

Coal
Utility
Environmental
Cost



CUECost WORKBOOK DEVELOPMENT DOCUMENTATION

Version 5.0

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Air Pollution Prevention and Control Division
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ABSTRACT

This document serves as a user's manual for the Coal Utility Environmental Cost (CUECost) workbook and documents its development and the validity of methods used to estimate installed capital and annualized costs. The CUECost workbook produces rough-order-of-magnitude (ROM) cost estimates (+/-30% accuracy) of the installed capital and annualized operating costs for air pollution control (APC) systems installed on coal-fired power plants to control emissions of sulfur dioxide (SO₂), nitrogen oxides (NO_x), particulate matter (PM), mercury (Hg), and carbon dioxide (CO₂). In general, system performance is an input requirement for the workbook user. The workbook was designed to calculate estimates of an integrated APC system or individual component costs for various APC technologies used in the utility industry. Twelve technologies are currently in the workbook: flue gas desulfurization (FGD)—limestone with forced oxidation (LSFO) and with dibasic acid and lime spray drying (LSD); particulate matter removal—electrostatic precipitator (ESP) and fabric filter (FF); NO_x control—selective catalytic reduction (SCR), selective non-catalytic reduction (SNCR), natural gas reburning, and low-NO_x burner (LNB); mercury control—powdered activated carbon (PAC) injection; and CO₂ control—monoethanolamine (MEA) process, chilled ammonia process (CAP) and sorbent injection (SI). It is expected that this manual will be useful to a broad audience, including: (1) individuals responsible for developing and implementing SO₂, NO_x, PM, Hg, and CO₂ control strategies at sources, (2) state authorities implementing pollution control programs, and (3) the interested public at large. Moreover, persons engaged in research and development efforts aimed at improving cost-effectiveness of air pollution control technology applicable to coal-fired plants may also benefit from this manual.

Note:

The original model was delivered by Raytheon Engineers & Constructors, Inc. for Eastern Research Group, Inc. under EPA Contract No. 68-D7-0001. Subsequent revision was completed by Andover Technology Partners for ARCADIS under EPA Contract No. EP-C-04-023. This version 5.0 was revised and accompanying documentation prepared by ARCADIS under EPA Contract No. EP-C-04-023.

Table of Contents

<u>Section</u>	<u>Page</u>
Abstract	i
Acronyms.....	vii
Introduction and Summary	1
Overview	1
Background.....	1
Workbook Description	2
CUECost WORKBOOK development and documentation document Contents	4
Project Approach.....	5
NO _x Control Estimate	5
Particulate Matter Control Estimates	6
SO ₂ Control Estimates	7
CO ₂ Control Estimates	7
Mercury Control Estimates	7
Default Plant Criteria	8
Results	8
Getting Started	10
Hardware and Software requirements / Internet Access	10
Getting Started	10
Workbook Layout and Methodology	12
Workbook Layout	12
Methodology.....	14
Input and Output Options	20
Input Data	20
Output Options	21
Worksheet Validation	22
FGD Worksheets – LSFO and LSD Technologies	22
Particulate Matter Control Worksheet	23
NO _x Control Worksheet	25
Mercury Control Worksheet.....	27
Carbon Dioxide Control Worksheet.....	28
Validation Summary	29
References	30
Appendix A Terminology Definitions, Abbreviations, Acronyms, and Range Names	33
A.1 Definition of Terms	33

Appendix B Technology Descriptions/Criteria	36
B.1 Limestone Forced Oxidation Design Criteria	36
B.2 Lime Spray Dryer Design Criteria	39
B.3 Particulate Matter Control Design Criteria	41
B.4 NO _x Control Technology Criteria.....	42
B.4.1 Selective Catalytic Reduction Design Criteria.....	43
B.4.2 Selective Non-catalytic Reduction Design Criteria	45
B.4.3 Natural Gas Reburning Design Criteria	47
B.4.4 Low-NO _x Burner Technology Design Criteria	49
B.5 Hg Control Technology Criteria	50
B.5.1 Mercury Removal Models	50
B.5.2 Mercury Removal by Existing Equipment, $f_{\text{existing equipment}}$	51
B.5.3 Mercury Reduction by PAC injection, $f_{\text{PAC injection}}$	54
B.5.4 PAC Injection Models Developed from Full-Scale Data	56
B.5.5 Mercury Speciation with SCR	62
B.5.6 Conclusions	69
B.6 CO ₂ Control Design Criteria	69
References	71
Appendix C Design/Economic Criteria	73
C.1 General Plant Design Criteria.....	73
C.2 Economic Criteria	76
Appendix D Cost Algorithm Development/Validation/Sources	77
D.1 FGD Cost Algorithm Development	77
D.2 Selective Catalytic Reduction.....	78
D.2.1 Performance Parameters.....	78
D.2.2 Capital Costs.....	79
D.2.3 Operating and Maintenance Costs	81
D.2.4 CUECost Validation	83
D.3 Selective Noncatalytic Reduction.....	86
D.3.1 Performance Parameters.....	86
D.3.2 Capital Costs.....	87
D.3.3 Operating and Maintenance Costs	89
D.3.4 CUECost Validation	91
D.4 Natural Gas Reburning	91
D.4.1 Performance Parameters.....	91
D.4.2 Capital Costs.....	94
D.4.3 Operating and Maintenance Costs	95
D.4.4 CUECost Validation	96
D.5 Low-NO _x Burner Technology.....	98
D.5.1 Capital Costs.....	98
D.5.2 Operating and Maintenance Costs	99
D.5.3 CUECost Validation	100

Table of Contents

D.6	Hg Control Technology	100
D.7	CO ₂ MEA Control System Cost Algorithm Development	104
D.7.1	Capital Cost	104
D.7.2	Operating and Maintenance Costs	105
D.8	CO ₂ Cap Control System Cost Algorithm Development.....	110
D.8.1	Capital Cost	110
D.9	CO ₂ SI Control System Cost Algorithm Development.....	112
D.9.1	Preconditioning.....	113
D.9.2	Absorber	115
D.9.3	Blower/ID Fan	119
D.9.4	Regenerator	120
	References	123
Appendix E INPUT WORKSHEET SCREENS		125
E.1	Getting Started	125
E.2	Inputs.....	126
E.2.1	Economic Inputs.....	126
E.2.2	Power Generation Technique Choices.....	127
E.2.3	APC Technology Choices	127
E.2.5	Particulate Control Inputs	129
E.2.6	SO ₂ Control Inputs	129
E.2.7	Mercury Control Inputs.....	130
E.2.8	CO ₂ Control Inputs.....	131
Appendix F Programs for Economic Parameters.....		133

List of Figures		Page
Figure 1.	CUECost Workbook Map	3
Figure 2.	CUECost Logic Diagram.....	15
Figure B-1.	Salem Harbor Mercury Removal without PAC Injection (Durham et al., 2001)	54
Figure B-2a.	Gaston Testing	57
Figure B-2b.	Gaston Testing	58
Figure B-3.	Deviation of the Gaston PAC Algorithm.....	59
Figure B-4.	PPPP Testing	60
Figure B-5.	Deviation from the PPPP PAC Algorithm	60
Figure B-6.	Brayton Point Testing	61
Figure B-7.	Deviation from the Brayton Point PAC Algorithm	62
Figure B-8.	Mercury Oxidation without a Catalyst as a Function of Residence Time, Gas Temperature, and HCl Content (Hocquel et al., 2002).....	64
Figure B-9.	Mercury Oxidation across SCR Catalysts and without SCR Catalyst (Hocquel et al., 2002)	64
Figure B-10.	Oxidation of Mercury across C-1 SCR Catalyst in PRB-derived Flue Gas (Richardson et al., 2002)	65
Figure B-11.	Effect of Flue Gas Exposure Time on C-1 SCR Catalyst Oxidation of Elemental Mercury: 700 °F and Space Velocity of 1,450 h ⁻¹ (Richardson et al., 2002)	65
Figure D-1.	PAC, Bituminous FF	101
Figure D-2.	PAC, Bituminous ESP	101
Figure D-3.	PAC, Subbituminous FF	102
Figure D-4.	PAC, Subbituminous ESP	102
Figure D-5.	BPAC.....	102
Figure D-6.	Cost of Mercury Reduction, LS Bituminous Coal and ESP	103
 List of Tables		 Page
Table 1.	British to Metric Conversion Factors	9
Table 2.	Total Capital Requirement Calculation Method	17
Table 3.	Annualized Cost Calculation Method	19
Table 4.	CUECost-FGD Cost Comparison to FGDCOST by EPRI for Phase 1 Acid Rain Installations	23
Table 5.	Comparison of CUECost ESP Sizing Estimates with Raytheon Model	25
Table 6.	Percent Difference between CUECost and Acid Rain Division Studies (Khan and Srivastava, 2004) for Retrofit Cases	27
Table 7.	Estimated Costs of ACI Control Systems according to CUECost	28
Table 8.	Estimated Costs of CO ₂ Control Technologies with CUECost and IECM model.....	28
Table B-1.	Specific Design Criteria for LSFO	38
Table B-2.	Input for LSD and the Default Values for the Inputs	41
Table B-3.	Inputs for Particulate Matter Control and Its Default Values.....	42
Table B-4.	Default Input Parameters for SCR.....	45

Table of Contents

Table B-5.	Default Input Parameters for SNCR.....	47
Table B-6.	Default Input Parameters for NGR	49
Table B-7.	Default Values for LNBT Input Parameters	50
Table B-8.	Predicted Collection of Mercury by ESP according to Eqs. B-19 and B-20	52
Table B-9.	Values of Constants Used in the PAC Injection from Eqs. B-21 and B-22	56
Table B-10.	Coefficients for Curve Fit Algorithms	58
Table B-11.	Summary of Results from Full-Scale SCR Mercury Oxidation Tests (Bustard et al., 2001)	67
Table B-12.	Default Values for Mercury Control Input Parameters	68
Table C-1.	Snapshot for a Specific Plant and Its Default Parameters	74
Table C-2.	Coal Analysis Library	75
Table C-3.	Economic Inputs	76
Table D-1.	Variable and Constant Parameters for Wet FGD Cost Algorithm.....	77
Table D-2.	Parameters for LSD Cost Algorithms	78
Table D-3.	Direct Capital Costs for Hot-side SCR (Installed equipment costs)	80
Table D-4.	Indirect Capital Costs for Hot-side SCR.....	81
Table D-5.	Operating and Maintenance Cost Equations for SCR (\$/year)	82
Table D-6.	CUECost with Acid Rain Division Study Design for SCR (1990 dollars)	84
Table D-7.	Acid Rain Division Study: SCR Applications	85
Table D-8.	Direct Capital Costs For SNCR (Installed Equipment Costs).....	88
Table D-9.	Indirect Capital Costs for SNCR	89
Table D-10.	Annual Operating and Maintenance Costs for SNCR.....	90
Table D-11.	CUECost with Acid Rain Division Study Cases for SNCR (1990 dollars)	92
Table D-12.	Acid Rain Division Study: SNCR Applications (1990 dollars)	93
Table D-13.	Direct Capital Costs for NGR (Installed equipment cost).....	94
Table D-14.	Indirect Capital Costs for NGR.....	95
Table D-15.	Annual Operating and Maintenance Costs and Savings for NGR.....	95
Table D-16.	CUECost with Acid Rain Division Study Cases for NGR (1990 dollars)	97
Table D-17.	Acid Rain Division Study: NGR Applications (1990 dollars)	98
Table D-18.	Total Capital Costs for LNBT Retrofit	99
Table D-19.	Annual Operating and Maintenance Costs for LNBT (\$/year)	100
Table D-20.	CUECost with Acid Rain Division Study Cases for LNBT (1990 dollars)	100
Table D-21.	Constants for Eqs. D-9 and D-10.....	101
Table D-22.	Indirect Capital Costs for CO ₂ Control.....	105

ACRONYMS

A/C	Air to Cloth Ratio
ACI	Activated Carbon Injection
AFDC	Allowance for Funds Used During Construction
APC	Air Pollution Control (equipment)
APPCD	Air Pollution Prevention and Control Division
ARD	Acid Rain Division (of EPA)
BPAC	Brominated Powdered Activated Carbon
CAP	Chilled Ammonia Process
CCF	Carrying Charge Factor
CE	Chemical Engineering (Magazine)
CEMS	Continuous Emissions Monitoring System
CEPCI	Chemical Engineering Plant Cost Index
CF	Capacity Factor
CFR	Code of Federal Regulations
CMU	Carnegie Mellon University
COHPAC	Compact Hybrid Particle Collector
CUECost	Coal Utility Environmental Cost (model)
CV	Catalyst Volume
DBA	Dibasic Acid
DC	Direct Capital
DCC	Direct Contact Cooler
DOE	Department of Energy
EPA	Environmental Protection Agency
EPAC	Enhanced Powdered Activated Carbon
EPRI	Electric Power Research Institute
ESP	Electrostatic Precipitator
ESPC	Electrostatic Precipitator (cold)
ESPh	Electrostatic Precipitator (hot)
FF	Fabric Filter
FGD	Flue Gas Desulfurization
GDP	Gross Domestic Product
GHG	Greenhouse Gas
HHV	Higher Heating Value
IAPCS	Integrated Air Pollution Control System

Acronyms, Units, and Symbols _____

ICR	Information Collection Request
ID	Induced Draft (fan)
IECM	Integrated Environmental Control Model
IGCC	Integrated Gasification Combustion Combination
IPM	Integrated Planning Model
k	Exponential constant used to related coal type, removal efficiency and ash resistivity to ESP size in terms of the Specific Collection Area (SCA)
L/G	Liquid-to-Gas Ratio
LNB	Low NO _x Burner
LNBT	Low NO _x Burner Technology
LNCFS	Low-NOX Concentric Firing Systems
LOI	Loss On Ignition
LR	Learning Rate
LS	Low Sulfur (bituminous coal)
LSD	Lime Spray Drying (flue gas desulfurization)
LSFO	Limestone (flue gas desulfurization)with Forced Oxidation
MEA	Monoethanolamine
MEL	Magnesium Enhanced Lime
MHI	Mitsubishi Heavy Industry
NEMS	National Energy Modeling System
NETL	National Energy Technology Laboratory (DOE)
NG	Natural Gas Combined Cycle
NGR	Natural Gas Reburning
NM	Not Measured
NRMRL	U.S. EPA National Risk Management Research Laboratory
O&M	Operation and Maintenance
OFA	Over-fire air (used to complete coal combustion in some LNBT applications)
ORD	U.S. EPA Office of Research and Development
PAC	Powdered Activated Carbon
PC	Pulverized Coal
PCI	Plant Cost Index
PJFF	Pulse-Jet Fabric Filter
PM	Particulate Matter
PPPP	Pleasant Prairie Power Plant
PR	Progress Ratio
PRB	Powder River Basin (coal)

Acronyms, Units, and Symbols _____

PV	Present Value
RAM	Random Access Memory
RF	Retrofit Factor
RLCS	Rubber-Lined Carbon Steel
ROM	Rough Order of Magnitude
SCA	Specific Collection Area (refers to ESP size in terms of plate area (ft ²)/1000 acfm)
SCPC	Supercritical Pulverized Coal
SCR	Selective Catalytic Reduction
SDA	Spray Dryer Absorber
SI	Sorbent Injection
SNCR	Selective Non-catalytic Reduction
SV	Space Velocity
TAG	Technical Assessment Guide
TCE	Total Cash Expended
TCR	Total Capital Requirement
TEG	Triethylene Glycol (dehydrator)
TOL	Technological Optimism Learning
TP	Tax Paid
TPC	Total Plant Cost
TPI	Total Plant Investment
TTN	Technology Transfer Network
TVA	Tennessee Valley Authority
WACC	Weighted Average Cost of Capital

Units

acfm	Actual Cubic Feet per Minute
gpm	Gallons Per Minute
GW	Gigawatt
h	Hour
hp	Horsepower
kW	Kilowatt
kWh	Kilowatt Hour
MB	Megabyte
MMacf	Millions of Actual Cubic Feet
MMBtu	Millions of British Thermal Units
MPa	Megapascal
MW	Megawatt
MW _e	Megawatt (electric)
MWh	Megawatt Hour
ppm	parts per million
scfm	Standard Cubic Feet Per Minute

See Table 1 in the Introduction for units not listed here

Chemical Symbols

CO	Carbon Monoxide
CO ₂	Carbon Dioxide
H ₂ O	Water
HCl	Hydrogen Chloride
HCO ₃ ⁻	Bicarbonate
N ₂ O	Nitrous Oxide
NH ₃	Ammonia
NH ₄ ⁺	Ammonium
NH ₄ HSO ₄	Ammonium Bisulfate
NO	Nitrogen Oxide
NO ₂	Nitrogen Dioxide
NO _x	Nitrogen Oxides
SO ₂	Sulfur Dioxide
SO ₃	Sulfur Trioxide
SO _x	Sulfur Oxides

INTRODUCTION AND SUMMARY

OVERVIEW

This document serves as a User's Manual for the CUECost workbook and documents its development and the validity of the methods used to estimate installed capital and annualized costs. The CUECost economic analysis workbook produces rough-order-of-magnitude (ROM) cost estimates ($\pm 30\%$ accuracy) of the installed capital and annualized operating costs for air pollution control (APC) systems installed on coal-fired power plants. Costs for utility APC systems are site-specific. These costs are subject to change with changes in technology, labor rates, and material costs. The costs estimated by the CUECost workbook come from a variety of sources. With that understanding, one may assume, but it is not guaranteed, that CUECost will produce estimates in the range of accuracy of $\pm 30\%$ of the actual cost, which was the goal of this project.

The CUECost workbook was developed in Microsoft Excel workbook format to provide users with complete insight into the equipment cost estimating methodology. All assumptions are readily accessible to the user by reviewing the specific equations and references for each cell in the worksheets. CUECost is composed of technology-specific worksheets with one common input worksheet for all technologies. This structure allows the workbook to be expanded to incorporate other technologies in the future.

The original model (1998) was developed by Raytheon Engineers & Constructors, Inc. for Eastern Research Group, Inc. under EPA Contract No. 68-D7-0001. Subsequent revision was completed by Andover Technology Partners for ARCADIS under EPA Contract No. EP-C-04-023. This version 5.0 was revised by ARCADIS under EPA Contract No. EP-C-04-023.

Background

The Air Pollution Prevention and Control Division (APPCD) of the National Risk Management Research Laboratory (NRMRL) contracted for development of a cost estimating workbook for APC systems on coal-fired power plants. This workbook was developed in Excel format to

provide the user with more flexibility in modifying the worksheet and outputs to meet the user's needs for site-specific applications.

The workbook was designed to calculate estimates of an integrated APC system or individual component costs for various APC technologies currently used in the utility industry to reduce emissions of sulfur dioxide (SO₂), particulate matter (PM), nitrogen oxides (NO_x), mercury (Hg) and (in the future) carbon dioxide (CO₂) generated by coal-fired boilers. Technologies currently included in the workbook are:

Flue Gas Desulfurization (FGD)	=	Limestone with Forced Oxidation (LSFO) Lime Spray Drying (LSD)
Particulate Matter Removal	=	Electrostatic Precipitator (ESP) Fabric Filter (FF)
Nitrogen Oxide Control	=	Selective Catalytic Reduction (SCR) Selective Non-Catalytic Reduction (SNCR) Natural Gas Reburning (NGR) Low NO _x Burners (LNB)
Mercury Control	=	Powdered Activated Carbon (PAC) injection
Carbon Dioxide Control	=	Monoethanoamine (MEA) Process Chilled Ammonia Process (CAP) Sorbent Injection (SI)

WORKBOOK DESCRIPTION

A map of the CUECost workbook is shown in Figure 1. This design allows the addition of future technologies by inserting new worksheets into the workbook. The workbook calculates both new and retrofit plant costs using a 1.0 factor for a new facility, a 1.3 factor for a moderately difficult retrofit, and a 1.6 factor for a difficult retrofit. The user is also given the option to input his own retrofit factor based on plant-specific information. Equipment sizing and variable operating costs are derived based on the calculated material balances for specific process criteria, including flue gas flow rate, pollutant removal rate, chemical consumption rate, waste production rate, etc.

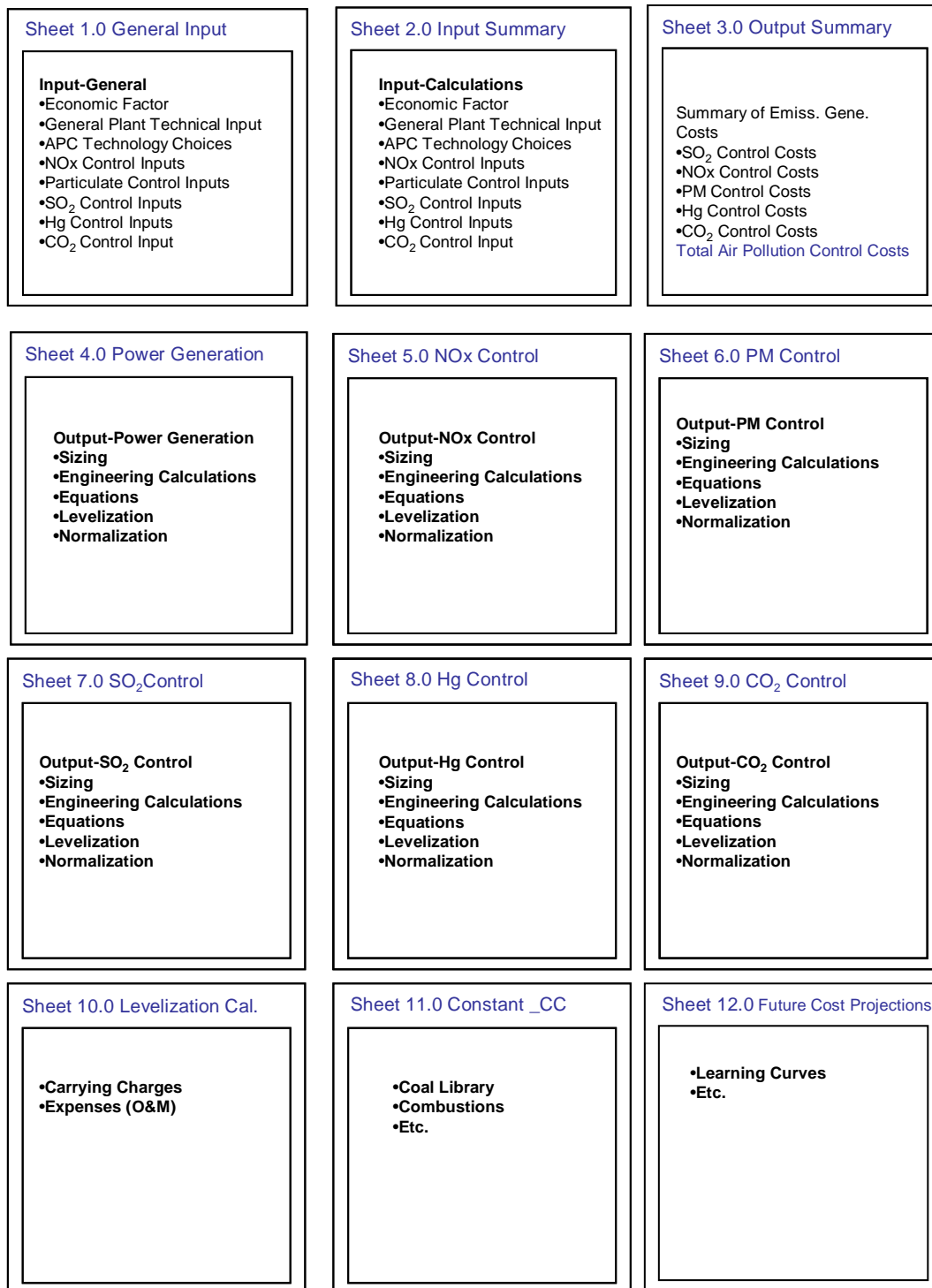


Figure 1. CUECost Workbook Map

The first sheet of the workbook functions as the menu of all sheets in the workbook. Users can follow the link, by clicking the icons, to the input or output for a specific air pollutant control technology. In this version 5.0, a toolbar was developed, including **Main Menu, Go to Top, User Input, Outputs, and Print Buttons**.

All inputs are integrated into one worksheet, and outputs for a specific control technology are listed separately in one worksheet. Economic-related outputs are first listed at the top of the outputs worksheet, with engineering-related calculations listed at the bottom. Version 5.0 of the CUECost workbook contains calculations of the carrying charges and levelizing factors for expenses in worksheet 10.0. In calculating the capital carrying charges and operation and maintenance (O&M) levelized cost, a 30-year plant duration was used. The calculation can be accessed from the worksheet "1.0 General Input" by clicking the calculator link.

CUECOST WORKBOOK DEVELOPMENT AND DOCUMENTATION DOCUMENT CONTENTS

This document consists of the following sections:

Overview of CUECost Workbook states the purpose and content of this document.

Getting Started presents an itemized listing of requirements for the user's computer system and is followed by a series of installation guidelines for use in installing the CUECost workbook to the user's hard disk. Instruction is also provided for the first-time user on how to get started producing a cost estimate using CUECost. These starting instructions include listings of the input sequence and other preliminary steps for the user to complete prior to using the CUECost workbook.

Workbook Layout and Methodology presents a detailed description of the contents of each worksheet and provides a layout diagram. This section provides a technical description of the workbook and discusses how the worksheets are integrated to minimize user input. The cost estimating methodology is also described, including a logic diagram to illustrate the calculation sequence that is used to develop capital and annualized cost estimates.

Input and Output Options provides a description of the input and output options available to the user for cost estimate development.

Worksheet Validation, the final section of the user's manual, summarizes the validation procedure that was followed during development and subsequent testing of the CUECost workbook.

Appendix A provides the definitions of terminology used in the text and worksheets.

Appendix B provides process criteria and technology descriptions of equipment included in each technology cost estimate.

Appendix C presents tabulations of the primary assumptions that served as the estimate basis for the default values included in the worksheets, including both plant design and economic criteria.

Appendix D discusses the data sources for the cost-versus-capacity algorithms. Previous publications, vendor quotations, and costs from recent APC installations served as the basis for all cost-versus-capacity curves used in the worksheets.

Appendix E provides a demonstration of the worksheets to show how the workbook is used. Pictures of the actual Excel screens are provided for easy reference to the screens shown when running CUECost.

Appendix F provides programs to calculate carrying charges and levelization of O&M costs.

PROJECT APPROACH

The workbook design allows the user to review all of the assumptions and equations contained in each worksheet and to adjust any of them to fit the user's particular needs. A multi-worksheet format was selected to allow the addition of other technologies if future expansion of the workbook is desired. A separate input worksheet was assembled, along with technology-specific Excel worksheets that perform equipment sizing and economic calculations for each APC system.

NO_x Control Estimate

NO_x control technology design and cost algorithms are based on research conducted for the EPA Acid Rain Division (ARD) (now the Clean Air Markets Division), the EPA Office of Research and Development (ORD), and the U.S. Department of Energy (DOE) National Energy Technology Laboratory (NETL) (Frey and Rubin, 1994). Design parameter calculations for SCR, SNCR, and NGR are taken from the Integrated Air Pollution Control System (IAPCS) model, Version 5.0 (Gundappa et al., 1995).

SCR capital cost components are based on algorithms developed for DOE (Frey and Rubin, 1994) as part of the Integrated Environmental Control Model (IECM).¹ For SNCR and NGR total capital equipment costs, ARD research was used to update IAPCS methodology. The ARD cost data used to update IAPCS are presented in the following:

¹ IECM is a computer-modeling program that performs a systematic cost and performance analysis of emission control equipment at coal-fired power plants. It is developed for the U.S. Department of Energy by Carnegie Mellon University and is available at <http://www.iecm-online.com> (accessed February 13, 2009).

- “Cost Estimates for Selected Applications of NO_x Control Technologies on Stationary Combustion Boilers and Responses to Comments,” (EPA, 1998) and
- “Investigation of Performance and Cost of NO_x Controls as Applied to Group 2 Boilers,” (EPA, 1997).

Low NO_x burner technology (LNBT) total plant costs are based on algorithms presented in another ARD report (EPA, 1996). The cost estimates presented in the ARD reports are being used in the NO_x-related rulemaking and have been reviewed by stakeholders associated with the rulemaking process. O&M cost algorithms for all technologies use IAPCS equations from IAPCS 5.0 (Gundappa et al., 1995). Operating costs are estimated in the workbook based on simplified material balances calculated within CUECost based on the inputs supplied by the user. The ultimate coal analysis, including weight percent sulfur, carbon, hydrogen, oxygen, nitrogen, moisture and ash, serves as the primary input for the combustion calculations performed by the worksheet. The resulting gas flow is the basis for the remaining material balance calculations.

In this manual, the default values for NO_x control devices were generally adopted from Integrated Planning Model (IPM)/IECM models (the IPM model can be found at <http://www.epa.gov/airmarkt/progsregs/epa-ipm/index.html>). For NO_x control technologies, CUECost results were compared to cost data reported by the EPA ARD (EPA, 1997; EPA, 1998) for NO_x controls applied to utility boilers and Chapter 5 of the IPM manual. The ARD reports are based on an EPA national database of boilers. Using CUECost’s default values for Retrofit Factor, General Facilities, etc., should produce capital cost results that are the same as or very close to the results that would be produced by the IPM source algorithms.

Particulate Matter Control Estimates

The particulate matter control technology cost estimates are based on IECM model constructed by Carnegie Mellon University (CMU) for the DOE (Berkenpas et al., 1999). This model was constructed based on a combination of theoretical equations for Electrostatic Precipitator (ESP) sizing. The theoretical equations were modified to incorporate the empirical data obtained from a series of ESP vendors for installations firing different coals. This framework taken from the CMU model served as the basis for the CUECost ESP design portion of the worksheet. The CMU worksheet was based on 150 to 200 actual installations firing a wider variety of fuels. Operating costs are calculated using the inlet-flow rate-versus-expected-power-consumption algorithms. Maintenance costs are calculated as a percentage of the installed equipment cost.

Fabric filter (FF) costs were also based on a set of cost equations developed by Berkenpas et al. (1999) to relate the FF size [calculated as a function of the volumetric flue gas flow rate times the air-to-cloth ratio (A/C) selected by the user] and the FF inlet flue gas volumetric flow rate to determine the expected cost for the installed system (Berkenpas et al., 1999). Operating costs are calculated using the inlet-flow rate-versus-expected-power-

consumption algorithms. Maintenance costs are calculated as a percentage of the installed equipment cost.

SO₂ Control Estimates

For the FGD technologies, cost-versus-capacity equations were based on the historical database (Keeth, 1991) of actual equipment costs incurred during Phase 1 of the utility Clean Air Act compliance programs, budgetary quotations for components as received from vendors during early 1998, and cost data obtained from industry database programs (Srivastava, 2000). These parametric equations serve as the basis for the FGD system capital costs calculated by CUECost. Operating cost equations were formulated based on the consumption rates estimated in the worksheets by the material balance calculations. A material balance is developed specifically for each FGD system and provides the chemical consumption rates, wastes production rates, and flow rates through process equipment that are used to estimate the system power consumption. Operating labor requirements are based on a formula that relates plant size to the number of operating staff needed to run the FGD equipment, and maintenance costs are calculated as a percentage of the installed costs for the system.

