2
100

✓ DOE/PC/91346--5

DE-AC22-92PC91346-5

M-1

MAR 02 1994

DEVELOPMENT OF THE INTEGRATED ENVIRONMENTAL CONTROL MODEL:

COST MODELS OF SELECTIVE CATALYTIC REDUCTION
(SCR) NO_x CONTROL SYSTEMS

October-December 1993

Quarterly Progress Report

IVE

MAR 02 1994

OSTI

to

Pittsburgh Energy Technology Center
U.S. Department of Energy
Pittsburgh, PA 15236

from

Center for Energy and Environmental Studies
Carnegie Mellon University
Pittsburgh, PA 15213-5903

Prepared by

H. C. Frey
E.S. Rubin

January 31, 1994

U.S. DEPARTMENT OF ENERGY

OFFICE OF ENERGY EFFICIENCY

ENERGY EFFICIENCY
PROGRAM

MASTER

870

DISTRIBUTION OF THIS DOCUMENT IS UNLIMITED

DISCLAIMER

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference herein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed herein do not necessarily state or reflect those of the United States Government or any agency thereof.

Table of Contents

1.0	Introduction to SCR Technology	1
	Figure 1. Alternative Placements of SCR Systems in Coal-Fired Power Plants.....	2
2.0	SCR Capital Cost Model.....	2
2.1	Reactor Housing.....	3
	Figure 2. Direct Capital Cost Model for SCR Reactor Housing.....	5
2.2	Ammonia Handling and Injection	5
2.3	Ductwork	6
	2.3.1 Hot-Side SCR Applications	6
	Figure 3. Direct Capital Cost Model for the Ammonia Handling and Injection System.....	7
	2.3.2 Tail-End SCR Applications	7
2.4	Air Preheater Modifications.....	8
	Figure 4. Simplified Schematic of a Counter-Current Flow Heat Exchanger	9
2.5	Gas-Gas Heat Exchanger.....	11
2.6	ID Fan and Booster Fan Costs.....	11
	Figure 5. Simplified Schematic of the Gas-Gas Heat Exchanger Employed in Cold-Side SCR Systems.....	12
2.7	Structural Support.....	13
2.8	Miscellaneous Other Direct Capital Costs.....	13
2.9	Total Direct Cost	14
2.10	Other Capital Costs.....	15
3.0	Annual Costs.....	16
	3.1 Fixed Operating Costs.....	16
	3.2 Variable Operating Costs.....	16
4.0	Nomenclature.....	17
5.0	References	20

1.0 INTRODUCTION TO SCR TECHNOLOGY

Selective catalytic reduction (SCR) is a process for the post-combustion removal of NO_x from the flue gas of fossil-fuel-fired power plants. SCR is capable of NO_x reduction efficiencies of up to 80 or 90 percent. SCR technology has been applied for treatment of flue gases from a variety of emission sources, including natural gas- and oil-fired gas turbines, process steam boilers in refineries, and coal-fired power plants. SCR applications to coal-fired power plants have occurred in Japan and Germany. Full-scale SCR systems have not been applied to coal-fired power plants in the U.S., although there have been small-scale demonstration projects.

Increasingly strict NO_x control requirements are being imposed by various state and local regulatory agencies in the U.S. These requirements may lead to U.S. SCR applications, particularly for plants burning low sulfur coals (Robie et al., 1991). Furthermore, implicit in Title IV of the 1990 Clean Air Act Amendment is a national NO_x emission reduction of 2 million tons per year. Thus, there may be other incentives to adapt SCR technology more generally to U.S. coal-fired power plants with varying coal sulfur contents. However, concern remains over the applicability of SCR technology to U.S. plants burning high sulfur coals or coals with significantly different fly ash characteristics than those burned in Germany and Japan. There is also concern regarding the application of SCR to peaking units due to potential startup and shutdown problems (Lowe et al., 1991).

In this report, new capital cost models of two SCR systems are developed. These are "hot-side high-dust" and "tail-end low-dust" options. These configurations are shown schematically in Figure 1. In a previous quarterly report (CMU, 1993), performance models for these two systems were developed.

