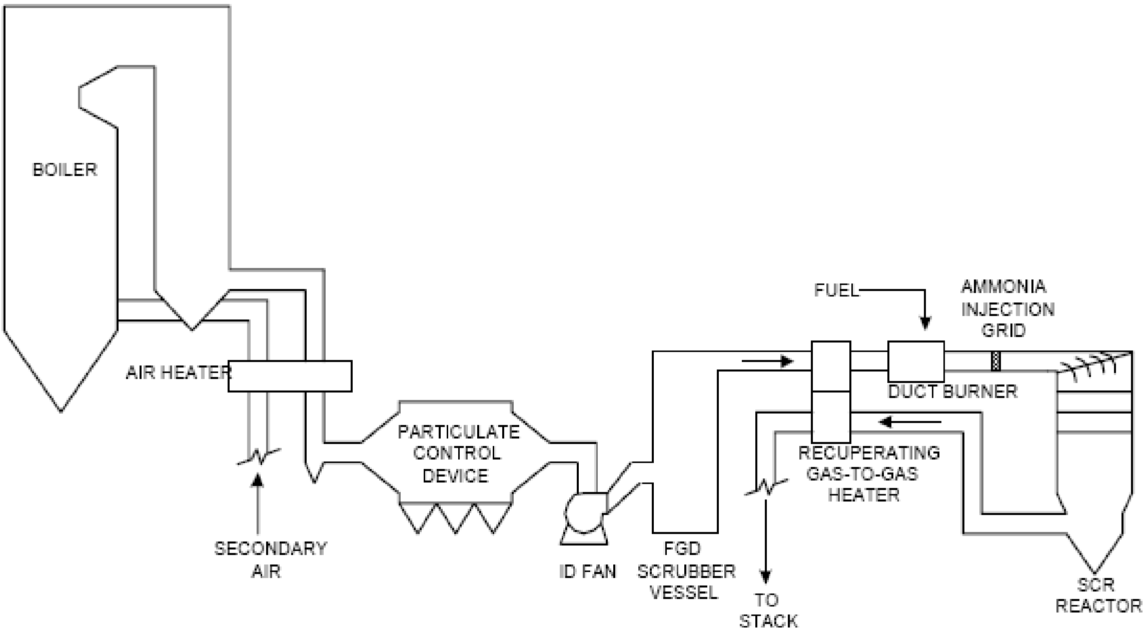

Figure 2.9: Tail-end SCR Arrangement [46]

Gas Turbines

Natural gas–fired turbine applications frequently use SCR technology for post-combustion NO_x control. There are two basic gas turbine configurations: combined cycle (cogeneration cycle) and simple cycle. The majority of SCR systems are installed as combined cycle applications. As shown in Figure 2.10, a typical combined-cycle SCR design places the reactor chamber within a cavity of the heat recovery steam generator system (HRSG), between the superheater and the evaporator. The flue gas temperature in this area is within the operating range for base metal catalysts. Most new HRSG units include a cavity designed to accommodate an SCR reactor. However, older HRSG units may not have sufficient space to house the SCR reactor within the HRSG. In these cases, a low-temperature SCR reactor may be installed after the HRSG. The high temperature SCRs used on simple-cycle turbines are generally not retrofitted to combined cycle turbines equipped with HRSG due to lack of space between the turbine and the HRSG [67, 82, 83]. Simple-cycle applications of SCR place the reactor chamber directly at the turbine exhaust, where the flue gas temperature is in the range of 850 to 1000°F (450 to 540°C). This requires the use of a high-temperature catalyst such as zeolite [46].

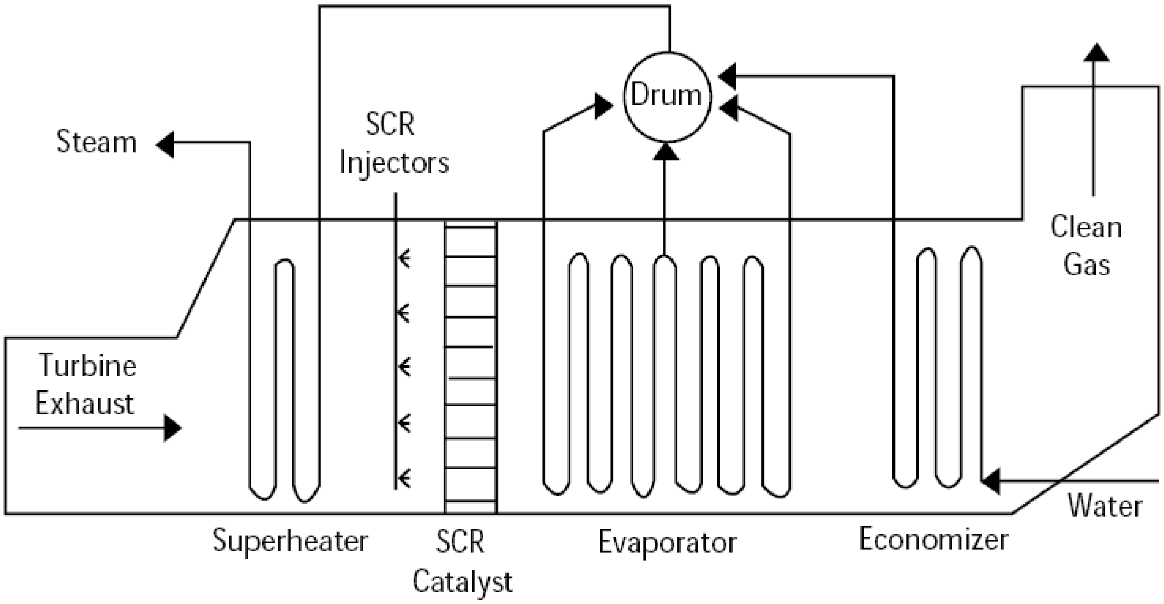

Figure 2.10: SCR Arrangement for a Combined-Cycle Gas Turbine [46]

Cement Kilns

SCR systems applied to cement kilns can have "tail-end", "low-dust", or "high-dust" configurations. Because of the potential for catalyst plugging, the "high-dust" configuration on cement kilns require catalyst cleaning mechanisms. The "low dust" and "tail-end" configurations avoid the costs of catalyst cleaning systems. Currently, three "high-dust" SCR systems[17], three "low-dust" SCRs[18], and one "tail-end" SCR[19] are known. The "high-dust" SCRs reportedly achieve control efficiencies of approximately 80 percent with inlet dust loading of up to 100 g/m^3. The "low-dust" SCRs are reported to have dust loadings less than 20 mg/m^3, while the inlet dust loading for the "tail-end" SCR is reported to be less than 10 g/m^3. [10, 11, 12]

SCR Reactor Designs

The reactor design affects the capital and operating costs of the SCR system and the CMP. There are two different types of SCR reactors: full SCR and in-duct SCR. Full SCR designs house the catalyst in a separate reactor chamber. The boiler flue gas must be ducted from the economizer outlet to the SCR reactor, then to the air heater inlet. A separate reactor allows a large volume of catalyst to be installed in layers, which increases NO_x reduction and catalyst

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[17] The first "high-dust" configuration SCR was installed on a preheater cement kiln at the Solnhofer Zementwerkes in Germany in 2001 and operated until 2006. Two other "high-dust" SCRs have been installed on preheater cement kilns at the Cementeria di Monselice plant in Bergamo, Italy in 2006 and the Mergelstetten plant in Germany in 2010.

[18] "Low-dust" configuration SCRs have been installed at the Sarche plant in Italy (2007), the Mannersdorf plant in Austria (2012), and the Joppa plant in the USA (2013). The Mannersdorf SCR is installed on a preheater cement kiln, while the Joppa SCR is installed on a long dry kiln. Both plants use an electrostatic precipitator to reduce particulate emissions entering the SCR. The Sarchi SCR is installed on a small Polysius Lepol kiln with no particulate controls, but low dust loading (reportedly less than 15 g/m^3).

[19] The Rohrdorf plant in Germany installed a "tail-end" SCR in 2011 on a preheater kiln.

lifetime. It also increases the duct length available for the mixing of reactants before entering the reactor chamber. However, a separate reactor requires a large amount of space adjacent the boiler to install the reactor and ductwork. The additional ductwork often necessitates upgrades to the draft fan system.

In-duct (inline) SCR systems house the reactor within the plant's existing ductwork rather than in a separate reactor chamber. The ductwork is generally enlarged to provide sufficient room for the catalyst. In-duct systems save on costs for the ductwork, reactor chamber, and induced draft (ID) fan. In-duct designs limit catalyst volume and mixing length; therefore, they are commonly used in conjunction with other NO_x control technologies [45]. Catalyst erosion is generally higher for in-duct systems. Installation and maintenance of in-duct systems typically require more boiler outages. Natural gas–fired boilers, which have low catalyst volumes, frequently employ in-duct systems. Coal-fired boilers frequently employ full SCR reactors but may apply in-duct SCR reactors where space limitations restrict the installation of a full reactor [45]. Cement kilns have also used full scale SCR reactors, rather than in-duct SCRs. The SCRs used for cement kilns have typically consisted of multiple catalyst layers and extensive catalyst cleaning systems. For example, the SCR systems installed at the Solnhofen cement plant in Germany and the Cementeria di Monselice plant in Italy used reactors with six catalyst layers, although only three layers were in use at a time [10].

2.2.4 SCR System Primary Equipment

The majority of SCR designs use Thermal $DeNO_x$®, an ammonia-based NO_x reduction system developed and patented by Exxon Research and Engineering Company in 1975. An SCR system consists of five basic steps:

- Receive and store the ammonia (or the urea reactant, followed by onsite ammonia production);
- Vaporize the ammonia and mix it with air;
- Inject the ammonia/air mixture at appropriate locations;
- Mix the ammonia/air with flue gas; and
- Diffuse the reactants into the catalyst and reduce the NO_x.

Although the basic steps in an SCR system are similar for all configurations, the system design and equipment specifications are somewhat different. A discussion of the SCR system design and equipment is given below for an ammonia reagent, high-dust, full reactor SCR for a 120 MW (approximately 1,200 MMBtu/hr) coal-burning utility boiler. These discussions are also pertinent to industrial applications. For example, cement kilns operating in the high-dust configuration would also require catalyst cleaning equipment [10]. The SCR process steps, related auxiliary equipment, and the potential impacts of SCR operation on existing plant equipment are also discussed. Simplified system flow schematics are presented in Figure 2.1 and Figure 2.7, and a list of equipment is presented in Table 2.3.

Table 2.3: Major Equipment List for an SCR Application

Item	Description/Size
SCR reactors (1–2)	Vertical flow type, 805,000 acfm capacity, 44 ft × 44 ft × 31 ft. high (excluding outlet duct and hoppers), equipped with 9,604 ft^3 of ceramic honeycomb catalyst, insulated casing, soot blowers or sonic horns, hoppers, and hoisting mechanism for catalyst replacement
Anhydrous ammonia tank (1 or more)	Horizontal tank, 250 pounds per square inch gauge (psig) design pressure, storage tanks 15,000 gal, 34-ton storage capacity
Air compressor (2)	Centrifugal type, rated at 3,200 acfm and 30 hp motor
Vaporizers (2)	Electrical type, rated at 80 kW
Mixing chamber	Carbon steel vessel for mixing or air and ammonia
Ammonia injection grid	Stainless steel construction, piping, valves and nozzles
Ammonia supply piping	Piping for ammonia unloading and supply, carbon steel pipe: 1.0-inch diameter, with valves and fittings
Soot blowing steam	Steam supply piping for the reactor soot-piping blowers, 2-inch diameter pipe with an on-off control valve and drain and vent valved connections
Air ductwork	Ductwork between air blowers, mixing chamber, and ammonia injection grid, carbon steel, 14-inch diameter, with two isolation butterfly dampers and expansion joints
Flue gas ductwork	Ductwork modifications to install the SCR modifications reactors, consisting of insulated duct, static mixers, turning vanes, and expansion joints
Economizer bypass	Ductwork addition to increase flue gas temperature during low loads consisting of insulated duct, flow control dampers, static mixers, turning vanes, expansion joints, and an opening in the boiler casing
Ash handling	Extension of the existing fly ash handling modifications system: modifications consisting of twelve slide gate valves, twelve material handling valves, one segregating valve, and ash conveyor piping
Induced draft fans	Centrifugal type, 650,000 acfm at 34 inches water gauge and 4,000 hp motor
Controls and instrumentation	Stand-alone, microprocessor-based controls for the SCR system with feedback from the plant controls for the unit load, NO_x emissions, etc., including NO_x analyzers, air and ammonia flow monitoring devices, ammonia sensing and alarming devices at the tank area, and other miscellaneous instrumentation
Electrical supply	Electrical wiring, raceway, and conduit to connect the new equipment and controls to the existing plant supply systems
Electrical equipment	System service transformer fans off (OA)/fans on (FA)/-60 Hz, 1,000/1,250 kilovolt-amperes (kVA) (65°C)
Foundations	Foundations for the equipment and ductwork/piping, as required
Structural steel	Steel for access to and support of the SCR reactors and other equipment, ductwork, and piping

Reagent Production, Storage, and Vaporization

As discussed previously, one of several reagents may be used in an SCR system, including anhydrous ammonia, aqueous ammonia, or urea. In the past, reagents have typically been purchased and stored before vaporization and use in the SCR. Ammonia (both anhydrous and aqueous) is the type of reagent most often used in SCR systems. Of about 230 utility boilers for which reagent type was reported in response to a survey in 2009, about 80 percent used ammonia, and 20 percent used urea [4]. Urea reagent is mostly used in SNCR systems [84], however, U.S. cement plants typically use 19 percent aqueous ammonia for SNCR systems and likely would use the same reagent for SCR applications. Another option that some facilities have recently adopted is to produce ammonia onsite from urea feedstock. The onsite ammonia production system may reduce or eliminate ammonia shipping, handling, and onsite storage. Load following by the onsite ammonia production system is extremely important for the proper operation of the SCR.

Several of the pros and cons of each ammonia system are shown in Table 2.4. In general, anhydrous ammonia is the least costly reagent; however, plant personnel and community safety, permitting, and other hazard planning concerns associated with its use may make this option less attractive and add to its cost. Aqueous ammonia is typically higher cost, given the energy required to vaporize or decompose the reagent, although some facilities have chosen this option over anhydrous ammonia to avoid some of the safety and planning concerns for anhydrous ammonia [41]. In general, as ammonia consumption increases, onsite urea-derived ammonia production is the most economical, while for lower consumption rates, aqueous ammonia may be the preferred economic option. For year-round operation, onsite urea-derived ammonia systems become economically competitive with 29 percent aqueous ammonia for plants around 800 MW and larger. For ozone season operation, onsite urea-derived ammonia systems become competitive with 29 percent aqueous ammonia at a plant size of 1,300 MW and larger [84]. The total cost of an SCR system with an onsite urea-derived ammonia system is approximately 2 to 5 percent more than an SCR system based on anhydrous NH_3 [14]. Another source reported a capital cost of $24 million for its onsite urea-derived ammonia system for a 1,300 MW unit delivering approximately 7,000 lb/hr NH_3, with a total capital investment of $175 million for the SCR system (not including the ammonia system) [85].

Table 2.4: Comparison of Ammonia Delivery Systems [84]

Measure	Anhydrous NH_3	Aqueous 19 percent NH_3	Aqueous 29 percent NH_3	Urea-derived NH_3
Risk level	Highest safety, hazard, permitting, and regulatory issues	Lower safety, hazard, permitting, and regulatory issues	Lower safety, hazard, permitting, and regulatory issues	Lowest safety, hazard, permitting, regulatory issues
Energy	Lowest energy use	High energy use	Medium energy use	Not known
Product deliveries	Fewest product deliveries	Large number of product deliveries	Medium number of product deliveries	Not applicable
Capital cost	Low capital cost	Low capital cost	Low capital cost	Highest capital cost
Annual cost	Lowest annual cost	High annual cost	Medium annual cost	Medium annual cost

Aqueous ammonia is typically available as a 19 or 29 percent solution in water. Anhydrous ammonia is nearly 100 percent pure ammonia and stored as a liquid under pressure. Table 2.2 presents the properties of aqueous ammonia (29 percent) and anhydrous ammonia.

If facilities receive anhydrous or aqueous ammonia from offsite, it is received via a tank-truck or rail car and pumped into one or more storage tanks. Ammonia is typically stored as a liquid in horizontal cylindrical tanks. An aqueous ammonia tank is an enclosed tank rated for only slightly elevated pressure, while an anhydrous ammonia tank is a pressure vessel rated for at least 250 psig. An anhydrous ammonia tank can be filled to only about 85 percent of its total volume to allow for a vapor space above the liquid level. The tanks are equipped with level and temperature indicators; a manway, vent, and access ladder; and other appurtenances. The applicability of heat tracing, insulation, and seismic design criteria are determined based on site-specific conditions. The tank should be mounted on a concrete pad and surrounded by a spill containment structure such as a dike.

SCR applications on large boilers generally require one to five tanks with volumes ranging from 10,000 to 20,000 gallons per tank to maintain sufficient volume for 1 to 3 weeks of SCR operations. The ammonia storage tank may be sized for 3 to 30 days of storage. The high end of the range would be used in conservative design practice. Alternatively, if ammonia distributors are located nearby and considered reliable, the plant owner might opt for a smaller tank, sized for fewer days of ammonia storage.

Aqueous ammonia is vaporized by pumping it to a vessel where it mixes with hot air. The air from the dilution air fan is heated in an electric heater or other heat exchanger (e.g., steam). In most aqueous ammonia applications, the ammonia-air mixture leaves the vaporizer vessel at about 300°F (150°C). The vaporization energy required for aqueous ammonia is much greater than that required for anhydrous ammonia because the water in the aqueous ammonia solution also must be vaporized.

If anhydrous ammonia is used, it is fed to the electrical vaporizer by gravity, and the vaporized gas is returned to the storage tank vapor space. Vapor is drawn from the vapor space and piped to the ammonia/air mixer. Alternatively, liquid anhydrous ammonia may be pumped to a vaporizer and piped to the ammonia/air mixer.

Ammonia (aqueous or anhydrous), diluted with air at a ratio of about 20:1 (air:NH_3), is transported to the ammonia injection grid. The high proportion of air helps ensure good mixing of air and ammonia and keeps the mixture below the flammable limit.

Onsite Urea-Derived Ammonia Production

Use of onsite ammonia production systems for feed into SCR reactors has increased, mostly as a result of safety concerns [41]. Several process types can produce ammonia from urea, including (1) systems that hydrolyze an aqueous urea solution to form ammonia and carbon dioxide (and water); (2) systems that melt solid urea and mix the liquor with steam, where it reacts the melted urea across a catalyst to form ammonia, carbon dioxide, and water vapor; and (3) systems that atomize an aqueous urea solution in a decomposition chamber containing a hot air stream or flue gas stream at 800 to 1200°F (430 to 650°C) to form ammonia and isocyanic acid (NHCO) [41]. The capital costs of these systems vary with design [41]. Urea feedstock is available in solid form or as urea solution in deionized water [84]. Almost all urea-to-ammonia systems use solid urea [84]. Urea-to-ammonia systems typically include dry urea unloading equipment, storage silo, dissolving tank using deionized water, feed tanks, feed pumps, a solution heater, and a hydrolyzing reactor or decomposition chamber, depending on the type of process used [84]. Descriptions of two types of these systems are provided.