CO₂ Control Estimates

The monoethanolamine (MEA) CO₂ control technology cost estimates are based mainly on a report from the U. S. DOE/NETL (2007). The cost of a bare erected plant was estimated based on 30% of MEA, currently the most practical concentration. For MEA islands, the same total plant cost (TPC) is assumed to occur for KS-1 and MEA solvents. The compressor island cost, however, depends upon the compressor stages and power consumption through a regression of cost and compressor power. In the CUECost design, the details of O&M cost, which are the major concern, are listed for the users. The absorption island of chilled ammonia process (CAP) is estimated to be 97% of the bare erected cost of the same size MEA-type island. Although the pressure of CO₂ out of the reflux drum is significantly higher than from the MEA process, the investment estimation of the compressor island follows the same algorithm as the MEA process, depending upon only the regressed relation of cost-power. Regarding the sorbent injection (SI) option, due to the lack of information for a full scale plant, the bare erected cost for absorption island is estimated with a same size of MEA island total cost. Detailed values for consumption of power, steam and cooling water are given in the engineering calculation section.

Mercury Control Estimates

The effects of existing equipment on mercury reduction were isolated from the effects of powdered activated carbon (PAC) injection on mercury reduction, and new algorithms were developed for PAC injection. The PAC injection algorithms include algorithms based on the results of two full-scale demonstrations (Bustard et al., 2001; Durham et al., 2001; Bustard et al., 2002; Sterns, 2002) as well as algorithms developed from pilot-scale data (EPA, 2000).

Economic criteria supplied by the user are used by CUECost to calculate the capital and annualized costs for the selected APC system. The user has the option to use the default values provided in the worksheet if some of the input data requested are not readily available. For the convenience of users, a levelization calculation worksheet is included in the CUECost workbook. This worksheet provides carrying charge and levelization factor for expenses (O&M) in terms of current dollars and constant dollars.

DEFAULT PLANT CRITERIA

The CUECost workbook includes default values for all input parameters. These criteria are specific to a generic 500 MW coal-fired power plant located in Pennsylvania. The specific design and economic criteria used as defaults are provided in Appendix C for reference. A coal library is also included in worksheet 11.0 so that the user can select a coal similar to that actually burned at the plant if an actual ultimate analysis is not readily available. User has the capability to adjust coal properties as desired to create "user-defined" coal in worksheet 11.0. The coal information was retrieved from the DOE Coal Sample Bank and Database at <http://datamine.ei.psu.edu/index.php>.

RESULTS

The CUECost workbook provides rough-order-of-magnitude (ROM) cost estimates ($\pm 30\%$ accuracy) for a wide variety of APC technology scenarios. Cost estimates for different combinations of control technologies can easily be compared in the results summaries presented in five parallel columns on the worksheets. Examples of the input sheets are shown in Appendix E.

CUECost is designed to produce ROM estimates for a wide range of plant sizes and coal types. However, appropriate ranges of plant size and operating conditions have been established based on the limits to the database used to construct the cost-versus-capacity algorithms. Range limits are provided in the worksheet for each input supplied by the user. The major criterion limitation for CUECost is the plant size range. Algorithms are based on the assumption that equipment options will be installed at a facility ranging from 100 to 2000 MW in net capacity. All other criteria are limited only by their technical validity. The suggested technical limits for each criterion are provided in the worksheets when applicable.

It is expected that this document will be useful to a broad audience, including: (1) individuals responsible for developing and implementing SO₂, NO_x, PM, Hg, and CO₂ control strategies at sources, (2) state authorities implementing pollution control programs, and (3) interested public at large. Moreover, persons engaged in research and development efforts aimed at improving cost-effectiveness of air pollution control technology applicable to coal-fired plants may also benefit from this document.

Note that the cost estimates provided in this study and generated by CUECost are dependent upon the various underlying assumptions, inclusions, and exclusions utilized in developing them. Actual project costs will differ and can be significantly affected by factors such as changes in the external environment, the manner in which the project is implemented, and other factors which impact the estimate basis or otherwise affect the project. Estimate accuracy ranges are only projections based upon cost estimating methods and are not guarantees of actual project costs.

EPA policy is to express all measurements in EPA documents in metric units. Values in this document are given in British units for the convenience of the engineers and other technical staff accustomed to using the British system. The following conversion factors presented in Table 1 can be used to provide metric equivalents.

Table 1. British to Metric Conversion Factors

Abbr.	British Unit		Conv. Factor	=	Abbr.	Metric Unit
ac	acre	×	0.405	=	ha	hectare
Btu	British thermal unit	×	0.252	=	kcal	kilocalories
°F	deg. Fahrenheit - 32	×	0.5556	=	°C	degrees Centigrade
ft	feet	×	0.3048	=	m	meters
ft ²	square feet	×	0.0929	=	m ²	square meters
ft ³	cubic feet	×	0.02832	=	m ³	cubic meters
ft/m	feet per minute	×	0.00508	=	m/s	meters per second
ft ³ /m	cubic feet per minute	×	0.000472	=	m ³ /s	cubic meters/second
gal	gallons (U.S.)	×	3.785	=	L	Liters
gpm	gallons per minute	×	0.06308	=	L/s	liters per second
gr	grains	×	0.0648	=	g	grams
gr/ft ³	grains per cubic foot	×	2.288	=	g/m ³	grams per cubic meter
hp	horsepower	×	0.746	=	kW	kilowatts
in.	inches	×	0.0254	=	m	meters
lb	pounds	×	0.4536	=	kg	kilograms
lb/ft ³	pounds per cubic foot	×	16.02	=	kg/m ³	kilograms/cubic meter
lb/h	pounds per hour	×	0.126	=	g/s	grams per second
mi	miles	×	1609	=	m	meters
psi	pounds per square inch	×	6895	=	Pa	Pascals (Newton/m ²)
rpm	revolutions per minute	×	0.1047	=	rad/s	radians per second
scfm	standard (60 °F) cubic feet/minute	×	1.6077	=	nm ³ /h	normal cubic meters/h
t/h	short tons per hour	×	0.252	=	kg/s	kilograms per second

GETTING STARTED

HARDWARE AND SOFTWARE REQUIREMENTS / INTERNET ACCESS

The CUECost workbook is written in Microsoft Excel 2003 format. The hardware and software requirements for CUECost and User's Manual are listed below:

Computer Hardware:	500 MHz Processor, 256 MB RAM and 100 MB hard drive Internet access required for down-loading the worksheet from web site
Operating System:	Windows 2000 or higher
Memory Requirements:	3 MB on hard drive for download of CUECost workbook
Installation Requirements:	Download from EPA's Technology Transfer Network (TTN) web site: http://www.epa.gov/ttn/catc/products.html Search for CUECost under heading: Software (Executables & Manuals) Download to hard drive
Commercial Support Software Required:	Microsoft Excel 5.0 or higher for Workbook Microsoft Word 6.0 or higher or Adobe Acrobat Reader (latest version available for free at: http://get.adobe.com/reader/) for User's Manual

GETTING STARTED

After accessing the workbook via the EPA web site and storing the files on the user's hard drive (note that the files may have to be decompressed using "WinZip"), the User's Manual may be called up in WordPerfect or Acrobat, depending upon the format downloaded, and then printed out for easy access. Each user should read the user's manual to become familiar with how CUECost works and where various input and technical data are provided within the workbook. After reviewing the user's manual, the user should then call up the workbook as an Excel file and begin review of the worksheets contained therein. The file will

be active when called up from the web site (after decompression). The user should go to the home site (cell A1) on the first sheet of the workbook to begin.

NOTE: The CUECost workbook can be modified by the user. To ensure its integrity, a copy of the original worksheet should be saved in a separate file in a new directory and all other copies saved under different file names.

The default values provided in the worksheet will allow the user to immediately run a test case and print output sheets to test the existing printer setup routine. Familiarity with Excel worksheet software is required to modify the workbook to correct printing problems.

The input requirements for the worksheet are itemized in the section titled "Input and Output Options" of this user's manual. The user should first obtain the necessary input data for all cases to be evaluated. Up to ten cases can be run simultaneously for direct on-screen comparison of results. Up to twelve site-specific coal analyses can be added to the ten columns available in the coal library for use in any series of estimating runs. This file can then be saved for use in the future. The existing default values can be deleted by entering values in the library cells, and then saving the new file for future use under a different file name. The input cells are colored blue for identification by the user.

When running the workbook for the first time, it would be best to save your input data to a separate file on a regular basis. The worksheet provides the capability to select from a variety of system options, picking alternate control technologies and combinations of the component options provided.

WORKBOOK LAYOUT AND METHODOLOGY

WORKBOOK LAYOUT

Figure 1 provides the basic layout of the various sheets currently included in the CUECost workbook. The following descriptions apply to the individual worksheets.

Worksheet Menu

Upon opening CUECost, the user can enter the menu by clicking the “Menu” tab. The menu provides easy links to specific control technologies. The user can also use the CUECost toolbar. The toolbar offers quick access to many functions while working in a particular worksheet.

This worksheet menu provides the primary user interface and basic instructions on how to proceed. The interface consists of a series of buttons the user selects based on the technology cost estimates desired and the part of the workbook to be reviewed at that time. Note that when the user selects a specific technology for evaluation, the inputs for preceding technologies must be fulfilled according to the selected air pollutant control sequence. For example, if both selective catalytic reduction (SCR) and limestone forced oxidation (LSFO) technologies are selected; inputs for SCR must be completed prior to the inputs for LSFO.

Worksheet 1.0 = General Input – This worksheet contains three subsections: Economic Factors, Power Generation Technology and APC Technology Choices (only for pulverized coal application). Through this worksheet, the user constructs the basis for all pollutant control technology-related estimates. The various columns in this worksheet are described below:

- Column B provides a text description of the cells in each row.
- Column C defines the units that should be used for the input to the cells in each row.
- Column D supplies a suggested range of input values based on technical limits or worksheet validity limitations.

- Column E is a listing of the default values included in the worksheet.
- Columns F through O provide entry points for values specific for up to ten simultaneous case evaluations.

Worksheet 2.0 = Input Summary - This sheet summarizes economic input, power generation technology choices, and air pollutant control technology choices specified in the general input worksheet.

Worksheet 3.0 = Output Summary – This worksheet summarizes all APC technology outputs from a specific technology evaluation. A filter function is provided in this worksheet for the user to select the preferred outputs easily.

Worksheet 4.0 = Power Generation – This worksheet is specifically designed to evaluate the investment and O&M costs of a specific type of power generation technology such as subcritical, supercritical, ultra-supercritical, integrated gasification combined cycle (IGCC), and Oxyfuel.

Worksheet 5.0 = NO_x Control - NO_x calculations are completed on this worksheet. The results of the combustion calculations provided in worksheet 1.0 “Constants_CC” are used to calculate the material balance for the NO_x systems. These values are then used to calculate the expected costs for the various cost areas using algorithms developed for the CUECost workbook.

Worksheets 6.0 = PMF, 7.0 = SO₂, 8.0 = Hg, 9.0 = CO₂ Control Technologies - These worksheets perform the same function as Sheet 5.0 for the other APC technologies.

Sheet 10.0 = Levelization Calculations – This worksheet is specifically designed for the user to calculate carrying charges and non-carrying expenses.

Worksheet 11.0 = Constants_CC - This worksheet contains range name definitions, tables of constants used by the workbook (such as the molecular weights of compounds), and other macros used by the CUECost workbook. This worksheet also contains the coal library and the combustion calculation sequence used for all of the material balances performed in the other process-specific worksheets.

In general, the methodology employed in the workbook for cost development follows the format used by the IAPCS model (Gundappa et al., 1995), providing installed capital and operating costs for the selected technologies. The calculation sequence takes advantage of the vertical arrangement of the worksheet. A series of tables presents the equations (and all variables used in these equations) contained in each cell and the units of the calculated results. Descriptive material is included in the documentation to define the purpose and method employed within various subsections of the worksheets.

METHODOLOGY

The calculation sequence used in the worksheets to estimate capital and annualized costs is summarized in the following material. Additional details regarding the specific equations and interrelationships between sections of the worksheets can be found in this documentation provided in Appendix D. The worksheet design will accommodate the addition of alternate APC technologies by inserting new worksheets for system cost estimation and technical calculations that will use the common input sections and common economic calculations. The cost worksheet allows the user to select the technologies of interest and calculates the associated costs for each control system based on the data that the user enters to define site-specific conditions. Figure 2 is the logic diagram for the workbook and illustrates how the capital and annualized costs for APC equipment are calculated. The methodology and the calculation sequences used by CUECost are described below in the following material.

Step 1

This step begins with input worksheets and can be split into two sub-steps.

Sub-step 1

The user is first asked to select the "1.0 General Input" worksheet. This worksheet provides a general description of the power plant and desired combination of APC technologies. Following the initial process selection, the user enters the necessary technical parameters specific to the project. Default values are provided for all inputs. The inputs are separated into the following distinct sections:

- Economic Factor

Inflation adjustment factors are used for cost adjustment from algorithm development years to current cost basis year. User can select either gross domestic product (GDP) price deflator or chemical engineering cost index (Chem Index) for cost adjustment. GDP price deflator can be obtained from the Bureau of Economic analysis, and the Chem Index can be obtained from the journal of Chemical Engineering. Carrying charges (current dollars, constant dollars, first year constant dollars and first year current dollars.) and levelization factors for expenses (current and constant dollar based on a 30-year lifecycle of plant) follow the definition in EPRI TAG Technical Assessment Guide (EPRI TR-102276-V1R7 volume 1, 1993). For the convenience of user, carrying charges and levelization factors for expenses can be calculated by clicking the calculator link. Once the inputs in worksheet 10.0 are given by the user, the outputs are automatically sent back to worksheet 1.0.

- General Plant Technical Inputs (boiler operation, coal analysis, excess air, etc.)

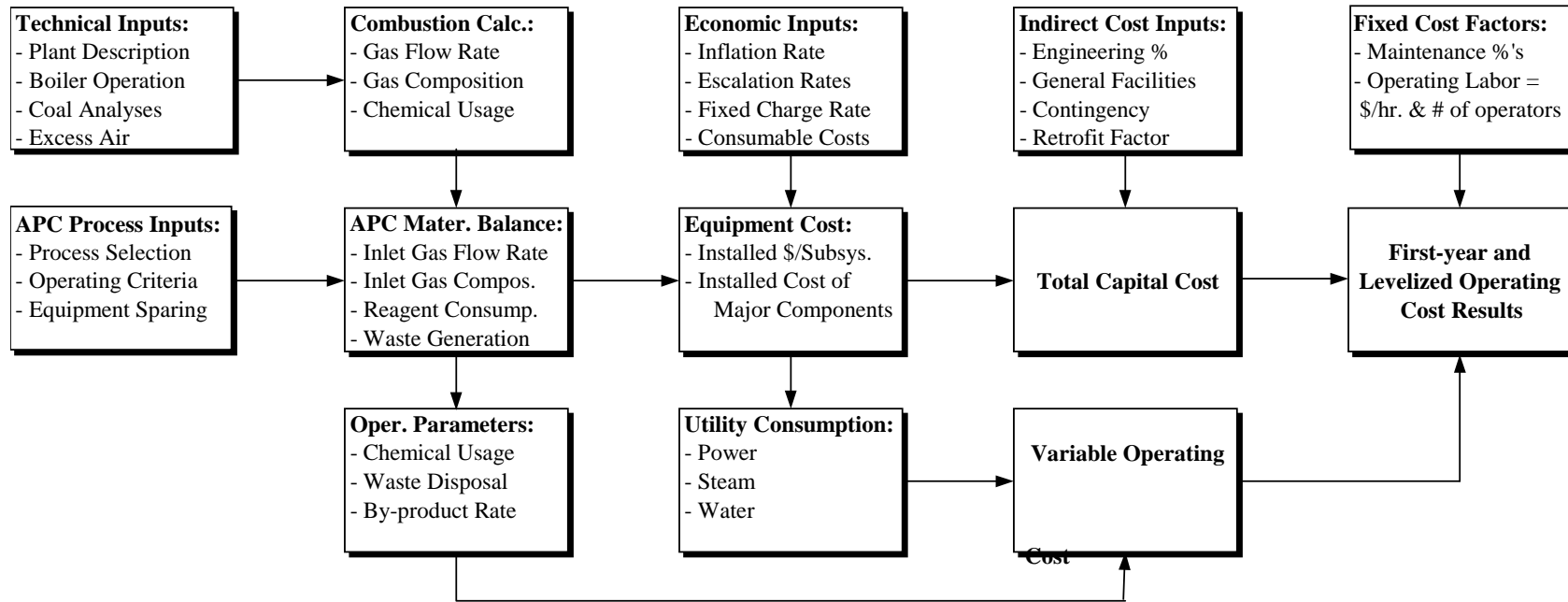


Figure 2. CUECost Logic Diagram

- For the power generation technology selection, the user can select sub-critical, super-critical, ultra super-critical, IGCC, or oxyfuel. When IGCC or oxyfuel power generation is selected, there will be no control technologies related to them in this version.
- APC Technology Choices(NO_x/SO₂/PM/Hg/CO₂)
This section is of importance because it tells the system what plant configuration is desired. As aforementioned, there will be no technology selected when IGCC or oxyfuel power generation technology is selected.

Sub-step 2

After users complete common input, the user then can access a specific control technology as described in the APC Technology Section for process specific input, fully exploiting the convenience of the toolbar attached to this workbook. Default values are provided for all inputs.

Step 2

After the user has entered the technical inputs, the workbook performs the combustion calculations in the Constants_CC worksheet. The flue gas flow rate and composition are calculated in this step. The results of these calculations are summarized in the Constants_CC worksheet.

Step 3

Using the results of the combustion calculation and the APC-specific technical inputs, the necessary material balance calculations are performed. Reagent consumption and waste generation are calculated based on the inlet gas flow and composition (see APC technology worksheets).

Step 4

Following the calculation of the material balance, the equipment costs associated with the specific equipment areas (APC worksheets) are calculated. The largest equipment components for each area [absorber, induced draft (ID) fan, etc.] are broken out and estimated separately. All capital costs are installed costs (i.e., they include all costs associated with the installation of the subsystem or component). These installation expenditures include the costs for the following:

- Earthwork
- Concrete
- Structural steel
- Piping
- Electrical

- Instrumentation and controls
- Painting
- Insulation
- Buildings and architectural.

Costs for demolition are treated as an input, assuming that the user can provide the expected costs for any demolition that might be required at a specific site. The items listed above, when added to the bare equipment cost, are equivalent to "A" in the calculation sequence for the capital cost shown in Table 2.

Table 2. Total Capital Requirement Calculation Method

Installed Process Capital Cost	=	A
General Facilities at % of A	=	B
Engineering and Home Office Fees at % of A	=	C
Contingency at % of (A + B + C)	=	D
Total Plant Cost (TPC)	=	A+B+C+D
Total Cash Expended (TCE)	=	TPC x Adjustment Factor*
Allowance for Funds During Construction (AFDC)	=	AFDC % (input) x TPC
Total Plant Investment (TPI)	=	TCE + AFDC
Preproduction Costs	=	F
Inventory Capital	=	G
Total Capital Requirement (TCR)	=	TPI + F + G

* Adjustment Factor is based on the years of construction, the inflation rate, and the escalation rate. The factor reduces the cost of the capital investment due to the purchase of components prior to the completion of the construction period, allowing the TCR to be expressed in a single-year dollar value.

Step 5

Adding the costs listed above to the uninstalled bare equipment costs results in the total direct field cost for the installed equipment (APC worksheets). The installed equipment costs (bare equipment cost multiplied by an installation factor composed of various cost accounts listed above--earthwork, steel, piping, etc.) for each component include the typical indirect field costs, such as field staff and legalities, craft fringes and insurance, temporary facilities, construction equipment and tools, and an allowance for start-up and testing. Allowances for taxes are also included in the final installed cost for each subsystem. The Total Installed Cost then serves as the basis for the calculation of the engineering and general facilities cost components and the contingency cost associated with the project capital cost. Escalation of the capital cost is then performed using the GDP Index or CE Index (see Economic Indicators found on the last page of each issue of the *Chemical Engineering* magazine) for the year selected by the user as the basis for the cost estimate.

For most equipment areas and components, a cost algorithm is supplied to relate installed component cost to the component capacity. The worksheet was constructed to allow the user to generate cost estimates for units ranging from 100 MW to 1000 MW and for facilities firing almost any coal.

Step 6

In addition to the equipment costs, the APC worksheets also calculate operating parameters (chemical usage, waste disposal, byproduct rate, etc.) after the calculation of the material balances (APC worksheets). The usage and production rates serve as the basis for the calculation of the variable operating cost components. The workbook uses the operating parameters and the calculated utility consumption (electrical energy, steam, water, etc.) to calculate the variable operating costs. The annualized cost calculation method is summarized in Table 3.

Step 7

Finally, the total capital and operating costs are used to calculate the levelized constant dollars and also first-year annualized current dollars. Operating costs belong to the non-carrying charge category. Operating costs will be levelized at the 30-year level (L30) at constant dollars. Both of these costs (for both capital and operating components) are represented in absolute (\$/year) and normalized terms (i.e., mills/kWh or \$/kW). These costs are summarized in the summary worksheet for direct comparison of case cost estimates and printing of output summaries.

Table 3. Annualized Cost Calculation Method

Fixed O&M Costs			
Operating Labor	=	Labor Rate × 8760 × Number of Operators Added	= A
Maintenance Labor/Materials	=	Maintenance Factor × Installed Capital Cost	= B
Administrative/Support Labor	=	0.3 × (Operating Labor + Maintenance Labor)	= C
Variable Operating Cost			
Chemicals	=	Chemical Cost × Consumption Rate/Year	= D
Solids Disposal	=	Waste Disposal Cost × Waste Production Rate/Year	= E
Water Cost	=	Water Cost × Water Consumption Rate	= F
Power	=	Power Cost × Power Consumption Rate	= G
Steam	=	Steam Cost × Steam Consumption Rate	= H
Carrying Charges			
Carrying Charges	=	Total Capital Requirement × Carrying Charge Rate	= I*
Annualized Constant Cost			
	=	(A+B+C+(D+E+F+G+H)×Capacity Factor)×L(30) +I	=
First Year Current Cost	=	(A+B+C)+(D+E+F+G+H)×Capacity Factor +I	=

*When calculating levelized/annualized constant cost, I =total capital requirement (TCR) × constant \$ carrying charge rate.

When calculating first year current cost, I =total capital requirement (TCR) × first year current \$ carrying charge rate.

L(30), levelization factor for 30 year service life, can be calculated with worksheet 10.0. Results are automatically sent back to worksheet 1.0.

INPUT AND OUTPUT OPTIONS

INPUT DATA

The worksheet “1.0 general inputs” starts with default, inputs, represented with “D”, in worksheet 1.0. The user can change the “D” to its defined value for a site-specific estimate. Input cells are colored blue for highlighting purposes. Each column is specific to an individual case. Duplicate data for each case can simply be copied over into the remaining columns rather than entered individually for each case.

The general input worksheet is divided into various sections for clarity. The detailed input requirements are listed below with a brief description of the content of each:

1. Economic Factors – These economic data define the basis for the cost estimates that are produced, including the basis year, inflation rates, escalation rates, capital carrying charges, non-carrying expense (O&M), operating labor rates, chemical costs, and utility costs. The economic factors apply to all the control technologies.
2. Power Generation Technology – These criteria define the operating conditions at the facility under investigation. Fuel characteristics, heat rate, location conditions, etc., are requested in this section. These data are then used as the basis for the combustion calculations and definition of the plant ambient conditions.
3. Air Pollution Control system definition – This is a section of the utmost importance where the user can select from among various APC technology configurations for a specific site. Each specific site can contain one or all of the APC subsystems.
 - 3.1 Nitrogen Oxides Control technology – All data required to define the NO_x control system are requested in this section. The user also selects the type of control system that is desired: selective catalytic reduction (SCR), selective non-catalytic reduction (SNCR), natural gas reburn (NGR) technology, or low-NO_x burner technology (LNBT) [including low NO_x burners (LNB) for pulverized coal boilers, and low-NO_x concentric firing systems (LNCFS) for tangentially fired boilers].

- 3.2 Particulate Matter Control Technology – All data required to define the particulate matter control system are entered in this section. The user also selects the type of control system that is desired, ESP or a FF, and what type of fabric filter is selected.
- 3.3 Sulfur Dioxide Control Technology – This section provides a series of inputs that define the operating conditions for the scrubber system. In this section the user can define conditions that are specific to vendor data, or the default values can be used to determine the generic costs for the FGD system. The option to use a dibasic acid (DBA) additive is also provided. The DBA acts as a buffer in the SO₂ absorption reaction, potentially reducing the operating costs for the FGD system and improving performance at some sites.
- 3.4 Lime Spray Dryer FGD Process Definition – The data inputs on this worksheet are similar to the inputs for the LSFO worksheet. Once again the process operating conditions are defined for each case being considered.
- 3.5 Mercury Control Technology – All data required to define the Hg control system are requested in this section. The user selects the type of sorbent that is desired: enhanced PAC (EPAC), PAC, or other. The user also specifies the design and operating conditions of the pulse-jet fabric filter (PJFF) downstream of PAC if desired.
- 3.6 Carbon Dioxide Control Technology – All data required to define the CO₂ control system are input in this section. The user selects the type of process: MEA, CAP, or SI.

OUTPUT OPTIONS

The input values are then summarized in the Input Summary worksheet. The Output Summary worksheet compiles the results generated by the technology-specific worksheets. The tables are constructed for use in printing the output sheets. A summary table is also available in the section of Summary of Emissions and Generation. The output summary table primarily provides the cost estimates generated for all of the control technologies selected for each case. More detailed breakdowns of each technology cost estimate are also generated in the technology-specific worksheets to identify the components of the estimates. To obtain outputs from the workbook, the user can return to the General Input Sheets, ensure that the workbook has been recalculated by pressing the F9 button, and then click on the Print buttons provided for printing. The user can also enter any worksheet of interest and click on the print icon. The workbook is set up to automatically print all of the cost related material in each worksheet.

The workbook also allows the user to select any specific portion of an individual worksheet that is of interest and print out that material only. A specific range can be selected and that section printed using the standard Excel methodology.

WORKSHEET VALIDATION

The CUECost workbook was constructed to allow the user to have the maximum flexibility to modify it to generate site-specific cost estimates without requiring an extensive amount of input data. The technology worksheets were developed using different sets of cost and design data. The basis for each set of parametric design and cost equations is described in the following section.

FGD WORKSHEETS – LSFO AND LSD TECHNOLOGIES

The equipment design parameters and cost data are based on a combination of vendor quotes and a historical database of installed power projects (Keeth et al., 1991). Cost-versus-capacity curves were constructed based on this historical information combined with vendor quotations from both installed FGD systems and budgetary quotes received specifically for this CUECost project. Many of the sources of information that were used in this development of the FGD system costs are not available to the public due to the proprietary nature of the information and the project-specific sensitivity of the cost data.

This equipment cost database was assembled based on the experience gained at FGD installations for 10-15 plants ranging in size from 300 to 2000 MW. Equipment cost data is produced by compiling data taken from these 10 to 15 actual installations, vendor quotations for construction contracts, and budgetary quotations obtained in 1998 specifically to support this CUECost project. The budgetary quotes for large equipment items were received from one to six vendors depending on the component. The accuracy of the CUECost is validated by comparing the results generated by the CUECost model to published cost data for many of the Phase 1 FGD systems installed in response to acid rain regulations. The validation of the data used in the development of these algorithms is described in Appendix D.

The cost estimates from CUECost were compared to the results generated by other models, including the comparison to the Electric Power Research Institute's (EPRI) FGDCOST model (Keeth et al., 1991). EPRI's FGDCOST model has been used throughout the utility industry

for the last seven years and has demonstrated its ability to estimate site-specific costs well within ROM accuracy requirements. The CUECost estimates were found to agree well with the results generated by the FGDCOST model when allowance was made for the changes in the technology that have occurred since the FGDCOST model was constructed, the escalation of costs, and the reduced level of design data that is required by the CUECost workbook.

CUECost was also used to calculate cost estimates for many of the Phase 1 FGD systems. Actual installed cost data have been published in various sources for these systems. These data were compared to the estimates generated by the CUECost workbook and CUECost reproduced these actual costs within an accuracy of $\pm 12\%$. Table 4 provides the results of this comparative analysis for previously installed FGD systems.

Table 4. CUECost-FGD Cost Comparison to FGDCOST by EPRI for Phase 1 Acid Rain Installations

Unit	Capacity, MW	Sulfur Content, %	Removal Efficiency, %	EPRI, \$/kW	CUECost, \$/kW	%Difference
Petersburg	657	3.50	95	317	291	-8.20
Cumberland	2600	4.00	95	200	187	-6.50
Conemaugh	1700	2.80	95	195	179	-8.20
Ghent	511	3.50	90	215	229	+6.5
Gibson	668	3.50	91	247	218	-11.70
Bailly	600	4.50	95	180	196	+8.9
Milliken	316	3.20	98	348	362	+4.0
Navajo	2250	0.75	92	236	213	-9.75

PARTICULATE MATTER CONTROL WORKSHEET

The particulate matter control sizing equations were based on previously published correlations developed by CMU (Berkenpas et al., 1999). This development process is described in Appendix D. The CMU model was constructed using design information supplied by multiple vendors.

The CMU model used a modified version of the Deutsch-Anderson equation (Edgar, 1983) to relate removal efficiency to collection area and gas flow rate for various coals as part of the ESP sizing calculations. The original Deutsch-Anderson equation was found to be inaccurate for removal efficiencies above 95%. Various empirical models were developed to overcome this inaccuracy, and the CMU model chose to use the White version (White, 1977) of the modified Deutsch equation provided below (Eq. 1):

$$h = 1 - \exp \{-A/V H w_k\}^k \quad (\text{Eq. 1})$$

where

h = collector removal efficiency

A = collector area, ft²

V = volumetric flue gas flow rate, actual cubic feet per minute (acfm)

w_k = precipitation rate parameter

k = constant varying with coal type.

The w_k and k values used in the CMU ESP sizing equations (Berkenpas, 1999) were correlated with the calculated total ash resistivity (based on the ash analysis provided by the user or the default database in worksheet 11.0), and separate k curves were developed for groups of coals that have similar sulfur content. The modified sizing worksheet provides the expected specific collection area (SCA) for the ESP, and a new set of cost equations was developed to relate the ESP size (calculated from the SCA and the ESP inlet flue gas volumetric flow rate) to the expected cost for the installed system.