Figure 1. Alternative Placements of SCR Systems in Coal-Fired Power Plants

2.0 SCR CAPITAL COST MODEL

The major equipment cost items for hot-side SCR systems include: (1) reactor housing; (2) catalyst; (3) ammonia storage and injection; (4) ductwork; (5) air preheater modifications; (6) induced draft fan modifications; (7) ash handling; and (8) water treatment. Additional costs may be incurred for flow modeling to properly design the ammonia injection system. For a tail-end system, there are no costs associated with air preheater modifications or ash handling. However, there are additional costs associated with the gas-gas heat exchanger and duct-burner used for flue gas reheat.

2.1 Reactor Housing

A reactor housing cost model was developed by Frey (1988). This model was later incorporated into the Integrated Air Pollution Control System (IAPCS) Version 4.0 computer program (Maibodi et al., 1990). The model was based on a statistical analysis of data from TVA (Maxwell and Humphries, 1985). The reactor housing costs include carbon steel reactor vessel with six inches of mineral wool insulation, vessel internals and supports, steam sootblowers, reactor crane and hoist, installation labor, foundations, structures, piping, and electrical equipment. The costs for the reactor housing exclude catalyst. The direct capital cost for the reactor housing is given by:

$$DC_R(1982) = 145,800 + 316.9 \left(\frac{V_{TOT}}{N_{TOT}} \right) \quad (1)$$

This model was based on regression analysis of 12 data points and has a coefficient of determination (R^2) of 0.998. The catalyst volume per housing ranged from 5,000 to 17,000 ft³.

A similar analysis was done for data reported by EPRI (Bauer and Spendle, 1984). A total of 18 cost estimates were reported with catalyst volumes ranging from roughly 4,000 to 14,000 ft³. However, the nature of the model was substantially different from the TVA model. It appears that the TVA model is predicated on the assumption of a simple linear relationship between reactor cost and catalyst volume. The EPRI data indicate a non-linear relationship, in which larger volumes yield a substantial economy of scale.

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

Robie and Ireland (1991) did not report the actual catalyst volumes for all 11 data points used in the regression model. Therefore, the catalyst volume was estimated based on the reported flue gas mass flow and flue gas molecular weight (which were used to calculate flue gas

volumetric flow rate), active catalyst space velocity, and the ratio of the number of active plus spare catalyst layers to the number of active layers. Thus, the estimated total catalyst volume is given by:

$$V_{TOT} = \left(\frac{G_{fg}^o}{SV_{act}^o} \right) \left(\frac{N_{act} + N_{sp}}{N_{act}} \right) \quad (2)$$

where:

$$G_{fg}^o = \frac{\left(\frac{m_{fg}}{MW_{fg}} \right) R T_{fg}^o}{P_{fg}^o} \quad (3)$$

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

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

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

$$DC_R = 18.65 N_{R,TOT} \left(\frac{V_{TOT}}{N_{R,TOT}} \right)^{0.489} \left(\frac{PCI}{357.3} \right) \quad (4)$$

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

Figure 2. Direct Capital Cost Model for SCR Reactor Housing.

From a statistical perspective, this regression model has significant limitations. The high coefficient of determination is influenced by the wide separation between two groups of data points. One set of data points are clustered for space velocities ranging from 10,000 to 20,000 cubic feet, whereas two other datapoints are close to 50,000 cubic feet. There is considerable scatter among the data points within the first cluster. However, because of the lack of reported detail regarding catalyst volumes and installation costs, it is not possible to reconcile these differences. Furthermore, there is no reported basis for the difference in values between the oil-fired and coal-fired data points. The oil-fired data points were excluded from the regression analysis, but are shown in Figure 2 for comparative purposes.