One of the urea-derived ammonia production systems converts urea by thermal hydrolysis to ammonia, carbon dioxide, and water vapor. As shown in Figure 2.11, the system consists of urea storage, handling, and dissolvers; a reactor feed tank; circulation; feed pump; condensate skids; a hydrolysis reactor; and ammonia flow control units. In the urea solution preparation, dry urea is fed batch-wise from delivery trucks directly to the dissolver along with deionized water, and the urea solution is stored. In the solution feed and control system, urea solution is transferred to the reactor feed tank, and the feed pump meters the urea solution to the reactor, which is heated using steam. The hydrolysis reactor is a kettle-reboiler type heat exchanger that operates at 80 psig and at a temperature in the range of 280 to 310°F (140°C to150°C). The urea-to-ammonia reaction occurs in two steps: the first reaction produces ammonium carbamate (NH_4COONH_2) from urea and water, and the ammonium carbamate breaks down into carbon dioxide and ammonia in the presence of heat. The temperature of the reactor drives the rate of ammonia production. Trace amounts of formaldehyde are present when formaldehyde-conditioned urea is used as feedstock; operating the hydrolysis reactor at a pH of 9 or higher limits formation of urea formaldehyde polymeric resins that can deposit on the reactor [86].

One of the urea-derived ammonia production systems decomposes urea to generate ammonia that is fed to the AIG, as shown in Figure 2.12. The system consists of a blower, decomposition chamber, urea storage, chemical pumping system, and process controls. In the urea storage and pumping system, dry urea from the storage tank is mixed in a solution tank with water and transferred to an aqueous urea solution storage tank. Filtered ambient air is fed into the decomposition chamber through the use of a blower with automatic dampers to control discharge flow and pressure. In the chamber, a burner is fired downstream of the dampers, and an aqueous urea solution that is supplied by the storage and pumping system is sprayed into the post combustion gases by injectors. The decomposition occurs under a specific temperature and residence time, with the decomposition temperature ranging from 600 to 1000°F (320 to 540°C), and the urea is decomposed to ammonia and isocyanic acid. The outlet ammonia stream from the decomposition chamber feeds into the AIG system for the SCR [87].

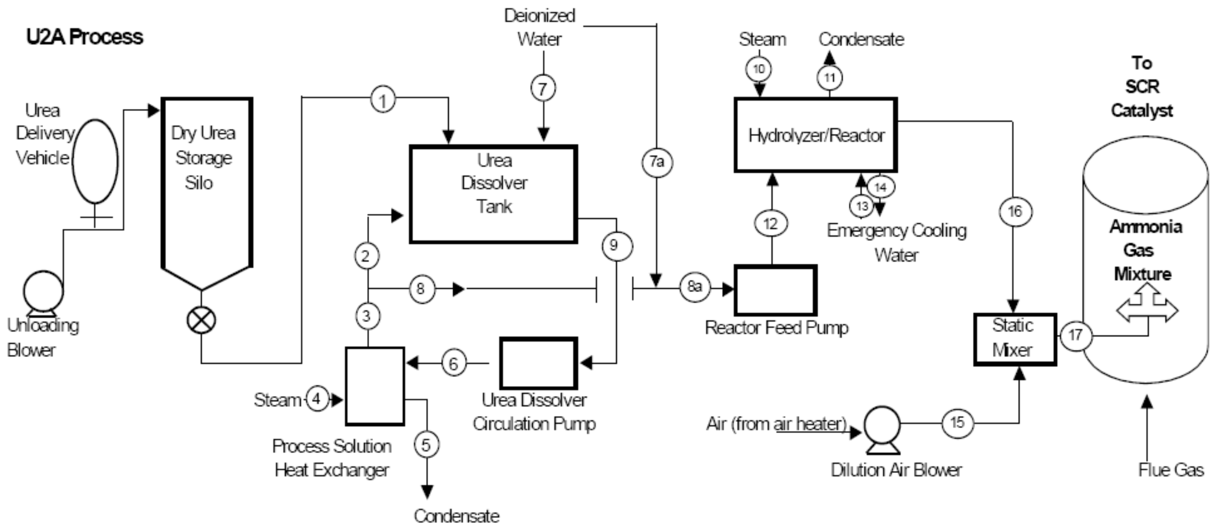

Figure 2-11: Urea-Derived Ammonia Production System Using U2A system [84]
Used by permission of R. Salib of URS Washington Division (formerly Washington Group International).

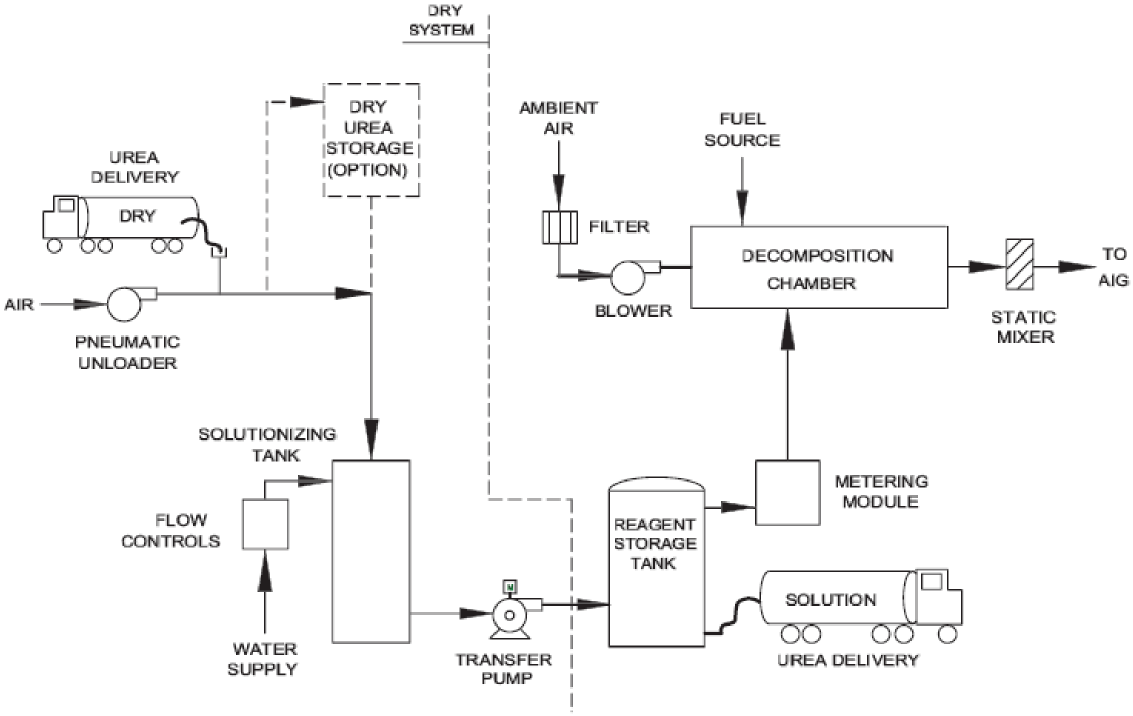

Figure 2.12: Urea-Derived Ammonia Production System Using NO_x ULTRA System [87]
Used by permission of K.R. Dougherty of Fuel Tech, Inc.

Ammonia Injection

For aqueous, anhydrous, or urea-derived ammonia, the ammonia-air mixture is directed through a flow-balancing skid to the AIG, where it is injected under pressure. The flow-balancing skid consists of flow meters and manual valves to adjust the flow to each part of the AIG.

The AIG consists of a network of pipes or lances connected in parallel and perforated with several holes or nozzles. The lances are placed in a grid formation across the width and height of the ductwork. The lances and holes are sized to distribute the ammonia uniformly into the flue gas. The spray angle and velocity of the injection control the trajectory of the ammonia. Injectors are subject to high temperatures and flue gas impingement, which cause erosion, corrosion, and degradation of structural integrity. Therefore, injectors are generally constructed of stainless steel and designed to be replaceable. Multiple injection zones may be used to increase the distribution of ammonia.

The ammonia can be injected with a low- or high-energy system. A low-energy system uses little or no pressurized air while a high-energy system uses large amounts of compressed air or steam to inject and vigorously mix the solution with the flue gas. AIG systems in large boilers typically use high-energy systems. High-energy systems are more expensive to build and operate because they require a larger compressor and a more robust injection system, and they consume more electric power.

Uniform distribution and mixing with flue gas is critical to maintain desired low levels of ammonia slip. Cold gas flow modeling and numerical flow modeling are generally performed for the AIG and SCR system to ensure uniform mixing and dispersion before the gases enter the SCR reactor. If duct length is inadequate to ensure thorough mixing or results from the model study indicate poor gas mixing characteristics, devices such as turning vanes or static gas mixers may be added.

An essential part of an AIG system is the controller used to regulate ammonia injection. Boiler load, inlet NO_x, and inlet gas temperatures set the feed-forward signal to establish the base ammonia injection rate. A feedback signal measuring the SCR outlet NO_x concentration is used to trim the base ammonia injection rate.

Although not necessarily required to achieve high NO_x removal efficiencies and low ammonia slip, most SCR installations today employ some type of static mixer to achieve good NH_3/NO mixing [41]. Static mixers provide a more uniform flux of NO_x and more uniform temperature and mixing of NH_3 and NO [41]. For example, achieving 90 percent NO_x removal and a 2-ppm ammonia slip typically requires NH_3/NO uniformity less than 5 percent and perhaps as low as 3 percent, as measured on a root mean square basis, and static mixers enable the SCR to achieve these levels [41]. The costs for static mixers vary (e.g., a 500 MW unit may have an installed cost of $750,000 and an additional 1 in. w.g. of flue gas pressure drop) [41]. One specific type of static mixer is the Delta Wing^{TM} mixer, which consists of an obstruction in the duct, usually a stationary disk or triangular plate, oriented at a slant to the flow direction. The Delta Wing mixer creates large vortices downstream of the device, which promotes mixing to a more homogenous gas. The ammonia injection nozzles are located in the vortex zone immediately downstream of the mixer [88]. The Delta Wing mixer costs for installing a new SCR typically are less than $500,000 and include the capital costs for the mixer and the modeling necessary to determine the location for maximum mixing effects. For an existing SCR that was installed without a static mixer, the costs to install the Delta Wing mixer, including capital, modeling, and other retrofit costs, could be up to $1,000,000 [89].

Another approach that may help retain good NH_3/NO uniformity is to conduct an annual tuning of the AIG, which can improve the NH_3/NO mixing (reduces NH_3/NO “unmixedness” by 2 to 5 percent on a root mean square basis) [41]. Data have shown that the ability of the AIG to achieve good mixing can decline over time and that annual tuning can return the AIG to startup or near-startup mixing uniformity [41]. Annual AIG tuning can cost from $30,000 to $50,000 depending on the unit size [41]. Depending on the type of mixer used, annual tuning may not be necessary, because some static mixers combine gas mixing and reagent injection in one application, with no moving parts in the gas stream; this type of mixer avoids much tuning during startup and commissioning and for annual maintenance [90].

Use of static mixers and annual tuning can either increase the NO_x efficiency at the same ammonia slip level or extend the catalyst life at the same NO_x removal efficiency [41]. Extending the catalyst life can significantly reduce operating costs, even after accounting for the outsourced tuning costs [41].

Catalytic Reduction of NO_x

The catalytic reduction of NO_x in the SCR reactor occurs when the NO_x and ammonia in the flue gas contact the catalyst layers. The catalyst itself is the key component of the SCR system. The catalyst composition, type (honeycomb, corrugated, or plate), and physical properties affect performance, reliability, catalyst quantity required, and cost. However, because the SCR system supplier and catalyst supplier must guarantee catalyst life and performance, most catalyst characteristics are selected by the SCR system supplier.

2.2.5 SCR System Auxiliary Equipment

SCR Inlet and Outlet Ductwork

In retrofit installations, new ductwork is required to integrate the SCR system with the existing equipment. In high-dust SCR systems for utility and industrial boilers, the reactor is located between the economizer outlet and the air heater inlet. In low-dust SCR systems for utility and industrial boilers, the SCR reactor is located between the outlet duct of the particulate control device and the air heater inlet duct. In tail-end SCR systems for utility and industrial boilers, the ductwork tie-ins are downstream of the FGD system and also require the integration of the flue gas reheating equipment.

See Section 2, Generic Equipment and Devices and Chapter 1, Hoods, Ductwork, and Stacks, for more details.

SCR Bypass Duct

Low-load boiler operations can decrease the temperature at the SCR reactor inlet below the SCR operating range. In addition, startup and shutdown of the boiler causes drastic temperature fluctuations. For these operating conditions, an SCR bypass may (but not necessarily) be required to route the flue gas around the reactor chamber. The bypass prevents catalyst poisoning and fouling during periods when flue gas stream conditions do not meet design specifications for proper SCR operation. The bypass system also must include zero-leakage dampers to prevent flue gas leakage from poisoning and fouling the catalyst while the SCR is not operating. A bypass system may also be considered for seasonal operation of the SCR system, such as for boilers that would require NO_x control during the ozone season (typically May to September), but not at other times of the year.

An SCR bypass may be needed for cement plant applications. During periods of startup and shutdown, the operating temperatures and constituents in the kiln gases may affect SCR operating conditions and may cause catalyst plugging or damage.

It should be noted that operational routines can be applied during SCR startup and shutdown that could preclude the need for a bypass, however, particularly for SCRs that operate year round [62]. Also, a Haldor Topsoe paper indicates that a bypass is not recommended for reasons including: complicated flue gas duct work, increased risk of dust depositing in horizontal parts, dust precipitation around dampers, and erosion of louver-type dampers that may result [91]. In fact, many of the SCRs being built in the late 2000’s for compliance with the Clean Air Interstate Rule (CAIR) do not include bypasses [92]. Recently built coal fired power plants in

Germany are designed without any SCR bypasses, according to a report by a major engineering services firm [93].

Soot Blower or Sonic Horn

In coal-fired boilers, soot blowers are usually installed in the SCR reactor to remove particulates that may mask or block active catalyst surfaces and gas passages. Soot blowing helps maintain acceptable flue gas pressure drop in the SCR reactor by keeping the catalyst gas passages free of particulate. Soot blowers also keep the air heater gas passages open and thereby reduce system pressure drop. This is especially true for SCR retrofits where the air heater plate spacing is generally narrow, making it more susceptible to fouling or clogging by ammonia-sulfur salts.

Retractable rake-type soot blowers that use steam or air for blowing are used in SCR designs. The soot blowers are typically located above each catalyst layer. Soot blowing is usually performed on one catalyst layer or part of one catalyst layer at a time. Soot blowing of all the catalyst layers takes 30 minutes to 2 hours but is usually done infrequently. In European SCR installations, soot blowing is done approximately once or twice a week [94]. Traveling-rake steam soot blowers can have installed costs of \$120,000 to \$160,000 [41].

Use of sonic horns has risen as an alternative to soot blowers [41]. Sonic horns require less preventive maintenance than soot blowers, but they are susceptible to moisture and fly ash, which cause plugging of the horn [41]. Sonic horns also cannot damage catalyst through either high-pressure operation or steam leaks, as can occur with soot blowers [41]. The capital cost for sonic horns can be \$40,000 to \$100,000 for each catalyst layer, depending on the size of the unit [41]. Sonic horn operating costs have been reported from \$1/day to <\$4/day for each catalyst layer, compared with approximately \$41/day for conventional soot blowers [41]. Although sonic horns may have some advantages over soot blowers, the demand for high NO_x removal efficiency requires extremely clean catalyst and thus the best cleaning system (regardless of cost); therefore, higher costs for soot blowers are justified for many applications [41].

For high dust loadings, the cement industry also reports use of soot blowers using heated, dry compressed air and use of sonic horns [10]. Improvements to cement dust cleaning for high dust loading and sticky dust have been made by changing the geometry and operating parameters of the dust blowers [95].

Large Particle Ash (LPA) Equipment

A significant concern for utility and industrial boiler SCR operation that was not evident in early applications is the role of the accumulation of LPA, also referred to as “popcorn ash,” on catalyst surfaces of high-dust SCR applications. LPA is defined as particles that are 4 to 7 mm in characteristic dimension and large enough to lodge in the openings of grid- or plate-type catalysts [41]. It is estimated that up to half of SCR units on coal-fired utility boilers are affected by LPA [9]. LPA is not an issue for natural gas-fired applications.

The cause or mechanism by which LPA is formed is unknown. A survey of 32 utility boiler operators found 23 had experienced significant plugging of catalyst modules, but there was no clear correlation of design or operating characteristics (e.g., coal type, boiler or reactor design, SCR cleaning method, or catalyst geometry) with either significant LPA problems or the lack of problems. Without knowledge of how LPA is generated, many facilities mitigate its impacts by removing some of the LPA from the flue gas before it reaches the catalyst. The most common mitigation method is the use of screens and/or baffles between the economizer exit and the SCR reactor that provide a barrier to the LPA and divert it to an ash hopper. This approach was used by operators of 21 of the 23 surveyed boilers that experienced LPA problems, and the other 2 were considering adding such equipment. Some of the 9 boilers that were not equipped with screens also had flow or deflector baffles. The frequency of economizer ash hopper evacuation has been suggested as a key variable affecting catalyst blockage. Increasing this frequency would be a low-cost operating change, but the survey of utility boiler operators did not show a correlation between evacuation frequency and catalyst blockage levels [96].

The survey report also identified the following recommendations for effective use of screens: (1) orient the screen at an angle to the flue gas flow, or use pleats; (2) maintain at least 50 percent to 60 percent open flow area; (3) conduct CFD or physical flow modeling and design the duct and screen to keep flue gas velocity below 50 actual ft/s (or preferably <45 actual ft/s) and eliminate peaks in velocity; and (4) use active cleaning systems. The open flow area and velocity recommendations are intended to minimize erosion of the screen material by fly ash, which was found to be significant regardless of the screen material used at the surveyed facilities [96]. The low velocity may also encourage LPA to drop out of the flue gas [41]. LPA screens can be modular to allow replacement, coated or uncoated depending on velocity, and rigid or flexible [41].

Capital costs for a simple rigid screen can be $200,000 to $500,000 for an erosion-tolerant design for high flue-gas velocity and exotic construction material [41]. The installed cost of screens for two SCRs on twin boilers that collectively generate 745 MW was $600,000 in 2004 [97]. A redesign and replacement of this screen along with CFD flow modeling in 2009 cost $806,000 [98, 99]. Operating costs can be a 1-in. w.g. pressure drop and require an additional $150,000 every 2 years to replace eroded screen panels. The costs associated with LPA mitigation methods can be recovered by avoiding an outage, by not accelerating an outage, and by maintaining clean catalyst [41].

Another option for LPA mitigation involves the use of targeted in-furnace injection, which models injections strategies to reduce SO_3 formation, and also minimizes slag and fouling. This slag and fouling control also reduces LPA formation. A suspended slurry of magnesium hydroxide ($Mg(OH)_2$) is used to change the slag formation by traveling into a furnace, becoming superheated and subsequently forming very small particles of magnesium oxide (MgO). The performance of targeted injection, and associated reduction of LPA, has been established in a case study on a 600 MW opposed wall-fired unit, where successful control of LPA has been demonstrated for more than six years. After successfully implementing the targeted in-furnace injection, the facility was able to remove their pre-existing LPA screens [100].

Economizer Bypass Duct

Although the SCR reaction occurs within a temperature window of 600 to 750°F (320 to 400°C), the catalyst for a given application is designed for a somewhat narrower range: the economizer outlet temperature at normal boiler operating load. Maintaining the flue gas temperature within the required window is essential for optimizing the NO_x reduction reaction.