The ESP equations provided in the CMU model were reviewed and compared to the expected ESP sizes in terms of SCA, evaluating the various types of coals listed in Table 5. The "Raytheon" SCA data provided in Table 5 were calculated using a series of parametric equations developed by Raytheon Co.² These equations were derived from SCA data for utility coal-fired installations over the past 25 years obtained by Raytheon Co. and incorporated into a proprietary model used for confirmation of vendor data and specification preparation. As can be seen in Table 5, the CUECost workbook calculates SCA values that are within ±12% of the values generated by the Raytheon model.

The costs generated by CUECost were compared to the current IAPCS results for the same plant sizes and coals. The results were within 30% of the IAPCS cost estimating model (Gundappa et al., 1995) over a range of SCA values from 300 to 600. The FF cost algorithms (one for pulse jet design and one for reverse gas) were developed from 10 to 12 firm price quotations (obtained during 1992-1997) for each FF design. The coal-fired boilers ranged in size from 50 to 500 MW.

² The Raytheon database is proprietary.

Table 5. Comparison of CUECost ESP Sizing Estimates with Raytheon Model

Coal Type	Sulfur Content, %	Removal Efficiency, %	Raytheon SCA*	CUECost SCA*	% Difference
Indiantown	1.09	99.4	385	429	+11.43
WV-EPRI	0.66	99.2	418	375	-10.29
Low S Bituminous	0.97	99.4	403	424	+5.21
Keystone	1.09	99.3	393	386	-1.78
India	0.5	99.9	965	883	-8.50
Logan, WV	0.89	99.7	569	502	-11.78
ND Lignite	0.94	99.4	376	411	+9.31
UT-EPRI	0.53	99.5	446	442	-0.90
UT-Alternate	0.66	99.6	435	482	+10.80
Rosebud, MT	0.56	99.5	482	459	-4.77
WY-PRB	0.37	99.3	558	558	0
Test Coal	2	99.1	287	283	-1.39
Pitts 8	2.13	99.2	272	285	+4.78
Carneys	2	99.1	288	281	-2.43
TX Lignite	1.16	99.8	549	549	0
OH Alternate	4.7	99.6	247	259	+4.86
IL #6	3.25	99.5	276	261	-5.43
Armstrong, PA	2.6	99.3	277	274	-1.08
Jefferson, OH	3.43	99.6	321	326	1.56

* SCA = square feet of plate area per 1000 actual cubic feet per minute of flue gas flow

NO_x CONTROL WORKSHEET

For NO_x control technologies, CUECost results were compared to cost data reported by the EPA Acid Rain Division for NO_x controls applied to utility boilers (Khan and Srivastava, 2004), and Chapter 5 of the IPM manual.³ The Acid Rain Division reports (EPA, 1997; EPA, 1998) are based on an EPA national database of boilers (Khan and Srivastava, 2004). The 1990 Clean Air Act Amendments required the EPA to examine NO_x control technology costs, and the resulting Acid Rain Division studies (EPA, 1997; EPA, 1998) were used and reviewed during the rule-making process. A comparison was made for four cases with various boiler types, boiler sizes (100 to 400 MW) and coals burned. The boiler design and operating parameters for each case were input into CUECost to obtain capital and operating and maintenance costs. In some cases the capital cost estimating algorithms in these sources

³ available at: <http://www.epa.gov/airmarkets/progsregs/epa-ipm/index.html> and can be downloaded at <http://www.epa.gov/airmarkets/progsregs/epa-ipm/docs/Section-5.pdf>.

already included provisions for retrofit factor, general facilities, engineering, contingency, and other factors that, using CUECost methodology, are in addition to equipment costs. So it was necessary to adjust for these additional costs in arriving at the equipment cost algorithms for CUECost. For this reason, the algorithms programmed into CUECost may be different from those shown in the IPM report or the Acid Rain Division reports (Khan and Srivastava, 2004). However, using CUECost's default values for retrofit factor, general facilities, etc., should produce capital cost results that are the same as or very close to the results that would be produced by the IPM source algorithms.

Different approaches were taken to verify or validate the costs predicted by CUECost for the various NO_x control technologies. For SCR, SNCR and NGR, design parameters used for the ARD study cases (EPA, 1997; EPA, 1998) were used to calculate preliminary operating parameters and costs with CUECost. Algorithms for SCR in CUECost were compared to the ARD study costs to validate the algorithms. However, the ARD data were incorporated into the algorithms for SNCR and NGR. As a result, the cost comparisons for these technologies were conducted to benchmark the algorithms and evaluate how well they track the ARD data. The percent differences found for the four boiler cases are presented in Table 6. Differences range in magnitude from 0 to 11% for total plant costs and from 0 to 22% for operating and maintenance costs.

SNCR capital costs are determined from the IPM and are documented in Chapter 5 of the IPM manual (<http://www.epa.gov/airmarkets/progsregs/epa-ipm/index.html>), which can be downloaded at <http://www.epa.gov/airmarkets/progsregs/epa-ipm/docs/Section-5.pdf>.

The algorithms used to estimate costs for LNBT in CUECost were taken from an Acid Rain Division study (EPA, 1996). The cost data upon which the algorithms were based represent actual LNBT retrofit cases. The capital cost comparison shows 0% difference, as expected, because the algorithms are based solely on ARD data. A comparison is not presented for operating and maintenance costs because these costs are highly boiler specific.

Table 6. Percent Difference between CUECost and Acid Rain Division Studies (Khan and Srivastava, 2004) for Retrofit Cases *

	Cyclone Fired		Wet Bottom	
			Vertical Fired	Wall Fired
	Midwestern Bituminous		Eastern Bituminous	
	Boiler Size (MW)			
	140	400	100	259
Total Plant Costs				
SCR (50% removal)	4%	0%	8%	-4%
SNCR (50% removal)	8%	0%	12%	4%
NGR (35% removal)	-11%	-7%	-12%	-12%
O&M Costs				
SCR (50% removal)	-12%	-18%	-16%	-22%
SNCR (50% removal)	8%	0%	12%	4%
NGR (35% removal)	-11%	-7%	-12%	-12%

* Note: Percent Difference = (Acid Rain Costs - CUECost Results) x 100 / Acid Rain Costs

MERCURY CONTROL WORKSHEET

Mercury control technologies included in CUECost are co-benefit controls from air pollution control technology used for other pollutants and sorbent-based mercury-specific controls. Mercury control technology cost and performance estimates are determined by algorithms described in U.S. EPA (2003), Staudt, Jozewicz, and Srivastava (2003), Srivastava, Staudt, and Jozewicz (2004), as well as in Appendix F of this manual.

In addition to the impact of sorbent cost on operating cost of mercury control technologies, calculations include estimates of the impact of parasitic load and filter replacement (if a fabric filter is retrofit) and the impact the sorbent may have on fly ash marketability when the sorbent and fly ash are collected in the same PM control device.

Capital cost estimating methodology was made consistent with other technologies with the sole exception that we included Process Contingency for Hg Control in addition to the other cost factors because of the relative newness of Hg control technologies. The user may input a Process Contingency percentage in the Input worksheet or accept the default Process Contingency value of 5%.

Detailed review of mercury control technologies, performance, and future improvements can be found elsewhere (Srivastava et al. 2004; EPA 2003). One of the most promising technologies to control mercury emissions from coal-fired power plants is SI, especially the activated carbon injection (ACI). Table 7 provides the estimated capital and O&M costs of an ACI control system from CUECost.

Table 7. Estimated Costs of ACI Control Systems according to CUECost*

APC Configuration	Capital Cost (2005\$/kW)	O&M Cost (2005\$/MWh)	Hg Removed by Sorbent Injection (lb/yr)	Control Cost (2005\$/lb Hg removed)
ACI +Cold-side ESP	19.41	4.06	240.7	53,380
ACI +Cold-side ESP + Wet FGD System	19.41	4.06	188.9	68,013
ACI+ Dry Scrubber + Fabric Filter	3.17	0.32	290.7	3,844

* Note: 500 MW, Wyoming Powder River Basin (PRB) coal, activated carbon injection, capacity factor = 65%, 80% Hg removal.

CARBON DIOXIDE CONTROL WORKSHEET

Carbon dioxide (CO₂) control technologies included in CUECost are based on MEA solvent, chilled ammonia and sorbent injection. These technologies, as described in Appendix B.6 of this manual, generally include absorption/regeneration island and compressor island. The capital cost of the CO₂ control facility in the CUECost is the lump-sum of the individual costs and given by a regressed equation based on currently available bare erected plant cost from the DOE report (2007). As there is no SI based CO₂ control technology adopted in a power plant, its capital cost is estimated to a comparable cost for MEA process. For all the control technologies, the compressor island cost is regressed on the basis of power consumption of compressors. Gas flow rate and specific variable costs are estimates and determined by algorithms described in Appendix D. Table 8 summarizes the estimated capital and O&M costs of CO₂-related control technologies from CUECost.

Table 8. Estimated Costs of CO₂ Control Technologies with CUECost and IECM model.

	CUECost			IECM
	MEA	CAP	SI	MEA*
Total Plant Cost (TPC), Million \$	350.8	322.1	350.8	232.2
Total Capacity Requirement (TCR), Million \$	388.3	353.1	380.2	273.3
O&M				
Fixed O&M, Million \$/yr	12.3	11.36	12.3	7.3
Variable O&M, Million \$/yr	76.4	42.6	39.0	103.7
\$ (constant)/ton CO ₂	51	34	34	37

Note: estimates are based on a 580 MW plant, firing Illinois bituminous coal. The plant capacity factor is 65% and demands a 90% CO₂ removal efficiency. Capital cost is calculated for the base year of 2006.

*The MEA data was calculated with the IECM model developed by CMU. The IECM program can be downloaded from http://www.iecm-online.com/iecm_dl.html.

VALIDATION SUMMARY

Costs for utility APC systems are site-specific. These costs are subject to change with changes in technology, labor rates, and material costs. The costs estimated by the CUECost workbook come from a variety of sources. With that understanding, one may assume, but it is not guaranteed, that CUECost will produce estimates in the range of accuracy of $\pm 30\%$ of the actual cost, which was the goal of the CUECost development. The operating cost estimates are more straightforward than the capital cost estimates, relying more on the accuracy of the input data supplied by the user. The calculation sequences for these estimates have been verified on a cell-by-cell basis during the course of the workbook development. The documentation provided in Appendix F also allows any user to verify a specific calculation sequence that might be in question at some point in the future. The economic calculation methods used have been well established for many years throughout the utility industry, and have been documented in the EPRI Technical Assessment Guide (Ramachandran, 1989).

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APPENDIX A TERMINOLOGY DEFINITIONS, ABBREVIATIONS, ACRONYMS, AND RANGE NAMES

A.1 DEFINITION OF TERMS

Allowance for Funds Used During Construction (AFDC) - Represents the time value of money during the construction period. AFDC is calculated based on the weighted cost of capital, compounded on an annual basis throughout the period, and applied to all funds spent during each year. This cost is added to the Total Cash Expended to obtain TPI. See Table 2 for the use of the AFDC factor. The AFDC factor is input by the user, and is a function of the years of construction and the discount rate.

Ammonia Slip = The un-reacted ammonia that exits an SCR or SNCR process, and exits the stack with the flue gas. Ammonia slip is expressed as a concentration in the exit gas or as a percentage of the mass of ammonia input to the process.

Battery Limits = The boundary limits within a plant used to define the equipment components contained in a subsystem.

Capacity Factor (CF) - Equivalent to the ratio of the total energy output over a time period divided by the total gross energy generating capacity of the unit. Typically the CF is input as the expected average value over the remaining plant life.

Carrying Charge Factor (CCF) - Amount of revenue per dollar of investment that must be collected from customers in order to pay the carrying charges on that investment. The CCF is expressed as a decimal that is multiplied by the original investment to obtain a carrying charge in terms of dollars. The carrying charge rate can be a present value or levelized quantity over a specified period of time (up to the book life), or an annual quantity in a specific year of life. The factor includes the return on debt, return on equity, income and property taxes, book depreciation, rate of return to shareholders, and insurance.

Constant Dollar - Cost estimate presented in terms of the base year dollars without including the impact of inflation over the plant life. However, real escalation is included in the calculation of future year costs. Constant dollar analysis requires the use of a discount rate that does not include inflation.

Contingency - A capital cost included in the estimate to cover the costs for additional equipment or other costs that are expected to be incurred during a project after the detailed design is completed. These are funds that are expected to be spent during implementation of the final project. The contingency is factored as a percent of process capital plus engineering, home office and general facilities.

Current Dollar - A cost analysis that includes the effects of inflation and real escalation. The discount rate used for current dollar analyses is equivalent to the return required to attract investment capital and is equivalent to the weighted average of the return on equity and return on debt.

Engineering and Home Office Costs - Derived as a percentage of the total direct capital cost. This indirect cost includes the costs for an architectural/engineering company and for home office engineering expenses by the user's company. This value typically ranges from 5 to 20% of the Process Capital, with the percentage varying based on the level of complexity for equipment installation (e.g., a new plant might have a value of 5 to 10% while a retrofit might experience engineering costs closer to 15-20%).

General Facilities - Includes costs for items such as roads, office buildings, maintenance shops, and laboratories. The indirect cost for these facilities typically ranges from 5 to 20% of the Process Capital.

Heat Rate - Equivalent to the fuel energy content (Btu) required to produce 1 kWh of electric energy. Fuel energy content is typically based on the higher heating value of the fuel.

Inflation Rate - Equivalent to the rise in prices caused by an increase in the available currency and credit without a proportionate increase in availability of goods and services of equal quality. The inflation rate does not include the effects of real escalation.

Operating Costs - Operating costs for each technology are expressed in terms of both \$/kW-year and mills/kWh. The \$/kW-year costs are considered to be an expression of annual costs and, therefore, include the capacity factor in the calculation. The mills/kWh values are considered instantaneous values, and, therefore, do not include the capacity factor in their calculation.

Present Value (PV) - Monetary equivalent to the amount of money at a point in time other than that at which the amount of money is paid or received.

Process Capital - Total installed cost of all process equipment.

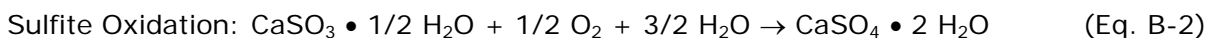
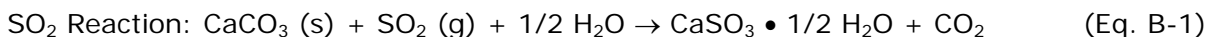
Total Capital Requirement (TCR) - Equivalent to the Total Plant Cost, AFDC, plant startup costs, and inventory capital.

Total Plant Cost (TPC) - Equivalent to the total installed cost for all plant equipment, including all direct and indirect construction costs, engineering, overheads, fees, and contingency.

APPENDIX B TECHNOLOGY DESCRIPTIONS/CRITERIA

B.1 LIMESTONE FORCED OXIDATION DESIGN CRITERIA

In a limestone with forced oxidation (LSFO) system, the flue gas is contacted with slurry containing approximately 15% calcium carbonate and sulfate solids. The aqueous sulfite formed by SO₂ absorption is oxidized to sulfate by forced air injection in the tower recirculation tank to produce slurry with essentially 100% conversion of calcium sulfite to sulfate. The series of chemical reactions that occur in an LSFO absorber and reaction tank is described in Eqs. B-1 and B-2:



The CUECost workbook requires that the user input new values for the slurry recycle rate (Liquid to Gas Ratio = L/G) whenever the SO₂ removal efficiency across the FGD system is changed versus the current 95% removal rate included as the base case default value. Typically the increase in removal efficiency above this 95% level will require significant increases in the recycle rate. A value of 140 gallons/1000 actual cubic feet (L/G) would be typical for a 97% removal system versus the 125 value for a 95% system. Therefore, the pump sizes and power consumption required in the FGD system would increase significantly. Values for the limestone feed rate (stoichiometric feed ratio default = 1.05 moles of CaCO₃ per mole of SO₂ removed) also remain constant with changes in the removal efficiency, but can be modified by the user if additional vendor information is available.

The slurry produced by the FGD system can be thickened and pumped directly to a gypsum stack for final disposal, vacuum filtered or centrifuged for landfill disposal, or washed and dewatered for commercial wallboard production.

The LSFO Process Equipment includes the Reagent Handling and Preparation, SO₂ Control System, and the Byproduct Handling.

Reagent Handling and Preparation includes the following:

- Reagent storage
- Reagent feed
- Ball mill and hydroclones
- DBA acid tank.

SO₂ Control System includes:

- SO₂ removal system
- Absorber tower
- Spray pumps, spray nozzles, associated piping.

Byproduct Handling includes:

- Waste/byproduct handling system
- Thickener system.

ID Fans and Ductwork are:

- Booster fans needed for the system
- Ductwork between components.

Chimney is:

- Cost of replacement chimney and associated foundations.

Support equipment is:

- Electrical support equipment and modifications not included elsewhere.

An alternative design option is provided in the LSFO system to include the addition of DBA. This additive helps to buffer the SO₂ absorption reaction, increasing the available alkalinity in the slurry. Addition of DBA allows the system to be designed with lower recycle rates and potentially a lower limestone feed rate while maintaining the removal efficiency.

Specific design criteria for LSFO are shown in Table B-1. The default values provided in the worksheet are considered typical for operating FGD systems recently installed in the U.S. Reagent costs are typically based on the costs stated in the journal *Chemical Marketing Reporter*.

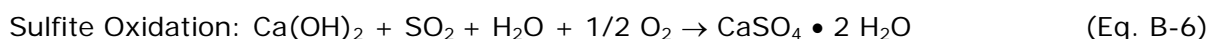
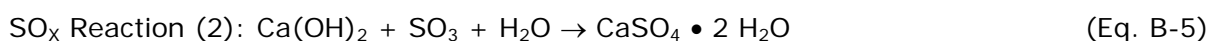
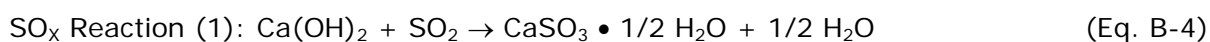
Table B-1. Specific Design Criteria for LSFO

Description	Units	Range	Default
Year equipment placed in service	year		2004
SO ₂ Removal Required	%	90-98%	95%
L/G Ratio	gal / 1000 acf	95-160	125
Design Scrubber with Dibasic Acid Addition? (1 = yes, 2 = no)	Integer	1 or 2	1
Adiabatic Saturation Temperature	°F	100-170	127
Reagent Feed Ratio (Mole CaCO ₃ / Mole SO ₂ removed)	Factor	1.0-2.0	1.05
Scrubber Slurry Solids Concentration	Wt. %		15%
Reheat Air Temperature	°F		440
Pressure	in. H ₂ O		1
Stacking, Landfill, Wallboard (1 = stacking, 2 = landfill, 3 = wallboard)	Integer	1,2,3	1
Number of Absorbers (Max. Capacity = 900 MW per absorber)	Integer	1-6	1
Absorber Pressure Drop	in. H ₂ O		6
Reheat Required? (1 = yes, 2 = no)	Integer	1 or 2	1
Amount of Reheat	°F	0-50	25
Reagent Bulk Storage	Days		60
Reagent Cost (delivered)	\$/ton		\$15
Landfill Disposal Cost	\$/ton		\$30
Stacking Disposal Cost	\$/ton		\$6
Credit for Gypsum Byproduct	\$/ton		\$2
Retrofit Factor			1.3
Maintenance Factor (% of TPC)	%		3%
Contingency (% of Installed Cost)	%		15%
General Facilities (% of Installed Cost)	%		5%
Engineering Fees (% of Installed Cost)	%		10%
Time for Retrofit to use for TCE and AFDC factors	years		2

B.2 LIME SPRAY DRYER DESIGN CRITERIA

In a lime spray dryer (LSD) process the flue gas exiting the air heaters enters a spray dryer vessel. Within the vessel, an atomized slurry of lime and recycled solids contacts the flue gas stream. The sulfur oxides in the flue gas react with the lime and fly ash alkali to form calcium salts.

The chemical reactions associated with the SO₂ removal from the flue gas are provided below (Eqs. B-3 through B-6):



The water entering with the slurry vaporizes, lowering the temperature and raising the moisture content of the scrubbed gas. A particulate matter control device downstream of the spray dryer removes the dry solids and fly ash that did not fall out in the vessel. A portion of the collected reaction products and fly ash solids is recycled to the slurry feed system. The remaining solids are transported to a landfill for disposal.

The CUECost workbook responds to changes in the removal efficiency and any other parameter by using the input values entered by the user and recalculating the material balance on that new basis. No other changes in the worksheet are done automatically in response to changes in parameters. The CUECost workbook does modify the solids recycle rate as the coal sulfur content is modified. The modification is done with a look-up tabulation of recycle values associated with various coal sulfur percentages. A look-up table is embedded in worksheet 11.0 Constants_CC of the CUECost workbook.

The LSD system incorporates five specific equipment areas:

- Reagent handling and preparation
- SO₂ control system
- Byproduct handling
- ID fans and ductwork
- Support equipment.

The Reagent Handling and Preparation includes the following:

- Lime storage and preparation
- Lime slaker.

SO₂ Control System includes:

- SO₂ removal system
- Absorber tower
- Spray pumps, spray nozzles, associated piping.

Byproduct Handling includes:

- If LSD system is installed upstream of existing ESP, this includes modifications to existing ESP due to increased solids handling and gas with more moisture
- Otherwise, SDA and new FF or SDA and new ESP need to be added. Their costs can further be calculated for ESP and FF calculations in the ESP and FF worksheet.

ID Fans and Ductwork are:

- Booster fans needed for the system
- Ductwork between components.

Support Equipment is:

- Electrical support equipment and modifications not included elsewhere

The annual Maintenance (component of the operating cost), additional General Facilities, and Engineering factors provided in Table B-2 are multiplied by the installed equipment capital cost to obtain an estimate of these costs to the utility. The Contingency factor is applied to the total bottom line cost (Equipment Installed Cost plus Site Facilities and Engineering) and represents an estimate of the capital that will be expended but not accounted for in the estimate due to the level of detail included in the system design for this cost worksheet.

Table B-2. Input for LSD and the Default Values for the Inputs

Description	Units	Range	Default
SO ₂ Removal Required	%	90-95%	90%
Is SDA being retrofit upstream of existing ESP? (0 = no, 1 = yes)	integer	0, 1	0
Adiabatic Saturation Temperature	°F	100-170	127
Flue Gas Approach to Saturation	°F	10-50	20
Recycle Slurry Solids Concentration	Wt. %	10-50	35%
Number of Absorbers (Max. Capacity = 300 MW per spray dryer)	integer	1-7	1
Absorber Material (1 = alloy, 2 = RLCS)	integer	1 or 2	2
Spray Cooler Pressure Drop	in. H ₂ O		1
Reagent Bulk Storage (days)	integer		30
Reagent Cost (delivered)	\$/ton		\$65
Dry Waste Disposal Cost	\$/ton		\$30
Retrofit Factor			1.3
Maintenance Factor (% of TPC)	%		2%
Contingency (% of Installed Cost)	%		15%
General Facilities (% of Installed Cost)	%		5%
Engineering Fees (% of Installed Cost)	%		10%
Project Duration (years)	integer		2

B.3 PARTICULATE MATTER CONTROL DESIGN CRITERIA

In a particulate control system, the flue gas exiting the air heaters enters an ESP or FF through the inlet manifold. In an ESP, the particulate matter is electrically charged by the electric fields generated. This charge helps to move the particles to the collecting plates' surfaces, and holds them in place until the collected material can be discharged into the collecting hoppers. ESPs are available in a wide variety of designs and construction materials; collecting plate design, size and spacing; electrode design; etc. These variations in design among vendors are not addressed in this worksheet, and are not expected to drive the final system cost estimates beyond the stated ROM estimate accuracy. The dry fly ash material is typically transferred to final disposal silos by a pneumatic conveying system.

Within the FF, the particulate matter is collected on filter bags suspended vertically within the FF vessel. The particulate matter is physically removed from the gas as it passes through the filter bags, by impacting both the bag fibers and the filter cake that collects on the surface of the bags. Periodically, individual FF compartments are mechanically cleaned by reversing the gas flow or using a pulse jet design that uses pressurized air to force the collected fly ash off the bags and into the collection hoppers. The two design options

(reverse gas and pulse jet) are available as options in the worksheet. The air-to-cloth ratio (square feet of cloth required per 1000 actual cubic feet per minute of flue gas flow) identifies the size of the FF required, quantifying the amount of cloth area required to treat a given gas flow rate. Once again, the ash is typically transferred to the waste silo by a pneumatic conveying system.

The CUECost workbook responds to changes in the removal efficiency and any other parameter by using the input values entered by the user and recalculating the material balance on that new basis. No other changes in the worksheet are done automatically in response to changes in parameters. The model does modify the solids collection rate as the coal ash content is modified.

Specific design criteria associated with particulate matter control are summarized in Table B-3 below:

Table B-3. Inputs for Particulate Matter Control and Its Default Values

Description	Units	Value
<i>Particulate Matter Control</i>		
Outlet Part. Matter Emission Limit	lbs/MMBtu	0.03
Particulate Matter Control Process (1 = Fabric Filter, 2 = ESP)	integer	1
<i>Fabric Filter</i>		
Fabric Filter Type (1 = Reverse Gas, 2 = Pulse Jet)	integer	2
Gas-to-Cloth Ratio	acfm/ft ²	1.8
Bag Life	years	5
<i>Electrostatic Precipitator</i>		
Specific Collection Area (SCA)	ft ² Collecting Plate/ 1000 acfm Gas	Calculated based on ash composition and collection efficiency

B.4 NO_x CONTROL TECHNOLOGY CRITERIA

Four NO_x control technologies are included in CUECost:

- Selective Catalytic Reduction (SCR)
- Non-Selective Catalytic Reduction (SNCR)
- Natural Gas Reburning (NGR)
- Low NO_x Burners (LNB).

The process design criteria and assumptions that serve as defaults within the worksheet are described in the following sections.

B.4.1 Selective Catalytic Reduction Design Criteria

Selective catalytic reduction (SCR) is a post-combustion nitrogen oxides (NO_x) reduction process where NO_x in the flue gas is reduced to nitrogen (N₂) and water (H₂O) using ammonia (NH₃) as a reductant. The reduction occurs in the presence of a catalyst at reaction temperatures between 600 and 750 °F. SCR systems are typically based on one of two designs:

- A hot-side, high-dust SCR where the SCR system is located between the economizer and air preheater
- A cold-side, low-dust SCR where the SCR is typically located downstream of the air heater and particulate control device
- In a variation of this design, the SCR system can be located further downstream, after the flue gas desulfurization (FGD) system (often called a tail-end SCR system).

The CUECost algorithms estimate costs for hot-side, high-dust systems, because hot-side systems have been used on most SCR applications (EPA, 1996).

An SCR system reduces NO_x concentrations in the flue gas using ammonia as the reducing agent in a series of gas-phase reactions in the presence of a catalyst to form nitrogen and water. The chemical reactions for these reduction reactions are provided below:



Small fractions of the ammonia can also be oxidized to alternate forms of nitrogen oxides:



Some of the residual ammonia will also react with trace concentrations of the sulfur oxides in the flue gas in the reactions shown below.



The solids formed in this reaction can contribute to catalyst fouling and contamination of fly ash.

The key operating parameters that affect the performance and, consequently, the capital and operating cost of SCR systems include the allowable NH₃ slip emissions, the space velocity, the NO_x reduction efficiency, and the NH₃/NO_x molar ratio. For SCR systems, these parameters are interrelated, and their values depend on the type of SCR application (high-dust or tail-end) and the desired performance levels. Ammonia slip emissions are controlled by the SCR system design. Typically SCR catalyst suppliers provide a guarantee of 2 ppm over the catalyst life. Since the 2 ppm NH₃ slip is guaranteed at the end of the catalyst's life, the initial NH₃ slip emissions will be very low (<1 ppm). For this reason, ammonia slip does not affect the catalyst volume calculations in CUECost.

The space velocity is the primary parameter used to specify catalyst volume. If the user does not input a value for space velocity, CUECost calculates space velocity based on the NO_x reduction efficiency and the NH₃/NO_x molar ratio. For SCR, NO_x reduction efficiency can range from approximately 60 to 95%, but systems are typically designed to achieve 70 to 90% removal. The NH₃/NO_x molar ratio generally ranges from about 0.7 to 1.0. Ammonia can be injected at a greater than 1:1 stoichiometric ratio to increase NO_x reduction efficiency, but NH₃ slip would also increase significantly.

CUECost estimates capital costs for reactor housing, initial catalyst, ammonia storage and injection system, flue gas handling including ductwork and induced draft fan modifications, air preheater modifications and miscellaneous direct costs, including ash handling and water treatment additions that typically are modified due to the increased concentrations of ammonium salts in the collected fly ash.

Operating and maintenance costs include NH₃, catalyst replacement and disposal, electricity, steam, labor and maintenance costs. Annual catalyst replacement costs are based on the catalyst life. For example, if the catalyst life is 3 years and there are three catalyst sections, then one-third of the catalyst is replaced each year. The catalyst disposal cost reflects the cost of disposing of the spent catalyst. A typical value of 48 lb/ft³ was used for the catalyst density to calculate the mass of the spent catalyst. Default input values for SCR are presented in Table B-4. The default inputs were taken from EPA's ARD studies (EPA, 1996) where available. Unit costs are escalated from 1995 dollars to 2004 dollars using *Chemical Engineering Magazine* cost indices.

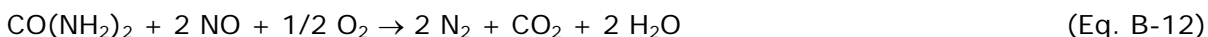
Table B-4. Default Input Parameters for SCR

Description	Units	Range	Default
Inlet NO _x level	lb/MMBtu		calculated
NH ₃ /NO _x Stoichiometric Ratio	decimal	0.7-1.0	0.9
NO _x Reduction Efficiency	decimal	0.60-0.90	0.90
Space Velocity (Calculated if zero)	1/h		0
Time to first catalyst replenishment (years)	integer	2-5	3
Ammonia Cost	\$/ton		400
Catalyst Cost	\$/m ³		5000
Solid Waste Disposal Cost	\$/ton		11.48
Maintenance (% of installed cost)	%		0.66%
Retrofit Difficulty Factor	decimal		1.5
Contingency (% of installed cost)	%		15%
General Facilities (% of installed cost)	%		5%
Engineering Fees (% of installed cost)	%		10%
Mercury Oxidation Rate - bituminous coal (removal if downstream)	%		90.0%
Mercury Oxidation Rate - subbituminous coal	%		0.0%
Duration of Project (years)	integer		2

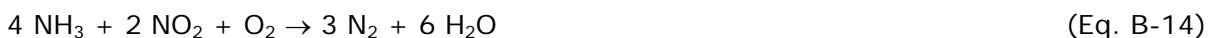
B.4.2 Selective Non-catalytic Reduction Design Criteria

The selective non-catalytic reduction (SNCR) process involves injection of a nitrogen-bearing chemical (usually NH₃ or urea) into boiler flue gases within a prescribed temperature range (typically 1600 to 2000 °F). The NH₃ or urea [CO(NH₂)₂] selectively reacts with NO_x in the flue gas to convert it to N₂. For the NH₃-based SNCR process, either aqueous or anhydrous NH₃ is injected into the flue gas where the temperature is between 1600 and 1900 °F. Most of the NH₃ reacts with NO and oxygen in the gas stream to form N₂ and H₂O. For the CO(NH₂)₂-based SNCR process, an aqueous solution of CO(NH₂)₂ is injected into the flue gas at one or more locations in the upper furnace and/or convective pass. The CO(NH₂)₂ reacts with NO_x in the flue gas to form N₂, H₂O, and carbon dioxide (CO₂). The chemical reactions for this conversion process are not well defined, consisting of a series of dissociation reactions at the elevated gas temperatures in the boiler gas path. The following summary equation describes the overall reaction that is occurring, while the actual reaction mechanism is a long series of dissociation and chemical reactions between various free radicals.