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

2.2 Ammonia Handling and Injection

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

dampers, and truck unloading station. The latter includes vapor recovery compressors, water deluge system, and transfer piping. The dilution air requirement is 20 parts air to one part ammonia. The regression model for ammonia process area direct costs is:

$$DC_{NH_3} = 50.8 (m_{NH_3,i})^{0.482} \left(\frac{PCI}{357.3} \right) \quad (5)$$

This model has a coefficient of determination (R^2) of 0.87 and a standard error of \$285,000. The regression model is shown graphically in Figure 3. Also shown in the figure is a regression model developed from data reported in an earlier EPRI report (Bauer and Spendle, 1984). The newer model yields costs that are lower than the previous model by a factor of approximately two. The reasons for the difference are not immediately clear. The earlier report used generally more conservative assumptions, reflecting higher levels of uncertainty perceived at that time for this technology.

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

2.3 Ductwork

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

2.3.1 Hot-Side SCR Applications

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

Figure 3. Direct Capital Cost Model for the Ammonia Handling and Injection System.

duct runs would be similar for differently sized systems. The parameters of this model, which include a multiplicative constant and an exponential coefficient, were estimated from the two data points. The resulting model is:

$$DC_D = 14.2 (G_{fg})^{0.7} \left(\frac{PCI}{357.3} \right) \quad (6)$$

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

2.3.2 Tail-End SCR Applications

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

are estimated as a multiplier of the costs for ductwork for a hot-side SCR application. The general equation is:

$$DC_{D,CS} = f_{D,CS} 14.2 (G_{fg})^{0.7} \left(\frac{PCI}{357.3} \right) \quad (7)$$

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

2.4 Air Preheater Modifications

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

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

Figure 4. Simplified Schematic of a Counter-Current Flow Heat Exchanger.

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

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

For a counter-current heat exchanger (see Figure 4), the heat transfer is given by:

$$q_{aph} = U_{aph} A_{aph} \Delta T_{LM,aph} \quad (8)$$

where

$$\Delta T_{LM,aph} = \frac{(T_{fg,i} - T_{a,o}) - (T_{fg,o} - T_{a,i})}{\ln \left(\frac{T_{fg,i} - T_{a,o}}{T_{fg,o} - T_{a,i}} \right)} \quad (9)$$

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

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

$$(UA)_{aph} = \frac{q_{aph}}{\Delta T_{LM,aph}} \quad (10)$$

$$q_{aph} = M_{fg} c_{p,fg} (T_{fg,o} - T_{fg,i}) \quad (11)$$

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

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

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

$$DC_{aph,mod} = 1370 N_{T,aph} \left(\frac{UA_{T,aph}}{4.4 \times 10^6 N_{T,aph}} \right)^{0.8} \left(\frac{PCI}{357.3} \right) \quad (12)$$

While this model is based on only two data points, it nonetheless appears to provide a qualitatively reasonable relationship between air preheater modification costs based on the size of the air preheater. The model suggests a modest economy of scale for modifications to larger air preheaters.

2.5 Gas-Gas Heat Exchanger

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

A simplified schematic of the GGH is shown in Figure 5. Gas exiting the flue gas desulfurization (FGD) system is heated by counter-current heat exchange with high temperature gas exiting the SCR reactor. The "untreated" gas entering the SCR system is heated by a duct burner.

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

$$DC_{GGH} = 9100 N_{T,GGH} \left(\frac{UA_{T,GGH}}{4.4 \times 10^6 N_{T,GGH}} \right)^{0.8} \left(\frac{PCI}{357.3} \right) \quad (13)$$

This model is based on a GGH design with 5.6 percent gas leakage, on a mass basis. For a system with no leakage, multiply the cost given in Equation (13) by a factor of 2.4.

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

$$DC_{DB} = 264 N_{T,DB} \left(\frac{Q_{NG}}{90 N_{T,DB}} \right)^{0.6} \left(\frac{PCI}{357.3} \right) \quad (14)$$

where

$$Q_{NG} = \frac{m_{NG} HHV_{NG}}{10^6} \quad (15)$$

2.6 ID Fan and Booster Fan Costs

For a new SCR installation, the ID fans must be sized to deal with the increased flue gas pressure drop resulting from the additional ductwork and the SCR reactor. Typically, the increase in flue gas pressure drop is approximately 11 inches of water. The size of an ID fan and motor is