When the economizer outlet flue gas temperature decreases because the plant is operating at reduced loads, the temperature can be raised using an economizer bypass.

The economizer bypass duct generally has a modulating damper to regulate the amount of hot bypass gas flow to be mixed with the cooler economizer outlet flue gas. The lower the boiler load, the more this damper opens, thus admitting more hot gas. The economizer outlet duct also needs a modulating damper to provide enough backpressure to allow the required volume of gas to flow through the bypass. The main design considerations for an economizer bypass involve maintaining the optimum gas temperature and ensuring uniform mixing of the two gas streams prior to entering the SCR reactor.

Upgraded or New Induced Draft (ID) Fan

The new ductwork and the SCR reactor's catalyst layers decrease the flue gas pressure. To maintain the same flow rate through the duct work, additional energy is required. The existing ID fan may be unable to provide the required increase in static pressure. In such cases, an upgraded or new ID fan is installed. This is also true for cement kiln applications due to the high dust loading and high pressure drop across the SCR. The existing fan and motor foundation may also need modification. Replacement involves installation of a new fan or booster fan. In all cases, additional electric power for the ID fan is needed to overcome the additional pressure drop through the SCR system. Based on typical values for the pressure drop through the additional duct work and the catalyst layers, the additional electric power needed (i.e., the heat rate penalty) is equivalent to approximately 0.3 percent of the plant's electric output for SCR on a utility boiler. Refer to Equations 2.57 and 2.58 for estimates of the total additional electric power needs for the ID fan as well as other equipment in the SCR system.

2.2.6 Other Considerations

Fuel Source

Utility and industrial boilers use coal, distillate oil, residual oil, natural gas, and a variety of other fuels such as biomass (e.g., wood, bark). The fuel type and grade affect the SCR design, and therefore, the capital costs of the SCR system. Fuels with high heating value have higher gas flow rates, which in turn increase the required SCR reactor size and catalyst volume. Coal-fueled applications are more costly than oil- and natural gas–fired boilers, due to their higher flue gas flow rates [58].

The quantity of nitrogen, fly ash, and pollutants in the flue gas stream varies according to the type and grade of the fuel. This affects the volume of catalyst required, as well as the catalyst design, composition, and rate of deactivation. Coal flue gas contains a greater amount of fly ash, SO_2, SO_3, arsenic, and other trace pollutants than does flue gas from burning oil or natural gas. Natural gas is the cleanest fuel and contains the least nitrogen; therefore, burning natural gas results in the least NO_x and pollutants in the flue gas stream. The amount of nitrogen and pollutants in oil-based fuels sources varies with the grade and type of oil, either refined or residual.

Formation of SO_x

SO_3 forms during the combustion of fuels that contain sulfur, and additional SO_3 is formed over the SCR catalyst. SO_3 is an emission concern for low sulfur coal. In cement plants, the fuel sulfur is incorporated into the clinker, however, the raw material sulfur can be a source of SO_2 and SO_3 in the gas stream. SO_3 reacts with ammonia in the flue gas downstream of the reactor (ammonia slip) to form ammonium bisulfate (NH_4HSO_4) and ammonium sulfate [$(NH_4)_2SO_4$]. The amount formed depends on the sulfur content of the fuel, the amount of ammonia slip, the SCR temperature, and the catalyst formulation and activity. Ammonium bisulfate condenses as the flue gas stream temperature lowers. It then deposits on the SCR catalyst and downstream equipment such as the air heater, ducts, and fans. Ammonia slip limits are generally imposed as part of the SCR design requirements to avoid impacts on downstream equipment.

There are several methods for limiting the impact of ammonia-sulfur salt deposition. Soot blowers can be installed between catalyst layers to remove surface deposits by blowing air or steam across the catalyst. Increased acid washing of the air preheater and other equipment may be required to remove deposits. However, more frequent acid washing generates additional wastewater, which must be disposed of or treated by the plant. The sulfur content of the flue gas can be decreased by coal desulfurization processes or fuel switching. Lastly, the flue gas temperature may be raised to prevent condensation.

Elevated SO_3 concentrations raise the acid dew point of the flue gas. This phenomenon potentially leads to more corrosion on the air heater's cold-end surfaces if the flue gas temperature is below the acid dew point. For Ljungström© preheaters, the cold-end baskets of the air heater can be replaced with enamel-coated baskets to protect against this possibility.

Ammonia sulfates also deposit on the fly ash. Ammonia content in the fly ash greater than 5 ppm can result in off-gassing, which would impact the salability of the ash as a byproduct and the storage and disposal of the ash by landfill [56].

Oxidation of SO_2 in the flue gas stream to SO_3 over the catalyst bed creates even more SO_3 (a $PM_{2.5}$ [20]precursor compound) and is an emission concern for SCR control devices only if the sulfur content of coal is sufficiently high. This can be a concern with SCR applied to boilers that use coal containing greater than 2 percent sulfur content or higher [101]. In some areas of the country, limits for SO_3 of 5 ppm or less have been implemented to control SO_3 plumes emitted from SCRs (i.e., blue plumes) [41]. Mitigation of SO_3 formation from oxidation of SO_2 on the catalyst has evolved to be of equal importance to NO_x control in SCR design [41]. Catalysts designed for low SO_2 oxidation have been developed, but depending on the catalyst supplier, their use may increase the catalyst volume needed to meet the same NO_x removal and ammonia slip performance and may also alter the catalyst management strategy [41]. Another approach to mitigating SO_3 formation that has been analyzed is use of reagents such as sodium, magnesium, or calcium-based sorbents [41]. However, the use of reagents alone to lower SO_3 concentration can rival the costs for annual SCR ammonia or urea reagent support and can exceed catalyst replacement charges, although other sources cite significant plant operation and maintenance cost savings due to eliminating corrosion and fouling of ductwork, ESPs, and the

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[20] $PM_{2.5}$ refers to fine particulate matter having a width of 2.5 micrometers or less.

blade housings of ID fans; eliminating modification or coatings for the air heater; and reducing the heat rate and fuel costs [41, 102]. Use of low-SO_2-oxidation catalyst can be used in combination with reagent-based SO_3 mitigation to lower costs of reagent-only approaches to reducing SO_3 [41].

Formation of Arsenic Oxide

Arsenic oxides (As_2O_3) formed during combustion of fuel containing arsenic cause catalyst deactivation by occupying active pore sites. Formation of arsenic oxides is a concern primarily for SCR units on coal-fired boilers because coal is the only fuel that may contain arsenic. Gaseous arsenic oxide can physically block active sites by condensing in the pores, and As_2O_3 can react with oxygen on the vanadium compounds in the catalyst to form arsenic pentoxide (As_2O_5), which chemically bonds to the site [103]. Catalyst manufacturers have developed arsenic-resistant catalysts that include control on porosity to minimize pore condensation and addition of compounds such as molybdenum or other materials to mitigate deactivation of active sites [103]. Deactivation of arsenic-resistant catalysts occurs at a slower rate, but these catalysts are not arsenic proof [103]. Arsenic-resistant catalyst will maintain adequate activity for approximately 14,000 hours of operation versus approximately 5,500 hours for a nonresistant catalyst [103].

The addition of small amounts of calcium to the fuel has been shown to be an effective method of controlling arsenic poisoning [104]. Limestone ($CaCO_3$) can be injected into the flue gas to generate the solid calcium arsenate ($Ca_3(AsO_4)_2$), which does not deposit on the catalyst and can be removed from the flue gas with a precipitator. Burning coal with higher calcium oxide (CaO) concentration can have the same effect as injecting limestone [103]. It should also be noted that CaO can be a catalyst poison via formation of $CaSO_4$ in the catalyst pores, effectively blocking catalyst reactive sites [104]. For the cement industry, a significant portion of the dust in cement kiln gases is CaO [10].

Mercury Oxidation

SCR catalysts have been found to oxidize a significant portion of elemental mercury to oxidized mercury (e.g., elemental mercury oxidized to mercuric chloride ($HgCl_2$). The oxidized mercury is water soluble and easier to remove in downstream wet FGD units or SO_2 scrubbers. Studies have suggested that the oxidation of elemental mercury by SCR may be affected by catalyst space velocity, reaction temperature, ammonia concentration, catalyst age, and concentration of chlorine in the gas stream. The type of coal burned and its associated chlorine content is another factor [105].

Studies on simulated flue gas streams and slip streams from actual electric generating units for multiple catalyst types have shown that mercury oxidation is in the range of 80–90 percent for fresh catalyst and space velocities of approximately 1,000 hr^{-1}, but that the oxidation rate declines to 30 to 80 percent with increasing space velocity, in the range 4,000 hr^{-1}, with other factors such as temperature, ammonia concentration, and so on more prominently influencing the rate when the space velocity increases. These studies also showed that lower temperatures increase the mercury oxidation rate. The oxidation of mercury was higher at temperatures of approximately 700°F (370°C) and were relatively lower at approximately 800°F

(430°C), which is consistent with the fact that mercury oxidation to $HgCl_2$ is greater at lower temperatures [105].

In earlier simulated flue gas and slip stream studies, it was noted that the ammonia reagent may suppress or interfere with the oxidation of elemental mercury, especially as the catalyst ages, although ammonia showed little or no effect on oxidation rates in studies on full-scale utility boiler SCR systems [105]. In recent studies, where the NH_3 to NO_x ratio was varied between 0 and 0.95, the suppressive nature of ammonia on mercury oxidation has been demonstrated [106]. In the reaction process on the catalyst sites, ammonia adsorption takes precedence over the site chlorination, therefore the NO_x reduction reaction has precedence over the mercury oxidation [51]. Because the favored ammonia adsorption on the catalyst sites minimizes the chlorinated sites, the NO_x reduction reaction may actually inhibit mercury oxidation [51]. In several studies, it has been observed that with NO_x reduction efficiencies up to 90 percent, the mercury oxidation readily occurs [51]. However, at NO_x reduction efficiencies beyond 90 percent, mercury oxidation is greatly reduced [51].

A higher chlorine concentration enhances mercury oxidation; bench-scale testing suggests that hydrochloric acid (HCl) is an important exhaust gas constituent that provides the chlorine for oxidation of mercury to $HgCl_2$ across the SCR catalyst [105]. Coal types, such as subbituminous coal, that tend to have lower chlorine levels will have lower HCl concentrations in the exhaust gas than bituminous coal, which has higher chlorine levels [105]. Significant mercury oxidation by SCR catalysts occurs with bituminous coal, and the oxidation rate is less certain with other types of coals [105]. In recent studies, HCl gas was injected directly into the flue gas and was varied from 0 to 150 ppmv Cl; the data demonstrated increased mercury oxidation across the SCR as chlorine was increased [106]. Chlorine appears to have the greatest effect on mercury oxidation in the range of 0 to 50 ppm [51]. The chlorine content does not appear to affect the NO_x reduction performance of the SCR [51].

There is a close correlation between mercury oxidation on the catalyst and the SO_2 to SO_3 conversion mechanism on the catalyst [51]. The mercury reaction with HCl and oxygen is considered to be a diffusion-controlled reaction that takes place in the gas phase [51]. The mercury oxidation rate is faster than the diffusion velocity of mercury through the SCR catalyst due to its molecular weight, while the reaction rate of SO_2 to SO_3 conversion is slower than the diffusion velocity of SO_2 through the catalyst [51]. As mentioned above in Section 2.2.1 (see Catalyst heading), by altering the catalyst to favor mercury oxidation based on this reaction mechanism, the active sites oxidize mercury instead of converting SO_2 to SO_3 [51]. Higher mercury oxidation rates of greater than 90 percent can be achieved while maintaining low SO_2 to SO_3 conversion rates in the range of less than 0.5 percent [51].

The age of the catalyst has also been shown to be a factor in mercury oxidation. One study compared the mercury oxidation rate for new, aged, and regenerated catalyst [106]. It was determined that there is little to no difference in the mercury oxidation capability of the new and regenerated catalysts [106]. While the operating conditions during testing of the aged catalyst were different from the new and regenerated catalysts (i.e., slightly higher temperature, which would decrease mercury oxidation; slightly higher chlorine content, which would increase mercury oxidation; and lower space velocity), the data showed lower mercury oxidation for the aged catalyst [106].

Dioxin/Furan Control

Dioxins and furans are emitted from combustion processes like waste incineration, burning fuels like wood, coal, or oil, and even from forest fires. Other processes, like chlorine bleaching of pulp and paper, for example, can result in small quantities of dioxins and furans [107]. The SCR systems for control of dioxins and furans were first used in the late 1990s in Europe and Japan [108].

Research has shown that commonly used metal oxide catalysts used in SCR for reducing NO_x emissions (e.g., vanadium and tungsten oxides on a titanium or platinum oxide-based substrate) from waste incineration systems can also reduce dioxin and furan emissions. This means SCR allows for an integrated approach to NO_x and dioxin/furan control, since only one pollution control device is needed to reduce the emission levels of both pollutants. The catalytic oxidation has been shown to occur in a temperature range of about 240 to 330°C. Dioxin/furan destruction efficiencies of 95 to 98 percent have been demonstrated [107].

Retrofit Versus New Design and Balance of Plant

Retrofit of SCR on an existing unit has higher capital costs than SCR installed on a new system. There is a wide range of SCR retrofit costs due to site-specific factors, scope differences, and site congestion [14]. Specific factors that impact the retrofit costs include the following: [14]

- The amount of available space between and around the economizer and air heater;
- Congestion downstream of the air heater (i.e., buildings, conveyors, existing ESPs, FGD system, ID fan, or stack);
- The age/vintage and manufacturer of the boiler;
- The design margin of the existing ID fan (i.e., the need to upgrade or replace fan impellers, replace ID fans, or add booster fans);
- The capacity, condition, and design margins of the electrical distribution system;
- The design margins of the existing structural steel support systems;
- The positive and negative design pressure of the furnace and existing ESP; and
- The number, nature, and type of existing items that must be relocated to accommodate the SCR and associated systems.

While not all of these components are specifically applicable to cement applications, each of these issues have an equal counterpart for cement kiln retrofits.

The primary balance of plant impacts include new ID fans and fan foundations or less expensive new booster fans, a new air preheater plus foundations or modifications to the existing air preheater, duct reinforcements, economizer duct bypass or modifications, elevated SCR, ESP reinforcement, and relocating the existing ESP flue gas conditioning system [14]. Retrofit costs for cyclone or wet bottom wall-fired boilers are somewhat higher than retrofit costs for dry bottom wall- or tangentially-fired boilers [46]. Differential retrofit costs for an SCR in Germany is approximately $200 per MMBtu/hr ($20/kW) [46]. However, a large part of the capital costs is not affected by a retrofit, including those for ammonia storage, vaporization, and injection equipment. The capital costs estimated by the equations in section 2.4 Cost Analysis are for retrofit of an existing SCR. The cost examples in section 2.5 represent an average retrofit difficulty.

Combustion Unit Design and Configuration

Boiler size is one of the primary factors that determines the SCR system capital costs. In addition, boiler configuration influences SCR costs. Boiler configurations that split the flue gas flow for two or more air preheaters or particulate removal systems require more than one SCR reactor. Additional reactors substantially increase capital costs. Boiler operations that have varying operating load, frequent startups/shutdowns, or seasonal operations require an SCR bypass. Additional ductwork, dampers, and control systems increase the SCR system capital costs. The SCR system may require modifications to draft fans and/or installation of additional fans. This increases both capital and operating costs of the SCR system. In addition, boiler and duct modifications may be required for implosion protection to accommodate increased draft requirements [58].

In a cement application, the kiln type, production rate, dust loading, the presence of catalyst poisons, and the magnitude and variability of raw material sulfur, SO_2, and SO_3 are all important factors impacting capital and operating costs for SCR systems.

Modeling of the SCR System

Computational fluid dynamics and chemical kinetic modeling are performed as part of the design process for SCR. In addition to computational fluid dynamics and chemical kinetic modeling, three-dimensional physical flow modeling, also referred to as cold flow modeling, is generally required. Cold flow modeling ensures that the flow through the SCR reactor provides adequate residence time, achieves uniform mixing of flue gas and ammonia, minimizes linear velocities to prevent catalyst erosion, and minimizes pressure drop across the catalyst layers. It involves constructing a model of the ammonia injection system, mixing area, and reactor chamber. In a high-dust configuration, this involves modeling from the economizer outlet to the inlet of the air heater. Typical model scales range from a 1:10 ratio to a 1:12 ratio for large electric utility boilers.

2.3 Design Parameters

SCR system design is a proprietary technology. Extensive details of the theory and correlations that can be used to estimate design parameters such as the required catalyst volume are not published in the technical literature [46]. Furthermore, the design is highly site-specific. In light of these complexities, SCR system design is generally undertaken by providing all of the plant- and unit-specific data to the SCR system supplier, who specifies the required catalyst volume and other design parameters based on prior experience and computational fluid dynamics and chemical kinetic modeling [37]. For some industrial applications, such as cement kilns where flue gas composition varies with the raw materials used, a slip stream pilot study can be conducted to determine whether trace elements and dust characteristics of the flue gas are compatible with the selected catalyst.

This section presents an approach to estimating design parameters that are elements in the costing equations used in EPA’s Integrated Planning Model (IPM) [9]. This section also presents an approach to estimating other design parameters that characterize an SCR system but that are not used directly in the costing procedure [37]. Although this approach is based on SCR data for utility boilers, it provides sufficient accuracy and detail to be of aid in developing capital and annual costs estimates for SCR as applied to industrial boilers and potentially other industrial sources.

2.3.1 Boiler Heat Input

The primary cost estimation parameter in the methodology presented in Reference [37] is the maximum potential heat released by the boiler or heat input rate, Q_B, expressed as MMBtu/hr. It is obtained from the higher heating value, HHV, of the fuel in Btu per pound (Btu/lb) and the maximum fuel consumption rate in pounds per hour (lb/hr), $\dot{m}_{fuel}$:

$$Q_B = HHV \times \dot{m}_{fuel} \times \frac{1}{10^6} \tag{2.3}$$

Where:

Q_B, = maximum heat rate input to the boiler, MMBtu/hr
HHV = higher heating value of the fuel, Btu/lb
$\dot{m}_{fuel}$ = maximum fuel consumption rate of the boiler, lb/hr
$1/10^6$ = conversion factor of 1 MMBtu/10^6 Btu.

Table 2.5 provides the HHV for various coals that may be used if the HHV for the specific coal used in the project is not available.[21]

Table 2.5: Higher Heating Values for Various Coals

Type of Coal	Energy Content (Btu/lb)
Lignite	5,000–7,500
Subbituminous	8,000–10,000
Bituminous	11,000–15,000
Anthracite	14,000

If the boiler produces electricity, its maximum heat input can be estimated using the boiler net plant heat rate (*NPHR*), in MMBtu per Megawatt-hour (MMBtu/MWh):

$$Q_B = B_{MW} \times NPHR \tag{2.4}$$

Where:

B_{MW} = boiler MW rating at full load capacity
NPHR = net plant heat rate, MMBtu/MWh.