Urea Reaction:



Ammonia Reactions:



CUECost allows the user to select either $\text{CO}(\text{NH}_2)_2$ or NH_3 as the SNCR reagent. The user is asked to specify the NO_x reduction efficiency and the stoichiometric molar ratio of reagent to NO_x . SNCR can achieve NO_x -reduction efficiencies ranging from 30 to 70%. Approximately 50% reduction is typical. The SNCR process requires stoichiometric reagent-to- NO_x ratios of greater than 1:1 to achieve significant NO_x removal. The ratio can range from about 0.5 to 2.5, but will typically fall within the range of 1 to 2. The NH_3 and $\text{CO}(\text{NH}_2)_2$ injection rates are then calculated based on the stoichiometric ratio, inlet NO_x and boiler heat input.

For the $\text{CO}(\text{NH}_2)_2$ -based SNCR process, the user chooses wall injectors, lances, or both. Wall injectors are nozzles installed in the upper furnace waterwalls. In-furnace lances protrude into the upper furnace or convective pass and allow better mixing of the reagent with the flue gas. In-furnace lances require either an air- or water-cooling circulation system. If the user enters values for both wall injectors and lances, then costs include both lances and wall injectors. If wall injectors are to be used alone, then the user enters zero for both the number of lance levels and the number of lances. Similarly, if lances are to be used alone, the user enters zero for both the number of injector levels and the number of wall injectors. CUECost uses input parameters for the number of injectors and lances unless the user wants these parameters to be calculated from the number of levels. If the user inputs zero for the number of injectors and also inputs the number of injector levels, CUECost will calculate the number of injectors. Similarly, if the user inputs zero for the number of lances, the number of lances will be calculated from the number of lance levels. For the NH_3 -based SNCR process, the user can choose either steam or air as the atomizing medium. Based on the user's choice, an annual operating cost for steam or electricity usage is calculated.

The main equipment areas in the battery limits for SNCR include the reagent receiving area, storage tanks, and recirculation system; the injection system, including injectors, pumps, valves, piping, and distribution system; the control system; and air compressors. In addition, NH_3 -based SNCR systems use electrically powered vaporizers to vaporize the NH_3 prior to injection.

Operating labor costs are based on two person-hours required per 8-hour shift of operation. The annual cost of the reagent is the major operating cost item for the process and is calculated as the product of the reagent usage in tons/year and the cost in dollars per ton of

pure reagent. Electricity, water, and steam requirements are based on vendor information. The cost of steam or air for atomization of reagent is included as an operating cost.

Default input values for SNCR are presented in Table B-5. The default inputs were taken from studies by EPA's Acid Rain Division (EPA, 1996; EPA, 1997; EPA, 1998) where available. Unit costs are escalated from 1995 dollars to 2004 dollars using *Chemical Engineering Magazine* cost indices.

Table B-5. Default Input Parameters for SNCR

Description	Units	Range	Default
Inlet NO _x level	lb/MMBtu		calculated
Reagent (1:Urea 2:Ammonia)	integer		1
Number of Injector Levels	integer		3
Number of Injectors	integer		18
Number of Lance Levels	integer		0
Number of Lances	integer		0
Steam or Air Injection for Ammonia (1: Steam, 2: Air)	integer		2
NO _x Reduction Efficiency	fraction	0.30-0.70	0.50
NH ₃ /NO _x Stoichiometric Ratio	decimal	0.8-2.0	1.2
Urea/NO _x Stoichiometric Ratio	decimal	0.8-2.0	1.2
Urea Cost	\$/ton		400
Ammonia Cost	\$/ton		300
Retrofit Factor	decimal		1.3
Maintenance (% of installed cost)	%		1.5%
Contingency (% of installed cost)	%		15%
General Facilities (% of installed cost)	%		5%
Engineering Fees (% of installed cost)	%		10%
Duration of Project (years)	integer		1

B.4.3 Natural Gas Reburning Design Criteria

Natural gas reburning (NGR) involves substituting natural gas for a portion of the pulverized coal supplied to the primary combustion zone and injecting the natural gas downstream of the primary combustion zone to form a reducing zone in which NO_x compounds are reduced to N₂. Combustion air for the reburning fuel (natural gas) is injected further downstream. Because the main combustion zone of furnaces employing this technology operates in its normal manner, gas reburning is applicable to a wide range of wall-, tangential-, and cyclone-fired boilers.

Boiler modifications for gas reburning involve installation of additional fuel injectors and associated piping and control valves. In the burnout zone, key components include overfire air (OFA) ports, a windbox, ductwork, and control dampers. Installation of the gas injectors and OFA ports requires waterwall modifications. Adequate residence time must be available both in the reburn zone and the burnout zone to maximize NO_x reduction and to minimize unburned carbon losses. Consequently, for retrofit applications, adequate space between the top burner row and the furnace exit must be available for appropriately locating the reburn fuel injectors and OFA ports.

The fraction of boiler heat input contributed by natural gas combustion (reburn fraction) depends on the desired NO_x removal efficiency. The relationship between the reburn fraction and NO_x reduction efficiency applies for NO_x reduction efficiencies from 55 to 65% and corresponding reburn fractions from 0.08 to 0.20. In CUECost, these are the valid input ranges for the NO_x removal efficiency and reburn fraction. If the user inputs both parameters within the valid ranges, the input values are used for cost calculations. If only one parameter is outside the valid range, that parameter is calculated using the other parameter. If both input values are outside of the valid ranges, a default reburn fraction of 0.15 is used with a corresponding 61% NO_x removal efficiency. The installed costs of gas injectors, OFA ports, and related equipment are included in the NGR cost worksheet. Also included in the NGR cost worksheet is the cost associated with piping natural gas to the boiler from the metering station located at the utility plant fence-line.

In general, natural gas reburning reduces the boiler operating costs associated with coal- and ash-handling process areas, including maintenance, electricity, and ash disposal. Fuel costs are generally higher, because the price of natural gas is typically higher than the price of coal. Maintenance costs for operating the NGR system are estimated at 2% of the total plant cost, plus a maintenance credit for operating the coal handling process at reduced coal feed rates. Savings from reduced fly ash disposal are estimated only for retrofit applications. The incremental fuel cost for firing gas is estimated by multiplying the amount of gas burned by the fuel price difference between gas and coal. Default values for NGR input parameters are presented in Table B-6. The default inputs were taken from ARD studies (EPA, 1996) where available. Unit costs are escalated using *Chemical Engineering Magazine* cost indices.

Table B-6. Default Input Parameters for NGR

Description	Units	Range	Default
Uncontrolled NO _x level	lb/MMBtu		calculated
NO _x Reduction Efficiency	fraction	0.55-0.65	0.61
Gas Reburn Fraction	fraction	0.08 - 0.20	0.15
Waste Disposal Cost	\$/ton		11.48
Natural Gas Cost	\$/MMBtu		9.00
Retrofit Factor			1.30
Maintenance (% of installed cost)	%		1.5%
Contingency (% of installed cost)	%		15%
General Facilities (% of installed cost)	%		2%
Engineering Fees (% of installed cost)	%		10%
Duration of Project (years)	integer		1

B.4.4 Low-NO_x Burner Technology Design Criteria

Low-NO_x burner technology (LNBT) limits NO_x formation by controlling both the stoichiometric and temperature profiles of the combustion process in each burner flame envelope. This control is achieved with design features that regulate the aerodynamic distribution and mixing of the fuel and air, yielding one or more of the following conditions:

- Reduced O₂ in the primary combustion zone, which limits fuel NO_x formation;
- Reduced flame temperature, which limits thermal NO_x formation; and
- Reduced residence time at peak temperature, which limits thermal NO_x formation.

Low NO_x burner designs for wall-fired boilers can be divided into two general categories: "delayed combustion" and "internally staged." Delayed combustion LNBT is designed to decrease flame turbulence (thus delaying fuel/air mixing) in the primary combustion zone, thereby establishing a fuel-rich condition in the initial stages of combustion. Internally staged LNBT is designed to create stratified fuel-rich and fuel-lean conditions in or near the burner. In the fuel-rich regions, combustion occurs under reducing conditions, promoting the conversion of fuel nitrogen to N₂ and inhibiting fuel NO_x formation. In the fuel-lean regions, combustion is completed at lower temperatures, thus inhibiting thermal NO_x formation.

Conventional tangentially-fired boilers consist of corner-mounted vertical burner assemblies from which fuel and air are injected into the furnace. The fuel and air nozzles are directed tangent to an imaginary circle in the center of the furnace, generating a rotating fireball in the center of the boiler. Each corner has its own windbox that supplies primary air through

the air compartments located above and below each fuel compartment. For tangentially-fired boilers, LNBT changes the air flow through the windbox by decreasing the amount of primary air and directing secondary air away from the fireball and toward the furnace wall.

Default input parameters for LNBT and suggested ranges are presented in Table B-7. The user selects the boiler type and the retrofit difficulty. CUECost calculates total capital cost as a function of boiler size. The NO_x reduction efficiency input does not affect the capital cost estimate, but is used to estimate emissions reduction.

Table B-7. Default Values for LNBT Input Parameters

Description	Units	Range	Default
Uncontrolled NO _x level	lb/MMBtu		calculated
Boiler Type (T:T-fired, W:Wall)	letter		T
Burner Type 1 = LNB or LNC1, 2=LNB and OFA or LNC2, 3=LNC3	integer		1
Retrofit Difficulty Factor	number		1.3
General Facilities	percent		5.0%
Engineering	percent		10.0%
Contingency	percent		15.0%
Duration of Project (years)	integer		1

B.5 Hg CONTROL TECHNOLOGY CRITERIA

Injection of powdered activated carbon (PAC) has been developed and tested at full scale on coal-fired utility boilers. Test programs have been performed on a utility boiler firing subbituminous coal with a downstream cold-side ESP, on utility boilers firing bituminous coal with a downstream cold-side ESP, and firing bituminous coal with a compact hybrid particle collector (COHPAC) arrangement (upstream hot-side ESP with downstream baghouse after the air preheater). Performance models were developed.

B.5.1 Mercury Removal Models

EPA's Information Collection Request (ICR)⁴ showed that mercury released from coal combustion may be partly removed from the exhaust gases by existing equipment without additional retrofit technology. The existing equipment may be one or more pieces of equipment that contribute to mercury removal.

⁴ Available at <http://www.epa.gov/icr/>

If $f_{\text{equipment}}$ is equal to the fraction of mercury removed from the boiler gases by a piece of equipment, then $(1 - f_{\text{equipment}})$ equals the fraction of mercury remaining in the gases after that piece of equipment. The fraction of mercury remaining after n pieces of equipment is equal to

Fraction of mercury remaining after n pieces of equipment =

$$[(1 - f_{\text{equipment } 1}) \times (1 - f_{\text{equipment } 2}) \times (1 - f_{\text{equipment } 3}) \times \dots \times (1 - f_{\text{equipment } n})] \quad (\text{Eq. B-15})$$

Therefore, the total mercury removal fraction = f_{Total}

$$f_{\text{Total}} = 1 - [(1 - f_{\text{equipment } 1}) \times (1 - f_{\text{equipment } 2}) \times (1 - f_{\text{equipment } 3}) \times \dots \times (1 - f_{\text{equipment } n})] \quad (\text{Eq. B-16})$$

If one of the pieces of equipment is PAC injection, then the total mercury removal fraction = f_{Total} =

$$1 - [(1 - f_{\text{equipment } 1}) \times (1 - f_{\text{equipment } 2}) \times (1 - f_{\text{equipment } 3}) \times \dots \times (1 - f_{\text{PAC injection}}) \times \dots \times (1 - f_{\text{equipment } n})] \quad (\text{Eq. B-17})$$

where

$f_{\text{PAC injection}}$ is the fraction of mercury removed by PAC injection.

If PAC injection is simply added to existing equipment and the removal effects of the existing equipment are combined into one term, then we can represent Eq. B-17 as

$$f_{\text{Total}} = 1 - [(1 - f_{\text{existing equipment}}) \times (1 - f_{\text{PAC injection}})] \quad (\text{Eq. B-18})$$

where

$f_{\text{existing equipment}}$ is the removal fraction of the existing equipment.

In this effort, data from full-scale tests of mercury reduction were used to formulate models for mercury reduction from existing equipment and from PAC injection. Full-scale data for mercury removal by existing equipment are available from the ICR data. Full-scale testing results for mercury reduction from PAC injection are available from the Department of Energy's field testing programs at Southern Company's Gaston plant, Wisconsin Electric Power Company's Pleasant Prairie power plant (PPPP) and at PG&E National Generating Group's Brayton Point and Salem Harbor plants.

B.5.2 Mercury Removal by Existing Equipment, $f_{\text{existing equipment}}$

Through statistical analysis of the ICR data, EPRI (2000) shows that mercury reduction is a function of both emission control equipment configuration and a function of chlorine content of the coal, and in some cases a function of the SO₂ emissions level from the boiler. EPRI (2000) provides algorithms to estimate mercury capture as a function of the plant configuration, the coal chlorine content, and the SO₂ emissions. These algorithms are:

Algorithm 1 (cold-side ESP):

$$f_{\text{existing equipment}} = C_1 \times \ln [(\text{coal Cl, ppm})/(\text{SO}_2, \text{ in lb/MMBtu})] + C_2 \quad (\text{Eq. B-19})$$

where

C_1 and C_2 = Algorithm 1 constants

Algorithm 2 (all other categories):

$$f_{\text{existing equipment}} = C_1 \times \ln (\text{coal Cl, ppm}) + C_2 \quad (\text{Eq. B-20})$$

where

C_1 and C_2 = Algorithm 2 constants

Minimum and maximum allowable values are set for the results of Equations B-19 and B-20. Values of C_1 and C_2 , minimum and maximum are shown in the left columns in Table B-8 for hot and cold side ESP operating conditions.

According to Eqs. B-19 and B-20, the predicted mercury reduction efficiencies for conditions at Gaston (Bustard et al., 2001; Durham et al., 2001; and Bustard et al., 2002), Pleasant Prairie power plant (PPPP) (Bustard et al., 2001; Durham et al., 2001), Brayton Point (Durham et al., 2001) and at Salem Harbor (Durham et al., 2001) are presented in Table B-8.

Table B-8. Predicted Collection of Mercury by ESP according to Eqs. B-19 and B-20

					Gaston	PPPP	Brayton Point	Salem Harbor
		Chlorine, % by weight in coal			0.03	0.0015	0.08	0.03
		Coal Chlorine, ppm			300	15	800	300
		Flue Gas SO₂, lb/MMBtu			0.650	0.360	0.820	0.500
	C₁	C₂	Min	Max	Predicted Mercury Reduction			
ESPc	0.1233	-0.3885	0.0%	55.0%		7.1%	46.0%	40.0%
ESPh	0.0927	-0.4024	0.0%	27.0%	12.6%			

ESPc = cold-side ESP

ESPh = hot-side ESP

Source: EPRI (2000)

Gaston fires bituminous coal and has a hot-side ESP followed by an air preheater and then a low-pressure pulse-jet FF for a COHPAC arrangement (Bustard et al., 2001; Durham et al., 2001; and Bustard et al., 2002). EPRI (2000) did not include algorithms for facilities with this arrangement. One might expect that the mercury reduction without PAC might

correspond approximately to the predicted mercury reduction in Table B-8 for a hot-side ESP (ESPh). Under the conditions at Gaston, predicted mercury reduction equals 12.6%. However, tests at Gaston showed negligible mercury removal. But considering the range of variability in the possible results, the difference may be reasonable. However, this example demonstrates that this algorithm will not give precise values, but reasonable estimates.

At the Pleasant Prairie power plant (PPPP) (Starns et al., 2002), a facility firing PRB coal with a cold-side ESP (ESPc), the test results showed about 5% actual mercury removal from existing equipment compared to about 7% as estimated by the algorithm of (EPRI 2000) for the conditions at PPPP, and shown in Table B-8 (Bustard et al., 2001; Durham et al., 2001). Therefore, the value estimated by the algorithm is approximately in the same range. The 15 ppm chlorine content of the coal used at PPPP (which is much lower than that of most other PRB sites) probably contributes to the low removal by the existing equipment. With chlorine content more typical of a PRB coal, around 100 ppm or more, the algorithm predicts that mercury would be reduced by a greater amount.

For Brayton Point, a facility firing bituminous coal and equipped with an ESPc, the algorithm of EPRI (2000) produces an estimated mercury reduction by existing equipment of about 46% (see Table B-8) versus an actual measured removal efficiency of 32% (Durham et al., 2001). These values, which are in about the same range, further illustrate that the algorithm of EPRI (2000) is not exact, but approximate, at estimating mercury removal by existing equipment.

At Salem Harbor, a facility firing bituminous coal and equipped with an ESPc, 87% mercury reduction from existing equipment was measured (Durham et al., 2001). This measured value compares to about 40% estimated from the algorithm of EPRI (2000) as shown in Figure B-1. The significant difference can be explained as follows: First, Salem Harbor operates with fly ash loss on ignition (LOI) in the range of 25-35%. According to Bustard et al. (2001), this fly ash loss is approximately equivalent to a carbon loading of 60-84 lb/MMacf in the exhaust stream. This value is higher than a typical plant's inject rate. So, the carbon present in the fly ash has likely contributed to a very high intrinsic capture of mercury. Second, temperature plays a role in intrinsic mercury capture. Because Salem Harbor has the ability to increase its ESP inlet temperature through operation of steam heaters, parametric tests of intrinsic mercury removal as a function of temperature could be performed. Figure B-1 shows the results of that testing under various firing conditions and also with data taken from another test using low sulfur bituminous coal (not the baseline coal). The trend is quite clear that increasing temperature reduces intrinsic mercury capture from around 90% down to around 10%. Thus, mercury absorption by fly ash is enhanced when flue gas is cooled. Cooling the flue gas can enhance mercury uptake by flash.

However, when PAC is injected, its large capacity for mercury absorption allows the sorbent to be operated at temperatures of 350 °F or higher. As such, spray cooling usually promotes little or nearly zero mercury absorption by PAC.

Because a facility's mercury reduction by existing equipment may be significantly different from what the algorithm of EPRI (2000) determines, this algorithm should be used with care and only for making approximate estimates. As the measurements at Salem Harbor clearly indicate, LOI or other ash qualities and gas temperature can have a very significant impact on the level of mercury being removed by existing equipment and may be worth including as parameters in this algorithm at some future date when more information is available. Therefore, the algorithm of Equation B-20 and EPRI (2000) may provide reasonable estimates in many cases. But there is a chance that actual mercury capture may differ significantly from what Equation B-20 predicts. For any specific facility, actual measurements of mercury removal, if available, should be used.

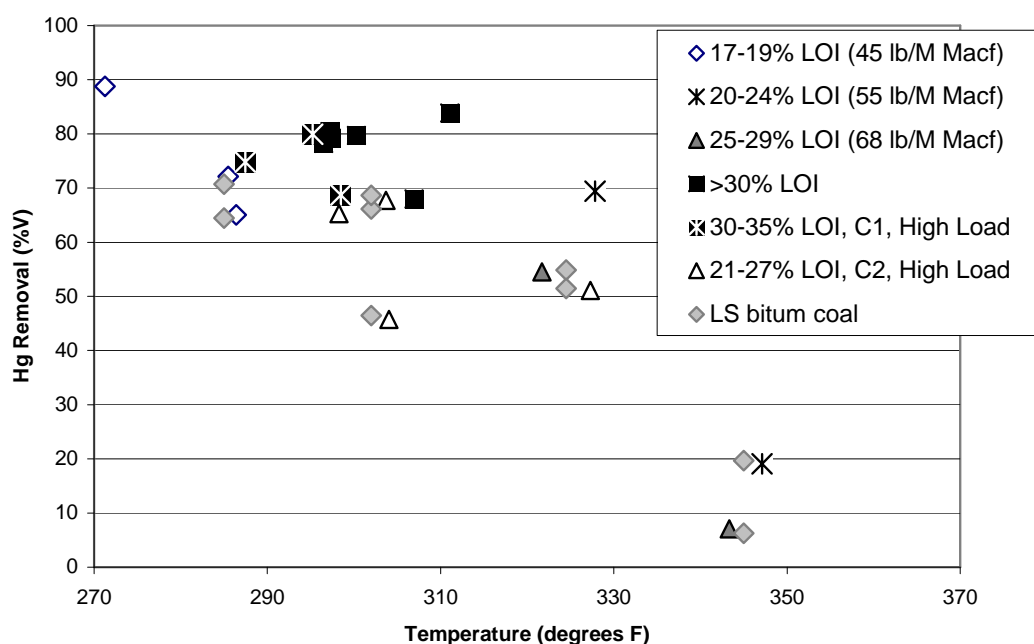


Figure B-1. Salem Harbor Mercury Removal without PAC Injection (Durham et al., 2001)

B.5.3 Mercury Reduction by PAC injection, $f_{PAC\ injection}$

EPA (2000) has algorithms developed from pilot-scale data for mercury reduction on boilers equipped with PAC injection. In this work, we have made the following model improvements:

1. The algorithms of EPA (2000) were developed from pilot-scale tests and characterize total mercury reduction from both PAC injection and from existing equipment as a function of PAC injection concentration. When using the algorithms of EPA (2000), it is necessary to have a different PAC injection algorithm for each type of equipment configuration, including upstream equipment. These PAC injection algorithms may have to be updated as new information regarding mercury control from existing equipment

becomes available. In the effort described in this paper, the mercury reduction from PAC injection was isolated from that of the other equipment. Therefore, as we gain more information on reduction of mercury from equipment other than PAC injection, it should not be necessary to perform new regressions on the PAC injection models and it will also be possible to assess the fate of mercury in equipment that is either upstream or downstream of the PAC injection system.

2. The algorithms of EPA (2000) are of a form where it is possible for Hg removal to approach 100% by injection of very high concentrations of PAC. As will be shown, experience at PPPP showed that under some circumstances it is not possible to achieve such extremely high reduction of mercury emissions with PAC injection. Therefore, the algorithm for mercury reduction from PAC injection was modified to permit an upper limit to mercury removal that may be less than 100%.
3. Because the algorithms of EPRI (2000) are based on the full-scale ICR data, it is desirable to use them to characterize mercury reduction from existing equipment. However, it is not possible to integrate the algorithms of EPRI (2000) into the approach used in EPA (2000). By treating the mercury reduction from PAC injection independently from mercury reduction from other equipment, it is possible to use the algorithms of EPRI (2000) to characterize mercury reduction from existing equipment.

In the case of PPPP, PAC injection test results demonstrated that mercury reduction behaved asymptotically with a maximum achievable mercury reduction from PAC that is well below 100%, regardless of PAC injection rate. For this reason, the equation that is used in EPA (2000) to characterize the relationship between mercury reduction and PAC injection

$$\% \text{ Hg reduction} = \eta = 100 \times f_{\text{from PAC injection}} = 100 - [A/(M+B)^C] \quad (\text{Eq. B-21})$$

where

M = the mass injection rate of PAC (in lb/MMacf) and A, B, C are curve-fit constants determined with available data.

$$\% \text{ Hg reduction} = \eta = 100 \times f_{\text{from PAC injection}} = 100 \times D - [A/(M+B)^C] \quad (\text{Eq. B-22})$$

where

D = the asymptotic fraction of mercury reduction that is approached but is not achieved.

Constants A, B, C, and D appearing in Eq. 8 are specified for a given plant configuration and gas temperature. At this time, these constants can only be developed for full-scale applications similar to the conditions where full-scale data exists. For some other boiler configurations there is test data available from pilot-scale (Bustard et al., 2001) tests that can be used until full-scale data becomes available. For other configurations where neither full-scale nor pilot-scale data exists, the constants can be developed as data becomes available from future tests. The constants A, B, C and D used in CUECost are listed in Table B-9.

Table B-9. Values of Constants Used in the PAC Injection Eqs. B-21 and B-22

Coal	FF retrofit	PAC Capacity	A	B	C	D
Bituminous	in flight	EPAC	0	1.207	-0.2277	100%
Bituminous	in flight	PAC	-0.6647	2.1232	-0.0665	100%
Bituminous	in flight	other				
Bituminous	FF	EPAC	0	2.5007	-2.2097	100%
Bituminous	FF	PAC	1.6944	-1.1267	-0.0009	100%
Bituminous	FF	other				
Subbituminous	in flight	EPAC	0.8837	0.4485	-0.575	100%
Subbituminous	in flight	PAC	3.308	0.754	-0.5925	70%
Subbituminous	in flight	other				
Subbituminous	FF	EPAC	0	2.5007	-2.2097	100%
Subbituminous	FF	PAC	-0.4318	1.9551	-0.8937	100%
Subbituminous	FF	other				

B.5.4 PAC Injection Models Developed from Full-Scale Data

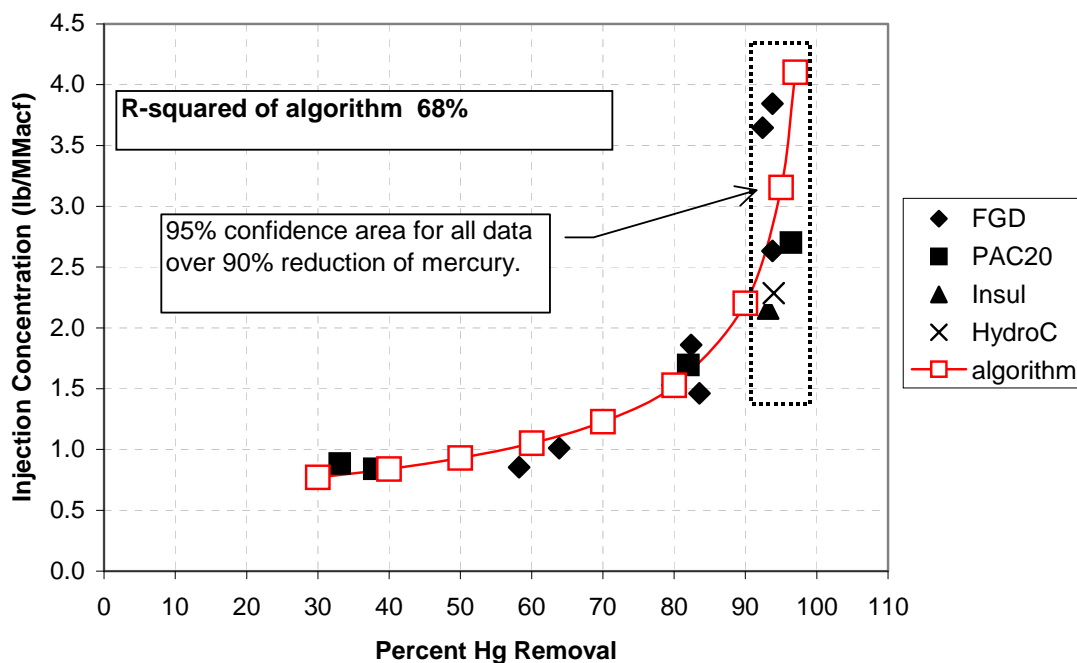
For the purpose of modeling, we are interested in estimating the necessary PAC injection rate to achieve a specified level of mercury control. Therefore, we developed algorithms of PAC injection rate as a function of desired mercury reduction by PAC. So, rather than plotting mercury reduction versus PAC injection concentration, as is done in Bustard et al. (2001; 2002) and Starns et al. (2002), we have reversed the axes from those shown in these references.

In these tests (Bustard et al., 2001; Bustard et al., 2002; Starns, 2002) several different PAC sorbents were tested. The different PAC sorbents will be designated on the legends of the figures. In this effort, we did not have specific information regarding the sorbent properties of the tested sorbents. Therefore, since we did not want to conjecture on the role of particular sorbent properties on mercury removal performance, we did not evaluate sorbent choice effects except to determine whether or not sorbent type has an effect on performance under a particular condition.

Gaston

Figure B-2a shows mercury collection results measured from an on-line mercury analyzer during testing conducted at Gaston. Data are plotted as PAC injection concentration versus mercury reduction percent. Data include results obtained with several different sorbent types (Bustard et al., 2001). Figure B-2a also shows a curve developed in the form of Equation B-22 to approximately correspond to the results achieved at Gaston. The coefficients for the algorithm are listed in Table B-10. At Gaston the choice of sorbent appeared to have little or no impact on performance. At mercury removal rates in the range

of 92-96%, mercury reduction is less sensitive to changes in PAC injection rate. Figure B-2b shows the data for 92-96% mercury reduction in greater detail. The enclosed region on Figure B-2b includes the estimated 95% confidence range for these mercury reduction data.⁵ Figure B-3, a plot of deviation of the predicted and measured PAC injection rate,⁶ demonstrates this trend in another way. For most mercury reduction levels, the deviation between model and actual PAC injection rates is only about 10%. For mercury reduction in excess of 90%, however, the deviation is higher on a percentage basis. While Figure B-3 shows that at high removal rates the deviation between the model and measured value expressed as a percent of predicted level is -30% to $+40\%$, in fact this range of values only corresponds to a range of under ± 1 lb/MMacf. The high percentage of the deviation is due to the actual values being relatively small at Gaston.



Data provided by Jean Bustard, ADA Environmental Services, September 16, 2002

Figure B-2a. Gaston Testing

⁵ 95% confidence range for Hg reduction and for PAC injection concentration are determined by ± 2 standard deviations from the arithmetic mean, with correction for sample size.