Figure 5. Simplified Schematic of the Gas-Gas Heat Exchanger Employed in Cold-Side SCR Systems.

proportional both to the flue gas flow rate and to the pressure drop. Therefore, the cost of the ID fan modifications is assumed here to be proportional to the difference in flue gas energy requirement necessary to overcome the flue gas pressure drop. This energy requirement is given by:

$$EC_{ID,dif} = \frac{\dot{Q}_{fg} \Delta P_{SCR}}{8,512 \eta_{fan}} \quad (16)$$

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

$$DC_{ID,dif} = 180 \left(\frac{EC_{ID,dif}}{4,600} \right)^{0.6} \left(\frac{PCI}{357.3} \right) \quad (17)$$

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

energy requirement as a function of the actual flue gas volumetric flow rate and the pressure drop, the cost of a cold-side SCR system booster fan is given by:

$$DC_{BF} = 810 N_{T,BF} \left(\frac{EC_{BF}}{3,000 N_{T,BF}} \right)^{0.6} \left(\frac{PCI}{357.3} \right) \quad (18)$$

There is typically one booster fan per SCR reactor train.

2.7 Structural Support

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

$$DC_s = f_s (DC_R + DC_D + DC_{APH}) \quad (19)$$

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

$$DC_s = f_s (DC_R + DC_D + DC_{GGH}) \quad (20)$$

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

2.8 Miscellaneous Other Direct Capital Costs

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

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

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

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

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

$$DC_{\text{misc}} = \left\{ 100 + 300 \left(\frac{MW_e}{550} \right)^{0.6} \right\} \left(\frac{PCI}{357.3} \right) \quad (21)$$

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

$$DC_{\text{misc,CS}} = \left\{ 100 + 650 \left(\frac{MW_e}{550} \right)^{0.6} \right\} \left(\frac{PCI}{357.3} \right) \quad (22)$$

2.9 Total Direct Cost

The total direct cost is the sum of all of the direct costs. The cost of the initial catalyst charge is also included here in the total direct cost, because it is such a large and integral part of the SCR system. One component of the direct cost not estimated in any of the above sections is the cost of general facilities associated with the SCR process area. Based on the estimates reported by Robie and Ireland, it appears that general facilities cost is approximately 4 percent of the sum of all other direct costs. The total direct cost for the hot-side SCR system is therefore:

$$TDC_{\text{HS}} = (1+f_{\text{GF}}) \left(V_{\text{act}} UC_{\text{cat}} + \sum_{i=1}^7 DC_i \right) \quad (23)$$

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

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

$$TDC_{CS} = (1+f_{GF}) \left(V_{act} UC_{cat} + \sum_{i=1}^8 DC_i \right) \quad (24)$$

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

2.10 Other Capital Costs

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

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

$$C_{EHO} = f_{EHO} TDC \quad (25)$$

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

$$C_{ProjC} = f_{ProjC} TDC \quad (26)$$

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

$$C_{ProcC} = \sum_i f_{ProjC} DC_i \quad (27)$$

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

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

$$TPC = TDC + C_{EHO} + C_{ProjC} + C_{ProcC} \quad (28)$$

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

The final measure of capital cost is the total capital requirement (TCR). The TCR includes the total plant investment plus costs for royalties, startup costs, and initial inventories of feedstocks. In this case, no costs are assumed for royalties. Preproduction costs typically include one month of both fixed and variable operating costs and two percent of total plant investment. In the case of an SCR system, by far the largest portion of the preproduction (startup) costs are represented by the two percent multiplier on TPI. Inventory capital is estimated as 0.5 percent of total process capital excluding catalyst. The costs for initial catalysts and chemicals is zero. The SCR catalyst is included in the process capital costs. Thus, for an SCR system, the total capital requirement is:

$$TCR = (1 + f_{pp}) TPI + f_{IC} \sum_i DC_i \quad (29)$$

3.0 Annual Costs

The annual costs for SCR systems include fixed and variable operating costs. Fixed operating costs include operating labor, maintenance labor and materials, and overhead costs associated with administrative and support labor. Variable operating costs include consumables, such as ammonia and catalyst replacement. Costs for steam and electricity consumed from within the plant may also be estimated.