Note that if *NPHR* is not known (e.g., a cogeneration unit), a conversion value for coal of 10,000 Btu/kWh (or 10 MMBtu/MWh) can be used as a reasonable estimate; a conversion value for

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[21] Another source of EGU fuels data is http://www.eia.gov/electricity/data/eia923/

petroleum of 11,000 Btu/kWh (11 MMBtu/MWh) and for natural gas of 8,200 Btu/kWh (8.2 MMBtu/MWh) can be used [109].[22] Using this value, the heat input rate, Q_B, for a coal-fired unit is:

$$Q_B = B_{MW} \times 10 \tag{2.5}$$

Where:

10 = estimated NPHR for coal, MMBtu/MWh.

2.3.2 Heat Rate Factor

The heat rate factor (HRF) is the ratio of actual heat rate of the boiler, in terms of the boiler NPHR in MMBtu/MWh, compared to a typical heat rate of 10 MMBtu/MWh. The developers of the costing methodology presented in section 2.4.1 determined that using this ratio in the equation for capital costs helped account for observed differences in actual costs for different coal-fired boilers. To maintain consistency with that approach, the same ratio (i.e., with 10 in the denominator) also has been used in the equations for oil and gas fired boilers in section 2.4.1. The NPHR is simply the amount of fuel energy that a boiler consumes to generate 1 MWh of electricity and is determined based on measurements of the electricity generation and fuel consumption over the same period of time. As noted above, if the NPHR is not known for a particular boiler, use 10 MMBtu/MWh.

$$HRF = \frac{NPHR}{10} \tag{2.6}$$

Where:

HRF = Heat rate factor
NPHR = net plant heat rate of the system to be costed, MMBtu/MWh
10 = the NPHR that is the basis of the SCR base cost module capital cost, MMBtu/MWh.

2.3.3 System Capacity Factor

The total system capacity factor, CF_{total}, is a measure of the average annual use of the boiler in conjunction with the SCR system. CF_{total} is given by:

$$CF_{total} = CF_{plant}\, CF_{SCR} \tag{2.7}$$

Where:

CF_{total} = total system capacity factor
CF_{plant} = boiler capacity, which is the ratio of the actual quantity of fuel burned annually to the potential maximum quantity of fuel burned annually

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[22] In recent years (2003 to 2011), the average NPHR for coal has increased slightly (likely due to aging of equipment), and the average NPHR for natural gas has decreased slightly (likely due to the increased use of natural gas fuel and the installation of new equipment).

CF_{SCR} = SCR system capacity factor, which is the ratio of the actual days of SCR operation annually to the total number of plant operating days per year.

For utility boilers, the capacity factor of the boiler, CF_{plant}, is the ratio of actual electric output annually to the potential maximum electric annual output, as shown in Equation 2.8a:

$$CF_{plant} = \frac{B_{output}}{(B_{MW} \times 8760)} \tag{2.8a}$$

Where:

B_{MW} = boiler MW rating at full load capacity, MWh
B_{output} = annual actual MW output, MW/year.

Alternatively, for industrial and utility boilers, the capacity factor of the boiler, CF_{plant}, is the ratio of actual quantity of fuel burned annually to the potential maximum quantity of fuel burned annually in pounds, i.e., $\dot{m}_{fuel}$ in lb/hr × 8,760 hr/yr. CF_{plant} is given by:

$$CF_{plant} = \frac{actual\ annual \dot{m}_{fuel}}{maximum\ annual\ \dot{m}_{fuel}} \tag{2.8}$$

Where:

$actual\ \dot{m}_{fuel}$ = annual actual fuel consumption rate of the boiler, lb
$maximum\ \dot{m}_{fuel}$ = annual maximum fuel consumption rate of the boiler, lb.

SCR systems can be operated year-round or only during the specified ozone season (commonly, May 1 to September 30). The capacity factor for the SCR system, CF_{SCR}, is the ratio of the actual number of SCR operating days, t_{SCR}, to the total number of plant operating days per year:

$$CF_{SCR} = \frac{t_{SCR}}{t_{plant}} \tag{2.9}$$

Where:

t_{SCR}, = actual days of SCR operation annually, days
t_{plant} = actual days of plant (or boiler) operation in a year, days.

2.3.4 Inlet NO_x and Stack NO_x

Inlet NO_x, represented as NO_{xin}, is the NO_x emission level in the flue gas exit stream from a boiler prior to the SCR system. Note that NO_{xin} also accounts for combustion controls if the boiler is equipped with such controls. The inlet NO_x emissions level, obtained from analyzing the boiler flue gas stream, is generally given in lb/MMBtu of NO_2 [37].

The stack NO_x, represented as NO_{xout} is the required NO_x emission limit at the stack outlet. It is generally set by the plant or regulatory limits and also given in lb/MMBtu of NO_2 [37].

2.3.5 NO_x Removal Efficiency

The NO_x removal efficiency, represented as η_{NOx}, is determined from the inlet or uncontrolled NO_x level of the boiler at maximum heat input rate, CF_{plant} =1.0, and the required stack emission limit. The equation for the NO_x removal efficiency is given by:

$$\eta_{NO_x} = \frac{NO_{x\,in} - NO_{x\,out}}{NO_{x\,in}} \tag{2.10}$$

Where:

η_{NOx} = NO_x removal efficiency, fraction
NO_{xin} = inlet NO_x level from the boiler, i.e., inlet NO_x rate to the SCR, lb/MMBtu (at maximum heat input rate, $CF_{plant} = 1.0$)
NO_{xout} = outlet NO_x rate from the SCR, lb/MMBtu.

The required NO_x removal efficiency is one of the most influential parameters on the overall SCR system cost [58]. Typically, the annual average outlet NO_x should not be less than 0.04 lb/MMBtu, or at a level that results in a removal efficiency greater than 90 percent, unless a guarantee has been obtained from a vendor. Additionally, if a facility is subject to an outlet limit over a time period shorter than annually (e.g., a 30-day rolling average), then that value should be used in the calculation of the NO_x removal efficiency. If a facility is subject to both an annual limit and a short-term limit, then the annual limit should be used in the calculation of the removal efficiency. It is noted that 0.05 lb/MMBtu outlet NO_x based on a 30-day (boiler operating) average should be obtainable by a power plant boiler with an SCR system.

2.3.6 NO_x Removal Rates

The tons of NO_x removed annually (ton/yr) are:

$$NO_x\ Removed/yr = NO_{xin}\ \eta_{NOx}\ Q_B\ t_{op} / 2{,}000 \tag{2.11}$$

Where:

$NO_x\ Removed/yr$ = annual mass of NO_x removed by the SCR, tons/yr
Q_B = maximum heat input rate to the boiler, MMBtu/hr
t_{op} = operating time per year (CF_{total} x 8760), hr/yr
2000 = conversion factor for lb/ton.

The pounds of NO_x removed per hour (lb/hr) are:

$$NO_x\ Removed/hr = NO_{xin}\ \eta_{NOx}\ Q_B \tag{2.12}$$

Where:

NO_x *Removed/hr* = hourly mass of NO_x removed by the SCR, lb/hr
NO_{xin} = inlet NO_x of the boiler, lb/MMBtu (at maximum heat input rate, $CF_{plant} = 1.0$)
η_{NOx} = NO_x removal efficiency of the SCR, expressed as a fraction
Q_B = maximum heat input rate to the boiler, MMBtu/hr.

2.3.7 Stoichiometric Ratio Factor

The stoichiometric ratio factor (SRF) indicates the actual amount of reagent needed to achieve the targeted NO_x reduction. Typical SRF values are higher than theoretical values due to the complexity of the reactions involving the catalyst and limited mixing. Higher SRF values generally result in increased NO_x reduction. The SRF is an important parameter in SCR system design because it establishes the reagent use of the SCR system.

The SRF is defined as:

$$SRF = \frac{moles\ of\ \ reagent\, injected}{moles\ of\ \ NO_x\ \ Removed} \tag{2.13}$$

In a design developed by a system supplier, the SRF would be adjusted to account for temperature, residence time, degree of mixing, catalyst activity, and allowable ammonia slip for a specific boiler. No equation for estimating SRF is available for SCR. The value for *SRF* in a typical SCR system, using ammonia as reagent, is approximately 1.05 [37]. This value incorporates design margins for ammonia slip and the small amount of NO_2 in the boiler flue gas, which requires two moles of NH_3 per mole of NO_2 instead of one mole of NH_3 per mole of NO as shown in Equation 2.1a. For an SCR system using urea as the reagent, 0.525 is a typical value for SRF [9].

2.3.8 Flue Gas Flow Rate

The full-load flue gas flow rate, including the typical design margin of 5 to 15 percent, is used to size the SCR reactors and associated catalyst inventory. This flow rate should be obtained from test data or a combustion calculation.

If flow rate data are not available, an approximation of the flue gas flow rate to each of the SCR reactors, $q_{fluegas}$, can be calculated using Equation 2.14.

$$q_{fluegas} = \frac{q_{fuel}\ Q_B\ (460 + T)}{(460 + 700)\ n_{SCR}} \tag{2.14}$$

Where:

$q_{fluegas}$ = volumetric flue gas flow rate through the SCR, actual cubic feet per minute (acfm)
q_{fuel} = base case flue gas volumetric flow rate factor, ft^3/min-MMBtu/hr
T = operating gas temperature at the inlet to the SCR, °F
n_{SCR} = number of SCR reactor chambers

700 = temperature at which the base case flow rate factor was determined, °F
460 = conversion from degrees Fahrenheit to Rankine.

"Base case" flue gas flow rate factors per unit of heat input for three types of coals are listed in Table 2.6. These factors were calculated using procedures in Reference [110] for typical coals, typical boiler excess air levels (i.e., 20 percent), and typical SCR flue gas conditions (–10 inches w.g. and 700° F). Note that similar flow rates are obtained using the oxygen-based F-factors, wet basis in Table 19-1 of EPA Method 19 in 40 CFR Part 60, Appendix A-7.

Table 2.6: Estimated Flue Gas Volumetric Flow Rate Factors for Various Coals

Coal Type	Estimated value of q_{fuel} (ft^3/min-MMBtu/hr)
Eastern Bituminous	484
Powder River Basin	516
Lignite	547

Note that in general, the number of reactors, n_{SCR}, is site specific. One SCR reactor per boiler unit is typically required in small high-dust system designs. However, two SCR reactors may be needed to treat flue gas from a larger boiler or a boiler equipped with two air preheaters. The system designs developed for the base and sensitivity cases of this report use one reactor. Study-level costs of a two-reactor system are expected to be similar to the cost of a corresponding one-reactor system because the catalyst, ammonia, economizer bypass, and ID fan costs are essentially identical.

2.3.9 Space Velocity and Area Velocity

The space velocity, V_{space}, is defined as the inverse of the residence time and is given by:

$$V_{space} = \frac{1}{ResidenceTime} \qquad (2.15)$$

Where:

V_{space} = the volumetric flow rate divided by the catalyst bed volume, hr^{-1}
$ResidenceTime$ = the time necessary for a volume of flue gas equal to the catalyst bed volume to pass through the catalyst bed, hr.

Space velocity is calculated from the experimentally measured flue gas volumetric flow rate at the reactor inlet, represented as $q_{fluegas}$, and the reactor/catalyst volume, represented as $Vol_{catalyst}$, given by:

$$V_{space} = \frac{q_{flue\,gas}}{Vol_{catalyst}} \qquad (2.16)$$

Where:

$Vol_{catalyst}$ = volume of the reactor or catalyst layers, ft^3.

SCR system designers and vendors use the concept of area velocity, V_{area}, to account for the reaction being limited to active catalyst sites. The area velocity is calculated from the specific surface area of the catalyst per catalyst volume, $A_{specific}$, as follows:

$$V_{area} = \frac{V_{space}}{A_{specific}} \tag{2.17}$$

Where:

V_{area} = the space velocity divided by the catalyst pore surface area, $ft^3/ft^2 \cdot hr$
$A_{specific}$ = the specific surface area of the catalyst divided by the catalyst volume, ft^2/ft^3.

$A_{specific}$ is given in units of $length^2/length^3$, is sometimes referred to as the contact surface area of the catalyst, and must be provided by the catalyst manufacturer.

2.3.10 Theoretical NO_x Removal Efficiency

Equation 2.10 defines the NO_x removal efficiency. However, in SCR, NO_x removal efficiency changes with catalyst activation. The following theoretical equation allows for estimation of removal efficiency, η_{NOx}, based on the catalyst activity constant, $K_{catalyst}$, at a specified time, t [37]. The theoretical NO_x removal efficiency is:

$$\eta_{NOx} = SFR\ (1-e^{a}) \tag{2.18}$$

where

$$a = \left(\frac{-K_{catalyst}\ A_{specific}}{V_{space}} \right) \tag{2.19}$$

Where:

$K_{catalyst}$ = constant for catalyst activity, changes over time (t).

Both $K_{catalyst}$ and $A_{specific}$ are typically provided by the catalyst manufacturer.

According to this equation, the NO_x removal efficiency increases with increasing NH_3/NO_x ratio and decreasing space velocity (i.e., increasing catalyst volume for a given gas flow rate). In addition, the equation shows that as the activity of the catalyst decreases over time, the NO_x removal also decreases.

2.3.11 Catalyst Volume

The theoretical catalyst volume required for the SCR system is based on the factors discussed in Section 2.2, Process Description. Equations 2.16, 2.18, and 2.19 can be combined and rearranged to determine the theoretical catalyst volume [62]. Substituting the definition of space velocity (Equation 2.16) into the definition of a (Equation 2.19), and then substituting that into Equation 2.18 and solving for the volume of the catalyst gives:

$$Vol_{catalyst} = \frac{-\left(q_{fluegas} \times \ln\left[1-\left(\frac{\eta_{NOx}}{SRF}\right)\right]\right)}{K_{catalyst} \times A_{specific}} \tag{2.21}$$

An empirical equation was developed in Reference [37] as a function of several sensitivity variables. The sensitivity variables were determined from catalyst volume estimates obtained from catalyst suppliers for base and sensitivity cases. Adjustment factors for these variables were then developed using regression techniques.

The empirical equation for catalyst volume is given below:

$$Vol_{catalyst} = 2.81 \times Q_B \times \eta_{adj} \times Slip_{adj} \times NO_{x\,adj} \times S_{adj} \times \frac{T_{adj}}{N_{SCR}} \tag{2.22}$$

where N_{SCR} is the number of SCR reactors and the adjustment factors include:

- NO_x efficiency adjustment factor, η_{adj}:

$$\eta_{adj} = 0.2869 + (1.058 \times \eta_{NOx}) \tag{2.23}$$

- Ammonia slip adjustment factor, $Slip_{adj}$, for ammonia slips between 2 and 5 ppm:

$$Slip_{adj} = 1.2835 - (0.0567 \times Slip) \tag{2.24}$$

- NO_x adjustment factor for inlet NO_x, NOx_{adj}:

$$NO_{x\,adj} = 0.8524 + (0.3208 \times NO_{x\,in}) \tag{2.25}$$

- Sulfur in coal adjustment factor, S_{adj}:

$$S_{adj} = 0.9636 + (0.0455 \times S) \tag{2.26}$$

where S is the sulfur content of the fuel by dry weight fraction.

- The temperature adjustment factor, T_{adj}, for gas temperatures other than 700°F (370°C):

$$T_{adj} = 15.16 - (0.03937 \times T) + (2.74 \times 10^{-5} \times T^2) \tag{2.27}$$

where T is the temperature of the flue gas at the reactor inlet in degrees Fahrenheit (°F).

2.3.12 SCR Reactor Dimensions

The cross-sectional area of the SCR reactor is sized for the flow rate of the flue gas, in acfm, and the superficial velocity. A typical value for the superficial velocity is 16 feet per second (ft/sec) (960 feet/minute (ft/min)). Using this value for velocity, the equation for the catalyst cross-sectional area, $A_{catalyst}$, is given by:

$$A_{catalyst} = \frac{q_{fluegas}}{\left(\frac{16\,ft}{\sec}\right) \times \left(\frac{60\sec}{\min}\right)} \tag{2.28}$$

Where:

$A_{catalyst}$ = cross-sectional area of the catalyst, ft^2.

The SCR reactor cross-sectional area, A_{SCR}, is approximately 15 percent greater than the catalyst cross-sectional area to account to the module geometry and hardware:

$$A_{SCR} = 1.15 \text{ x } A_{catalyst} \tag{2.29}$$

Where:

A_{SCR} = cross-sectional area of the SCR reactor, ft^2.

The actual dimensions of the SCR depend on the module arrangement in the catalyst layer. The typical cross-sectional dimensions of a module are 3.3 feet wide by 6.6 feet long (1 m × 2 m). Therefore, the SCR plan dimensions are approximately multiples of these dimensions. Depending on the number of modules in width and in length, the SCR reactor may be square or rectangular. For the purposes of this report, the SCR reactor can be treated as a square. The screening costs are valid for rectangular SCR reactors as long as the aspect ratio (length divided by width) is not too large. Industry standard aspect ratios are between 1.0 and 1.5. For a square reactor, the length, l, and width, w, are estimated by:

$$l = w = \left(A_{SCR}\right)^{1/2} \tag{2.30}$$

Where:

w = width of the SCR reactor, ft
l = length of the SCR reactor, ft.

An initial value for the number of catalyst layers is estimated first. This estimate is then checked by calculating the catalyst height for each layer. The initial estimate for the number of catalyst layers can be determined from the total catalyst volume, the cross-sectional area of the catalyst, and the estimated height of the catalyst element. A nominal height for the catalyst, h'_{layer}, is 3.1 feet [37].[23] A first estimate for the number of catalyst layers, n_{layer}, is:

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[23] The specified nominal value is one value within a range of values for h'_{layer} that will give the same values for n_{layer} in Equation 2.31 and h_{layer} in Equation 2.32 for a particular system. The optimum range of values for h'_{layer} differs depending on the inlet NO_x rate. For example, for a relatively high inlet NO_x rate of 0.86 lb/MMBtu, the optimum range of h'_{layer} is 3.1 feet to 4.2 feet. The optimum range shifts to smaller values when the inlet NO_x rate is lower, but the range is expected to include 3.1 feet for all inlet NO_x rates greater than 0.1 lb/MMBtu. Higher values of h'_{layer} than those in the optimum range result in a layer height greater than 5.0 feet, which is outside the standard industry range. Lower values of h'_{layer} result in a lower, but still acceptable, value of h_{layer}. Low values of h'_{layer} also slightly decrease the amount and cost of catalyst but increase the electricity costs for the ID fan because the number of layers and pressure drop both increase. Thus, a nominal value of 3.1 feet is expected to provide optimum results for most SCR analyses.

$$n_{layer} = \frac{Vol_{catalyst}}{h'_{layer} \times A_{catalyst}} \tag{2.31}$$

Where:

n_{layer} = number of catalyst layers
h'_{layer} = nominal height of each catalyst layer, ft.

This value of n_{layer} is then rounded to the nearest integer. In addition, there must be at least two catalyst layers.

The height of each catalyst layer is calculated using the estimated number of layers. This must result in the height of a catalyst layer, h_{layer}, to be within the standard industry range of 2.5 to 5.0 feet. The height of a catalyst layer is calculated from the following equation:

$$h_{layer} = \left[\frac{Vol_{catalyst}}{n_{layer} \times A_{catalyst}} \right] + 1 \tag{2.32}$$

where 1 foot is added to account for space required above and below the catalyst material for module assembly.

The number of catalyst layers calculated above does not include any empty catalyst layers for the future installation of catalyst. An empty catalyst layer is recommended for use with a CMP. The total number of catalyst layers, n_{total}, includes all empty catalyst layers that will be installed:

$$n_{total} = n_{layer} + n_{empty} \tag{2.33}$$

Where:

n_{total} = total number of catalyst layers
n_{empty} = number of empty catalyst layers, included for future catalyst installation.