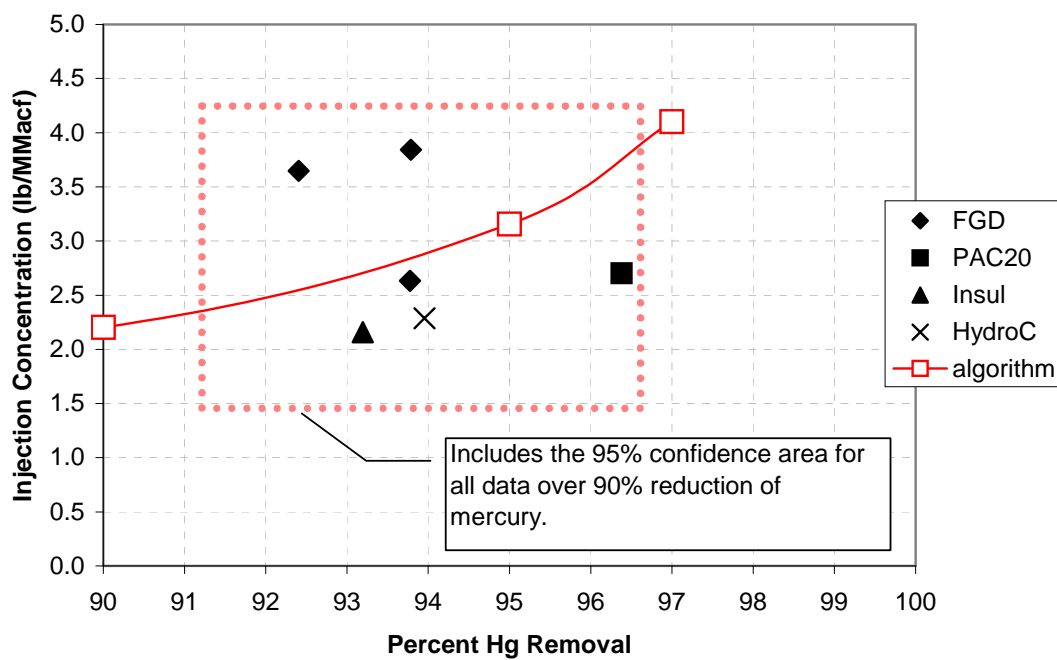
⁶ Calculated as $(\text{actual rate} - \text{predicted rate}) / \text{predicted rate}$ and expressed in percent.

Table B-10. Coefficients for Curve Fit Algorithms

Plant	Algorithm	A	B	C	D	Sorbent Type
Gaston		53	0.1	2	1.00	FGD, PAC 20
PPPP	a	150	5	1	0.72	FGD (14 micron fraction)
	b	140	1	1	0.69	FGL
	c	145	3	1	0.705	Insul
Brayton Point	a	300	3	0.8	1.13	FGD and FGD I
	b	300	0	0.8	1.05	FGL
	c	300	1.5	0.8	1.09	Insul

Note 1: Algorithms were developed for a specific plant and a specific sorbent.

Note 2: A, B, C, and D are coefficients for curve fit.



Data provided by Jean Bustard, ADA Environmental Services, September 16, 2002

Figure B-2b. Gaston Testing

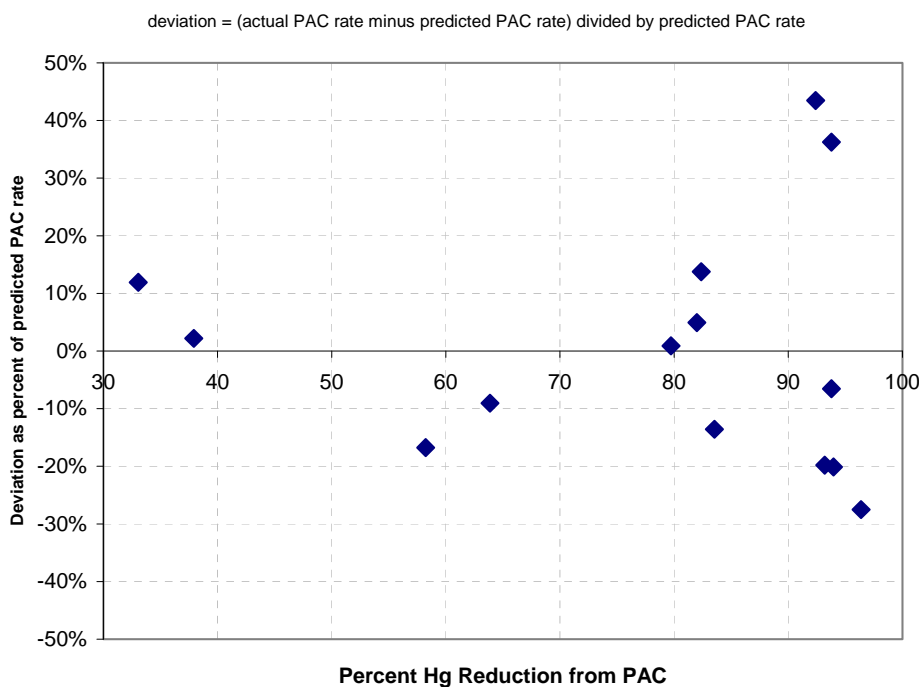


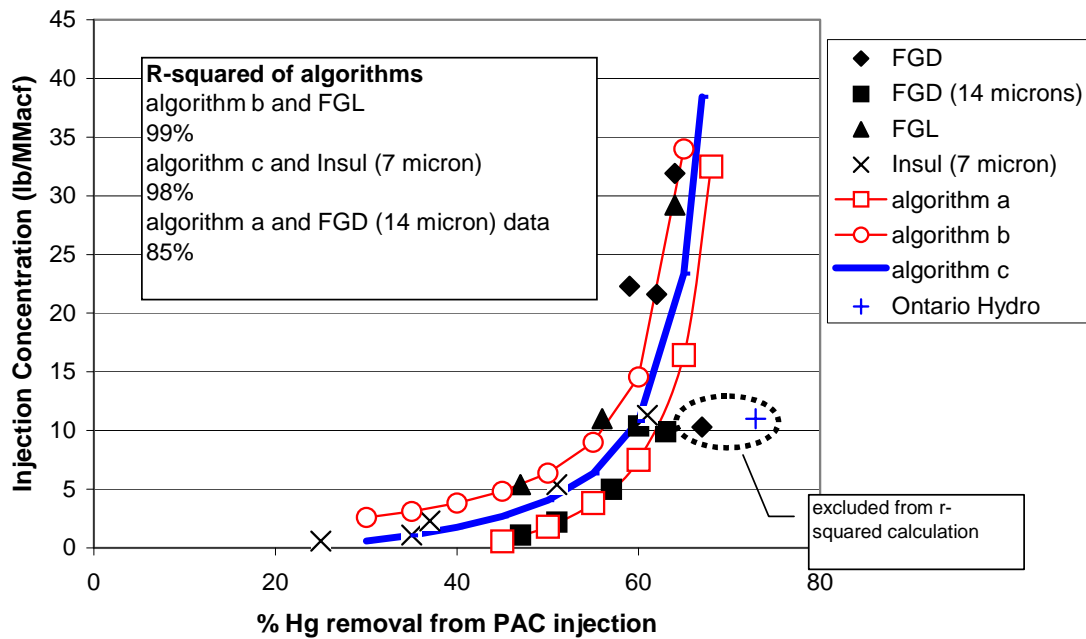
Figure B-3. Deviation of the Gaston PAC Algorithm

Pleasant Prairie Power Plant

Figure B-4 shows mercury collection results measured from an on-line mercury analyzer during testing conducted at PPPP. Data include results with several different sorbent types.⁷ Figure B-4 also shows a data point for the total mercury removal as measured by the Ontario Hydro method. The Ontario Hydro method shows a somewhat higher, but nevertheless a similar, mercury removal as the on-line mercury analyzer used for the testing. Two curves were developed in the form of Equation 9 to correspond to specific sets of data and are plotted on Figure B-4. The coefficients of these algorithms (A, B, C, D) are listed in Table B-10. Unlike the results at Gaston, at PPPP the choice of sorbent has a significant effect, possibly a result of the fact that at Gaston there is a downstream fabric filter, which provides improved sorbent-gas contact, while at PPPP all of the mercury absorption had to occur in the duct. Figure B-5 is a plot of deviation of the predicted and measured PAC injection rate.⁸ Had one algorithm been used for all of the sorbents, the deviations would have been very high in some cases. Nevertheless there is enough scatter in some of the data that, even with different algorithms for each sorbent, deviation can be on the order of 40%. Note that the one data point with very high percent deviation (over 70%) was actually at a low removal rate and the absolute difference between the algorithm results and measured results was quite small. For other plants with conditions similar to those at PPPP (sub-bituminous coal), some consideration should be made for the sorbent type.

⁷ Data provided by Jean Bustard, ADA Environmental Services, September 16, 2002.

⁸ Calculated as (actual rate – predicted rate) / predicted rate.



Data provided by Jean Bustard, ADA Environmental Services, September 16, 2002

Figure B-4. PPPP Testing

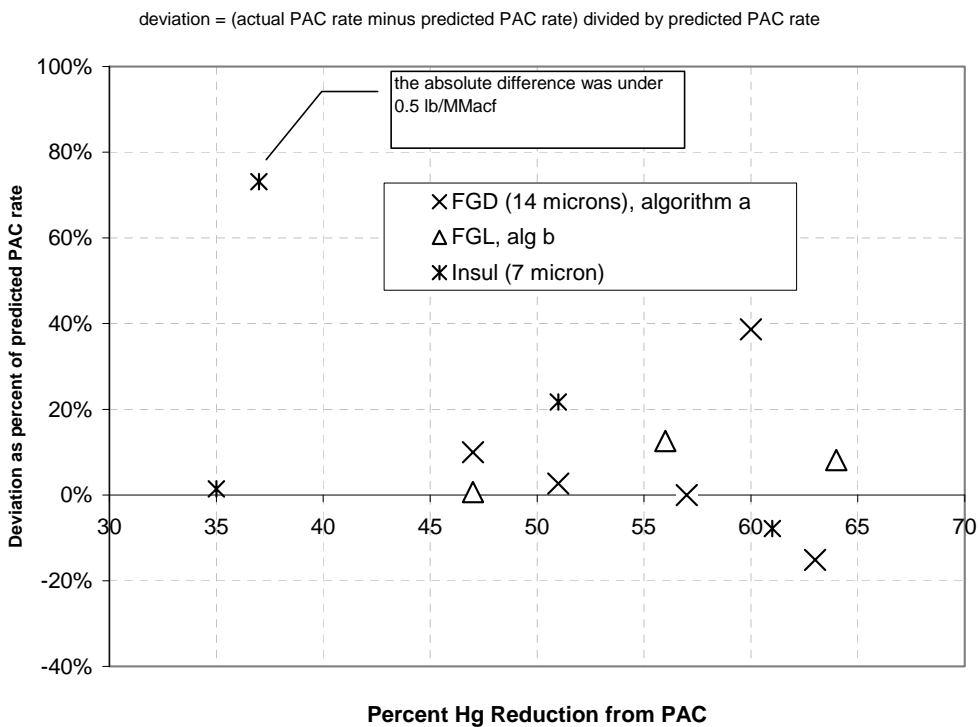


Figure B-5. Deviation from the PPPP PAC Algorithm

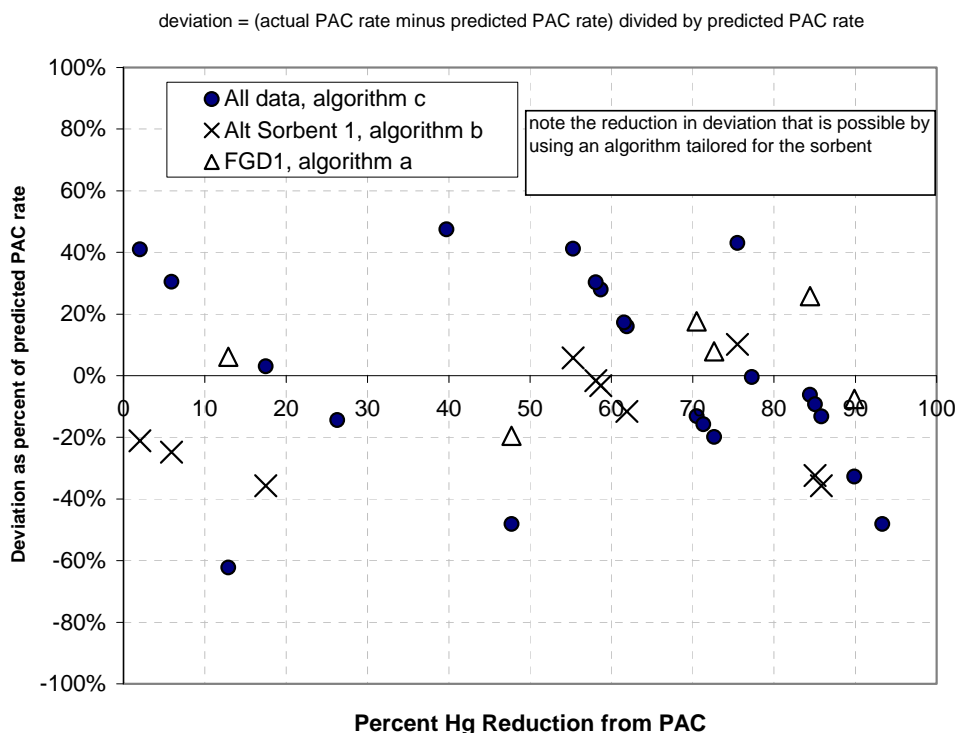


Figure B-7. Deviation from the Brayton Point PAC Algorithm

Salem Harbor

According to DOE (2003), long-term testing of PAC injection with baseline coal indicated about 90% reduction of mercury at a PAC injection concentration of 10 lb/MMacf. Because of the high level of intrinsic mercury reduction at Salem Harbor and the sensitivity of the measuring methods, the increased mercury reduction from PAC is difficult to assess. Therefore, mercury reduction from PAC was not analyzed in this effort. However, Salem Harbor information provided useful insights on the effects of unburned carbon and gas temperature on intrinsic levels of mercury reduction as previously discussed in this paper.

B.5.5 Mercury Speciation with SCR

The speciation of mercury is known to have a significant impact on the ability of air pollution control equipment to capture it. In particular, the oxidized form of mercury, mercuric chloride (HgCl_2 , a solid), has been shown to be far easier to capture in wet FGD systems, dry scrubbers or by dry sorbent injection than the elemental form of mercury (a liquid). Some data also indicate that SCR catalysts may act to oxidize a significant portion of the elemental mercury, which will make the oxidized mercury far easier to remove by downstream equipment such as wet FGD.

Studies have found that oxidation of elemental mercury by SCR catalyst may be affected by the following:

- The space velocity of the catalyst
- The temperature of the reaction
- The concentration of ammonia
- The age of the catalyst
- The concentration of HCl in the flue gas stream.

Bustard et al. (2001) showed that, in tests on a laboratory combustor, mercury oxidation without a catalyst was enhanced with higher Cl concentration (higher HCl in the flue gas) and that oxidation increased with residence time and at lower temperatures, as shown in Figure B-8. Hocquel et al. (2002) also describe the results of laboratory tests of oxidation of mercury across SCR catalysts. The results of these tests, shown in Figure B-9, demonstrated that the catalyst significantly increased the amount of mercury that oxidized to mercuric chloride.

In Richardson et al. (2002), tests of mercury oxidation by SCR catalyst were conducted using simulated flue gas and slip-streams from actual units. Results showed similar trends for both simulated flue gas and slip-streams from actual units with the exception that the effect of increasing space velocity appeared somewhat more significant with the slip-streams. Multiple catalyst types were tested with similar results obtained. According to Bustard et al. (2001), at space velocities in the range of 400 h^{-1} , mercury oxidation was in the range of about 80% to 90% for fresh catalyst. However, the oxidation rate falls off quickly with increased space velocity; oxidation might be in the range of 30-80% at a space velocity of 4000 h^{-1} . The wide range of oxidation performance at a space velocity of 4000 h^{-1} is the result of the influence of other factors – temperature, ammonia and possibly other effects.

As shown in Figure B-10, Richardson et al. (2002) showed that oxidation of mercury across fresh SCR catalyst was highest at temperatures in the range of 550 °F and lowest in the range of 800 °F, consistent with the fact that oxidation of mercury to mercuric chloride occurs mostly at lower temperatures.

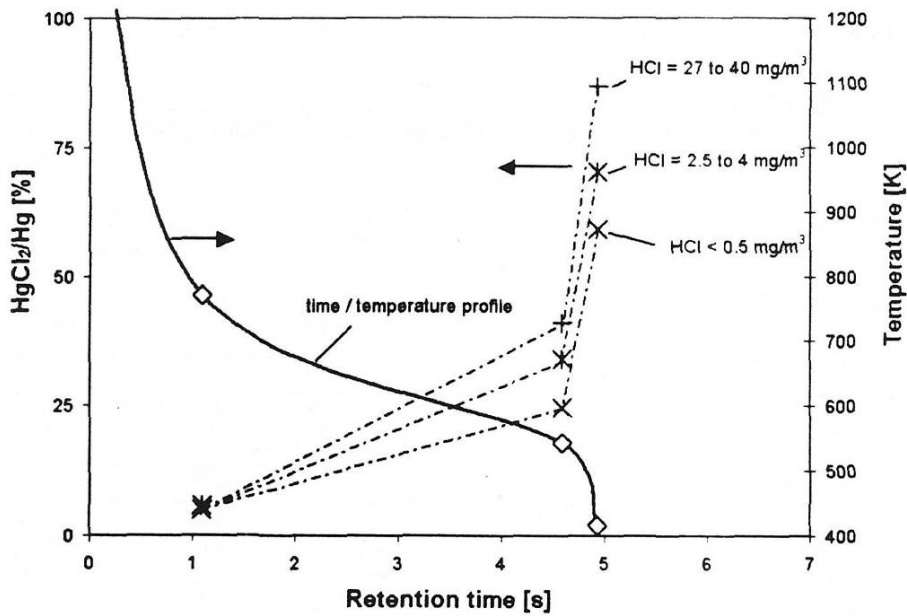


Figure B-8. Mercury Oxidation without a Catalyst as a Function of Residence Time, Gas Temperature, and HCl Content (Hocquel et al., 2002)

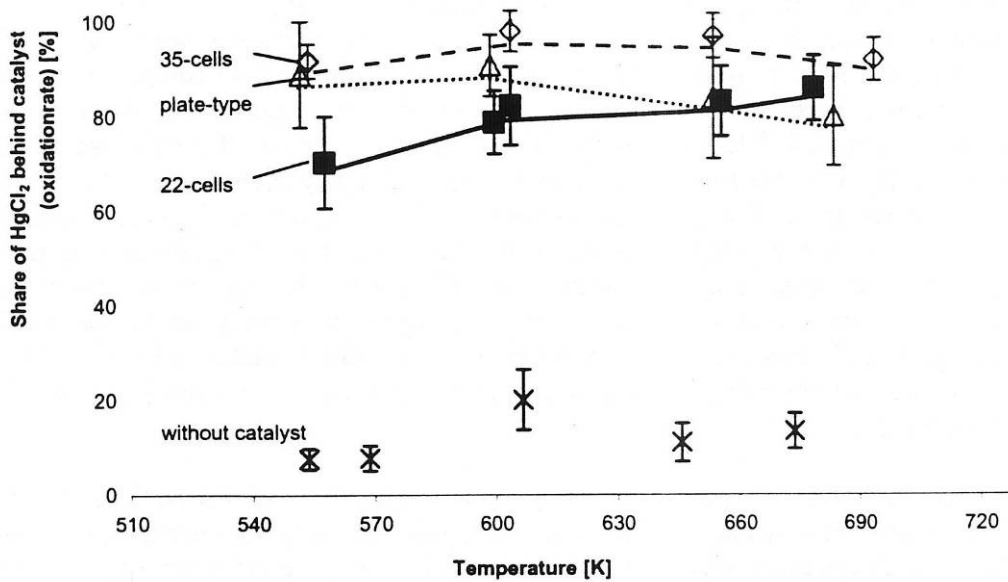


Figure B-9. Mercury Oxidation across SCR Catalysts and without SCR Catalyst (Hocquel et al., 2002)

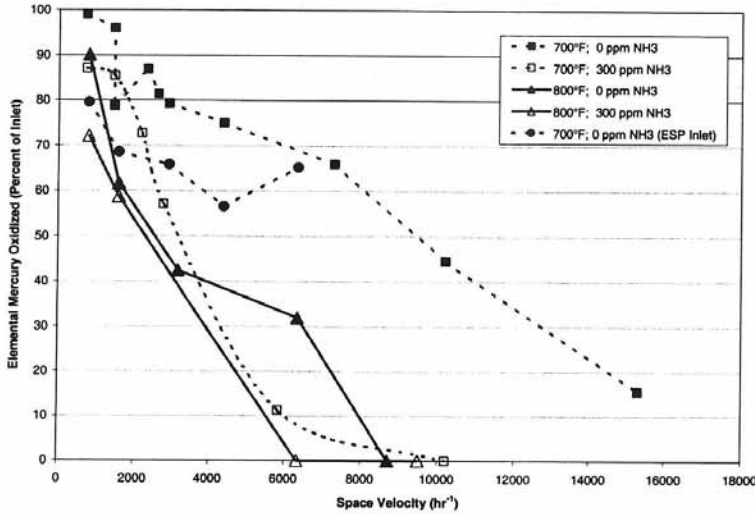


Figure B-10. Oxidation of Mercury across C-1 SCR Catalyst in PRB-derived Flue Gas (Richardson et al., 2002)

The presence of ammonia, which is the NO_x reducing reagent normally used in SCR systems, was shown by Richardson et al. (2002) to inhibit the oxidation of elemental mercury. This effect is most pronounced with catalyst that has been exposed to boiler exhaust gases for a number of months. As shown in Figure B-11, mercury oxidation without ammonia present remained between 80% and 90% after 4200 hours (about six months) of exposure to boiler gases at a space velocity of 1450 h⁻¹. When exposed to 300 ppm of ammonia, fresh catalyst continued to oxidize 80-90% of the elemental mercury. However, after 4200 hours of exposure no oxidation was measured across the catalyst when ammonia was present.

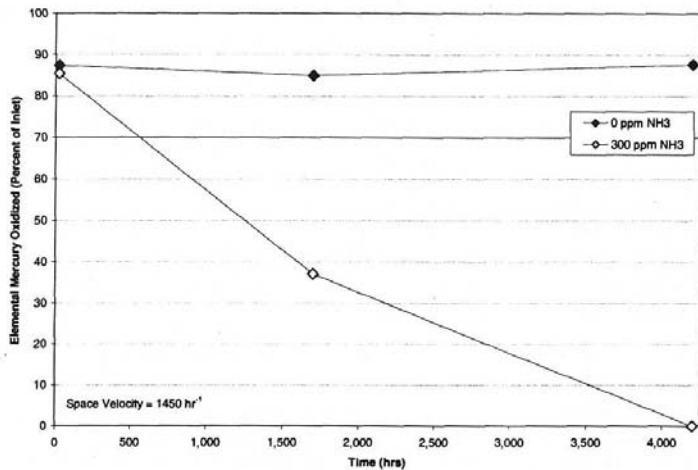


Figure B-11. Effect of Flue Gas Exposure Time on C-1 SCR Catalyst Oxidation of Elemental Mercury: 700 °F and Space Velocity of 1,450 h⁻¹ (Richardson et al., 2002)

Oxidation of elemental mercury to mercuric chloride across an SCR catalyst, therefore, may be a function of: space velocity, temperature, ammonia concentration, and catalyst life. Other factors, such as fly ash characteristics, are also believed to play a role.

Bustard et al. (2002) describe the results of a program that evaluated mercury oxidation across full-scale utility boiler SCR systems. A summary of the results of these tests is shown in the first four entries in Table B-11. Testing was performed at four coal-fired electric utility plants having catalyst age ranging from around 2500 hours to about 8000 hours. One plant fired subbituminous coal and three other plants fired eastern bituminous coal. The test results showed high levels of mercury oxidation in two of the three plants firing eastern bituminous coal and insignificant oxidation at the other two plants (one firing bituminous coal and the other subbituminous). However, for both of the plants where little or no mercury oxidation was measured (S1 and S3) over 85% of the mercury at the particle control device inlet was already in the non-elemental (oxidized) form. For the one bituminous coal fired plant with low mercury oxidation (S3), over 50% of the mercury at the SCR inlet was already in the oxidized form. At the plant firing subbituminous coal (S1), mercury oxidation was fairly low. But, due to the high carbon in that plant's fly ash, the elemental mercury was apparently adsorbed onto the ash, resulting in high particulate mercury levels. Finally, in contrast with the studies of Hocquel et al. (2002) and Richardson et al. (2002), ammonia appeared to have little or no effect on mercury oxidation on these actual, full-scale facilities.

Subsequent tests on sister units at those plants and at other plants are shown in the second four entries in Table B-11. All of the units fired bituminous coal and showed that mercury oxidation was generally enhanced to high levels of oxidized mercury at the SCR outlet. In each case where a scrubber was installed the mercury removal was high. For the unit with an ESP and no scrubber, mercury removal was not improved by the SCR.

At this point in time, the understanding of the effects of SCR catalyst on mercury oxidation is fairly limited. Clearly, mercury oxidation is substantial under some conditions, but less significant under others. However, significant mercury oxidation by SCR catalyst appears to occur with bituminous coal and oxidation may be less certain with PRB coals. Where bituminous coal was fired with an SCR and an FGD, high levels of mercury removal generally occurred.

Default values for mercury control input parameters are shown in Table B-12.

Table B-11. Summary of Results from Full-Scale SCR Mercury Oxidation Tests (Bustard et al., 2001)

Power Plant	Catalyst Vendor, Type, SV (h ⁻¹)	Catalyst Age	Coal Type	S in Coal (%)	Cl in Coal (ppm)	NH ₃ Slip (ppm)	SO ₃ ppm	Cl ppm	Oxidized mercury content, SCR in/out	Oxidized mercury content, w/o and w/ SCR, PM inlet	Total Hg Removal across PM+FGD, w/o and w/ SCR	Effect of NH ₃ on Hg Oxidation (SCR in/out)
S1, 650 MW gross Cyclone, ESP	Cormetech Honeycomb 1800	8000 h	PRB	0.2	<60	2 *	0.4 *	1.5 *	Unit 2: 8% - 18%; net 10%; small increase; 1 OH sample	5% - 8%; net 3%; small increase; 1 OH sample each	60% - 65%; net 5%; small increase- within experimental error; 1 OH sample each	No effect
S2, 1360 MW gross, Wall, ESP+FGD (MEL)	Siemens Westinghouse Plate 2125	3.5 months	OH Bit	3.9	1640	0.1	33 *	108 *	48% - 91%; net 43%; significant increase; 2 OH samples No effect of alkali injection (Unit 1)	73% - 97%; net 24%; significant increase; 2 OH samples each	51% - 88%; net 37%; significant increase; FGD removed 94% of oxidized Hg; 2 OH samples each	Not tested
S3, 750 MW gross, Tangential, ESP	KWH Honeycomb ~3930	1 ozone season	PA Bit blend	1.7	1150	0.8	24	81 #	55% - 65%; net 10%; small increase; 2 OH samples 35% - 61%; net 26%; for 2nd coal in sister unit; 2 OH samples	77% - 67%; net -10%; possible filter effects due to reactive ash Not tested in 2 nd coal/sister unit	16% - 13%; net -3%; within experimental error; 2 OH samples each Not tested in 2 nd coal/sister unit	Small neg. effect. Not tested in 2 nd coal/sister unit
S4, 704 MW gross, Cyclone, Lime venturi scrubber	Cormetech Honeycomb 2275	1 ozone season	KY Bit	2.9	360	0.2	16 #	19	9% - 80%; net 71%; significant increase; 2 OH samples	56% - 87%; net 31%; significant increase; 2 OH samples each	46% - 90%; net 44%; significant increase; 2 OH samples each	Small negative effect
684 MW gross, Wall, ESP+FGD	Halder-Topsoe "corrugated" ~3750?	2 months	PA/WV Bit	3.6	470	0.3	10.6	Not Measured (NM)	Oxidation to 80+%; Net +38% increase	Oxidation to 95%; Net +15% (using data from sister unit w/o SCR)	Significant increase to 90+%; net +40%	Not tested
800 MW gross, Tangential, ESP	Cormetech Honeycomb 3800	2 seas. 2 layers repl. after 1 st season	KY/WV Bit	1	1000	0.1	14	NM	Oxidation to 80+%; Net +21% increase	Oxidation to 89%; Net -0% (using data from sister unit w/o SCR)	No effect; actually lower Hg removal in ESP (-6% vs 23%)	Not tested
1360 MW gross, Wall, ESP+FGD (MEL)	C&I Ceramics Plate 2125	2 ozone seasons	OH Bit	3.9	520	0.5	30	NM	Oxidation to 80+%; Net +33% increase	Oxidation to 95+%; Did not test w/o SCR	-85% Hg removal; Did not test w/o SCR	Not tested
Cyclone, Lime venturi scrubber	Cormetech Honeycomb 2275	2 ozone seasons	KY Bit	3.1	750 bypass 250 w/SCR	0.1	12	NM	Oxidation to 60+%; Net +20% increase; "More" oxidation if 1 outlier data not used	Oxidation to 90+%; Net +39% Cl in coal changed between tests	Significant increase to 90+%; net 47%	Not tested

NH₃, Cl, SO₃ – Sampled at SCR outlet unless noted (* - ESP outlet, # - Particulate control inlet)

Table B-12. Default Values for Mercury Control Input Parameters

Description	Units	Range	Default
<i>Sorbent Injection Inputs</i>			
Hg CEMS (0=no, 1=yes)	integer	0 or 1	1
Hg Reduction Required from Coal	percent		80.0%
Sorbent Type, 1=EPAC, 2=PAC, 3=other			2
Maximum Temperature before Spray Cooling	deg F	up to 325 F	325
Sorbent Recycle Used?	yes/no		no
Spray Cooling Desired?	yes/no		no
EPAC Cost (delivered cost of brominated PAC)	\$/ton		\$1,500
PAC Cost (delivered)	\$/ton		\$1,000
Other Sorbent Cost (delivered)	\$/ton		\$1,000
Does sorbent adversely impact fly ash sales? (0=no, 1=yes)	integer	0 or 1	1
Before Sorbent Injection, Fly Ash Sold (1) or Disposed of (2)	1 or 2		2
Revenue from Fly Ash Sales	\$/ton	0 to 35	\$6.00
Dry Waste Disposal Cost	\$/ton	1 to 25	\$6
Retrofit Factor			1.30
Maintenance Factors (% of Installed Cost)	%		5%
Process Contingency, % of process capital	%		5%
General Facilities (% of Installed Cost)	%		5%
Engineering Fees (% of Installed Cost)	%		10%
Project Contingency	%		15%
Duration of Project (years)	integer		1
<i>PJFF downstream of PAC Inputs</i>			
PJFF to COHPAC (i.e., TOXECON), 0=no, 1=yes	0 or 1		1
Cost of Bags, installed (\$/bag)	\$/bag		\$80
Estimated Number of Bags/MW	integer		20
Average bag life	years		5
Pressure Drop	iwc		8
Outlet Emissions	lb/MMBtu		0.012
Retrofit Difficulty Factor			1.30
Process Contingency, % of process capital	%		5%
General Facilities, % of Process Capital	% of process capital		5.0%
Engineering, Home Office, etc. % of Process Capital and General Facilities	%		10.0%
Project Contingency, % of Process Capital and Gen Facilities			15%
Owner's Overhead and costs			5.0%
Inventory Capital and Prepaid Royalties, etc.	% of process capital		1.0%
Maintenance, % of process capital and excluding bags			1%
Period of construction, yrs			1

B.5.6 Conclusions

Correlations for mercury removal from coal-fired power plants have been developed in the CUECost model, incorporating information on mercury removal from existing equipment that was developed from the ICR data in EPRI (2000). CUECost also incorporates mercury removal from injection of PAC, as developed from full-scale demonstrations of PAC injection where data are available. Algorithms developed with CUECost should be continuously updated and modified as more information becomes available on experience with mercury removal.