3.1 Fixed Operating Costs

The fixed operating costs are predicated on 4 hours of labor per day per SCR reactor train for operating labor, and 8 hours of labor per day per SCR train for maintenance labor. Maintenance materials are estimated as one percent of the total process capital cost, excluding catalyst. Administrative and support labor is estimated as 30 percent of operating and maintenance labor costs. Therefore, the total fixed operating cost is given by:

$$FOC = 1.3 (4 + 8) 365 N_{T,R} UC_L + 0.01 \sum_i DC_i \quad (30)$$

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

3.2 Variable Operating Costs

The major component of the variable operating cost is for catalyst replacement. The analytical models for estimating catalyst replacement are reported by Frey (1993) and are not repeated here. Similarly, the ammonia mass flow requirement, the steam requirement for ammonia injection, and the electricity consumption for the SCR systems are also reported by Frey (1993).

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

$$VOC = V_{C,a} UC_{cat} + 8,760 c_f \left[\left(\frac{M_{NH_3,A,i}}{117.6} \right) UC_{NH_3} + \left(\frac{18 M_{steam}}{1,000} \right) UC_{steam} + E_{scr} UC_{elec} \right] \quad (31)$$

The unit cost for catalyst is highly variable. Robie and Ireland (1991) used a value of \$660/ft³, but noted that at the time of their study, catalyst costs in Europe were as low as \$330/ft³. The unit cost of ammonia is typically \$150/ton. The unit cost of steam is approximately \$3.00/1,000 lb. The cost of power is often assumed to be 5 cents per kWh.

4.0 Nomenclature

c_f	=	Plant capacity factor (fraction of the year at full load)
C_{ProcC}	=	Cost of project contingencies, \$1,000
$DC_{aph,mod}$	=	Direct capital cost of air preheater modifications, \$1,000
DC_{BF}	=	Direct capital cost of the booster fan for cold-side SCR, \$1,000.
DC_D	=	Direct capital cost of ductwork, \$1,000
$DC_{D,CS}$	=	Direct capital cost of ductwork for a cold-side SCR system, \$1,000
DC_{DB}	=	Direct cost of duct burners for a cold-side SCR system, \$1,000.
DC_{GGH}	=	Direct capital cost of the gas-gas heat exchanger for cold-side SCR, \$1,000.
DC_{misc}	=	Direct capital cost of miscellaneous items, \$1,000.
DC_{NH_3}	=	Direct capital cost of the ammonia injection systems, \$1,000
DC_R	=	Direct capital cost of reactors, \$1,000.
$EC_{ID,dif}$	=	Electricity consumption associated with SCR flue gas pressure drop, kWh.
E_{scr}	=	Electricity requirement for SCR, kW
$f_{D,CS}$	=	Factor for cold-side SCR duct costs.
f_{IC}	=	Factor for inventory capital costs (0.005 of non-catalyst direct costs)
f_{pp}	=	Factor for preproduction costs (0.02 of TPI)
$f_{ProjC,i}$	=	Project contingency factors for each process area.