The height of the SCR reactor, h_{SCR}, including the initial and future catalyst layers, the flow-rectifying layer, space for soot blowers and catalyst loading, but excluding the inlet and outlet ductwork and hoppers, is determined from the equation:

$$h_{SCR} = n_{total}\,(c_1 + h_{layer}) + c_2 \tag{2.34}$$

Where:

h_{SCR} = height of the SCR reactor, ft
c_1 = constant based on common industry practice, i.e., 7 ft, ft
c_2 = constant based on common industry practice, i.e., 9 ft, ft.

where the constants are based on common industry practice of c_1 = 7 ft and c_2 = 9 ft.

2.3.13 Estimating Reagent Consumption and Tank Size

The rate of reagent consumption or mass flow rate of the reagent, $\dot{m}_{reagent}$, generally expressed as pounds per hour (lb/hr), can be calculated using the inlet NO_x in lb/MMBtu and heat input rate, Q_B, in MMBtu/hr.

$$\dot{m}_{reagent} = \frac{NO_{x_{in}} \times Q_B \times \eta_{NOx} \times SRF \times M_{reagent}}{M_{NO_x}} \tag{2.35}$$

Where:

$\dot{m}_{reagent}$ = mass flow rate, or consumption rate, of the reagent, lb/hr
$M_{reagent}$ = the molecular weight of the reagent (60.06 pounds per mole [lb/mole] for urea, 17.03 lb/mole for ammonia)
M_{NOx} = the molecular weight of NO_2 (46.01 lb/mole).
η_{NOx} = NO_x removal efficiency of the SCR, expressed as a fraction.

The molecular weight of NO_2 is used because the NO_x emissions, $NO_{x_{in}}$, are given in lb/MMBtu of NO_2.

For ammonia, the mass flow rate of the aqueous reagent solution, $\dot{m}_{sol}$, is given by:

$$\dot{m}_{sol} = \frac{\dot{m}_{reagent}}{C_{sol}} \tag{2.36}$$

Where:

$\dot{m}_{sol}$ = mass flow rate of the aqueous reagent solution, lb/hr
C_{sol} = the concentration of the aqueous reagent solution, by weight fraction.

The solution volume flow rate, q_{sol}, generally expressed as gallons per hour (gph), is:

$$q_{sol} = \frac{\dot{m}_{sol}}{\rho_{sol}} \times 7.4805 \tag{2.37}$$

Where:

q_{sol} = solution volume flow rate, gph
ρ_{sol} = the density of the aqueous reagent solution, lb/ft^3
7.4805 = conversion factor of 7.4805 gal/1 ft^3.

The ρ_{sol} is 56.0 lb/ft^3 for a 29 percent solution ammonia and 71.0 lb/ft^3 for a 50 percent urea solution at 60°F.

The total volume stored in the tank, or tanks, is based on the volume that the SCR system requires for operating a specified number of days. The volume stored onsite for the number of operating days, $t_{storage}$, is:

$$Vol_{tank} = q_{sol} \times t_{storage} \times 24 \tag{2.38}$$

Where:

Vol_{tank} = total volume of aqueous solution stored in the tank(s), gallons (gal)
$t_{storage}$ = number of operating days the SCR is required to operate between solution delivery, days
24 = conversion factor of 24 hr/1 day.

Note that the tank volume is typically based on full-load operation, so the capacity factor is not included in Equation 2.38. A common onsite storage requirement is for 14 days of SCR operation.

2.4 Cost Analysis

The cost-estimating methodology presented here provides a tool to estimate study-level costs. Actual selection of the most cost-effective option should be based on a detailed engineering study and cost quotations from the system suppliers. The costs presented here are expressed in 2016 dollars (2016$).[24]

The cost equations are based on the EPA Clean Air Markets Division (CAMD) IPM [9]. In the costing method for SCR from the IPM, the purchased equipment cost, the direct installation cost, and the indirect installation cost are estimated together. This methodology is different from the *EPA Air Pollution Control Cost Manual* methodology, which estimates equipment costs and installation costs separately. Due to the limited availability of equipment cost data and installation cost data, the IPM equations for SCR capital costs were not reformulated for this analysis.[25] One difference between the IPM methodology and the methodology presented here is that the IPM methodology includes owner's costs (for owner activities related to engineering, management, and procurement) and financing mechanisms (i.e., allowance for funds used during construction [AFUDC]). As stated in the cost methodology in this Manual (Section 1, Chapter 2), owner's costs and AFUDC costs are capital cost items that are not included in the EPA Control Cost Manual methodology, and thus are not included in the total capital investment (TCI) estimates in this section.

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[24] For cost escalation or de-escalation, one suggested index is the Chemical Engineering Plant Cost Index (CEPCI). More information on CEPCI values and the indexing procedure can be found at http://www.chemengonline.com/pci-home. Other cost indexes are also available. For more information on cost escalation or de-escalation, please refer to the cost methodology chapter in the Cost Manual (Section 1, Chapter 2).

[25] The EPA CAMD IPM methodology for estimating capital costs is based on an engineering and design firm's in-house databases of actual SCR projects. The documentation indicates that the current industry trend is to retrofit high-dust hot-side SCR, and cold-side tail-end SCRs encompass a small minority of units and were not considered in the evaluation. Thus, the SCR cost equations are likely most representative of high-dust SCR, and qualitative differences in equipment and costs are noted in the text for tail-end units.

Capital cost equations are provided for both coal-fired and oil- or gas-fired units. The capital cost equations are applicable to coal-fired utility boilers and to oil- or gas-fired utility boilers at facilities with generating capacity greater than or equal to (≥) 25 MW. Equations are also provided for coal-fired and oil- or gas-fired industrial boilers with a heat input capacity greater than approximately 250 MMBtu/hr. The capital costs estimated by the equation represent average retrofit costs; costs for new construction are typically 20 to 30 percent less than for average retrofits for units of the same size and design. The TCI equations include a retrofit factor (RF). A retrofit factor of 0.8 should be used for new construction and a retrofit factor of 1 should be used for average retrofits. The equations may overestimate costs for some simple retrofits of existing plants. For retrofits that are more complicated than average, a retrofit factor of greater than 1 can be used to estimate capital costs provided the reasons for using a higher retrofit factor are appropriate and fully documented.

The SCR system design shown in the discussion below is a high-dust configuration with one SCR reactor per combustion unit. It uses aqueous ammonia as the reagent with an allowed ammonia slip in the range of 2 to 5 ppm.[26] The catalyst is a ceramic honeycomb with an operating life of 3 years at full load operations [37]. The cost equations are sufficient for NO_{xout} emission levels as low as 0.07 lb/MMBtu for bituminous coal and 0.05 lb/MMBtu for both PRB and lignite coal [9]. In general, there are differences in capital cost, operating cost, and process risk for high-dust and tail-end SCR units [57]. High-dust units tend to have higher capital costs when there are space constraints while tail-end units tend to have higher operating costs [57]. Differences in the cost elements related to tail-end units are pointed out in the discussions for total capital investment and total annual costs in the sections that follow.

The cost information presented in this report is based on using ceramic honeycomb catalyst for the base case. In general, more catalyst volume is required for an SCR system using plate catalyst, although the unit cost of plate catalyst is lower than honeycomb. Thus, any difference in cost is expected to be within the accuracy of a study-level cost estimate.

The annual cost procedures in the IPM categorize annual cost elements as either “fixed O&M” costs or “variable O&M” costs. These elements are reclassified in this report as variable direct annual costs, semivariable direct annual costs, or indirect annual costs to be consistent with EPA Control Cost Methodology. The procedures in this report also include capital recovery, which is not included in the IPM procedures.

2.4.1 Total Capital Investment (TCI)

TCI includes direct and indirect costs associated with purchasing and installing SCR equipment. Costs include the equipment cost for the SCR system itself, the cost of auxiliary equipment, direct and indirect installation costs, additional costs due to installation such as asbestos removal, costs for buildings and site preparation, offsite facilities, land, and working capital. In general, SCR does not require buildings, site preparation, offsite facilities, land, and working capital. A more detailed discussion of capital costs can be found in Section 1, Chapter 2 of this Manual. The total project cost or TCI for the SCR is based on the approach used by EPA

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[26] While the EPA CAMD IPM cost method is based on use of urea-derived ammonia injection, the system design and example problems here are based on use of aqueous ammonia because the majority of SCR use ammonia as the reagent. This approach may slightly overestimate the capital costs for an ammonia-based system, as discussed in the “Reagent Production, Storage, and Vaporization” subsection of section 2.2.4.

CAMD in the Integrated Planning Model [9], and this approach includes both the direct capital costs and the indirect capital costs. The methods presented in sections 2.4.1.1 and 2.4.1.2 for utility boilers are identical to the methods in v6 of the IPM, except that two elements have been excluded, as noted above. The IPM does not include methods for estimating impacts to industrial boilers. Thus, the methods presented in sections 2.4.1.3 and 2.4.1.4 for industrial boilers are based on modified IPM equations; the equations were modified by replacing electricity production ratings with the corresponding typical boiler heat input capacities, as calculated using typical NPHRs, and assuming that SCR costs for industrial boilers and utility boilers that have the same heat input capacity would be the same. The capital cost equations included in the manual reflect a process contingency of 5 to 10 percent and a project contingency of 15 percent.

The SCR costs and the balance of plant costs are impacted by the unit's elevation with respect to sea level. These cost calculations have been developed for SCR systems located within 500 feet of sea level. For SCR systems located at higher elevations, the base SCR unit cost and balance of plant cost should be increased based on the ratio of the atmospheric pressure between sea level and the location of the system, i.e., atmospheric pressure at sea level divided by atmospheric pressure at the elevation of the unit [9]. The elevation factor is calculated as follows:

$$ELEVF = \frac{P_0}{P_{ELEV}} \tag{2.39a}$$

Where:

$ELEVF$ = elevation factor
P_0 = atmospheric pressure at sea level, 14.7 pounds per square inch absolute (psia)
P_{ELEV} = atmospheric pressure at elevation of the unit, psia (see Table 2.7 for atmospheric pressures for various elevations).

Table 2.7 presents atmospheric pressures for elevations up to 6000 feet above sea level.

Table 2.7. Atmospheric Pressure at Different Elevations.

Elevation above sea level, ft	Atmospheric pressure, psia
0	14.7
500	14.4
1000	14.2
1500	13.9
2000	13.7
2500	13.4
3000	13.2
3500	12.9
4000	12.7
4500	12.5
5000	12.2
6000	11.8

Alternatively, P_{ELEV} can be calculated using Equation 2.39b [111]:

$$P_{ELEV} = 2116 \times \left[\frac{59 - (0.00356 \times h) + 459.7}{518.6} \right]^{5.256} \times \frac{1}{144} \qquad (2.39b)$$

Where:

P_{ELEV} = atmospheric pressure at elevation of the unit, psia
h = altitude, feet.

2.4.1.1 Utility Boilers (Coal-fired)

Utility, coal-fired units. The capital cost equation for coal-fired units (applicable for ≥25 MW) is:

$$TCI = 1.3 \times \left(SCR_{Cost} + RPC + APHC + BPC \right) \qquad (2.40)$$

Where:

TCI = total capital investment for a SCR on a coal-fired boiler, \$
SCR_{Cost} = cost of the SCR, \$
RPC = reagent preparation cost, \$
$APHC$ = air pre-heater cost, \$
BPC = balance of plant costs, \$.

This TCI calculation includes a factor of 1.3 to estimate engineering and construction management costs, labor adjustment for installation (e.g., per diem and premium for work shifts of 10 hr), and contractor profit and fees. (For retrofits that are more complicated than average, the terms SCR, APHC, and BPC would be adjusted with a retrofit factor of greater than one.)

SCR costs, utility, coal-fired units ≥25 MW. The capital costs for the SCR base unit includes costs for the inlet ductwork, the reactor, and the bypass equipment [9]. The SCR costs are calculated as follows:

$$SCR_{Cost} = 310{,}000 \times \left(NRF \right)^{0.2} \times \left(B_{MW} \times HRF \times CoalF \right)^{0.92} \times ELEVF \times RF \qquad (2.41)$$

Where:

SCR_{Cost} = SCR unit costs, \$
$310{,}000$ = constant in the equation
NRF = NO_x removal factor ($\eta_{NOx}/80$)
B_{MW} = electric generating capacity of the unit supplied by the boiler, MW
HRF = heat rate factor
$CoalF$ = coal factor (CoalF=1 if bituminous; CoalF=1.05 if PRB; CoalF=1.07 if Lignite)
$ELEVF$ = elevation factor

RF = retrofit factor (RF = 0.8 for new construction; RF = 1 for retrofits of average difficulty).

The NO_x Removal Factor, NRF, is an adjustment factor that helps the equations more accurately reflect the actual costs in the database; it is expressed as the NO_x removal efficiency (η_{NOx}) divided by 80. The CoalF is 1 for bituminous coal, is 1.05 for PRB coal, and is 1.07 for lignite coal.

<u>Reagent Preparation costs, utility, coal-fired units ≥25 MW</u>. The costs for equipment to prepare reagents for injection into the SCR are based on the NO_x removal rate. As noted in previous sections, ammonia (either aqueous or anhydrous) is typically used as the reagent for SCR. The RPC equation is applicable for all types of reagent systems because it reflects the actual mix of types of reagent systems in the underlying database. As a result, it likely slightly overstates costs for anhydrous ammonia systems, and slightly understates costs for urea to ammonia systems. The reagent preparation costs are calculated as follows:

$$RPC = 564{,}000 \times \left(\left(NOx_{in} \times B_{MW} \times NPHR \right) \times \eta_{NOx} \right)^{0.25} \times RF \tag{2.42}$$

Where:

RPC = Reagent preparation cost, \$
564,000 = constant in the equation
NO_{xin} = inlet NO_x level from the boiler, i.e., inlet NO_x rate to the SCR, lb/MMBtu
$NPRH$ = net plant heat rate, MMBtu/MWh
η_{NOx} = NO_x removal efficiency, fraction
RF = retrofit factor (RF = 0.8 for new construction; RF = 1 for retrofits of average difficulty).

<u>Air Pre-Heater Modification costs, utility, coal-fired units ≥25 MW</u>. Air pre-heater modification costs are included only where SO_3 control is necessary. An air pre-heater modification is necessary for the control of SO_3 for boilers that burn bituminous coal where the SO_2 content in the outlet stream from the boiler is 3 lb/MMBtu or greater. Such modifications can include the use of steels resistant to corrosion, sootblowers and nozzles specifically designed to minimize SO_3 formation. If lower sulfur content coal types are used, then no air pre-heater modification is needed. The air pre-heater modification costs are calculated as follows:

$$APHC = 69{,}000 \times \left(B_{MW} \times HRF \times CoalF \right)^{0.78} \times AHF \times RF \tag{2.43}$$

Where:

$APHC$ = Air pre-heater cost, \$
69,000 = constant in the equation
AHF = air heater factor (AHF=1 if bituminous coal and SO_2 ≥3 lb/MMBtu; if not true, AHF=0)
RF = retrofit factor (RF = 0.8 for new construction; RF = 1 for retrofits of average difficulty).

The AHF is 1 for bituminous coal and where the SO_2 content of the coal is 3 lb/MMBtu or greater. If the boiler burns other coal types, then the AHF is 0 and the air pre-heater term drops out of the overall TCI equation for the SCR system.

Balance of plant costs, utility, coal-fired units ≥25 MW. The BPC include cost items such as ID and booster fans, piping, and auxiliary power modifications necessary for the SCR unit [9]. The BPC are calculated as follows:

$$BPC = 529{,}000 \times \left(B_{MW} \times HRF \times CoalF\right)^{0.42} \times ELEVF \times RF \tag{2.44}$$

Where:

BPC = Balance of plant cost, $

529,000 = constant in the equation

$ELEVF$ = elevation factor

RF = retrofit factor (RF = 0.8 for new construction; RF = 1 for retrofits of average difficulty).

2.4.1.2 Utility Boilers (Oil- and Gas-fired)

Utility, oil- and gas-fired units (≥25 MW to 500 MW). The capital cost equation for oil- and gas-fired units, applicable to ≥25 MW to 500 MW, is:

$$TCI = 86{,}380 \times \left(\frac{200}{B_{MW}}\right)^{0.35} \times B_{MW} \times ELEVF \times RF \tag{2.45}$$

Where:

TCI = total capital investment for an SCR unit on an oil-fired or gas-fired boiler, $

86,380 = installed cost of an SCR system in 2016$ for a 200 MW oil- or gas-fired boiler, $/MW

B_{MW} = electric generating capacity of the unit supplied by the boiler, MW

$ELEVF$ = elevation factor

RF = retrofit factor (RF = 0.8 for new construction; RF = 1 for retrofits of average difficulty).

The ELEVF was applied to the TCI equation for oil- and gas-fired units, however, because the ELEVF would apply only to the base SCR cost and the BPC but not to the RP costs, or to the APHC if applicable, including the factor may somewhat overestimate the TCI. For utility oil- and gas-fired units, unlike the utility coal-fired units, the TCI equation is a single equation for all capital costs that does not allow discrimination in applying the factor.

Utility, oil- and gas-fired units (>500 MW). For oil- and gas-fired units >500 MW, the normalized costs in $/kW are assumed to be equivalent to the costs for a 500 MW boiler unit. Thus, the equation for units >500 MW reduces to:

$$TCI = 62{,}680 \times B_{MW} \times ELEVF \times RF \tag{2.46}$$

With respect to high-dust and tail-end SCR units, high-dust units typically require larger catalyst volume that increases capital costs. Tail-end units require less catalyst volume and therefore lower capital costs, due to minimal ash and catalyst poisons in the flue gas following the ESP and wet scrubber. A rule of thumb for SCR catalyst volume is high-dust units (on cyclone-fired boilers) require approximately 1.5 m^3/MW, and tail-end units require less than half the catalyst volume of a high dust unit (or less than 0.75 m^3/MW) [112]. The lower catalyst volume for tail-end units helps reduce initial catalyst capital cost (and catalyst replacement operating costs) [72]. The capital costs for tail-end SCR units must include the equipment cost for reheating the flue gas. Reheating may be conducted using steam coils or natural gas firing. Capital costs for these reheating options are similar, however steam supply piping, supports, and valves may increase the steam coil reheating capital costs [72]. In a case study for a tail-end SCR on a 600 MW burning bituminous coal, one source cites SCR capital costs of $205 million for an SCR with steam coil reheating and $205 million for an SCR with a natural gas burner (2008$) [72].

As noted earlier, applying the ELEVF to the TCI equation for utility oil- and gas-fired units may overestimate the costs.

2.4.1.3 Industrial Boilers (Coal-fired)

Industrial, coal-fired units. The capital cost equation for coal-fired units (applicable for ≥250 MMBtu/hr) uses the utility equations to estimate the industrial boiler SCR costs. Use of the utility equations may overestimate the costs for industrial boilers since current retrofits of utility boilers are likely more complex than for industrial boilers. The capital cost equation is:

$$TCI = 1.3 \times \left(SCR_{Cost} + RPC + APHC + BPC \right) \tag{2.47}$$

Where:

TCI = total capital investment for a SCR on a coal-fired boiler, $
SCR_{Cost} = cost of the SCR, $
RPC = reagent preparation cost, $
$APHC$ = air pre-heater cost, $
BPC = balance of plant costs, $.