The following summarize some important findings that influence modeling mercury removal:

- The CUECost workbook that permits isolation of the effects of different air pollution control equipment on the fate of mercury will facilitate modeling combined effects with PAC injection over a wide range of boiler configurations and scenarios without the need for new regressions of PAC injection test data. Impact of a specific piece of equipment can be estimated with models best suited for that equipment.
- PAC injection followed by a fabric filter results in much lower injection concentrations being necessary for a given level of mercury reduction than for PAC injection followed by a cold-side ESP. Thus, economic modeling may show that in some cases the additional capital cost of a fabric filter may be justified by reduced operating costs associated with PAC consumption.
- Sorbent selection appears to have little effect on performance when PAC injection is followed by a fabric filter. But sorbent choice appears to have a significant effect when PAC injection is followed by an ESP.
- As demonstrated by the Salem Harbor test results, LOI and temperature can have a significant effect on the mercury removal by existing equipment. For this reason, the correlations of EPRI (2000), which do not include these effects, do not always provide an accurate indication of mercury removal by existing equipment.
- In some cases PAC injection without a downstream fabric filter may not be able to achieve very high mercury removal rates of 90% or more, regardless of PAC injection concentration.

B.6 CO₂ CONTROL DESIGN CRITERIA

In a monoethanolamine (MEA)-based CO₂ control system, a continuous scrubbing system is used to separate CO₂ from the flue gas. The system consists of an MEA island and a compressor island. The temperature of flue gas coming out the wet scrubber system is often higher than the temperature required by the MEA process. Therefore, in order for CO₂ to be efficiently scrubbed and to reduce solvent losses, the flue gas must be cooled down below 50 °C. As SO₂ reacts with MEA, the concentration of SO₂ prior to the absorber should be low

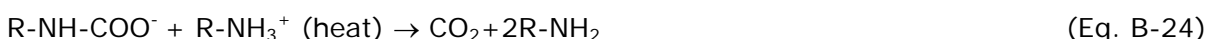
(<10 ppm) to reduce the losses and degradation of MEA by SO₂. NaOH scrubbing is often required before the absorber.

Flue gas then flows through the absorber where CO₂ binds to MEA. The CO₂-rich solution leaves the absorber and passes through the heat exchanger and finally enters the regenerator where CO₂ is released from MEA by external heat from steam supply or natural gas burning. The hot CO₂ lean solvent then flows back to the heat exchangers where it is cooled, and then is sent back to the absorber. To supplement the MEA losses, fresh MEA is added. Eqs. B-23 and B-24 show this cycle.

CO₂ absorption:



MEA regeneration:



The regeneration of MEA consumes a great deal of energy when the MEA concentration is low in the solvent. Inhibitors are therefore added to the solvent to increase the MEA concentration. In the worksheet, a typical MEA concentration is 30% with the addition of inhibitors.

In sorbent injection (SI) to capture CO₂, a continuous scrubbing system is used to separate CO₂ from the flue gas. The system consists of a sorbent absorption and regeneration island and a compressor island. The flue gas coming out of the wet scrubber system is cooled to relatively low temperatures (30 to 35 °C) for the easy capture of CO₂ by the sorbent. When SO₂ reacts with sorbent to degrade the sorbent, the concentration of SO₂ prior to the absorber should be lowered, in general <10 ppm, to minimize sorbent consumption. As such, additional scrubbing is required before the absorber. Flue gas then flows through the absorber where CO₂ binds to sorbents. The CO₂-rich sorbent leaves the absorber and passes through the heat exchanger in the regenerator where CO₂ is released from sorbent under the assistance of external heat.

In the CAP, CO₂ is absorbed in an ammoniated solution at 32 °F. Cooling the flue gas to such a low temperature is a necessary step within the process. As the result of flue gas cooling, moisture in the flue gas is also condensed, leading to less actual flue gas flow rate through the booster fan. In the absorption process, the formation of aqueous ammonium carbonate [(NH₄)₂CO₃] with the precipitation of ammonium bicarbonate [(NH₄)HCO₃] solids at low temperatures optimizes the energy demand, improves CO₂ removal efficiency, and reduces ammonia slip. The formation of ammonium bicarbonate solids is a reversible reaction. With heat in the regenerator, the ammonium bicarbonate solids are dissolved with eventual evolution of ammonia, water and CO₂ gases. The CO₂ stream leaves the regeneration vessel from the CAP at a higher pressure than the other two CO₂ processes

(MEA and SI) which results in fewer stages of downstream CO₂ compression. The ammonia and water reaction products are stripped and condensed from the resulting gas stream for reuse as reagent and flue gas wash solvent (Sherrick, B. 2008).

Gas exiting the regenerator must be compressed and dehydrated to accommodate transport and disposal. Moist CO₂ from the CO₂ regenerator's reflux drum enters the compressor at 21 °C (69 °F) and nominally 160 kPa (23 psi). CO₂ is compressed in a six-stage integrally geared compressor. Intercoolers between stages cool the gas using chilled water from the plants' cooling tower. After exiting the compressor, and presumably a final heat exchanger, the CO₂ is dried to < 20 ppm water in a triethylene glycol (TEG) dehydrator. Dry gas exiting the dehydrator is at 15.27 MPa (2215 psi) and 51 °C (124 °F) (DOE 2007).

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APPENDIX C DESIGN/ECONOMIC CRITERIA

C.1 GENERAL PLANT DESIGN CRITERIA

The plant design and operating default values provided below were taken from the criteria established by EPA's Integrated Air Pollution Control System (IAPCS) model (Gundappa et al., 1995) and were generally replaced with IPM/IECM values (Table C-1). The user can override any default value as long as the value input is within the range of the parameter stated on the worksheet. Table C-2 lists the coal analysis embedded in Sheet 11.0 Constant_CC (Coal Analysis Library). More information for coal analysis can be found from DOE coal bank and database (<http://datamine.ei.psu.edu/index.php>).

Table C-1. Snapshot for a Specific Plant and Its Default Parameters

Plant Information	Units	Range	Default
Cost Basis -Year (For Power Generation Estimation only)			2005
Location - State		All States	PA
Power Generation Technologies			1
<i>General Plant Factors</i>			
Gross Plant output	MW	500-800	580
Net Plant Output	MW	500-750	500
Plant Heat Rate	Btu/kWh		10,500
Plant Capacity Factor	%	40-90%	65%
Coal Type			5
Price of Coal	\$/MMBtu		2.05
<i>Other Operating Information</i>			
Percent Excess Air in Boiler	%		120%
Uncontrolled NO _x from Boiler			algorithm
Air Heater Inleakage	%		12%
Air Heater Outlet Gas Temperature	°F		300
Inlet Air Temperature	°F		80
Ambient Absolute Pressure	in. Hg		29.4
Pressure After Air Heater	in. H ₂ O		-12
Moisture in Air	lb/lb dry air		0.013
Ash Split:			
Fly Ash	%		80%
Conversion of SO ₂ to SO ₃			1.0%

Table C-2. Coal Analysis Library

		Coal 1, Wyoming PRB: 8,227 Btu, 0.37% S, 5.32% ash	Coal 2, Armstrong, PA: 13,100 Btu, 2.6% S, 9.1% ash	Coal 3, Jefferson, OH: 11,922 Btu, 3.43% S, 13% ash	Coal 4, Logan, WV: 12,058 Btu, 0.89% S, 16.6% ash	Coal 5, No. 6 Illinois: 10,100 Btu, 4% S, 16% ash	Coal 6, Rosebud, MT: 8,789 Btu, 0.56% S, 8.15% ash	Coal 7, Lignite, ND: 7,500 Btu, 0.94% S, 5.9% ash	Coal 8, DOE HS: 12,676 Btu, 3% S, 9% ash	Coal 9, DOE LS: 14,175 Btu, 0.6% S, 3.8% ash	Coal 10, DOE PRB: 8,304 Btu, 0.48% S, 6.4% ash	Coal 11, K Fuel: 11,718 Btu, 0.38% S, 6.42% ash	Coal 12, Med S: 11,570 Btu, 1.5% S, 8.15% ash
Go Back to Input Sheet													
COAL ANALYSIS LIBRARY													
Index Number		1	2	3	4	5	6	7	8	9	10	11	12
Coal Name		Wyoming PRB	Armstrong, PA	Jefferson, OH	Logan, WV	No. 6 Illinois	Rosebud, MT	Lignite, ND	DOE HS	DOE LS	DOE PRB	K Fuel	Med S
Coal Cost	\$/MMBtu	1.50	1.50	1.50	1.50	1.50	1.50	1.50	1.50	1.50	1.50	1.50	1.50
PROXIMATE ANALYSIS (ASTM, as rec'd)													
Moisture - Enter below in Ultimate Analysis													
Volatile Matter	wt%	31.39	36.20	37.20	35.40	33.00	36.40	42.00	40.40	44.00	30.79	40.20	36.19
Fixed Carbon	wt%	33.05	48.70	44.80	43.00	39.00	30.30	20.10	47.50	50.00	32.41	45.50	43.80
Ash - Enter below in Ultimate Analysis													
		100.00	100.00	100.00	100.00	100.00	100.05	100.00	100.00	100.00	100.00	99.62	100.00
COAL ULTIMATE ANALYSIS (ASTM, as rec'd)													
Moisture	wt%	30.24	6.00	5.00	5.00	12.00	25.20	32.00	3.10	2.20	30.40	7.50	11.86
Carbon	wt%	48.18	71.55	65.72	65.99	55.35	51.52	45.06	69.82	78.48	47.85	66.70	65.12
Hydrogen	wt%	3.31	4.88	4.53	4.75	4.00	3.29	2.80	5.00	5.50	3.40	4.80	4.22
Nitrogen	wt%	0.70	1.40	1.21	0.70	1.08	0.69	1.50	1.26	1.30	0.62	1.00	1.33
Chlorine	wt%	0.003	0.000	0.100	0.100	0.100	0.100	0.100	0.120	0.120	0.003	0.030	0.380
Sulfur	wt%	0.37	2.60	3.43	0.89	4.00	0.56	0.94	3.00	0.60	0.48	0.38	1.50
Ash	wt%	5.32	9.10	13.00	16.60	16.00	8.15	5.90	9.00	3.80	6.40	6.42	8.15
Oxygen	wt%	11.87	4.47	7.01	5.97	7.47	10.49	11.70	8.70	8.00	10.82	13.20	7.44
TOTAL	wt%	99.99	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	99.97	100.03	100.00
Mercury	mg/kg	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.07	0.04	0.10
Modified Mott Spooner HHV (B	Btu/lb	8,227	13,100	11,922	12,058	10,100	8,789	7,500	12,676	14,175	8,304	11,718	11,570
COAL ASH ANALYSIS (ASTM, as rec'd)													
SiO ₂	wt%	35.51	46.92	51.35	50.68	50.82	27.00	29.80	29.00	51.00	31.60	28.40	51.35
Al ₂ O ₃	wt%	17.11	21.00	30.00	29.00	19.06	19.00	10.00	17.00	30.00	15.30	17.30	30.00
TiO ₂	wt%	1.26	2.40	1.80	1.70	0.83	1.08	0.40	0.74	1.50	1.10	1.60	1.80
Fe ₂ O ₃	wt%	6.07	20.20	9.00	9.00	20.00	9.00	9.00	36.00	5.60	4.60	6.00	9.00
CaO	wt%	26.67	3.25	4.50	5.50	3.43	18.50	21.40	6.50	4.20	22.80	23.50	4.50
MgO	wt%	5.30	2.65	2.00	1.00	3.07	2.40	10.50	0.83	0.76	4.70	4.00	2.00
Na ₂ O	wt%	1.68	0.90	0.40	0.40	0.60	2.80	4.40	0.20	1.40	1.30	1.40	0.40
K ₂ O	wt%	2.87	0.30	0.20	0.90	0.37	0.45	0.49	1.20	0.40	0.40	0.27	0.20
P ₂ O ₅	wt%	0.97	0.00	0.16	0.60	0.17	0.42	0.00	0.22	1.80	0.80	2.43	0.16
SO ₃	wt%	1.56	1.38	0.59	1.22	1.22	18.85	14.01	7.30	2.60	16.60	13.63	0.59
Other Unaccounted for	wt%	1.00	1.00	0.00	0.00	0.43	0.50	0.00	1.01	0.74	0.80	1.47	0.00
TOTAL	wt%	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
		PRB	Bituminous	Bituminous	Bituminous	Bituminous	Subbituminous	Lignite	Bituminous	Bituminous	PRB	PRB	Bituminous

C.2 ECONOMIC CRITERIA

Economic inputs for CUECost workbook calculations are shown in Table C-3.

Table C-3. Economic Inputs

Description	Units	Range	Default
Cost Basis, Year Dollars	date		2008
Service Life (Levelization Period, years)	integer		30
Sales Tax Rate	%		6%
Escalation/Inflation Adjustment (GDP or Chem Index*)		C or D	GDP
Economic Factors During Construction Period			
Construction Labor Rate	\$/h		\$35
Prime Contractor's Markup	%		3%
Current Inflation Rate	%		2%
Current Escalation Rates	%		3%
After Tax Discount Rate (Current \$'s)	%		9%
Capital Carrying Charges			
First-year Carrying Charge (Current \$'s)	%	Calculator	16%
Levelized Carrying Charge (Constant \$'s)	%		8%
Non-Carrying Expense (O&M)			
Levelizing Factor (L30) (Constant \$'s)			1.48
Variable Cost Factors			
Operating Labor Rate (include benefit)	\$/h		\$25.0
Power Cost	Mills/kWh		60
Steam Cost	\$/1000 lbs		3.5
Demineralized Water	\$/lb		\$0.0030
Makeup Water	\$/1000 lb		\$0.05

* Chem Index = *Chemical Engineering Magazine* - Plant Index updated in each issue. This is the user input value for the year selected. The model divides the input value by the January 1998 index value to determine the escalation factor that is needed.

APPENDIX D

COST ALGORITHM DEVELOPMENT/VALIDATION/SOURCES

D.1 FGD COST ALGORITHM DEVELOPMENT

The cost algorithms associated with the flue gas desulfurization processes were developed based on historical data and new equipment quotations received by Raytheon during 1998 for some of the major equipment items. Algorithm development began with derivations from Raytheon's in-house historical database. These data sets were then modified by adding the additional data points from the new budgetary quotations, and then deriving new equations to represent the costs for equipment areas and for specific large pieces of equipment.

Performance data were sent to multiple vendors for one or two of the major equipment components identified in each cost area. These vendor contacts included a minimum of four vendors in each case. Responses to cost data requests were received from a minimum of one and normally three or more of the vendors solicited. Where vendor responses were limited due to refusals or delayed responses, additional data sources were obtained from recent projects to add to the data base of cost information for specific components. The cost data requests were made over the expected range of component sizes that could be used in the CUECost estimating workbook. LSFO capital cost algorithms are shown in Table D-1.

Table D-1. Variable and Constant Parameters for Wet FGD Cost Algorithm

	x =	Equation	A=	B=
LSFO Process Equipment	MW	$(x \times 1000 \times A \times x^{1.3})/1.3$	4456.5	-0.6442
ID Fans and Ductwork	Chimney acfm	$(A \times x + B)/1.3$	1.6225	3,000,000
Chimney	Chimney acfm	$(A \times x + B)/1.3$	3.4736	5,000,000
Support Equipment	MW	$= (0.0003 \times x^3 - 1.0667 \times x^2 + 1993.8 \times x + 1177674) \times 1.22$		

LSFO Process Equipment includes the Reagent Handling and Preparation, SO₂ Control System, and the Byproduct Handling. The capital costs for these equipments are described in Appendix B.1.

LSD capital cost algorithms are shown in Table D-2.

Table D-2. Parameters for LSD Cost Algorithms

	X=	Equation	A=	B=
Reagent Handling and Preparation	pph lime	$(Ax+B)/1.3$	136.84	3,000,000
SO₂ Control System	Inlet gas acfm	$(Ax+B)/1.3$	9.262	5,000,000
Byproduct Handling	pph byproduct (if upstream of existing ESP)	$(Ax+B)/1.3$	31.124	2,000,000
	<i>If new ESP or FF</i>	Determined by ESP and FF Worksheets		
Flue Gas Handling and ID Fans	acfm	$(Ax+B)/1.3$	2.9232	3.00E+06
Chimney Modification	KW	$(Ax+B)/1.3$	3.4	0
Support Equipment	MW	$(-1.211 \times x^2 + 2704.2 \times x + 1354716.2) \times 1.22$		

The capital costs for LSD Process Equipments are described in Appendix B.2.

D.2 SELECTIVE CATALYTIC REDUCTION

D.2.1 Performance Parameters

The key operating parameters that affect the performance and, consequently, the capital and operating costs of SCR systems include the allowable NH₃ slip emissions, the space velocity, the NO_x reduction efficiency, and the NH₃/NO_x molar ratio. For SCR systems these parameters are interrelated, and their values depend on the type of SCR application (high-dust or tail-end) and the desired performance levels. Ammonia slip emissions are controlled by the SCR system design. Typically SCR catalyst suppliers provide a guarantee of 2 ppm over the catalyst life. Since the 2 ppm NH₃ slip is guaranteed at the end of the catalyst's life, the initial NH₃ slip emissions will be very low (<1 ppm). Ammonia slip is not taken into consideration in the catalyst volume determination. The space velocity is the primary parameter used to specify catalyst volume. If the user does not input a value for space velocity, CUECost calculates it based on the NO_x reduction efficiency and the NH₃/NO_x molar ratio (molecular weight of NO_x = molecular weight of NO₂):

Space Velocity

$$SV = 6131.06 / 3 \times (n)^{-0.241} \times (\text{NH}_3:\text{NO}_x \text{ ratio})^{-2.306} \quad (\text{Eq. D-1})$$

where

SV = space velocity, 1/h

n = NO_x reduction efficiency, fraction

NH₃:NO_x ratio = stoichiometric ratio of NH₃ to NO_x.

The NO_x reduction efficiency (n) and molar ratio of NH₃ to NO_x (NH₃/NO_x ratio) are user-specific input values. The gross catalyst volume and NH₃ injection rate are determined from the following equations taken from IAPCS sources (Gundappa et al., 1995):

Ammonia Injection Rate

$$\text{NH}_3 = 3.702 \times 10^{-4} \times \text{NH}_3:\text{NO}_x \text{ ratio} \times \text{BSIZE} \times \text{HTR} \times \text{NO}_x \quad (\text{Eq. D-2})$$

where

NH₃ = ammonia injection rate, lb/h

BSIZE = boiler size, MW

HTR = net heat rate, Btu/kWh

NO_x = inlet NO_x emissions, lb/MMBtu

CV = gross catalyst volume, ft³

Q = flue gas volume flow rate, SCFH.

Gross Catalyst Volume

$$CV = Q / SV \quad (\text{Eq. D-3})$$

where

SV = space velocity.

D.2.2 Capital Costs

CUECost estimates capital costs for reactor housing, initial catalyst, ammonia storage and injection system, flue gas handling including ductwork and induced draft fan modifications, air preheater modifications and miscellaneous direct costs, including ash handling and water treatment additions. CUECost equations for SCR direct capital costs are shown below.

For all items except flue gas handling, cost algorithms are based on regression models developed for the Integrated Environmental Control Model (IECM) (Frey and Rubin, 1994). The IECM regression models were developed from cost data for 12 coal-fired power plants (Robie and Ireland, 1991). The flue gas handling cost algorithm is taken from the Integrated Air Pollution Control System (IAPCS) model, version 5.0 (Gundappa et al., 1995). Costs derived from the IAPCS equations for flue gas handling were found to be on

the same order of magnitude as costs reported by the Acid Rain Division study (EPA, 1997; EPA, 1998). IECM equations (Frey and Rubin, 1994) were used for the other direct capital cost items because they are based on more current cost data than IAPCS (Gundappa et al., 1995). Installation costs for items such as structural supports, foundations, concrete, earthwork are accounted for in the cost data used to develop the IECM and IAPCS equations and, therefore, are not a separate item in CUECost. Plant cost indices from *Chemical Engineering Magazine* are included in the equations to update direct capital costs. Direct capital costs for hot-side SCR are shown in Table D-3.

Table D-3. Direct Capital Costs for Hot-side SCR (Installed equipment costs)

<p><u>Reactor Housing</u></p> $DC_r = 18.65 \times N_{r,tot} \times (CV / N_{r,tot})^{0.489} \times 1000 \times RF \times PCI / 357.3$
<p><u>Ammonia Storage and Injection System</u></p> $DC_{NH_3} = 50.8 \times (NH_3)^{0.482} \times 1000 \times RF \times PCI / 357.3$
<p><u>Flue Gas Handling: Ductwork and Fans</u></p> $DC_{fgh} = 143.66 \times [G_{fg} \times (750+460) / (70+460)]^{0.694} \times RF \times PCI / 314.0$
<p><u>Air Preheater Modifications</u></p> $DC_{aph,mod} = 1370 \times N_{t,aph} \times (UA_{t,aph} / 4.4 / 10^6 / N_{t,aph})^{0.8} \times 1000 \times RF \times PCI / 357.3$
<p><u>Miscellaneous Direct Costs</u></p> $DC_{misc} = [100 + 300 \times (BSIZE / 550)^{0.6}] \times 1000 \times RF \times PCI / 357.3$

where

G_{fg} = flue gas volumetric flow rate for SCR ductwork, scfm

$N_{r,tot}$ = Number of SCR reactors

$N_{t,aph}$ = total number of air preheaters

RF = retrofit factor

PCI = chemical engineering plant cost index from *Chemical Engineering Magazine*
= 388 for 1998 dollars, 314.0 for 1982 dollars and 357.3 for IECM base year dollars

$UA_{t,aph}$ = product of universal heat transfer coefficient and heat exchanger surface area

$$= \frac{q_{aph}}{dt_{LM,aph}}, \text{ Btu/}^\circ\text{R}$$

$$q_{aph} = \text{heat transfer} = \frac{\text{Flue gas scfm} \times 60 \times 7.9 \times (T_{\text{flue gas, out}} - T_{\text{flue gas, in}})}{0.7302 \times 530}$$

$dt_{LM,aph}$ = log-mean temperature difference

$$= \frac{(T_{\text{flue gas, in}} - T_{\text{air, out}}) - (T_{\text{flue gas, out}} - T_{\text{air, in}})}{\text{LN}[(T_{\text{flue gas, in}} - T_{\text{air, out}}) / (T_{\text{flue gas, out}} - T_{\text{air, in}})]}$$

The flue gas inlet temperature ($T_{\text{flue gas, in}}$) and the outlet temperature ($T_{\text{flue gas, out}}$) are assumed to be the respective typical values of 725 and 600 °F.

Capital costs for instruments and controls, sales tax and freight are calculated from percentages of the equipment cost subtotal. The equipment cost subtotal is the sum of the equations listed above. For instruments and controls and freight, the respective default percentages are 2% and 5%. The sales tax rate is a user input value. The total direct cost is determined by applying the retrofit factor to the capital equipment cost subtotal, which is the sum of the equipment costs listed above as well as instruments and controls, sales tax and freight. The retrofit factor is a user input value that ranges from one for new applications to three for the most difficult retrofit cases. Equations for indirect capital costs are given in Table D-4.

Table D-4. Indirect Capital Costs for Hot-side SCR

$\text{General Facilities} = \text{Total Direct Cost with Retrofit} \times \text{General Facilities (\% of installed cost)}$
$\text{Engineering fees} = \text{Total Direct Cost with Retrofit} \times \text{Engineering Fees (\% of installed cost)}$
$\text{Contingency} = \text{Total Direct Cost with Retrofit} \times \text{Contingency (\% of installed cost)}$
$\text{Total Plant Investment} = \text{Sum of Total Direct Cost with Retrofit, General Facilities, Engineering fees, Contingency taking into account allowance for funds during construction}$
$\text{Preproduction} = \text{Total Plant Investment} \times 0.02 + \text{One month fixed operating costs} + \text{One month variable operating costs (at full capacity)}$
$\text{Initial Ammonia (60 days)} = \text{NH}_3 \times 24 \times \text{CF} \times 60 \times \text{UC}_{\text{NH}_3} / 2000$
$\text{Initial Catalyst} = \text{CV} \times \text{UC}_{\text{CAT}}$

where

CF = capacity factor, fraction

UC_{NH_3} = ammonia cost rate, \$/ton

UC_{CAT} = unit cost of catalyst, \$/ft³

CV = gross catalyst volume, ft³

NH_3 = Ammonia injection rate, lb/h.

D.2.3 Operating and Maintenance Costs

Operating and maintenance costs include NH_3 , catalyst replacement and disposal, electricity, steam, labor and maintenance costs. The CUECost operating and maintenance cost equations presented below are based on IAPCS equations (Gundappa et al., 1995). IAPCS equations were selected instead of IECM equations (Frey and Rubin, 1994) for operating and maintenance costs because the level of detail required for IAPCS input parameters was closer to that of other CUECost inputs. Additionally, the parameters affecting operation and maintenance costs are not likely to have changed significantly since the IAPCS equations were developed. With the exception of catalyst replacement costs, the equations from

IAPCS were derived from data reported by TVA for the high-dust system (Maxwell and Humphries, 1985). Annual catalyst replacement costs are based on the catalyst life. For example, if the catalyst life is 3 years, then one-third of the catalyst is replaced each year. The catalyst disposal cost reflects the cost of disposing of the spent catalyst. Catalyst disposal is typically included in the purchase cost of the catalyst. As a result, the recommended default for this line item is zero. However, an equation is included to allow the user to estimate a disposal cost, if applicable. A typical value of 48 lb/cubic foot was used for the catalyst density to calculate the mass of the spent catalyst. Operation and maintenance cost equations for SCR are shown Table D-5.

Table D-5. Operating and Maintenance Cost Equations for SCR (\$/year)

Ammonia Cost = $(8,760/2,000) \times (\text{NH}_3 \times \text{CF} \times \text{UC}_{\text{NH}_3})$
Catalyst Replacement Cost = $\text{CV}/\text{N} \times \text{UC}_{\text{CAT}}$
Catalyst Disposal Cost = $\frac{48 \times \text{Catalyst Replacement Cost} \times \text{UC}_{\text{WASTE}}}{2,000 \times \text{UC}_{\text{CAT}}}$
Electricity = $(-545,133 + 5.801 \times \text{G}) \times (\text{CF} / 0.628) \times \text{UC}_{\text{ELEC}}$
Steam = $(-14.91 + 33.29 \times \text{NH}_3 \times \text{CF}) \times \text{UC}_{\text{STEAM}}$
Operating Labor = $(1,341 + 5.363 \times \text{BSIZE}) \times \text{UC}_{\text{OL}}$
Maintenance Costs = $\text{Maintenance (\%)} \times \text{TPC}$

where

BSIZE= boiler size, MW_e

CF= capacity factor, fraction

CV= gross catalyst volume, ft³

G = flue gas flow rate, acfm

N = overall catalyst life, years

Maintenance (%) = annual maintenance cost as a percent of total plant cost

TPC = total direct and indirect capital costs, \$

UC_{CAT}= catalyst cost, \$/ft³

UC_{ELEC} = electricity rate, \$/kWh

UC_{OL} = operating labor wage, \$/person-h

UC_{NH3}= ammonia cost rate, \$/ton

UC_{STEAM} = steam rate, \$/MMBtu

UC_{WASTE} = solid waste disposal rate, \$/ton

NH₃= Ammonia injection rate, lb/h.

D.2.4 CUECost Validation

Total plant costs and operating and maintenance costs estimated by CUECost algorithms were compared to current cost data developed and validated by EPA's ARD. Cost and design information for four applications of SCR on various boiler types, boiler sizes and coals was taken from a 1996 Acid Rain Division (ARD) study (EPA, 1996) (Tables D-6 and D-7). The design information for these SCR applications was used to evaluate equations from CUECost. Total plant capital costs include the reactor housing, initial catalyst, ammonia storage and injection system, flue gas handling including ductwork and induced draft fan modifications, air preheater modifications and miscellaneous direct costs, including ash handling and water treatment additions. Other direct capital costs for taxes, freight, instruments and controls and initial inventory are included in the comparison of direct capital costs. The total plant cost includes direct costs listed above as well as indirect capital costs for engineering, general facilities and contingencies. Chemical engineering plant cost indices from *Chemical Engineering Magazine* were used to normalize costs in consistent year dollars.

The percent difference between ARD study costs and the CUECost estimates for total plant costs ranged from -4% to +8% for the cases evaluated. Operation and maintenance costs estimated by CUECost are 23 to 31% lower than those estimated by the ARD study. The largest difference appears to be the catalyst replacement cost.