G_{fg}^o	=	Flue gas volumetric flow rate, referenced to a standard temperature (32 °F) and pressure (1 atm), ft ³ /hr.
G_{fg}	=	Flue gas volumetric flow rate, at actual temperature and pressure, ft ³ /hr.
HHV_{NG}	=	Higher heating value of natural gas, BTU/lb
m_{fg}	=	Mass flow rate of flue gas, lb/hr.
m_{NG}	=	Mass flow rate of natural gas, lb/hr
$m_{NH_3,i}$	=	Mass flow rate of ammonia injected into the flue gas, lb/hr
$M_{NH_3,A,i}$	=	Molar flow rate of ammonia injected into the SCR system, lbmole/hr
M_{steam}	=	Molar flow rate of steam required for ammonia injection, lbmole/hr
MW_{fg}	=	Equivalent molecular weight of flue gas, lb/lbmole.
N_{act}	=	Number of active catalyst layers at plant startup.
N_{sp}	=	Number of spare catalyst layers at plant startup.
$N_{R,TOT}$	=	Total number of reactor housings.
$N_{T,aph}$	=	Total number of air preheaters
$N_{T,DB}$	=	Total number of duct burners
$N_{T,GGH}$	=	Total number of Gas-Gas Heat Exchangers.
P_{fg}^o	=	Reference pressure of the flue gas at standard conditions, 1 atm.
PCI	=	<i>Chemical Engineering</i> Plant Cost Index. For Dec. 1989, PCI = 357.3.
$\dot{Q}_{fg}$	=	Flue gas volumetric flow rate, actual ft ³ /min.
Q_{NG}	=	Heating value of natural gas used in duct burners, BTU/hr.
R	=	Ideal Gas Constant, 0.730 (ft ³ atm)/(lbmole °R).
SV_{act}^o	=	Active catalyst space velocity, referenced to a flue gas temperature of 32 °F, 1/hr.
$T_{a,i}$	=	Temperature of air entering a heat exchanger, °R.
$T_{a,o}$	=	Temperature of air exiting a heat exchanger, °R.

T_{fg}^o	=	Reference temperature of the flue gas at standard conditions, 32 °F.
$T_{fg,i}$	=	Temperature of flue gas entering a heat exchanger, °R.
$T_{fg,o}$	=	Temperature of flue gas exiting a heat exchanger, °R.
UA area,	=	Product of universal heat transfer coefficient and heat exchanger surface BTU/°R
UC_{cat}	=	Unit cost of catalyst, \$/ft ³ .
UC_{elec}	=	Unit cost of electricity, \$/kWh
UC_L	=	Unit cost of labor, \$/hour
UC_{NH_3}	=	Unit cost of ammonia, \$/ton
UC_{steam}	=	Unit cost of steam, \$/1,000 lb
$V_{C,a}$	=	Volume of catalyst replaced annually, ft ³ /yr.
V_{TOT}	=	Total volume of the catalyst, including active and spare layers, ft ³ .

5.0 References

- Bauer, T.K., and R.G. Spendle (1984), "Selective Catalytic Reduction for Coal-Fired Power Plants: Feasibility and Economics," CS-3603, Electric Power Research Institute, Palo Alto, CA, October.
- CMU (1993), "Development of the Integrated Environmental Control Model: Performance Models of Selective Catalytic Reduction (SCR) NO_x Control Systems," Quarterly Progress Report prepared by Carnegie Mellon University for U.S. Department of Energy, June 1993.
- EPRI (1986). TAG - Technical Assessment Guide. Vol 1. P-4463-SR. Electric Power Research Institute. Palo Alto, CA. December.
- Frey, H.C. (1988), "Selective Catalytic Reduction Cost Algorithm for the Advanced Utility Simulation Model", Technical Memorandum to J.A. Martinez, Radian Corporation, July 8.
- Maibodi, M., A.L. Blackard, and R.J. Page (1990), "Integrated Air Pollution Control System Version 4.0, Volume 2: Technical Documentation Manual," Prepared by Radian Corporation for U.S. Environmental Protection Agency, EPA-600/7-90-022b, December.
- Maxwell, J.D., and L.R. Humphries (1985), "Economics of Nitrogen Oxides, Sulfur Oxides, and Ash Control Systems for Coal-Fired Utility Power Plants," Prepared for U.S. Environmental Protection Agency, EPA-600/7-85-006, February.
- Lowe, P.A., W. Ellison, and M. Perlswieg. "Understanding the German and Japanese Coal-Fired SCR Experience," presented at the 1991 Symposium on Stationary Combustion Nitrogen Oxide Control, Washington, DC, March 25-28, 1991.
- Robie, C.P., and P.A. Ireland (1991), "Technical Feasibility and Cost of Selective Catalytic Reduction (SCR) NO_x Control," GS-7266, Prepared by United Engineers and Constructors, Inc. for the Electric Power Research Institute, Palo Alto, CA, May.

DATE

FILMED

6 / 1 / 94

END