This TCI calculation includes a factor of 1.3 to estimate engineering and construction management costs, labor adjustment for installation (e.g., per diem and premium for work shifts of 10 hr), and contractor profit and fees.

SCR costs, industrial, coal-fired units ≥250 MMBtu/hr. The capital costs for the SCR base unit includes costs for the inlet ductwork, the reactor, and the bypass equipment [9]. The SCR costs are calculated as follows:

$$SCR_{Cost} = 310{,}000 \times \left(NRF \right)^{0.2} \times \left(0.1 \times Q_B \times CoalF \right)^{0.92} \times ELEVF \times RF \tag{2.48}$$

Where:

SCR_{Cost} = SCR unit costs, $
310,000 = constant in the equation

NRF = NO$_x$ removal factor (η_{NOx}/80)

Q_B = maximum heat rate input to the boiler, MMBtu/hr

$CoalF$ = coal factor (CoalF=1 if bituminous; CoalF=1.05 if PRB; CoalF=1.07 if Lignite)

$ELEVF$ = elevation factor

RF = retrofit factor (RF = 0.8 for new construction; RF = 1 for retrofits of average difficulty).

The NO$_x$ Removal Factor, NRF, is an adjustment factor that helps the equations more accurately reflect the actual costs in the database; it is expressed as the NO$_x$ removal efficiency (η_{NOx}) divided by 80. The CoalF is 1 for bituminous coal, is 1.05 for Powder River Basin (PRB) coal and is 1.07 for lignite coal.

Reagent Preparation costs, industrial, coal-fired units ≥250 MMBtu/hr. The costs for equipment to prepare reagents for injection into the SCR are based on the NO$_x$ removal rate. As noted in previous sections, ammonia (either aqueous or anhydrous) is typically used as the reagent for SCR. The RPC equation is applicable for all types of reagent systems because it reflects the actual mix of types of reagent systems in the underlying database. As a result, it likely slightly overstates costs for anhydrous ammonia systems, and slightly understates costs for urea to ammonia systems. The reagent preparation costs are calculated as follows:

$$RPC = 564{,}000 \times \left(\left(NOx_{in} \times Q_B\right) \times n_{NOx}\right)^{0.25} \times RF \qquad (2.49)$$

Where:

RPC = Reagent preparation cost, \$

564,000 = constant in the equation

NO_{xin} = inlet NO$_x$ level from the boiler, i.e., inlet NO$_x$ rate to the SCR, lb/MMBtu

η_{NOx} = NO$_x$ removal efficiency, fraction

RF = retrofit factor (RF = 0.8 for new construction; RF = 1 for retrofits of average difficulty).

Air Pre-Heater Modification costs, industrial, coal-fired units ≥250 MMBtu/hr. Air pre-heater modification costs are included only where SO$_3$ control is necessary. An air pre-heater modification is necessary for the control of SO$_3$ for boilers that burn bituminous coal where the SO$_2$ content in the outlet stream from the boiler 3 lb/MMBtu or greater. If other coal types are used, then no air pre-heater modification is needed. The air pre-heater modification costs are calculated as follows:

$$APHC = 69{,}000 \times \left(0.1 \times Q_B \times CoalF\right)^{0.78} \times AHF \times RF \qquad (2.50)$$

Where:

$APHC$ = Air pre-heater cost, \$

69,000 = constant in the equation

$CoalF$ = coal factor (CoalF=1 if bituminous; CoalF=1.05 if PRB; CoalF=1.07 if Lignite)

AHF = air heater factor (AHF=1 if bituminous coal and SO$_2$ ≥3 lb/MMBtu; if not true, AHF=0)

RF = retrofit factor (RF = 0.8 for new construction; RF = 1 for retrofits of average difficulty).

The AHF is 1 for bituminous coal and where the SO_2 content of the coal is 3 lb/MMBtu or greater. If the boiler burns other coal types, then the AHF is 0 and this term drops out of the overall TCI equation for the SCR system.

Balance of plant costs, industrial, coal-fired units ≥250 MMBtu/hr. The BPC include cost items such as ID and booster fans, piping, and auxiliary power modifications necessary for the SCR unit [9]. The BPC are calculated as follows:

$$BPC = 529{,}000 \times \left(0.1 \times Q_B \times CoalF\right)^{0.42} \times ELEVF \times RF \qquad (2.51)$$

Where:

BPC = Balance of plant cost, $
529,000 = constant in the equation
ELEVF = elevation factor
RF = retrofit factor (RF= 0.8 for new construction; RF= 1 for retrofits of average difficulty).

2.4.1.4 Industrial Boilers (Oil- and Gas-fired)

Industrial, oil-fired units (≥275 to ≤5,500 MMBtu/hr). The capital cost equation for oil-fired industrial boilers was developed by modifying Equation 2.45. The 200 MW rating of the generating unit for the base utility boiler in Equation 2.45 was converted to a heat input capacity of 2,200 MMBtu/hr for an oil-fired boiler by using Equation 2.4 with a *NPHR* of 11,000 Btu/kwh. Similarly, dividing the base cost of $80/kW in Equation 2.45 by the *NPHR* of 11,000 Btu/kwh gives an estimated cost of 7,270 $/MMBtu/hr. The resulting equation, applicable to oil-fired industrial boilers with a heat input capacity of ≥275 to ≤5,500 MMBtu/hr, is:

$$TCI = 7{,}850 \times \left(\frac{2{,}200}{Q_B}\right)^{0.35} \times Q_B \times ELEVF \times RF \qquad (2.52)$$

Where:

TCI = total capital investment for SCR on an oil-fired boiler, $
7,850 = estimated installed cost of an SCR unit in 2016$ for an oil-fired boiler that has a 2,200 MMBtu/hr rating at full load capacity, $/MMBtu/hr
Q_B = maximum heat rate input to the boiler, MMBtu/hr
ELEVF = elevation factor
RF = retrofit factor (RF = 0.8 for new construction; RF = 1 for retrofits of average difficulty).

As noted earlier, applying the ELEVF to the TCI equation for industrial oil-fired units may overestimate the costs.

Industrial, gas-fired units (≥205 to ≤4,100 MMBtu/hr). The capital cost equation for gas-fired industrial boilers was developed by modifying Equation 2.44 in a manner similar to that described above for oil-fired boilers, except that the *NPHR* used in the conversions was 8,200 Btu/kwh. The resulting equation, applicable to gas-fired industrial boilers with a heat input capacity of ≥205 to ≤4,100 MMBtu/hr, is:

$$TCI = 10{,}530 \times \left(\frac{1{,}640}{Q_B} \right)^{0.35} \times Q_B \times ELEVF \times RF \tag{2.53}$$

Where:

TCI	=	total capital investment for SCR on a gas-fired boiler, $
10,530	=	estimated installed cost of an SCR unit in 2016$ for a gas-fired boiler that has a 1,640 MMBtu/hr rating at full load capacity, $/MMBtu/hr
Q_B	=	maximum heat rate input to the boiler, MMBtu/hr
ELEVF	=	elevation factor
RF	=	retrofit factor (RF = 0.8 for new construction; RF = 1 for retrofits of average difficulty).

As noted earlier, applying the ELEVF to the TCI equation for industrial gas-fired units may overestimate the costs.

Industrial, oil-fired units (>5,500 MMBtu/hr). For oil-fired industrial boilers >5,500 MMBtu/hr, the normalized costs in $/MMBtu/hr are assumed to be equivalent to the costs for a 5,500 MMBtu/hr boiler unit. Thus, the equation for units >5,500 MMBtu/hr reduces to:

$$TCI = 5{,}700 \times Q_B \times ELEVF \times RF \tag{2.54}$$

As noted earlier, applying the ELEVF to the TCI equation for industrial oil-fired units may overestimate the costs.

Industrial, gas-fired units (>4,100 MMBtu/hr). For gas-fired industrial boilers >4,100 MMBtu/hr, the normalized costs in $/MMBtu/hr are assumed to be equivalent to the costs for a 4,100 MMBtu/hr boiler unit. Thus, the equation for units >4,100 MMBtu/hr reduces to:

$$TCI = 7{,}640 \times Q_B \times ELEVF \times RF \tag{2.55}$$

As noted earlier, applying the ELEVF to the TCI equation for industrial gas-fired units may overestimate the costs.

2.4.2 Total Annual Costs

Total annual costs (TAC) consist of direct costs, indirect costs, and recovery credits. Direct annual costs are those proportional to the quantity of waste gas processed by the control system. Indirect (fixed) annual costs are independent of the operation of the control system and would be incurred even if it were shut down. No byproduct recovery credits are included because there are no salvageable byproducts generated from the SCR [113]. Each of these costs is discussed in the sections below. A more detailed discussion of annual costs can be found in Section 1, Chapter 2 of this Cost Manual.

Design parameters are estimated using the maximum annual heat input rate of the boiler to ensure adequate sizing of the SCR system. Annual costs are calculated using the annual heat input rate of the boiler and SCR system using CF_{total}.. This ensures that annual costs are based on the actual operating conditions rather than the design case. No escalation of annual costs is included in this procedure in order to be consistent with the cost methodology followed in this Cost Manual as described in Section 1, Chapter 2.

Direct Annual Costs

Direct annual costs, *DAC*, include variable and semivariable costs. Variable direct annual costs account for purchase of reagent and electrical power. Semivariable direct annual costs include operating and supervisory labor cost, maintenance cost, and catalyst replacement cost. These costs are discussed individually below. Equations for these variable cost items were derived in Reference [9].

Operating costs also result from small decreases in boiler efficiency due to operation of the economizer bypass. The economizer bypass operation depends on the flow rate of gas bypassed at full and partial loads and the boiler's capacity factor. Another operating cost is incurred for the steam or electric power used for compressed air, as required for the relatively infrequent operation of soot blowers. These operating costs are generally small and site-specific. Therefore, they are not discussed in this report.

$$DAC = \begin{pmatrix} Annual \\ Maintenance \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Reagent \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Electricity \\ Cost \end{pmatrix} + \begin{pmatrix} Annual \\ Catalyst \\ Cost \end{pmatrix} \tag{2.56}$$

Operating and Supervisory Labor

In general, operation of an SCR system requires only minimal, operating or supervisory labor. The SCR reactor is a stationary device with no moving parts. Further, the SCR system incorporates only a few pieces of rotating equipment (e.g., pumps, motors). The IPM [9] estimates operating labor time as 4 hours per day.

Maintenance

The annual maintenance labor and material cost in dollars per year ($/yr), including nozzle tip replacement for the injectors, is assumed to be 0.5 percent of the *TCI* in dollars [8, 9][27]:

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[27] Reference [9] applies the 0.5 percent factor for units smaller than 300 MW and applies 0.3 percent for larger units, and the factor is applied to the "Base Module" cost rather than the TCI (i.e., the equipment and installation cost before adding 30 percent for engineering and construction management, labor adjustment, and contractor profit and fees). Since the capital cost estimating procedure for oil- and gas-fired units does not include estimation of a Base Module cost, the procedures in this report use the more conventional approach of scaling the annual maintenance cost from the TCI.

$$\textit{Annual Maintenance Cost} = 0.005 \text{ x TCI} \quad (2.57)$$

Reagent Consumption

The annual cost of reagent purchase in $/yr is estimated using the reagent volume flow rate, q_{sol}, the operating time per year, t_{op}, and the cost of reagent in dollars per gallon, $Cost_{reag}$:

$$\textit{Annual Reagent Cost} = q_{sol} \text{ x } Cost_{reag} \text{ x } t_{op} \quad (2.58)$$

where q_{sol} is in gallons per hour (gph). The example here is for use of an aqueous ammonia solution, i.e., q_{sol}, however, if the cost of the ammonia is available in $/lb of ammonia, then the equation to estimate reagent costs would be based on $m_{reagent}$ from Equation 2.35 instead of q_{sol}.

The operating time per year, t_{op}, is estimated using the capacity factor, CF_{total}:

$$t_{op} = CF_{total} \times 8760 \quad (2.59)$$

Utilities

Power consumption for utility boilers. The electrical power consumption, P, in kW is estimated for SCR equipment, ammonia vaporization, water vaporization, and additional ID fan power [37]. The total additional auxiliary power required is estimated using Equation 2.60 [9]:

$$P = B_{MW} \times (1{,}000) \times (0.0056) \times (CoalF \times HRF)^{0.43} \quad (2.60)$$

Where:

P = electrical power consumption of the SCR system, kW
1,000 = conversion factor for 1,000 kW/MW
0.0056 = adjustment or scaling factor, dimensionless.

For oil- and gas-fired boilers, replace the coal factor with “1”.

Power consumption for industrial boilers. The electrical power consumption, P, in kW is estimated for SCR equipment, ammonia vaporization, water vaporization, and additional ID fan power [37]. It is estimated by converting Equation 2.60 to use the heat input to the boiler instead of the electric output as shown in Equation 2.61*:*

$$P = (0.1 \times Q_B) \times (1{,}000) \times (0.0056) \times (CoalF \times HRF)^{0.43} \quad (2.61)$$

The annual cost of electricity is estimated from the equation:

$$\textit{Annual Electricity Cost} = P \text{ x } Cost_{elect} \text{ x } t_{op} \quad (2.62)$$

Where:

$Cost_{elect}$ = cost of electricity, dollars per kWh ($/kWh).

In general, the power consumption for operating a high-dust SCR is lower than for low-dust SCR (data are not available for a tail-end unit). A relative comparison of power consumption for high-dust and low-dust SCR is shown in Table 2.8. As an example, one source cites a total power consumption of 3,500 kW for a high-dust SCR and 7,000 kW for a low-dust SCR on a 440 MW coal-fired boiler, showing that the power consumption for a low-dust unit is twice that of the high-dust unit [112].

Depending on the site configuration and space constraints, application of high-dust SCR to a boiler where there is significantly large retrofit costs may actually favor use of a tail-end unit with lower capital costs and more simple retrofit even though operating costs are higher [112].

Table 2.8: Comparison of Power Consumption for High-Dust and Low-Dust SCR [112]

Power component	High-dust SCR	Low-dust SCR
Induced draft fans, kW (accounts for largest portion of power needs, at 80 to 90 percent of total power)	Lower	Higher (approximately 2x higher)
Ammonia system power, kW	Higher (approximately 20 percent higher)	Lower
Dilution air blower, kW (second largest portion)	Higher (approximately 20 percent higher)	Lower
Dilution air heaters, kW	Higher (approximately 25 percent higher)	Lower
Ammonia pump, kW	Higher	Lower
Seal Air fans, kW	Same	Same
Electrical and control power consumption, kW	Lower	Higher (approximately 2x higher)
Total power consumption, kW	Lower	Higher

Catalyst Replacement

The catalyst life is a function of the catalyst activity and ammonia slip. As the catalyst activity decreases with time, the ammonia slip increases until it reaches the design limit and new catalyst must be added. Catalyst life is usually specified when purchasing the catalyst. For the most common SCR design, the high-dust SCR, a catalyst layer is typically guaranteed for 16,000 –24,000 operating hours based on information from catalyst vendors. For oil- and gas-fired units, the SCR catalyst life is assumed to be 40,000 hours, and the catalyst life for some gas-fired units has been reported to be up to 60,000 hours.

Two methodologies for estimating the annual catalyst replacement cost are presented in this chapter. One methodology is based on estimating the total volume of catalyst, the total number of catalyst layers, the number of layers replaced annually, and the future worth of the catalyst. This cost methodology assumes a guaranteed catalyst life of 24,000 hours or approximately 3 years for close to full time operation. The second methodology is an empirical equation that is part of the S&L cost methodology employed for power plants in the IPM [9].

Under catalyst replacement cost methodology 1, if the SCR does not have an empty catalyst layer, one approach is to replace all of the catalyst layers at the end of 24,000 operating hours. This very conservative assumption (i.e., likely overestimates the control cost) has been used in the SCR costs developed in the References [113] and [114]. If the SCR includes a spare catalyst layer, then only one catalyst layer is replaced at the end of 24,000 hours). Most SCR designs include a spare catalyst layer. The cost for catalyst replacement in all the SCR reactors for a given boiler is given by:

$$Catalyst\ Replacement\ Cost = n_{SCR}\ Vol_{catalyst}\ \frac{CC_{replace}}{R_{layer}} \tag{2.63}$$

Where:

$Catalyst\ Replacement\ Cost$ = cost to replace the SCR catalyst, either replacing all catalyst or replacing 1 layer at a time, $

$Vol_{catalyst}$ = volume of the reactor or catalyst layers, ft^3

n_{SCR} = number of SCR reactor chambers

$CC_{replace}$ = cost of catalyst, dollars per cubic foot ($\$/ft^3$)

R_{layer} = catalyst replacement factor (R_{layer} = 1 for full replacement and $R_{layer} = n_{layer}$ for replacing 1 layer at a time).

The catalyst cost, $CC_{replace}$, should reflect the current costs for the catalyst, including the costs associated with installing the new catalyst and removing and disposing of the old catalyst. Where a plant intends to use only regenerated catalyst, the catalyst cost used in Equation 2.63 should reflect the current cost of regenerated catalyst, which is typically lower than that for new catalyst. For situations where a plant may use regenerated catalyst and periodically purchase new catalyst, then a weighted average of current costs for new and regenerated catalyst should be used in Equation 2.63.

Because the catalyst is replaced every few years, the annual catalyst cost for all reactors is a function of the future worth of the catalyst, *FWF*, and is given by:

$$Annual\,Catalyst\,Replacement\ Cost = \begin{pmatrix} Catalyst \\ Replacement \\ Cost \end{pmatrix} \times FWF \tag{2.64}$$

where *FWF* is the future worth factor. Future worth is used because the annual catalyst replacement cost is accrued starting in the first year of operation, while catalyst replacement purchases occur every few years. To account for the time value of money, the *FWF* amortizes the catalyst cost over the years preceding the actual catalyst purchase [115]. Because the money is allocated in advance of the purchase, the sum of the annual catalyst replacement costs is less than the purchase price of the catalyst. The future worth factor, *FWF* is given by:

$$FWF = i\left[\frac{1}{(1+i)^{Y} - 1}\right] \tag{2.65}$$

Where:

i = interest rate, fraction

Y = term, years.

The term, Y, is given by the equation:

$$Y = \frac{h_{catalyst}}{h_{year}} \tag{2.66}$$

Where:

$h_{catalyst}$ = operating life of the catalyst, hours
h_{year} = number of hours per year the SCR is operated, hr/yr.

The value of Y estimated from the equation is then rounded to the nearest integer.

Under catalyst replacement cost methodology 2, the cost for catalyst replacement and disposal for a given boiler is part of the S&L cost methodology employed for power plants in this chapter given by [9]:

$$\begin{pmatrix} Annual\,Catalyst \\ Replacement\;Cost \end{pmatrix} = (B_{MW}) \times (0.4) \times (CoalF)^{2.9} \times (NRF)^{0.71} \times (CC_{replace}) \times 35.3 \tag{2.67}$$

Where:

Annual Catalyst Replacement Cost = cost to replace the SCR catalyst, \$/yr
CCreplace = cost of catalyst, dollars per cubic meter (\$/ft3)
35.3 = conversion factor for \$/ft^3 to \$/m^3.