Table D-6. CUECost with Acid Rain Division Study Design for SCR (1990 dollars)*

	Cyclone-Fired		Wet-Bottom		
			Vertical-Fired	Wall-Fired	
	Midwestern		Eastern Bituminous		
	Boiler Size (MW)				
Selective Catalytic Reduction	150	400	100	259	
CUECost with Acid Rain Division Design Parameters					
<u>Input Parameters Taken from Acid Rain Division Study</u>					
NO _x Reduction Efficiency	fraction	0.50	0.50	0.50	0.50
NH ₃ /NO _x Molar Ratio	fraction	0.50	0.50	0.50	0.51
Inlet NO _x	lbs/MMBtu	1.4	1.3	0.95	0.92
<u>Design Parameters Calculated by CUECost</u>					
Ammonia Injection Rate	lb/hr	340	884	155	399
Gross Catalyst Volume	ft ³	1,385	3,883	935	2,485
Flue Gas at Air Heater Outlet	SCFM	273,571	766,250	182,280	482,464
<u>Capital Costs Using Acid Rain Division Design Parameters (\$ 1000)</u>					
Reactor Housing and		1,188	1,967	981	1,582
Ammonia Handling and		1,097	1,739	752	1,185
Flue Gas Handling:Ductwork and Fans		2,238	4,574	1,689	3,318
Air Preheater Modifications		481	1,096	348	757
Misc. Other Direct Capital Costs		309	453	270	379
Initial Catalyst		485	1,359	327	870
Total Capital Equipment Cost		5,798	11,188	4,367	8,090
Freight, Sales Tax and Inst. & Controls		691	1,278	525	939
Total Plant Cost (TPC)		8,590	16,353	6,489	11,884
TPC (\$/kW)		57.3	40.9	64.9	45.9
<u>% Difference from Acid Rain Division Study</u>		4%	0%	8%	-4%
<u>O&M Costs using Acid Rain Division Design Parameters (\$1000/year)</u>					
Ammonia		157	407	72	184
Catalyst Replacement		162	453	109	290
Catalyst Disposal		0.10	0.28	0.07	0.18
Electricity		112	366	66	220
High-dust SCR Steam		34	88	15	40
<u>Maintenance</u>		122	225	92	165
O&M Total		586	1,539	354	899
<u>% Difference from Acid Rain Division Study</u>		-23%	-24%	-31%	-30%

* Source: EPA, 1997; EPA, 1998

Table D-7. Acid Rain Division Study: SCR Applications*

	Cyclone-Fired		Wet-Bottom Vertical- Wall-Fired Fired		
	Midwestern Bituminous		Eastern Bituminous		
	Boiler Size (MW)				
Selective Catalytic Reduction	150	400	100	259	
Acid Rain Division Costs and Design Parameters					
<u>Design Parameters from Acid Rain Division</u>					
NO _x Reduction Efficiency	fraction	0.50	0.50	0.50	0.50
NH ₃ /NO _x Molar Ratio	fraction	0.50	0.50	0.50	0.51
Inlet NO _x	lbs/MMBtu	1.4	1.3	0.95	0.92
Ammonia Injection Rate	lb/hr	339	882	155	398
Gross Catalyst Volume	ft ³	3,690	10,020	2,571	6,675
Flue Gas at Air Heater Outlet	SCFM	292,924	821,164	191,279	498,215
<u>Acid Rain Division Capital Costs (\$ 1000)</u>					
SCR Reactors/Ammonia Storage		3,180	7,040	2,150	4,921
Piping/Ductwork		945	1,600	860	1,528
Electrical/PLC		450	720	460	803
Draft Fans		1,065	1,760	650	1,166
Platform/Insulation/Enclosure		180	440	100	285
<u>Air Preheater Modifications</u>		<u>285</u>	<u>520</u>	<u>250</u>	<u>466</u>
Total Capital Equipment Cost		6,105	12,080	4,470	9,169
Total Plant Cost (TPC)		8,242	16,308	6,035	12,378
TPC (\$/kW)		55.05.0	40.8	60.4	47.8
<u>Acid Rain Division O&M Costs (\$ 1000/year)</u>					
Power Consumption		56	200	55	140
Ammonia Consumption		156	408	72	184
Catalyst Consumption		430	1,168	300	779
<u>General Maintenance</u>		<u>123</u>	<u>246</u>	<u>89</u>	<u>183</u>
O&M Total		764	2,023	516	1,286

* Source: EPA, 1997; EPA, 1998

D.3 SELECTIVE NONCATALYTIC REDUCTION

D.3.1 Performance Parameters

The CUECost workbook allows the user to select either urea [$\text{CO}(\text{NH}_2)_2$] or ammonia (NH_3) as the SNCR reagent. The user is asked to specify the NO_x reduction efficiency and the stoichiometric ratio of reagent to NO_x (molecular weight of NO_x = molecular weight of NO_2). The NH_3 and $\text{CO}(\text{NH}_2)_2$ injection rates in pounds of pure reagent per hour are then calculated based on the stoichiometric ratio, inlet NO_x and boiler heat input:

Urea Injection Rate

$$\text{Urea} = 6.5 \times 10^{-4} \times \text{UREA:NO}_x \text{ ratio} \times \text{BSIZE} \times \text{HTR} \times \text{NO}_x \quad (\text{Eq. D-4})$$

Ammonia Injection Rate

$$\text{NH}_3 = 3.702 \times 10^{-4} \times \text{NH}_3:\text{NO}_x \text{ ratio} \times \text{BSIZE} \times \text{HTR} \times \text{NO}_x \quad (\text{Eq. D-5})$$

where

Urea = $\text{CO}(\text{NH}_2)_2$ injection rate, lb/h

NH_3 = NH_3 injection rate, lb/h

BSIZE = boiler size, MW_e

HTR = net heat rate, Btu/kWh

$\text{NH}_3:\text{NO}_x$ ratio = stoichiometric ratio of NH_3 to NO_x

NO_x = inlet NO_x emissions, lb/MMBtu

UREA: NO_x ratio = normalized stoichiometric ratio of $\text{CO}(\text{NH}_2)_2$ to NO_x (i.e., moles of reagent nitrogen to moles of uncontrolled NO_x).

For the $\text{CO}(\text{NH}_2)_2$ -based SNCR process, the user may select to use wall injectors, lances, or both. Wall injectors are nozzles installed in the upper furnace waterwalls. In-furnace lances protrude into the upper furnace or convective pass and allow better mixing of the reagent with the flue gas. In-furnace lances require either an air- or water-cooling circulation system. Additionally, since the location of the temperature window changes with load, multiple levels of injectors and/or lances will be required for effective NO_x reduction over the operating load range of the boiler. If the user specifies a number of injector lance levels, but inputs zero for the number of injectors or lances, CUECost calculates the number of injectors or lances using the equations below:

$$N_I = (8.6 + 0.03 \times \text{BSIZE} - 0.013 \times \text{Red}) \times N_{IL} \quad (\text{Eq. D-6})$$

$$N_L = (2 + 0.013 \times \text{BSIZE}) \times N_{LL} \quad (\text{Eq. D-7})$$

where

N_I = number of wall injectors

Red = NO_x reduction efficiency, %

N_{IL} = number of injector levels

N_L = number of lances

BSIZE = boiler size, MW

N_{LL} = number of lance levels.

If the user enters values for both wall injectors and lances, then costs include both lances and wall injectors. If wall injectors are to be used alone, then the user enters zero for both the number of lance levels and the number of lances. Similarly, if lances are to be used alone, the user enters zero for both the number of injector levels and wall injectors. For the NH_3 -based SNCR process, the user can choose either steam or air as the atomizing medium. Based on the user's choice, an annual operating cost for steam and/or electricity usage is calculated.

D.3.2 Capital Costs

The main equipment areas in the battery limits for SNCR include the reagent receiving area, storage tanks, and recirculation system; the injection system, including injectors, pumps, valves, piping, and distribution modules; the control system; and air compressors. In addition, NH_3 -based SNCR systems use vaporizers to vaporize the NH_3 prior to injection. The capital costs are estimated using modified equations from IAPCS v.5.0 (Gundappa, 1995). The IAPCS equations were modified to incorporate the extensive current cost data developed and validated by EPA's ARD. IAPCS is a computer model developed for the EPA NRMRL-RTP (formerly the Air and Energy Engineering Research Laboratory) to estimate costs and performance for emission control systems applied to coal-fired utility boilers. IAPCS was developed in the 1980s and has been updated over the years. Documentation for the latest revision to IAPCS (Gundappa, 1995), completed in 1995, presents equations in 1982 dollars, with adjustments made using cost indices to normalize costs to other-year dollars.

Cost and design information was available in a 1996 ARD study (EPA, 1996) for six applications of urea-based (50% solution) SNCRs on various boiler types and sizes. The design information for these cases was input to the IAPCS model, and the capital cost estimates from IAPCS were compared to the ARD study estimates (EPA, 1996). The ratio of the ARD study costs to costs calculated using IAPCS equations was determined for each case. The ratios were then averaged, and the resulting average ratio was incorporated into each IAPCS capital cost equation. The ratios were determined for Total Direct Capital Cost. Itemization of equipment in major equipment areas varied between IAPCS and the ARD study so that unique ratios could not be established for each equipment area. As a result, the same ratio was added to each equipment cost equation. This approach was applied for both urea- and ammonia-based SNCR, because the capital costs do not vary significantly between the two processes (EPA, 1996). The algorithms for SNCR direct capital costs are presented below. Plant cost indices from *Chemical Engineering Magazine* are included in the equations to update direct capital costs. Direct capital costs for SNCR are shown in Table D-8.

Table D-8. Direct Capital Costs For SNCR (Installed Equipment Costs)

<u>Urea-Based SNCR Process</u>
Urea Storage & Handling = $38,143 \times (\text{Urea}/8.7)^{0.417} \times 0.915 \times \text{PCI} / 357.6$
Urea Injection = $(117,809 + 10,477 \times N_i + 53,111 \times N_L) \times 0.915 \times \text{PCI} / 357.6$
Misc. = $(96,082 + 106 \times \text{BSIZE} + 898 \times N_i + 2,433 \times N_L) \times 0.915 \times \text{PCI} / 357.6$
Air Heater Modifications = $11.2 \times (\text{acfm})^{0.772} \times 0.915 \times \text{PCI} / 357.6$
<u>Ammonia-Based SNCR Process</u>
Ammonia Storage = $63,822 \times (\text{BSIZE})^{0.6} \times 0.655 \times \text{PCI} / 357.6$
<u>Handling, Injection, Controls</u>
Air Heater Modifications = $11.2 \times (\text{acfm})^{0.772} \times 0.655 \times \text{PCI} / 357.6$

where

Urea = urea injection rate, lb/h

NH₃ = ammonia injection rate, lb/h

N_i = number of wall injectors

N_L = number of lances

acfm = flue gas volumetric flow rate at air heater inlet, ft³/min.

PCI = chemical engineering plant cost index from *Chemical Engineering Magazine*
= 388 for 1998 dollars and 357.6 for 1990 dollars.

Capital costs for instruments and controls, sales tax and freight are assumed to be included in the algorithms listed above because they are updated with ARD costs that include these items. The total direct cost with retrofit is determined by applying the retrofit factor to the capital equipment cost subtotal, which is the sum of the equipment costs listed above. The retrofit factor is a user input value that ranges from one for new applications to three for the most difficult retrofit cases. Equations for indirect capital costs are given in Table D-9.

Table D-9. Indirect Capital Costs for SNCR

<p>General Facilities = Total Direct Cost with Retrofit × General Facilities (% of installed cost)</p> <p><u>Engineering fees</u> = Total Direct Cost with Retrofit × Engineering Fees (% of installed cost)</p> <p><u>Contingency</u> = Total Direct Cost with Retrofit × Contingency (% of installed cost)</p> <p><u>Total Plant Investment</u> = Sum of Total Direct Cost with Retrofit, General Facilities, Engineering fees, Contingency taking into account allowance for funds during construction</p> <p><u>Preproduction</u> = Total Plant Investment × 0.02 + One Month Fixed Operating Costs + One Month Variable Operating Costs (at full capacity)</p> <p><u>Initial Ammonia (60 days)</u> = $\text{NH}_3 \times 24 \times \text{CF} \times 60 \times \text{UC}_{\text{NH}_3} / 2000$</p> <p><u>Initial Urea (60 days)</u> = $\text{NH}_3 \times 24 \times \text{CF} \times 60 \times \text{UC}_{\text{UREA}} / 2000$</p>
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where

CF = capacity factor, fraction

UC_{NH_3} = ammonia cost rate, \$/ton

UC_{UREA} = $\text{CO}(\text{NH}_2)_2$ cost rate, \$/ton.

D.3.3 Operating and Maintenance Costs

The operating and maintenance cost equations for SNCR, taken from IAPCS v.5.0 (Gundappa, 1995), are shown below. Equations for the urea- and ammonia-based processes are shown separately in the table. As in IAPCS, the operating labor costs are based on 2 person-hours required per 8-hour shift of operation. The default for maintenance labor and materials costs is 4% of the total direct and indirect capital cost. The annual cost of the reagent is the major operating cost item for the process and is calculated as the product of the reagent usage in tons/year and the cost in dollars per ton of pure reagent. Electricity, water, and steam requirements are based on vendor information. The increase in the energy requirement for steam or air atomization is included in the operating cost algorithms. Annual operating and maintenance costs for SNCR are shown in Table D-10.

Table D-10. Annual Operating and Maintenance Costs for SNCRUrea-Based SNCR Process (\$/year)

$$\text{Operating and Supervisory Labor} = 0.25 \times 8,760 \times UC_{OL}$$

$$\text{Maintenance Labor and Materials} = \text{Maintenance (\%)} \times \text{TPC}$$

$$\text{Reagent Requirement} = \text{Urea} \times 8760 \times \text{CF}/2,000 \times UC_{UREA}$$

$$\text{Electricity Requirement} = (5.97 + 0.29 \times N_I + 0.87 \times N_L) \times 8760 \times \text{CF} \times UC_{ELEC}$$

$$\text{Water Requirement} = (1.0 \times N_I + 2.5 \times N_L) \times 60 \times 8760 \times \text{CF}/1,000 \times UC_{H2O}$$

Ammonia-Based SNCR Process (\$/year)

$$\text{Operating and Supervisory Labor Requirement} = 0.25 \times 8,760 \times UC_{OL}$$

$$\text{Maintenance Labor and Materials Cost} = \text{Maintenance (\%)} \times \text{TPC}$$

$$\text{Reagent Requirement} = \text{NH}_3 \times 8760 \times \text{CF}/2,000 \times UC_{NH3}$$

$$\text{Steam Requirement (for steam atomization)} = \text{BSIZE} \times 99.2 \times 8,760 \times \text{CF}/1,000 \times UC_{STEAM}$$

$$\text{Electricity Requirement (for steam atomization)} = \text{BSIZE} \times 0.12 \times 8,760 \times \text{CF} \times UC_{ELEC}$$

$$\text{Electricity Requirement (for air atomization)} = \text{BSIZE} \times 4.23 \times 8,760 \times \text{CF} \times UC_{ELEC}$$

where

TPC = total direct and indirect capital costs, \$ (see Table 3-1)

UC_{ELEC} = electricity rate, \$/kWh

UC_{H2O} = unit cost water, \$/1,000 gallon

UC_{NH3} = NH_3 cost rate, \$/ton

UC_{STEAM} = steam rate, \$/MMBtu

UC_{UREA} = $\text{CO}(\text{NH}_2)_2$ cost rate, \$/ton.

D.3.4 CUECost Validation

To determine how successfully the IAPCS algorithms were modified using the ARD data, CUECost was run using the design information upon which the ARD cases were based. Total plant costs and operating and maintenance costs estimated using CUECost were compared to the costs developed by ARD (EPA, 1996). Results from this comparison are presented in Tables D-11 and D-12.

Total plant costs presented below include reagent storage and handling, injection system, air heater modifications, and miscellaneous direct capital costs. Total plant costs also include indirect capital costs such as engineering, general facilities and contingencies. *Chemical Engineering Magazine* plant cost indices were used to report costs in consistent year dollars.

The percent difference between ARD study costs and the CUECost estimates for total plant costs ranged from -15% to +7% for the cases evaluated. Operation and maintenance costs estimated by CUECost are 0 to 12% greater than those estimated by the ARD study (EPA, 1996).

D.4 NATURAL GAS REBURNING

D.4.1 Performance Parameters

The fraction of boiler heat input contributed by natural gas (reburn fraction) depends on the desired NO_x removal efficiency. The relationship between the reburn fraction and NO_x reduction efficiency, taken from IAPCS v.5.0, is based on vendor information and review of NGR performance data:

$$\text{RBFAC} = (\text{NO}_x\text{EFF} - 0.48)/0.86 \quad (\text{Eq. D-8})$$

where

RBFAC = boiler heat input contributed by natural gas (fraction)

NO_x EFF = NO_x reduction efficiency (fraction).

The relationship applies for NO_x reduction efficiencies from 55 to 65% and yields reburn fractions from 0.08 to 0.20. In CUECost, these are the valid input ranges for the NO_x removal efficiency and reburn fraction. If the user inputs both parameters within the valid ranges, the input values are used for cost calculations. If only one parameter is outside of the valid range, that parameter is calculated using the other parameter. If both input values are outside of the valid ranges, a default reburn fraction of 0.15 is used with a corresponding 61% NO_x removal efficiency.

Table D-11. CUECost with Acid Rain Division Study Cases for SNCR (1990 dollars) *

		Cyclone-Fired		Wet-Bottom		
				Vertical-Fired	Wall-Fired	
		Midwestern Bituminous		Eastern Bituminous		
				Boiler Size (MW)		
Selective Noncatalytic Reduction		150	400	100	259	
CUECost with Acid Rain Division Design Parameters						
Default Input Parameters						
Number of Injectors	integer	18	36	18	36	
Number of Lances	integer	0	0	0	0	
Urea/NOX Stoichiometric Ratio	fraction	0.90	0.90	0.90	0.90	
<u>Design Parameters calculated by CUECost</u>						
Urea Injection Rate	lb/hr	2,139	5,297	973	2,439	
Air Heater Inlet ACFM	ACFM	611,455	1,712,635	407,633	1,078,935	
<u>Capital Costs using Acid Rain Division Design Parameters (\$ 1000)</u>						
Urea Storage & Handling		451	658	324	476	
Urea Injection		364	589	364	589	
Controls/Miscellaneous		152	203	146	185	
<u>Air Heater Modifications</u>		<u>391</u>	<u>865</u>	<u>286</u>	<u>605</u>	
Total Capital Equipment Cost		1,358	2,314	1,120	1,855	
Total Plant Cost (TPC)		1,833	3,124	1,513	2,505	
TPC (\$/kW)		12.2	7.81	15.1	9.67	
% Difference from Acid Rain Division Cost Study		-7	7	-15	6	
<u>O&M Costs using Acid Rain Division Design Parameters (\$1000/year)</u>						
Operating and Supervisory Labor		46	46	46	46	
Maintenance Labor and Materials		27	47	23	38	
Reagent		1,102	2,730	501	1,257	
Electricity		3	5	3	5	
<u>Water</u>		<u>2</u>	<u>5</u>	<u>2</u>	<u>5</u>	
O&M Total		1,181	2,832	575	1,350	
<u>% Difference from Acid Rain Division Cost Study</u>		8	0	12	4	

* Source: EPA, 1996

Table D-12. Acid Rain Division Study: SNCR Applications (1990 dollars)*

		Cyclone-Fired		Wet-Bottom	
				Vertical-Fired	Wall-Fired
		Midwestern Bituminous		Eastern Bituminous	
Boiler Size (MW)					
Selective Noncatalytic Reduction		150	400	100	259
Acid Rain Division Costs and Design Parameters					
<u>Design Parameters from Acid Rain Division</u>					
Number of Injectors	integer	18	36	18	36
Number of Lances	integer	0	0	0	0
Urea/NOX Stoichiometric Ratio	fraction	0.90	0.90	0.90	0.90
Economizer Outlet	ACFM	648,029	1,812,657	416,969	1,085,858
<u>Acid Rain Division Capital Costs (\$ 1000)</u>					
Tanks, Pumps & Injectors		615	1,000	480	673
Pipes/Valves/Heat Tracing		510	680	530	725
Electrical/PLC		180	160	180	155
Platform/Insulation/Enclosure		135	280	90	155
Total Capital Equipment Cost		1,440	2,120	1,280	1,709
Total Plant Cost (TPC)		1,980	2,920	1,770	2,357
	TPC (\$/kW)	13.2	7.3	17.7	9.1
<u>Acid Rain Division O&M Costs (\$ 1000/year)</u>					
Coal Consumption		74	198	36	97
Power consumption		19	59	7	31
Ash Disposal		3	7	1	3
General Maintenance		31	48	27	37
Urea Consumption		961	2,494	437	1,119
Water Consumption		7	18	3	7
O&M Total		1,094	2,824	512	1,295

* Source: EPA, 1996

D.4.2 Capital Costs

Direct capital cost equations for NGR are presented below. The first equation includes the installed costs of gas injectors, OFA ports, and related equipment. This equation was developed by modifying the IAPCS equation for the same equipment area [Cost = $6,644,400 \times (\text{BSIZE}/500)^{0.214}$] to reflect recent cost estimates from an ARD study (EPA, 1996). The ARD study estimated NGR costs for four different boiler sizes. To bring the IAPCS model up to date, the constant in the equation (6,644,400) was replaced with a variable. Then the equation was set equal to each of the ARD cost cases, and the equation was solved to determine a new constant. The results showed that the new "constant" varied linearly with boiler size. Therefore, the constant in the IAPCS equation was replaced with an expression that is a function of boiler size ($\text{BSIZE} \times 3238 + 1504675$).

The second equation shown includes the costs associated with piping natural gas to the boiler from the metering station located at the utility plant fence line. The equation was derived by fitting an exponential curve to ARD costs for natural gas piping. Plant cost indices from *Chemical Engineering Magazine* are included in the equations to update direct capital costs. Direct capital costs for NGR are shown in Table D-13.

Table D-13. Direct Capital Costs for NGR (Installed equipment cost)

<p>Fuel injectors, overfire air ports, associated piping, valves, windbox, and control dampers</p> $= (\text{BSIZE} \times 3238 + 1504675) \times \left(\frac{\text{BSIZE}}{500} \right)^{0.217e} \times \frac{\text{PCI}}{357.6}$ <p>Gas pipeline from fence line to boiler = $372 \times \exp(2.64 \times 10^{-3} \times \text{BSIZE}) \times \frac{\text{PCI}}{357.6}$</p>

where

BSIZE = Boiler capacity (MW)

PCI = chemical engineering plant cost index from *Chemical Engineering Magazine*
 = 388 for 1998 dollars and 357.6 for 1990 dollars.

Capital costs for instruments and controls, sales tax and freight are assumed to be included in the algorithms listed above because they are updated with ARD costs that include these items. The total direct cost with retrofit is determined by applying the retrofit factor to the capital equipment cost subtotal, which is the sum of the equipment costs listed above. The retrofit factor is a user input value that ranges from 1 for new applications to 3 for the most difficult retrofit cases. Equations for indirect capital costs are given in Table D-14.

Table D-14. Indirect Capital Costs for NGR

<u>General Facilities</u> = Total Direct Cost with Retrofit × General Facilities (% of installed cost)
<u>Engineering fees</u> = Total Direct Cost with Retrofit × Engineering Fees (% of installed cost)
<u>Contingency</u> = Total Direct Cost with Retrofit × Contingency (% of installed cost)
<u>Total Plant Investment</u> = Sum of Total Direct Cost with Retrofit, General Facilities, Engineering fees, Contingency taking into account allowance for funds during construction
<u>Preproduction</u> = Total Plant Investment × 0.02 + One Month Fixed Operating Costs + One Month Variable Operating Costs (at full capacity)

D.4.3 Operating and Maintenance Costs

In general, natural gas reburning reduces the boiler operating costs associated with coal- and ash-handling process areas, including maintenance, electricity, and ash disposal. Fuel costs are generally higher, because the price of natural gas is typically higher than the price of coal per unit of energy. The equations used by CUECost and taken from IAPCS for estimating operating costs and savings are given below. The electricity requirement for coal and ash handling processes decreases in proportion to the amount of reburn fuel used. The default for maintenance costs for operating the NGR system is 1.5% of the total plant cost. The empirical equation for estimating waste disposal savings includes a reduction of bottom and fly ash as a result of firing gas. As in IAPCS, savings from reduced fly ash disposal are estimated only for retrofit applications. The incremental fuel cost for firing gas is estimated by multiplying the amount of gas burned by the fuel price difference between gas and coal. Annual operating and maintenance costs and savings for NGR are shown in Table D-15.

Table D-15. Annual Operating and Maintenance Costs and Savings for NGR

Electrical Consumption Savings (\$/year) $ELEC = 9.51 \times 10^7 \times Q_{in} \times CF \times RBFAC/HHV \times UC_{ELEC}$
<u>Maintenance Cost (\$/year)</u> $MAINT = \text{Maintenance (\%)} \times TPC - 1387.5 \times RBFAC \times (BSIZE/500)^{0.6}$
<u>Waste Disposal Savings (\$/year)</u> $WASTE = [BA \times RBFAC + (NR - 1) \times 4.336 \times RBFAC \times PPHPRT \times CF] \times UC_{WASTE}$
<u>Natural Gas Consumption Cost (\$/year)</u> $GAS = Q_{in} \times RBFAC \times 8,760 \times CF \times (UC_{GAS} - UCCO_{AL})$

where

Q_{in} = boiler heat input, MMBtu/h

CF = capacity factor, dimensionless

HHV = higher heating value of coal, Btu/lb

UC_{ELEC} = electricity rate, \$/kWh

TPC = total plant capital costs, \$

BA = bottom ash rate, tons/year estimated from:

$$BA = BAF \times ASH \times 500/HHV \times Q_{in} \times 8,760 \times CF/2,000$$

where

BAF = bottom ash factor, dimensionless

ASH = percent ash in coal, wt.%

NR = retrofit status, 1 for new "grass root" installation (retrofit factor = 1) and 2 for retrofit application (retrofit factor > 1)

PPHPRT = fly ash rate, lb/h

UC_{WASTE} = waste disposal rate, \$/ton

UC_{GAS} = gas rate, \$/MMBtu

$UCCO_{AL}$ = cost for coal, \$/MMBtu.

D.4.4 CUECost Validation

Total plant costs and operating and maintenance costs estimated by CUECost algorithms were compared to current cost data developed and validated by EPA's ARD (See Tables D-16 and D-17). Four applications of NGR for various boiler types, boiler sizes and coals were evaluated with CUECost. The design information provided by ARD for the four NGR applications was used to evaluate the direct capital cost equations from CUECost.

Total plant costs presented below include the fuel injectors, overfire air ports, associated piping, valves, windbox, and control dampers and the gas pipeline from the fence line to boiler. The total plant costs include direct costs listed above as well as indirect capital costs for engineering, general facilities and contingencies. *Chemical Engineering Magazine* plant cost indices were used to report costs in consistent year dollars.

The percent difference between ARD study costs and the CUECost estimates for total plant costs ranged from 0 to 11% for the cases evaluated. Operation and maintenance costs estimated by CUECost are 7 to 12% lower than those estimated by the ARD study.

Table D-16. CUECost with Acid Rain Division Study Cases for NGR (1990 dollars) *

	Cyclone-Fired		Wet-Bottom	
			Vertical-Fired	Wall-Fired
	Midwestern Bituminous		Eastern Bituminous	
	Boiler Size (MW)			
Natural Gas Reburning	150	400	100	259

CUECost with Acid Rain Division Design ParametersDesign Parameters from Acid Rain Division

Gas Reburn Fraction

Capital Costs using Acid Rain Division Design Parameters (\$ 1000)

Gas Pipeline from Fenceline to Boiler

Fuel Injectors, Overfire Air Ports and Associated Piping, Valves, Windbox and Control Dampers

	2,000	3,470	1,684	2,646
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Total Capital Equipment Cost	2,720	4,863	2,315	3,606
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Total Plant Cost (TPC)	3,590	6,419	3,056	4,760
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TPC(\$/kW)	23.9	16.1	30.6	18.4
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<u>% Difference from Acid Rain Division Cost Study</u>	11	6	0	2
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O&M Costs using Acid Rain Division Design Parameters (\$1000/year)

Electrical Consumption Savings	(54)	(152)	(34)	(89)
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Maintenance	54	96	46	71
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Waste Disposal Savings	(43)	(122)	(23)	(61)
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<u>Natural Gas Consumption</u>	<u>1,467</u>	<u>4,110</u>	<u>866</u>	<u>2,290</u>
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O&M	1,423	3,933	855	2,212
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<u>% Difference from Acid Rain Division Cost Study</u>	-11	-7	-12	-12
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* Source: EPA, 1996

Table D-17. Acid Rain Division Study: NGR Applications (1990 dollars)*

	Cyclone-Fired		CHAPTER 2 Vertical- Fired	WET- Wall-Fired
	Midwestern Bituminous		Eastern Bituminous	
	Boiler Size (MW)			
Natural Gas Reburning	150	400	100	259
Acid Rain Division Costs and Design Parameters				
<u>Design Parameters from Acid Rain Division</u>				
Gas Reburn Fraction	0.16	0.16	0.16	0.16
<u>Acid Rain Division Capital Costs (\$ 1000)</u>				
Fuel Piping System	510	1040	500	803
Burners/OFA	585	1840	540	1191
Electrical/BMS Modifications	735	1000	750	907
Windbox/Duct/Modifications	165	120	60	104
<u>Platform/Insulation/Demolition</u>	<u>405</u>	<u>520</u>	<u>410</u>	<u>466</u>
Total Capital Equipment Cost	2,400	4,520	2,260	3,471
Total Plant Cost (TPC)	3,225	6,080	3,050	4,662
TPC (\$/kW)	21.5	15.2	30.5	18.0
<u>Acid Rain Division O&M Costs (\$ 1000/year)</u>				
Coal Consumption	(1,630)	(4,564)	(1,201)	(3,184)
Ash Disposal	(50)	(141)	(27)	(71)
General Maintenance	50	93	47	71
<u>Natural Gas Consumption</u>	<u>3,239</u>	<u>8,848</u>	<u>2,150</u>	<u>5,694</u>
O&M Total	1,607	4,236	969	2,510

* Source: EPA, 1996

D.5 LOW-NO_x BURNER TECHNOLOGY

D.5.1 Capital Costs

CUECost estimates capital costs for retrofitting tangentially-fired and wall-fired boilers with LNBT. The cost algorithms are based on a study of LNBT by ARD (EPA, 1996). The study obtained information from 56 boilers--35 wall-fired and 21 tangentially-fired. The information provided for these retrofit cases was used to develop empirical equations that estimate total capital cost for LNBT retrofits as a function of boiler size. CUECost only addresses retrofit installations because most new boilers include LNBT in their base design.

The "bottom-line" costs include direct capital costs and indirect costs such as engineering, general facilities, and contingencies. The scope of direct costs collected for the ARD study includes (1) for the burner portion: burners or air and coal nozzles, burner throat and

waterwall modifications and windbox modifications; (2) for applicable combustion air staging: waterwall modifications or panels, windbox modifications, and ductwork; and (3) scope adders or supplemental equipment such as replacement or additional fans, dampers, or igniters necessary for the LNBT. The scope of installed LNBT retrofit capital costs includes materials, construction and installation labor, engineering, and overhead costs (40 CFR, Part 76, Appendix B).

The ARD study found that capital costs vary greatly depending on the scope of the retrofit and the degree of modification necessary. As a result, the cost data were statistically separated into subsets of high and low cost cases for each boiler type. Cost equations were then developed by ARD for the high and low cost subsets, as well as for the entire set of cost data. The CUECost user selects from any of the three ARD cost equations based on the estimated retrofitting difficulty: high, average or low. The equations are given in 1995 dollars and include the user input Chemical Engineering Magazine plant cost index (PCI) to escalate to the desired cost year. Total capital costs for LNBT retrofit are shown in Table D-18.

Table D-18. Total Capital Costs for LNBT Retrofit

<p><u>Tangential-fired Boilers</u></p> <p>High Cost: $57.04 \times (300/BSIZE)^{0.679} \times 1000 \times BSIZE \times PCI / 357.6$ Average Cost: $21.20 \times (300/BSIZE)^{0.35} \times 1000 \times BSIZE \times PCI / 357.6$ Low Cost: $11.71 \times 1000 \times BSIZE \times PCI / 357.6$</p> <p><u>Wall-fired Boilers</u></p> <p>High Cost: $27.72 \times (300/BSIZE)^{0.573} \times 1000 \times BSIZE \times PCI / 357.6$ Average Cost: $15.37 \times (300/BSIZE)^{0.35} \times 1000 \times BSIZE \times PCI / 357.6$ Low Cost: $6.53 \times (300/BSIZE)^{0.857} \times 1000 \times BSIZE \times PCI / 357.6$</p>

where

BSIZE = boiler size, MW

PCI = *Chemical Engineering* Plant Cost Index for desired cost basis year.