Because high-dust units typically require larger catalyst volume, the replacement costs for the catalyst are also higher. Tail-end units require not only less catalyst volume but also less frequent catalyst replacement, due to minimal ash and catalyst poisons in the flue gas at this point in the equipment train. Lower levels of fly ash and catalyst poisons in the flue gas increase the catalyst life and decrease operating costs related to replacement [57]. In addition, concentrations of SO_2 in the flue gas are low following the wet scrubber and there are fewer concerns related to SO_3 formation and ammonium salt deposition [57].

While catalyst vendors typically provide a 24,000 hour (or 3 year) guarantee for catalysts, catalysts in tail-end units may last for extended periods. One source cites tail-end SCR units in Europe that continue to operate using the initial catalyst that was installed in the 1980’s and have up to 130,000 operating hours [116], and another source reports tail-end catalysts that lasted for 100,000 operating hours [57].

Indirect Annual Costs

In general, as mentioned in the Cost Manual Methodology chapter in Section 1 of the Control Cost Manual, indirect annual costs (fixed costs) include the capital recovery cost, property taxes, insurance, administrative charges, and overhead. Capital recovery cost is based on the anticipated equipment lifetime[28] and the annual interest rate employed.[29] For the purposes of this cost example, the equipment lifetime of an SCR system is assumed to be 30 years for power plants and 20 to 25 years for industrial boilers. These assumptions are based on several sources, including estimates by six petroleum refiners that SCR for fluidized catalytic cracking units and other process units would be between 20 and 30 years [26]; results from a survey conducted by the South Coast Air Quality Management District that shows equipment life for SCRs at refineries to be 20 to 25 years [117], an expert report in the North Carolina (NC) lawsuit against the Tennessee Valley Authority (TVA) coal-fired electric generation units indicated expected useful life of an SCR is 30 years [118]; a 2002 study of the economic risks from SCR operation at the Detroit Edison Monroe power plant used 30 years as the anticipated lifetime [119]; and a design lifetime of 40 years was used for an SCR at the San Juan Generating Station [120]. Thus, broadly speaking, a representative value of the equipment life for SCR at power plants can be considered as 30 years. For other sources, the equipment life can be between 20 and 30 years. The remaining life of the boiler may also be a determining factor for the system lifetime.

In many cases, property taxes do not apply to capital improvements such as air pollution control equipment; therefore, for this analysis, taxes are assumed to be zero [45]. The cost of overhead for an SCR system is also considered to be zero. An SCR system is not viewed as risk-increasing hardware (e.g., a high-energy device such as a boiler or a turbine). Consequently, insurance on an SCR system is on the order of a few cents per thousand dollars annually [45]. Finally, there are two categories of overhead, payroll and plant. Payroll overhead includes expenses related to labor employed in operation and maintenance of hardware, whereas plant overhead accounts for items such as plant protection, control laboratories, and parking areas. Because this procedure assumes that no additional labor is needed in operation of an SCR system, payroll overhead is zero and plant overhead is considered to be negligible.

Using these assumptions, indirect annual costs, *IDAC*, in $/yr, consist of both administrative charges and capital recovery, which can be expressed as:

$$Indirect\ Annual\ Cost = \begin{pmatrix} Administrative \\ Charges \end{pmatrix} + \begin{pmatrix} Capital \\ Recovery \end{pmatrix} \tag{2.68}$$

Administrative Charges

Administrative charges may be calculated as:

$$Administrative\ Charges = 0.03 \times \left(\left(\frac{Operator}{Labor\ Cost} \right) + 0.4 \times \left(\frac{Annual\ Maintenance}{Cost} \right) \right) \tag{2.69}$$

Where

---

[28] The term “equipment life” as used here in this chapter and through the Control Cost Manual refers to operational or design life. See Section 1, Chapter 2 for more explanation.

[29] The interest rate recommended by EPA can vary by firm or industry, but the bank prime rate is a default rate that can be used for annualization of capital costs. This rate is 5.25 – 5.5 percent as of January 2019. For more information, please consult the cost estimation chapter of this Control Cost Manual (Section 1, Chapter 2).

Operator Labor Cost = t_{SCR} x Operator Hours/day x Labor Rate.

In general, the operating labor cost in this equation will be small because operation of an SCR system requires only minimal, operating or supervisory labor.

Capital Recovery

Capital recovery is estimated as:

$$CR = CRF \times TCI \tag{2.70}$$

where *TCI* is the total investment, and *CRF* is the capital recovery factor and defined by:

$$CRF = \frac{i(1+i)^n}{(1+i)^n - 1} \tag{2.71}$$

where *i* is the interest rate, and *n* is the equipment life of the SCR system.

Total Annual Cost

The total annual cost (*TAC)* for owning and operating an SCR system is the sum of direct and indirect annual costs as given in the following equation:

$$Total\ Annual\ Cost = \begin{pmatrix} Direct \\ Annual \\ Cost \end{pmatrix} + \begin{pmatrix} Indirect \\ Annual \\ Cost \end{pmatrix} \tag{2.72}$$

Cost Effectiveness

The cost in dollars per ton of NO_x removed per year is:

$$Cost\ Effectiveness = \frac{TAC}{NO_x\ Removed/yr} \tag{2.73}$$

Where:

Cost Effectiveness = the cost effectiveness, \$/ton
NO_x *Removed/yr* = annual mass of NO_x removed in the SCR, ton/yr.

2.5 Example Problem #1 – Utility Boiler

An example problem that calculates both the design parameters and capital and annual costs for an SCR system applied to a 120 MW utility boiler firing bituminous coal is presented below. The following assumptions are made to perform the calculations:

Fuel High Heating Value, *HHV*	12,000 Btu/lb
Net Plant Heat Rate, *NPHR*	10 MMBtu/MWh
Maximum Actual Output	102 MW

Total Annual Fuel Consumption, *actual* m_{fuel}	7.45×10^8 lb/yr
Number of plant (boiler) operating days, t_{Plant}	365 days
Number of SCR operating days, t_{SCR}	365 days
Inlet NO Level, NO_{xin}	0.35 lb/MMBtu
Required Annual Average Controlled NO_x Emission Level, NO_{xout}	0.05 lb/MMBtu
Acceptable Ammonia Slip, *Slip*	2.0 ppm
Base Case Flue Gas Flow Rate Factor, Eastern Bituminous, q_{fuel}	484 ft^3/min per MMBtu/hr
Fuel Sulfur Content, *S*	1.0 percent by weight
Stoichiometric Ratio Factor for Ammonia, *SRF*	1.05
Stored Ammonia Concentration, C_{sol}	29 percent
Number of Days of Storage for Ammonia, *t*	14 days
Pressure Drop for SCR Ductwork, ΔP_{duct}	3 inches w.g.
Pressure Drop for each Catalyst Layer, $\Delta P_{catalyst}$	1 inch w.g.
Temperature at SCR Inlet, *T*	650°F
Plant elevation, P_{ELEV}	1,500 ft

In addition to these assumptions, the estimated economic factors for the cost equations are:

Cost year	2016$
Equipment Life	30 years
Annual Interest Rate	5.5 percent
Catalyst Cost[30]	\$8,000/m^3 (\$227/ft^3) [9]
Electrical Power Cost[24]	\$0.0361/kWh [121]
29 percent Ammonia Solution Cost[24]	\$0.293/gallon [average for 2016][31]
Operating Life of Catalyst	24,000 hours
Number of hours of operator labor	4 hours/day [9]
Labor Rate	\$60/hour (including benefits) [9]
Retrofit Factor	1 (average level of difficulty)

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[30] The electricity, catalyst, and reagent unit costs used in this example are based on data for 2016. These values are provided here for demonstration purposes only. When estimating direct annual operating costs, the current price of these commodities reflecting the year in which the cost estimate is made should be used. Catalyst and reagent prices can be obtained from vendors. Industrial plants should use the electricity price from their latest utility bill, while electricity generators should use the busbar rate.

[31] U.S. Geological Survey, *Minerals Commodity Summaries*, January 2017. Available at https://minerals.usgs.gov/minerals/pubs/commodity/nitrogen/mcs-2017-nitro.pdf.

2.5.1 Design Parameter Example #1[32]

Boiler Calculations

The boiler annual heat input rate, Q_B, is calculated from the High Heating Value for bituminous coal (see Table 2.5 for typical values if the actual value is unknown) and the maximum fuel consumption rate, $\dot{m}_{fuel}$:

$$Q_B = \frac{12{,}000 \frac{Btu}{lb} \times 100{,}000 \frac{lb}{hr}}{10^6 \frac{Btu}{MMBtu}} = 1{,}200 \frac{MMBtu}{hr}$$

The plant capacity factor is calculated from the maximum and annual MW output using Equation 2.8a:

$$CF_{plant} = \frac{102 MW}{120 MW} = 0.85 = 85 percent$$

The SCR system capacity factor is calculated from the months of SCR operation, 12 months, using Equation 2.9:

$$CF_{SCR} = \frac{365\,days}{365\,days} = 1.0 = 100 percent$$

The total capacity factor including both plant and SCR capacity factors is given by:

$$CF_{total} = 0.85 \text{ x } 1.0 = 0.85 = 85 \text{ percent}$$

The flue gas flow rate using Equation 2.14 is:

$$q_{flue\,gas} = \frac{\frac{484\,ft^3}{min(\frac{MMBtu}{hr})} \times 1{,}200 \frac{MMBtu}{hr} \times (460 + 650^\circ F)}{(460 + 700^\circ F) \times 1} = 556{,}766 acfm$$

The NO_x removal efficiency, η_{NOx}, is calculated from the inlet NO_x level and the required controlled NO_x emission level using Equation 2.10:

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[32] Note: Results of all parameter calculations are shown rounded to an acceptable number of significant figures. However, the full, unrounded value is used in subsequent parameter and cost calculations that use the parameter as an input. Thus, the results shown for subsequent calculations often differ from what would be calculated using the shown rounded inputs. The use of extra significant figures in the subsequent calculations does not imply greater accuracy of the numbers.

$$\eta_{NOx} = \frac{0.35\frac{lb}{MMBtu} - 0.05\frac{lb}{MMBtu}}{0.35\frac{lb}{MMBtu}} = 0.857 = 85.7\,percent$$

SCR Reactor Calculations

The catalyst volume using Equation 2.22 and the equations for each adjustment factor is:

$Vol_{catalyst}$ = 2.81 x 1,200 MMBtu/hr

× [0.2869 + (1.058 x 0.857)] (η_{adjx})

× [0.8524 + (0.3208 x 0.35)] ($NO_{x_{adj}}$)

× [1.2835 – (0.0567 x 2.0)] ($Slip_{adj}$)

× [0.9636 + (0.0455 x 1.0)] ($Sulfur_{adj}$)

× [15.16 – (0.03937 x 650) + (0.0000274 x 650^2)] ($Temperature_{adj}$)

= 5,254 ft^3

The catalyst and SCR cross-sectional areas using Equations 2.28 and 2.29 are:

$$A_{catalyst} = \frac{555{,}766\,acfm}{\left(16\frac{ft}{s}\right)\left(\frac{60\,s}{\min}\right)} = 579\,ft^2$$

$$A_{SCR} = 1.15 \times 579\,ft^2 = 666\,ft^2$$

The length and width of the reactor using Equation 2.30 is:

$$l = w = (666)^{1/2} = 25.8\,ft$$

The first estimate of the number of catalyst layers using Equation 2.31 is:

$$n_{layer} = \frac{5{,}300}{3.1 \times 579} = 2.9$$

Rounding this value gives, n_{layer} = 3.

Checking the actual catalyst height using Equation 2.32:

$$h_{layer} = \frac{5{,}300\,ft^3}{3 \times 579\,ft^2} + 1 = 4.0$$

This value is within the design height limits of 2.5 to 5 feet.

The total number of catalyst layers is determined by Equation 2.33 with one empty catalyst layer:

$$n_{total} = 3 + 1 = 4$$

The SCR height, excluding the outlet duct and hoppers using Equation 2.34 is:

$$h_{SCR} = 4 \times (7 + 4.0) + 9 = 53\ ft$$

Reagent Calculations

The mass flow rate of the reagent is calculated using the molecular weight of the reagent, 17.03 g/mole and NO_2, 46.01g/mole. For an SRF of 1.05, the reagent mass flow rate is given by Equation 2.35:

$$\dot{m}_{reagent} = \frac{0.35\frac{lb_{NO_x}}{MMBtu} \times 1{,}200\frac{MMBtu}{hr} \times 0.857 \times 1.05 \times 17.03\frac{lb}{mole}}{46.01\frac{lb}{mole}} = 140\frac{lb}{hr}$$

The mass flow rate of 29 percent aqueous ammonia solution is given by Equation 2.36:

$$\dot{m}_{sol} = \frac{140\frac{lb}{hr}}{0.29} = 482\frac{lb}{hr}$$

The solution volume flow rate can then be calculated from Equation 2.37 where Δ_{sol} is the density of the 29 percent aqueous ammonia solution, 56.0 lb/ft^3 at 60^{o}F, and the conversion factor is 7.481 gal/ft^3:

$$q_{sol} = \frac{482\frac{lb}{hr} \times 7.481\frac{gal}{ft^3}}{56.0\frac{lb}{ft^3}} = 64gph$$

The total volume stored in the tank(s) is based on the volume that the SCR system requires for 14 days of operation. The onsite storage requirement is given by Equation 2.38:

$$Vol_{tank} = 64gph \times 14(days) \times \left(\frac{24hr}{day}\right) = 21{,}700gal$$

The onsite storage requirement for ammonia solution is approximately 22,700 gallons per 14 days of operation (rounded to the nearest 100 gallons).

Capital Cost Elevation Factor Calculation

The elevation factor for use in calculating the SCR base unit cost and the balance of plant costs is given by Equation 2.39 with the atmospheric pressure at 1500 ft above sea level (13.9 psia):

$$ELEVF = \frac{14.7\,psia}{13.9\,psia} = 1.06$$

2.5.2 Cost Estimation Example

Once the SCR system is sized, the capital and annual costs for the SCR system can be estimated. The TCI is estimated using Equation 2.40. The SCR$_{\text{cost}}$, RPC, APHC and BPC must be calculated individually using equations 2.41, 2.42, 2.43, and 2.44, respectively. These calculations are shown below.

$$SCR_{Cost} = 310{,}000 \times \left(\frac{85.7\%}{80}\right)^{0.2} \times \left(120MW \times \frac{10MMBtu/MWh}{10} \times 1.00\right)^{0.92} \times 1.06 \times 1 = \$27{,}136{,}058$$

$$RPC = 564{,}000 \times \left(\left(0.35\frac{lb}{MMBtu} \times 120MW \times 10\frac{MMBtu}{MWh}\right) \times 0.857\right)^{0.25} \times 1 = \$2{,}456{,}715$$

$$APHC = 69{,}000 \times (120MW \times 1 \times 1)^{0.78} \times 0 \times 1 = \$0$$

$$BPC = 529{,}000 \times (120MW \times 1 \times 1)^{0.42} \times 1.06 \times 1.4 = \$4{,}169{,}246$$

$$TCI = 1.3 \times (\$27{,}136{,}058 + \$2{,}456{,}715 + \$0 + \$4{,}169{,}246) = \$43{,}890{,}624$$

Annual costs are based on the economic factors listed above. In addition, the SCR system in this example is assumed to operate the entire year with a boiler loading of 85 percent, resulting in a total capacity factor of 85 percent. As discussed in section 2.5.1, the direct annual costs consist of the variable direct annual costs (reagent and electricity) and semivariable direct annual costs (maintenance and annual catalyst cost).

The annual maintenance costs are estimated using Equation 2.57:

$$\text{Annual Maintenance Cost} = 0.005 \times \$43{,}890{,}624 = \$219{,}453/yr$$

An estimate for power consumption is given by Equation 2.60:

$$P = (120\,MW) \times (1{,}000\,kW/MW) \times (0.56/100) \times (1 \times 1)^{0.43} = 672\,kW$$

Electricity cost can then be estimated from Equation 2.62:

$$Annual\ Electricity\ Cost = 672kW \times \frac{8{,}760hr}{yr} \times 0.85 \times \frac{\$0.0361}{kWh} = \frac{\$180{,}634}{yr}$$

Reagent cost is estimated using Equation 2.58:

$$Annual\ Reagent\,Cost = 64\frac{gal}{hr}\times 8{,}760\frac{hr}{yr}\times 0.85\times 0.293\frac{\$}{gal} = \frac{\$140{,}601}{yr}$$

A portion of the catalyst is replaced every few years as discussed earlier in this chapter. The annual catalyst replacement cost can be estimated using Equations 2.63 through 2.66 for catalyst replacement cost methodology 1 or using Equation 2.67 for catalyst replacement cost methodology 2:

Using catalyst replacement cost methodology 1, the actual price at the time of purchase is estimated using Equation 2.63. This example assumes that one layer is replaced at a time, and based on the calculation in section 2.5.1, there are 3 catalyst layers. Thus, R_{layer} = 3, and the total catalyst replacement cost is given by:

$$Catalyst\,Replacement\,Cost = \frac{5{,}300\,ft^3\times\frac{\$227}{ft^3}}{3} = \$401{,}033$$

To account for the time value of money, these periodic costs are amortized over the years preceding the actual catalyst purchase using the future worth factor as calculated using Equation 2.62. The term, Y, in Equation 2.65 is estimated using Equation 2.66. Assuming the boiler operates continuously all year means h_{year} is 8,760 hr/yr, and Y is given by:

$$Y = \frac{24{,}000\,hours}{8{,}760\,hours} = 2.7 \approx 3$$

And the future worth factor for the catalyst replacement is given by:

$$FWF = 0.055\times\frac{1}{(1+0.055)^3 - 1} = 0.3157$$

The annual catalyst replacement cost can then be estimated using Equation 2.64:

$$Annual\,Catalyst\,Replacement\,Cost = 0.3157\times \$401{,}033 = \frac{\$125{,}519}{yr}$$

Alternatively, using catalyst replacement cost methodology 2, the annual catalyst replacement cost can be estimated using Equation 2.67:

$$\begin{pmatrix} Annual\,Catalyst \\ Replacement\ Cost \end{pmatrix} = (120\,MW)\times(0.4)\times(1)^{2.9}\times\left(\frac{85.7}{80}\right)^{0.71}\times\left(\frac{\$227}{ft^3}\right)\times\left(\frac{35.3\,ft^3}{m^3}\right) = \frac{\$403{,}939}{yr}$$

The total direct annual cost is given by the sum of the variable direct annual costs and semi-variable direct annual costs. If using cost methodology 1 for the annual catalyst replacement cost, the total direct annual cost is:

$$Total\ Direct\ Annual\ Cost = \frac{\$219{,}453}{yr} + \frac{\$180{,}634}{yr} + \frac{\$140{,}601}{yr} + \frac{\$125{,}519}{yr} = \frac{\$666{,}207}{yr}$$

Alternatively, if using cost methodology 2 for the catalyst replacement cost, the total direct annual cost is:

$$\text{Total Direct Annual Cost} = \frac{\$219{,}453}{yr} + \frac{\$180{,}634}{yr} + \frac{\$140{,}601}{yr} + \frac{\$403{,}939}{yr} = \frac{\$944{,}627}{yr}$$

As discussed in section 2.4.2, property taxes and overhead are both assumed to be zero, and insurance costs are assumed to be negligible. Thus, administrative charges and capital recovery are the only components of indirect annual costs estimated in this analysis. Administrative charges are calculated using Equation 2.69 as:

$$\text{Administrative Charges} = 0.03 \times \left(\left(\frac{365 days}{year} \times \frac{4 hours}{day} \times \frac{\$60}{hr} \right) + 0.4 \times (\$219{,}453) \right) = \$5{,}261/\,yr$$

The capital recovery factor, *CRF*, is defined by Equation 2.71 as:

$$CRF = \frac{0.055(1 + 0.055)^{30}}{(1 + 0.055)^{30} - 1} = 0.0688$$

and the capital recovery is calculated from Equation 2.70:

$$Capital\ Recovery = 0.0688 \times \frac{\$43{,}890{,}624}{yr} = \frac{\$3{,}019{,}675}{yr}$$

The total indirect annual costs (IDAC) are calculated in Equation 2.68:

$$IDAC = \frac{\$5{,}261}{yr} + \frac{\$3{,}019{,}675}{yr} = \frac{\$3{,}024{,}936}{yr}$$

The total annual cost is the sum of the direct annual and indirect annual costs given by Equation 2.72. If using cost methodology 1 for the catalyst replacement cost, the total annual cost is:

$$Total\ Annual Cost = \frac{\$666{,}207}{yr} + \frac{\$3{,}024{,}936}{yr} = \frac{\$3{,}691{,}143}{yr}$$

Alternatively, if using cost methodology 2 for the catalyst replacement cost, the total annual cost is:

$$Total\ Annual Cost = \frac{\$944{,}627}{yr} + \frac{\$3{,}024{,}936}{yr} = \frac{\$3{,}969{,}564}{yr}$$

The annual cost in terms of NO_x removed can be calculated using the total annual cost and the tons of NO_x removed annually. The annual reduction in NO_x emissions is given by Equation 2.11:

$$NO_x\ Removed / yr = \frac{\frac{0.35 lb}{MMBtu / hr} \times 0.857 \times 1{,}200 \frac{MMBtu}{hr} \times 0.85 \times 8{,}760 \frac{hr}{yr}}{2{,}000 \frac{lb}{ton}} = 1{,}340\ tons / yr$$

and the cost effectiveness is estimated using Equation 2.73. If using cost methodology 1 for the catalyst replacement cost, the cost effectiveness is:

$$Cost\ of\ NO_x\ Removal = \frac{\frac{\$3{,}691{,}143}{yr}}{1{,}340 tons} = \frac{\$2{,}754}{ton}$$

Alternatively, if cost methodology 2 is used for the catalyst replacement cost, then the cost effectiveness is:

$$Cost\ of\ NO_x\ Removal = \frac{\frac{\$3{,}969{,}564}{yr}}{1{,}340 tons} = \frac{\$2{,}962}{ton}$$

2.6 Example Problem #2 – Industrial Boiler

An example problem that calculates both the design parameters and capital and annual costs for an SCR system applied to a 550 MMBtu/hr industrial boiler firing bituminous coal is presented below. The following assumptions are made to perform the calculations:

Fuel High Heating Value, HHV	12,000 Btu/lb
Maximum Fuel Consumption Rate, $\dot{m}_{fuel}$	4.58 x 10^4 lb/hr
Total Annual Fuel Consumption, *actual* m_{fuel}	3.30 x 10^8 lb/yr
Number of plant boiler operating days	333 days
Number of SCR operating days, t_{SCR}	333 days
Inlet NO Level, NO_{xin}	0.35 lb/MMBtu
Required Annual Average Controlled NO_x Emission Level, NO_{xout}	0.05 lb/MMBtu
Acceptable Ammonia Slip, *Slip*	2.0 ppm
Base Case Flue Gas Flow Rate Factor, Eastern Bituminous, q_{fuel}	484 ft^3/min per MMBtu/hr
Fuel Sulfur Content, S	1.0 percent by weight
Stoichiometric Ratio Factor for Ammonia, SRF	1.05
Stored Ammonia Concentration, C_{sol}	29 percent
Number of Days of Storage for Ammonia, t	14 days
Pressure Drop for SCR Ductwork, ΔP_{duct}	3 inches w.g.

Pressure Drop for each Catalyst Layer, $\Delta P_{catalyst}$	1 inch w.g.
Temperature at SCR Inlet, T	650°F
Plant elevation, P_{ELEV}	<500 ft above sea level

In addition to these assumptions, the estimated economic factors for the cost equations are:

Cost year	2016$
Equipment Life	25 years
Annual Interest Rate	5.5 percent
Catalyst Cost[33]	$8,000/m^3 ($227/ft^3) [9]
Electrical Power Cost[26]	$0.0676 [122]
29 percent Ammonia Solution Cost[26]	$0.293/gallon [average for 2016][34]
Operating Life of Catalyst	24,000 hours
Number of hours of operator labor	4 hours/day
Labor Rate	$60/hour (including benefits)
Retrofit Factor	1.0

2.6.1 Design Parameter Example #2[35]

Boiler Calculations

The boiler annual heat input rate, Q_B, is calculated from the High Heating Value for bituminous coal (see Table 2.5 for typical values if the actual value is unknown) and the maximum fuel consumption rate, $\dot{m}_{fuel}$:

$$Q_B = \frac{12{,}000 \frac{Btu}{lb} \times 45{,}800 \frac{lb}{hr}}{10^6 \frac{Btu}{MMBtu}} = 550 \frac{MMBtu}{hr}$$

The plant capacity factor is calculated from the maximum and annual average fuel consumption:

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[33] The electricity, catalyst, and reagent unit costs used in this example are based on data for 2016. These values are provided here for demonstration purposes only. When estimating direct annual operating costs, the current price of these commodities reflecting the year in which the cost estimate is made should be used. Catalyst and reagent prices can be obtained from vendors. Industrial plants should use the electricity price from their latest utility bill, while electricity generators should use the busbar rate.

[34] U.S. Geological Survey, *Minerals Commodity Summaries*, January 2017. Available at https://minerals.usgs.gov/minerals/pubs/commodity/nitrogen/mcs-2017-nitro.pdf

[35] Note: Results of all parameter calculations are shown rounded to an acceptable number of significant figures. However, the full, unrounded value is used in subsequent parameter and cost calculations that use the parameter as an input. Thus, the results shown for subsequent calculations often differ from what would be calculated using the shown rounded inputs. The use of extra significant figures in the subsequent calculations does not imply greater accuracy of the numbers.

$$CF_{plant} = \frac{3.30 \times 10^{8} \, \frac{lb}{yr}}{4.58 \times 10^{4} \left(\frac{\text{lb}}{\text{hr}} \right) \times 8{,}760 \frac{hr}{yr}} = 0.82 = 82 percent$$

The SCR system capacity factor is calculated from the fraction of boiler operating time during which the SCR also operates:

$$CF_{SCR} = \frac{333\, days}{333\, days} = 1.0 = 100 percent$$

The total capacity factor including both plant and SCR capacity factors is given by:

CF_{total} = 0.82 x 1.0 = 0.82 = 82 percent

The flue gas flow rate using Equation 2.14 is:

$$q_{flue gas} = \frac{\frac{484\, ft^{3}}{min(\frac{MMBtu}{hr})} \times 550 \frac{MMBtu}{hr} \times (460 + 650^{\circ} F)}{(460 + 700^{\circ} F) \times 1} = 255{,}000 acfm$$

The *NO*NO_x removal efficiency, η_{NOx}, is calculated from the inlet NO_x level and the required controlled NO_x emission level using Equation 2.10:

$$\eta_{NOx} = \frac{0.35 \frac{lb}{MMBtu} - 0.05 \frac{lb}{MMBtu}}{0.35 \frac{lb}{MMBtu}} = 0.857 = 85.7 percent$$

SCR Reactor Calculations

The catalyst volume using Equation 2.22 and the equations for each adjustment factor is:

$Vol_{catalyst}$ = 2.81 x 550 MMBtu/hr

× [0.2869 + (1.058 x 0.857)] (η_{adjx})

× [0.8524 + (0.3208 x 0.35)] ($NO_{x_{adj}}$)

× [1.2835 – (0.0567 x 2.0)] ($Slip_{adj}$)

× [0.9636 + (0.0455 x 1.0)] ($Sulfur_{adj}$)

× [15.16 – (0.03937 x 650) + (0.0000274 x 650^2)] ($Temperature_{adj}$)

= 2,408 ft^3

The catalyst and SCR cross-sectional areas using Equations 2.28 and 2.29 are:

$$A_{catalyst} = \frac{255{,}000\ acfm}{\left(16\frac{ft}{s}\right)\left(\frac{60\,s}{\min}\right)} = 265\,ft^2$$

$$A_{SCR} = 1.15 \times 265\,ft^2 = 305\,ft^2$$

The length and width of the reactor using Equation 2.30 is:

$$l = w = \left(305\right)^{1/2} = 17.5\,ft$$

The first estimate of the number of catalyst layers using Equation 2.31 is:

$$n_{layer} = \frac{2{,}408}{3.1 \times 265} = 3.0$$

Rounding this value gives, $n_{layer} = 3$.

Checking the actual catalyst height using Equation 2.32:

$$h_{layer} = \frac{2{,}408\,ft^3}{3 \times 265\,ft^2} + 1 = 4.0$$

This value is within the design height limits of 2.5 to 5 feet.

The total number of catalyst layers is determined by Equation 2.33 with one empty catalyst layer:

$$n_{total} = 3 + 1 = 4$$

The SCR height, excluding the outlet duct and hoppers using Equation 2.34 is:

$$h_{SCR} = 4 \times (7 + 4.0) + 9 = 53\,ft$$

<u>Reagent Calculations</u>

The mass flow rate of the reagent is calculated using the molecular weight of the reagent, 17.03 g/mole and NO_2, 46.01g/mole. For an SRF of 1.05, the reagent mass flow rate is given by Equation 2.35:

$$\dot{m}_{reagent} = \frac{0.35\frac{lb_{NO_x}}{MMBtu} \times 550\frac{MMBtu}{hr} \times 0.857 \times 1.05 \times 17.03\frac{lb}{mole}}{46.01\frac{lb}{mole}} = 64\frac{lb}{hr}$$

The mass flow rate of 29 percent aqueous ammonia solution is given by Equation 2.36:

$$\dot{m}_{sol} = \frac{64\dfrac{lb}{hr}}{0.29} = 221\frac{lb}{hr}$$

The solution volume flow rate can then be calculated from Equation 2.37 where Δ_{sol} is the density of the 29 percent aqueous ammonia solution, 56.0 lb/ft^3 at 60^{o}F, and the conversion factor is 7.481 gal/ft^3:

$$q_{sol} = \frac{221\dfrac{lb}{hr} \times 7.481\dfrac{gal}{ft^3}}{56.0\dfrac{lb}{ft^3}} = 30gph$$

The total volume stored in the tank(s) is based on the volume that the SCR system requires for 14 days of operation. The onsite storage requirement is given by Equation 2.38:

$$Vol_{tank} = 30gph \times 14(days) \times \left(\frac{24hr}{day}\right) = 9{,}924gal$$

The onsite storage requirement for ammonia solution is approximately 10,000 gallons per 14 days of operation.

Capital Cost Elevation Factor Calculation

The elevation factor for use in calculating the SCR base unit cost and the balance of plant costs is given by Equation 2.39 with the atmospheric pressure at <500 ft above sea level (14.7 psia):

$$ELEVF = \frac{14.7\,psia}{14.7\,psia} = 1.0$$

2.6.2 Cost Estimation Example #2

Once the SCR system is sized, the capital and annual costs for the SCR system can be estimated. The TCI is estimated using Equation 2.47. The SCR$_{cost}$, RPC, APHC and BPC must be calculated individually using equations 2.48, 2.49, 2.50 and 2.51, respectively. These calculations are shown below.

$$SCR_{Cost} = 310{,}000 \times \left(\frac{85.7\%}{80}\right)^{0.2} \times \left(0.1 \times \frac{550MMBtu}{hr} \times 1.00\right)^{0.92} \times 1 \times 1 = \$12{,}545{,}501$$

$$RPC = 564{,}000 \times \left(\left(0.35\frac{lb}{MMBtu} \times 550\frac{MMBtu}{hr}\right) \times 0.857\right)^{0.25} \times 1 = \$2{,}021{,}390$$

$$APHC = 69{,}000 \times (0.1 \times 550MMBtu/hr \times 1)^{0.78} \times 0 \times 1 = \$0$$

$$BPC = 529{,}000 \times \left(0.1 \times 550 MBtu/hr \times 1\right)^{0.42} \times 1 \times 1 = \$2{,}847{,}135$$

$$TCI = 1.3 \times (\$12{,}545{,}501 + \$2{,}021{,}390 + \$0 + \$2{,}847{,}135) = \$22{,}638{,}233$$

Annual costs are based on the economic factors listed above. In addition, the SCR system in this example is assumed to operate all 333 days that the boiler operates, and the boiler loading is 82 percent, resulting in a total capacity factor of 82 percent. As discussed in section 2.6.1, the direct annual costs consist of the variable direct annual costs (reagent and electricity) and semivariable direct annual costs (maintenance and annual catalyst cost).

The annual maintenance costs are estimated using Equation 2.57:

$$Annual\,Maintenance\ Cost = 0.005 \times \$22{,}638{,}233 = \$113{,}191/\,yr$$

An estimate for power consumption is given by Equation 2.61:

$$P = \left(\frac{0.1MWh}{MMBtu}\right) \times \left(\frac{550MMBtu}{hr}\right) \times \left(\frac{1{,}000kW}{MW}\right) \times \left(\frac{0.56}{100}\right) \times (1 \times 1)^{0.43} = 308\,kW$$

Electricity cost can then be estimated from Equation 2.62:

$$Annual\,Electricity\ Cost = 308kW \times \frac{8{,}760hr}{yr} \times 0.82 \times \frac{\$0.0676}{kWh} = \frac{\$149{,}910}{yr}$$

Reagent cost is estimated using Equation 2.58:

$$\text{Annual Reagent Cost} = 30\frac{gal}{hr} \times 8{,}760\frac{hr}{yr} \times 0.82 \times 0.293\frac{\$}{lb} = \frac{\$62{,}313}{yr}$$

A portion of the catalyst is replaced every few years as discussed earlier in this chapter. The annual catalyst replacement cost can be estimated using Equations 2.63 through 2.66 for catalyst replacement cost methodology 1 or using Equation 2.67 for catalyst replacement cost methodology 2:

Using catalyst replacement cost methodology 1, the actual price at the time of purchase is estimated using Equation 2.63. This example assumes that one layer is replaced at a time, and based on the calculation in section 2.5.1, there are 3 catalyst layers. Thus, R_{layer} = 3, and the total catalyst replacement cost is given by:

$$Catalyst\,Replacement\ Cost = \frac{2{,}400\,ft^3 \times \dfrac{\$227}{ft^3}}{3} = \$182{,}228$$

To account for the time value of money, these periodic costs are amortized over the years preceding the actual catalyst purchase using the future worth factor as calculated using Equation

2.62. The term, Y, in Equation 2.65 is estimated using Equation 2.66. Assuming the boiler operates continuously all year means h_{year} is 8,760 hr/yr, and Y is given by:

$$Y = \frac{24{,}000\,hours}{8{,}760\,hours} = 2.7 \approx 3$$

And the future worth factor for the catalyst replacement is given by:

$$FWF = 0.055 \times \frac{1}{(1+0.055)^3 - 1} = 0.3157$$

The annual catalyst replacement cost can then be estimated using Equation 2.64:

$$Annual\ Catalyst\ Replacement\ Cost = 0.3157 \times \$182{,}228 = \$57{,}529/year$$

Alternatively, using catalyst replacement cost methodology 2, the annual catalyst replacement cost can be estimated using Equation 2.67:

$$\begin{pmatrix} Annual\ Catalyst \\ Replacement\ Cost \end{pmatrix} = 0.1 \times \left(\frac{550MMBtu}{hr}\right) \times (0.4) \times (1)^{2.9} \times \left(\frac{85.7}{80}\right)^{0.71} \times \left(\frac{\$227}{ft^3}\right) \times \left(\frac{35.3\,ft^3}{m^3}\right) = \frac{\$185{,}139}{yr}$$

The total direct annual cost is given by the sum of the variable direct annual costs and semivariable direct annual costs. If using catalyst replacement cost methodology 1, the total direct annual cost is:

$$Total\ Direct\ Annual\ Cost = \frac{\$113{,}191}{yr} + \frac{\$149{,}910}{yr} + \frac{\$62{,}313}{yr} + \frac{\$57{,}529}{yr} = \frac{\$382{,}943}{yr}$$

Alternatively, if using catalyst replacement cost methodology 2, the total direct annual cost is:

$$\text{Total Direct Annual Cost} = \frac{\$113{,}191}{yr} + \frac{\$149{,}910}{yr} + \frac{\$62{,}313}{yr} + \frac{\$185{,}139}{yr} = \frac{\$510{,}139}{yr}$$

As discussed in section 2.4.2, property taxes and overhead are both assumed to be zero, and insurance costs are assumed to be negligible. Thus, administrative charges and capital recovery are the only components of indirect annual costs estimated in this analysis. Administrative charges are calculated using Equation 2.69 as:

$$\text{Administrative Charges} = 0.03 \times \left(\left(\frac{333\,days}{year} \times \frac{4\,hours}{day} \times \frac{\$60}{hour}\right) + 0.4 \times (\$113{,}191)\right) = \$3{,}756/\,yr$$

The capital recovery factor, *CRF*, is defined by Equation 2.71 as:

$$CRF = \frac{0.055(1+0.055)^{25}}{(1+0.055)^{25} - 1} = 0.0745$$

and the capital recovery is calculated from Equation 2.70:

$$Capital\ Recovery = 0.0745\ \times \frac{\$22{,}638{,}233}{yr} = \frac{\$1{,}686{,}548}{yr}$$

The total indirect annual costs (IDAC) are calculated in Equation 2.68:

$$IDAC = \frac{\$3{,}756}{yr} + \frac{\$1{,}686{,}548}{yr} = \frac{\$1{,}690{,}304}{yr}$$

The total annual cost is the sum of the direct annual and indirect annual costs given by Equation 2.72. If using catalyst replacement cost methodology 1, the total annual cost is:

$$Total\ Annual\ Cost = \frac{\$382{,}943}{yr} + \frac{\$1{,}690{,}304}{yr} = \frac{\$2{,}073{,}248}{yr}$$

Alternatively, if using catalyst replacement cost methodology 2, the total annual cost is:

$$Total\ Annual\ Cost = \frac{\$510{,}553}{yr} + \frac{\$1{,}690{,}304}{yr} = \frac{\$2{,}200{,}857}{yr}$$

The annual cost in terms of NO_x removed can be calculated using the total annual cost and the tons of NO_x removed annually. The annual reduction in NO_x emissions is given by Equation 2.11:

$$NO_x\ Removed/yr = \frac{\frac{0.35 lb}{MMBtu/hr} \times 0.857 \times 550 \frac{MMBtu}{hr} \times 0.82 \times 8{,}760 \frac{hr}{yr}}{2{,}000 \frac{lb}{ton}} = 594\, tons/yr$$

and the cost effectiveness is estimated using Equation 2.73. If using catalyst replacement cost methodology 1, the cost effectiveness is:

$$Cost\ of\ NO_x\ Removal = \frac{\$2{,}073{,}265}{594\ tons} = \frac{\$3{,}490}{ton}$$

Alternatively, if using catalyst replacement cost methodology 2, the cost effectiveness is:

$$Cost\ of\ NO_x\ Removal = \frac{\$2{,}200{,}857}{594\ tons} = \frac{\$3{,}705}{ton}$$

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