A cost comparison between CUECost and IAPCS cost algorithms was not possible because design and economic parameters were not given in the ARD study of NGR technology.

D.5.2 Operating and Maintenance Costs

The only direct operating costs associated with LNBT are for maintenance labor and materials. No energy penalty is assumed to be incurred with this technology. Costs for the controls, administration and support labor, including overhead, are 30% of the maintenance labor costs. Annual operating and maintenance costs for LNBT are shown in Table D-19.

Table D-19. Annual Operating and Maintenance Costs for LNBT (\$/year)

Maintenance Labor = TPC (\$) × Maintenance Labor (0.8%)
Maintenance Materials = TPC (\$) × Maintenance Materials (1.2%)
Administration/Overhead = Maintenance Labor (\$/year) × 30%

where

Maintenance Labor = Annual maintenance labor cost, \$/year

Maintenance Materials = Annual maintenance materials cost, \$/year

Administration/Overhead = Annual costs, \$/year

TPC = Total Plant Costs (\$).

D.5.3 CUECost Validation

Total plant costs estimated by CUECost for the four boiler sizes examined for the other NO_x technologies are shown in Table D-20. The CUECost algorithm for total plant cost is identical to the cost function presented by the ARD study of LNBT (EPA, 1996). A comparison is not presented for operating and maintenance costs because these costs are highly boiler-specific.

Table D-20. CUECost with Acid Rain Division Study Cases for LNBT (1990 dollars)*

	Boiler Size (MW)			
	150	400	100	259
Low NO_x Burner Technology	Average Case			
<u>CUECost Total Plant Cost (\$ 1000)</u>				
Wall-Fired	2,938	5,559	2,258	4,191
T-Fired	4,053	7,668	3,114	5,781
<u>% Difference from Acid Rain Division Study</u>				
Wall-Fired	0	0	0	0
T-Fired	0	0	0	0

* Source: EPA, 1996

D.6 HG CONTROL TECHNOLOGY

The algorithm of PAC control cost has the form

$$x = \text{MIN}(X, D) \quad (\text{Eq. D-9})$$

$$y = \text{Log}_{10}(\text{Injection Rate}) = A x^2 + B x + C \quad (\text{Eq. D-10})$$

where X is the mercury reduction fraction desired and the injection rate is expressed in lb/MMacf. A, B, and C are provided in the table below. D is used to specify the maximum fraction of mercury that can be removed, essentially an upper limit. In CUECost, D is actually multiplied by 0.99 so that the maximum removal that can be calculated equals 99% of D. Calculation results are shown in Figures D-1 through D-5.

Constants for Eqs. D-9 and D-10 are shown in Table D-21.

Table D-21. Constants for Eqs. D-9 and D-10

	A	B	C	D
PAC, Bituminous FF	1.6944	-1.1267	-0.0009	1.0
PAC, Bituminous ESP	-0.6647	2.1232	-0.0665	1.0
PAC, Subbituminous FF	-0.4318	1.9551	-0.8937	1.0
PAC, Subbituminous ESP	3.308	0.754	-0.5925	0.7
Treated PAC, Subbituminous FF	0.0	2.5007	-2.2097	1.0
Treated PAC, Subbituminous ESP	0.8837	0.4485	-0.575	1.0
Treated PAC, Bituminous ESP	0.0	1.207	-0.2277	1.0

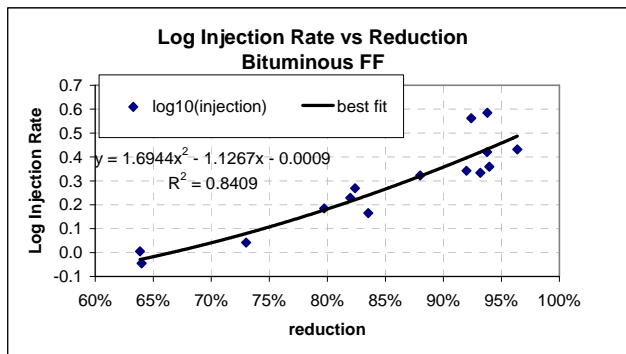


Figure D-1. PAC, Bituminous FF

$$y = 1.6944x^2 - 1.1267x - 0.0009$$

$$R^2 = 0.8409$$

$$x \leq 100\%$$

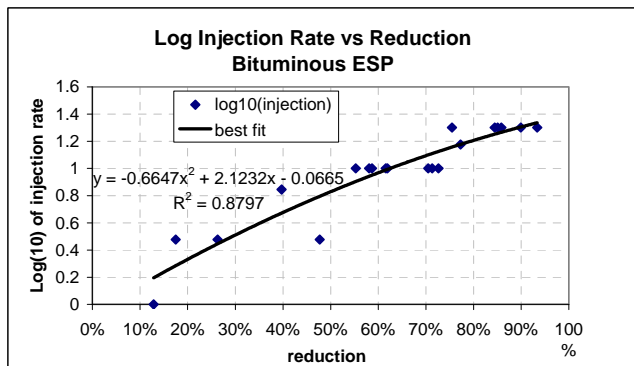


Figure D-2. PAC, Bituminous ESP

$$y = -0.6647x^2 + 2.1232x - 0.0665$$

$$R^2 = 0.8797$$

$$x \leq 100\%$$

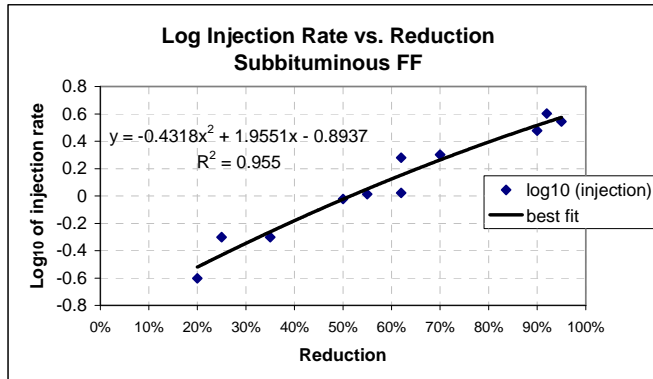


Figure D-3. PAC, Subbituminous FF

$$y = -0.4318x^2 + 1.9551x - 0.8937$$

$$R^2 = 0.955$$

$$x \leq 100\%$$

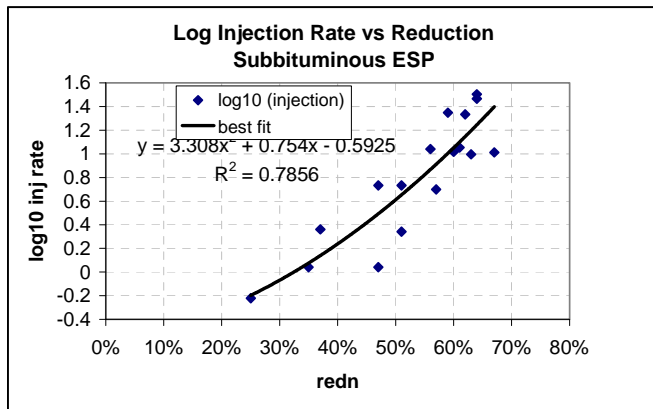


Figure D-4. PAC, Subbituminous ESP

$$y = 3.308x^2 + 0.754x - 0.5925$$

$$R^2 = 0.7856$$

$$x \leq 70\%$$

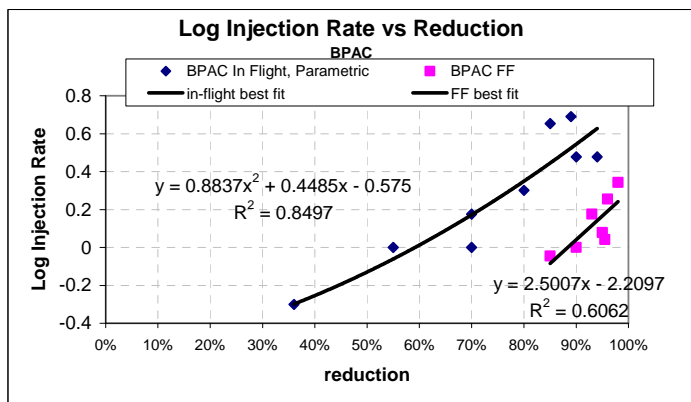


Figure D-5. BPAC

Treated PAC, Subbituminous FF

$$y = 2.5007x - 2.2097$$

$$R^2 = 0.6062$$

$$x \leq 100\%$$

Treated PAC, Subbituminous ESP

$$y = 0.8837x^2 + 0.4485x - 0.575$$

$$R^2 = 0.8497$$

$$x \leq 100\%$$

The algorithm of PAC control cost was incorporated into CUECost and a lookup table, located in the Constants_CC worksheet, and developed to ease users' selection. To use the lookup table in the CUECost workbook, a three-digit index key is constructed by summing the following digits:

- 100 or 200 for bituminous or subbituminous coal, respectively
- 10 or 20 for in-flight or filter capture, respectively
- 1, 2, or 3 for enhanced PAC (denote EPAC, such as brominated PAC), standard PAC, or other sorbent, respectively.

For example, a PRB coal fired boiler with a cold-side ESP and using enhanced PAC would have an index key of 211. If the same boiler were retrofit with a PJFF for a TOXECON arrangement and standard PAC were used, the index key would then be 222.

For the purpose of CUECost, subbituminous and lignite coals are treated the same way. For all practical purposes, the two categories are bituminous and low rank. In reality some bituminous coals with very low chlorine levels may behave more like low rank coals and some low rank coals with unusually high chlorine may behave more like bituminous coals. This issue will be addressed in the future.

Sample Calculations

To see how the new algorithms worked, some calculations of control cost were made. Figure D-6 shows comparison calculations for cost of controlling mercury for various situations on a 500 MW low sulfur bituminous coal-fired boiler equipped with an ESPc as a function of total mercury removal.

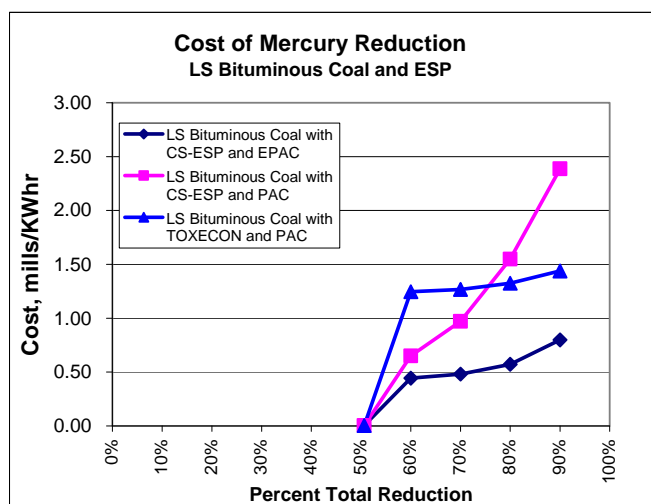


Figure D-6. Cost of Mercury Reduction, LS Bituminous Coal and ESP

In all calculations, addition of sorbent is assumed to cause fly ash in contact with sorbent to be disposed of at a differential cost of \$30/ton. Also, the cost of sorbent is assumed to be \$1000/ton for standard PAC and \$1500/ton for EPAC.

As shown in Figure D-6, EPAC incurs the lowest cost while maintaining the same level of Hg removal efficiency. For this reason, a TOXECON retrofit with PAC is not cost effective compared to EPAC. However, if fly ash is currently land filled, the differential disposal cost is negligible and an estimated 0.38 mills/kWh could be deducted from the cost of controlling mercury with sorbent injection upstream of an ESP (Staudt et al., 2003).

D.7 CO₂ MEA CONTROL SYSTEM COST ALGORITHM DEVELOPMENT

The cost algorithms associated with the CO₂ MEA control processes were developed based on DOE/NETL dataset in 2007 (DOE 2007). Algorithm development began with derivations from DOE/NETL database by running a series of data regressions and identified suitable equations. These datasets were then utilized to predict the cost by assuming the MEA concentration was at 30%. The derived regression equations represent a typical MEA operating plant for equipment areas and for specific O&M costs.

D.7.1 Capital Cost

The MEA island contains a pretreatment unit, a CO₂ absorber, and a CO₂ stripper. Costing has been based on the most recent DOE/NETL (2007) cost analysis of MEA CO₂ capture. MEA mainly reacts with two moles of amine and one mole of CO₂. Mitsubishi Heavy Industry (MHI) has developed a new solvent (named KS-1) primarily reacting with one mole of amine and one mole of CO₂. Little information has been found for the specific cost of the KS-1 based system. The capital costs found in MHI presentations did not provide sufficient detail to determine a comparable basis for Bare Erected Costs. Further, these cost estimates aggregated the recovery island and the compression island costs. As the KS-1 contains the same processes as the MEA, bare erected costs for the KS-1 island are therefore calculated in the same manner as the MEA island bare erected costs. As a regression of DOE dataset, the model MEA island cost will be:

$$Y = 69,412,748 \times X^{0.5741} \quad (\text{Eq. D-11})$$

where

Y = bare erected cost, 2007 \$

X = CO₂ capture, metric ton/h.

Gas exiting the CO₂ stripper must be compressed and dehydrated to accommodate transport and disposal. In DOE's model (DOE/NETL, 2007), moist CO₂ from the CO₂ stripper's reflux drum enters the compressor at 21 °C (69 °F) and nominally 160 kPa (23 psi). CO₂ is compressed in a 6-stage integrally geared compressor. Intercoolers between stages cool the gas using chilled water from the plant's cooling tower. After exiting the

compressor and presumably a final heat exchanger, the CO₂ is dried to <20 ppmv water in a TEG dehydrator. Dry gas exiting the dehydrator is at 15.27 MPa (2215 psi) and 51 °C (124 °F). Regression of the two coal-fired plants and one natural gas combustion plant cases presented results in a power law model for capital costs scaled to the power used for compression raised to 0.5429 power. This regression is based on a very limited data set. The uncertainty of the capital estimate increases as conditions deviate from those used in model development. As the result, the model compressor island cost will be

$$Y = 103,045 \times X^{0.5429} \quad (\text{Eq. D-12})$$

where

Y = bare erected cost, 2007 \$

X = compressor power, kW.

Indirect capital costs for CO₂ control are shown in Table D-22.

Table D-22. Indirect Capital Costs for CO₂ Control

General Facilities = Total Direct Cost with Retrofit × General Facilities (% of installed cost)
<u>Engineering fees</u> = Total Direct Cost with Retrofit × Engineering Fees (% of installed cost)
<u>Contingency</u> = Total Direct Cost with Retrofit × Contingency (% of installed cost)
<u>Total Plant Investment</u> = Sum of Total Direct Cost with Retrofit, General Facilities, Engineering fees, Contingency taking into account allowance for funds during construction
<u>Preproduction</u> = Total Plant Investment × 0.02 + One month fixed operating costs + One month variable operating costs (at full capacity)
<u>Inventory</u> = 0.5% Total Plant Cost (TPC)

D.7.2 Operating and Maintenance Costs

Steam

Steam is used in the reboiler of the CO₂ stripper to reverse the CO₂ reactions that took place in the CO₂ absorber. In addition to the heat required for CO₂ regeneration, some steam is used evaporating water in the stripper. In the recent DOE analysis (DOE/NETL, 2007), 1529 Btu/lb CO₂ were required to regenerate CO₂ in most cases for coal combustion while 1590 Btu/lb CO₂ were required for natural gas combined cycle (NGCC) (DOE, 2007). The major portion of this difference arises from the lower concentration of CO₂ in the NGCC gas. Rao (2002) reported a range of steam use of 3800-4000 kJ/kg CO₂ (1636-1723 Btu/lb CO₂). Steam use at the KS-1 installation at a Malaysia urea plant was 3270 kJ/kg CO₂ (1409 Btu/lb CO₂) with a feed gas containing 8% CO₂ on a dry basis. Data presented by DOE (2007) and Rao (2002) suggest that the steam requirement decreases with increasing CO₂ concentration. Steam consumption in the reboiler is estimated in this model based on a power law curve fit re-created from MHI's presentation of steam use. Assuming the CO₂

concentration is on a dry volumetric basis, this model predicts 3140 kJ/kg CO₂ steam consumption for an 8% CO₂ flue gas as documented for the Malaysia facility, a 4% error. Assuming there is no significant difference for different MEA processes, the steam use for regenerating MEA solvent in the worksheet will be regressed as:

$$Y = 4109.2 \times X^{-0.13} \quad (\text{Eq. D-13})$$

where

Y = energy demand, kJ/kg CO₂

X = CO₂ concentration, %.

Cooling Water Makeup

Cooling water will be used to remove heat from the direct contact cooler (DCC) during pretreatment, remove heat generated in the absorber, condense steam in the reflux drum of the CO₂ stripper, remove heat from the lean solvent returned from the CO₂ stripper, and remove heat generated by the compressor. As a budgetary estimation of cooling water makeup, we simplify the total use of cooling water makeup as:

Cooling water makeup = Loss due to DCC + Lump-sum Loss from MEA island + Loss from compressor island.

Enthalpy of the flue gas entering the direct contact cooler is calculated based on mass flows and temperatures exiting from the previous unit operation, for example, wet scrubbers. Enthalpy of the gas flow exiting the direct contact cooler is based on the mass flow exiting the direct contact cooler, assuming the exiting gas is saturated with water. Heat loss from the direct contact cooler is then calculated as the difference between the above two enthalpies. Although the pretreatment may also involve SO₂ polishing, this heat duty is expected to be inconsequential in comparison with the heat duty of condensing water vapor from the flue gas.

The MEA island cooling water requirement is estimated based on the steam requirement for the CO₂ stripper reboiler. The heat supplied to the reboiler is sufficient to reverse the CO₂ absorption, evaporate water and increase the enthalpy of the stripper effluent. The heat of reaction is removed in heat exchangers associated with the absorber. Steam is condensed in the stripper reflux drum and returned as reflux. Enthalpy of the stripper effluent in excess of the heat transferred to the stripper influent must be removed in a heat exchanger associated with the absorber; the lean solvent from the stripper is cooled to a lower temperature than the rich solvent effluent from the absorber. For simplicity, the heat input from steam will be equal to the heat rejected through cooling water evaporation.

Intercoolers are heat exchangers located between compressor stages with an intention to reduce the temperature of the gas, and, in turn, to protect the compressor from heat damage and reduce the power requirements. Chilled water is required for this purpose. The heat duty is assumed to be a fraction of power used by the compressors, as shown in Eq.

(D-14). The fraction is equal to the overall compressor efficiency which is equal to the isentropic efficiency of the compressor multiplied by the efficiency of the drive. The energy losses from the drives are assumed to flow to the surrounding environment, not to the CO₂ compression. In general, the isentropic efficiency is assumed to be 84% and the drive efficiency of the electric motor is 95%. Consequently, the overall efficiency is 80%.

$$Q = \frac{\eta w}{0.84} = 0.8 W \quad (\text{Eq. D-14})$$

The makeup cooling water flow rate (gpm) is equal to evaporation rate of water (approximately 2 gpm per 1 million Btu/h of heat) multiplied by an appropriate correction factor: 1/(cycle of concentration-1). The cycles of concentration = chlorides in tower water / chlorides in makeup water.

Power

The power used in the MEA Island is primarily consumed by an induced draft fan after the direct contact cooler. Pumps used to recirculate condensate and the MEA solvent represent the remainder of the MEA island power demand.

Power used in the induced draft fan is estimated based on the average volumetric flow entering and exiting the fan and the pressure differential across the fan. The recent DOE model (DOE, 2007) indicates a pressure differential of 0.014 MPa (2 psi) across this fan to overcome the pressure drop in the absorber. Gas is assumed to enter the fan saturated with water at 32 °C from the direct contact absorber. The recent DOE model (DOE, 2007) indicates a temperature rise across the fan of 17 °C for PC cases; an outlet temperature of 49 °C will be used in all cases. The pressure difference across the absorber with the MHI design using structured packing is substantially less than the power required with a randomly packed column. MHI claimed the pressure differential is 1/7 that of conventional MEA technology. Assuming isentropic compression and a k of 1.4, a temperature rise of 2 °C is estimated across the fan using MHI's design. The fan inlet temperature is therefore assumed to be 47 °C. The flue gas entering the fan is expected to be at nearly atmospheric pressure. An overall efficiency of 80% will be used to calculate the expected power requirement of the fan. Consequently, the power required by fan will be:

$$\text{Power (HP)} = \text{Gas flow (acfm)} \times \Delta P(\text{psi})/229/\text{efficiency} \quad (\text{Eq. D-15})$$

At this stage of estimation, the power for all the remaining pumps is estimated at 0.006 kWh/kg CO₂ removed, the average MEA island power use for PC units in the 2000 and 2004 analysis. Since the DOE 2007 analysis reflects similar steam requirements for PC units and NG-fired units, the loading of the MEA and the parasitic power is assumed to be similar for PC and NG-fired units. No new power requirements could be assessed from the DOE 2007 analysis since the fan power is added to the rest of the MEA system power requirements. Though a higher recirculation rate in the DCC is anticipated for the NG-fired units, this power consumption was disregarded at this level of analysis.

The larger share of the power consumed in the compressor island is associated with compression of the CO₂. At the study level of estimation, power required to run the TEG dehydrator is ignored. Power consumed by the intercoolers is contained in the power required for compression as pressure loss across the intercooler. The efficiency of the drive is included in the overall efficiency of the compressor.

Each stage of the compressor is nearly isentropic as there is limited surface area within the stage to remove heat in large compressors. Heat is, therefore, largely removed between stages by heat exchangers. Power consumption for compression is assumed to be isentropic in each stage with an efficiency factor applied to correct for non-ideal behavior of the compressor. Due to high pressures involved, the power estimate must account for the non-ideal behavior of the gas as well. The deviation from ideal gas behavior is corrected with a compressibility factor (Z). The estimation procedure used for Z is shown in Appendix F. An overall efficiency of 0.80 has been used consistent with the DOE (2007) analysis. The overall compression work is then calculated for each stage of compression as follows (Ulrich, 1984).

$$\frac{\eta w}{m} = \frac{RT_1 k \bar{Z}}{k-1} \left[\left(\frac{P_2}{P_1} \right)^{\frac{k-1}{k}} - 1 \right] \quad (\text{Eq. D-16})$$

where

P₁ = inlet pressure of the compressor

P₂ = outlet pressure of the compressor

T₁ = inlet gas temperature, K

R = 188.9 J/(kg × K) for CO₂,

k = 1.28 for CO₂. Ratio of constant pressure to constant volume heat capacity

m = flow rate of gas, kg/s.

η = 0.8, overall efficiency

Z = compressibility factor (See appendix F for calculation).

The power requirement for compression appears to decrease with each additional stage of compression. This work advantage is offset somewhat by the cost of the interstage coolers and the pressure drop between each of the stages. Because of the pressure drop across the interstage cooler, the pressure of the gas exiting a stage is slightly higher than the pressure of the gas entering the successive stage. At this level of estimation, a constant pressure drop of 0.01 MPa (1.5 psi) per cooler is assumed.

Generally, the power required and costs of a compressor are minimized when the same amount of compression work is accomplished in each stage. Since compressibility can vary from nearly 1 to 0.5 in these compressors, the compression ratio in installed equipment is likely to be different in each stage. This effect is most dramatic in the supercritical region where compressibility will be managed by controlling the stage inlet temperature. For computational simplicity, the pressure ratio of the final stage of compression is estimated to

be 30% higher than the other stages if the final pressure is supercritical. The pressure ratio for each of the preceding stages is evenly distributed prior to considering the pressure drop of the interstage cooler.

$$\frac{P_2}{P_1} = \left(\frac{P_f + 1.5n}{P_i} \right)^{1/n} \quad (\text{Eq. D-17})$$

The overall compressor efficiency includes the isentropic efficiency of the compressor and the efficiency of the drive. The work lost to the drive efficiency is assumed lost to the environment and not transferred to the CO₂. The isentropic efficiency is assumed to be 84% while the overall efficiency is assumed to be 80%; the resulting drive efficiency of the electric motor is 95%.

$$Q = \frac{\eta w}{0.84} \quad (\text{Eq. D-18})$$

During estimation, the average Z is 0.8 for the first n-1 stage compressors, and equal P₂/P₁ is also assumed for the first n-1 stage compressors.

Cost of MEA

MHI estimated MEA consumption at 0.45 kg MEA/metric ton CO₂ for its CO₂ capture technology. For the other MEA based processes, Rao (2002) reported a range of 0.5-3.1 kg/metric ton CO₂ and a typical value of 1.5 kg MEA/metric ton CO₂. For estimation purposes, the consumption rate of MEA employed in the worksheet will be at the typical value of 1.5 kg MEA/metric ton CO₂ for MEA process. Inhibitors are added to the absorber to prevent corrosion. The cost of inhibitors is estimated at 20% cost of MEA.

Cost of NaOH

Sodium hydroxide is used to bring down the SO₂ concentration in the influent gas to less than 10 ppm and to regenerate the MEA from the sulfate salts. The consumption of NaOH for SO₂ removal is based on the removal of SO₂ from the influent gas. For the consumption of NaOH in the reclaimer, no data on sodium hydroxide use were found in MHI papers or presentations for KS-1. Rao (2002) reported a typical value of 0.13 kg NaOH/metric ton CO₂. This value will be used in the worksheet.

Cost of Activated Carbon

Activated carbon is used to remove high molecular weight products. Rao (2002) reported a typical value of 0.075 kg carbon/metric ton CO₂. MHI claims a consumption of 0.06 kg carbon/metric ton CO₂. As the difference is not significant and the impact on the total cost is minor, a typical value of 0.075 kg carbon/metric ton CO₂ will be used in the worksheet.

D.8 CO₂ CAP CONTROL SYSTEM COST ALGORITHM DEVELOPMENT

The algorithm developed for CO₂ CAP control was based on the system description and data sets in the DOE/NETL report (DOE 2007). The algorithm for capital cost is developed through the comparisons with the MEA process. O&M cost is estimated based upon the system description by Sherrick et al. (2008).

D.8.1 Capital Cost

As with the MEA process, the CO₂ CAP control system contains a pretreatment unit, a CO₂ absorber, and a CO₂ stripper in the absorption island. The only cost information found to date for the CAP is based on analysis presented at Lyon, France in 2007 (EPRI, 2007). This analysis aggregates the total plant cost for a 3891 MMBtu/h supercritical pulverized coal (SCPC) plant with and without CAP and with MEA CO₂ capture.

For modeling purposes, the bare erected cost of the CAP separation island will be approximated as a fixed fraction of the modeled cost for a comparable MEA island. The total cost of the CAP CO₂ capture, CAP separation island and compression island is estimated based on the difference in TPC of the SCPC with CAP and the SCPC without CAP to be \$120 million. Though this approach includes plant effects other than the CO₂ capture system costs such as steam takeoffs, turbines, condenser, and chilled water systems, this shortcut is considered expedient in lieu of replicating the entire plant economic analysis. To return these costs to the bare erected cost basis, a constant escalation factor, 23% in aggregate, is applied. The estimated bare erected cost of the CAP CO₂ capture is \$97.6 million. The bare erected cost of the MEA system analyzed in the DOE/NETL report (2007) is \$111.8 million.

The capital costs associated with a CAP and MEA CO₂ capture are expected to be distributed differently between the separation island and the compression island. The compression island for the CAP is expected to be significantly cheaper than the compression island for MEA due to the high pressure, >400 psi, output of the CAP regenerator; the MEA regenerator is evaluated at 27.2 psia. The cost of the CAP separation island and the MEA separation island is estimated by subtracting the estimated cost of the respective compression islands from their respective bare erected costs of the CO₂ capture. The power requirement of the CAP compression island is estimated assuming 400 psia inlet pressure, 1217 psia outlet pressure, and 69 °F (21 °C) inlet temperature using a single stage compressor. Using the correlation developed from the DOE/NETL report (2007), the 6277 KW compressor estimated for the CAP CO₂ capture would have a bare erected cost of \$11.9 million in 2007, January 2000 bare erected costs are estimated at \$9.1 million. Using the correlation developed in the DOE/NETL report (2007), the 29730 kW compressor specified for the MEA CO₂ capture would have a bare erected cost of \$27.6 million in 2007. January 2000 bare erected costs are estimated at \$21.2 million.

The estimated bare erected cost of the CAP separation island is \$88.5 million. The estimated bare erected cost for the MEA separation island in the 2000 Parsons study is \$90.6 million.

The CAP separation island bare erected costs are estimated at 97.7% of the comparable cost of an MEA separation island.

Electricity

The main power consumers in the CAP separation island are expected to be the blower upstream of the absorber, recirculation pumps, and chiller. The Alstom analysis provides parasitic power for the overall study plant without providing details on power requirements for compressor, blower, pumps, or chillers. The relative power consumption of each function is expected to change with inlet gas composition and temperature. Since detailed information is not available, process conditions will be assumed to fix power consumption.

Power used in the induced draft fan is estimated based on the average volumetric flow entering and exiting the fan and the pressure differential across the fan. The recent DOE model (2007) indicates a pressure differential of 0.014 MPa (2 psi) across this fan to overcome the pressure drop in the absorber. Gas is assumed to enter the fan saturated with water at 32 °C from the direct contact absorber. The recent DOE model indicates a temperature rise across the fan of 17 °C for PC cases; an outlet temperature of 49 °C will be used in all cases. An 80% overall efficiency is assumed for the fan.

The power for pumping is assumed to be 0.006 kWh/kg CO₂ removed. This value was derived from the DOE/NETL report (2007) for MEA CO₂ capture.

The chiller is used to cool the flue gas exiting the ID fan prior to the absorber and to remove the heat of reaction. The flue gas is assumed to exit the fan at 49 °C (saturated at 32 °C) and will be cooled to saturation at 2 °C with the chiller. The heat load at the absorber will be approximated using the steam heating duty of the regenerator, 267 Btu/lb CO₂. Power use by the chiller will be approximated as ¼ total cooling duty of the chiller. The chiller is a mechanical chiller for removing heat.

Steam

Steam is used in the CAP CO₂ capture to reverse ammonium bicarbonate back to NH₃ and CO₂ (NH₃HCO₃+heat=NH₃+CO₂+H₂O). The Alstom analysis (DOE/NETL, 2007) uses 179,500 lb/h low pressure steam in the recovery of 710,423 lb/h of CO₂, 0.253 lb steam/lb CO₂. Assuming 1058 Btu/lb steam can be utilized, 267 Btu/lb CO₂ is required. The heat duty is very close to the heat of reaction suggesting minimal reflux in the regenerator and extraordinarily efficient heat transfer in the cross flow heat exchanger between the absorber and regenerator.

Cooling Water

The cooling water duty is obtained by subtracting the sum of the other energy flows out of the CAP island from the energy flows into the CAP island; the energy flows out of the CAP island must balance the energy flows into the CAP island.

Specific enthalpy of the flue gas entering the CAP island will be calculated based on temperature and composition information resulting from operation of the prior unit. Much of the water in the flue gas will be condensed in pretreatment contributing enthalpy to a CAP water balance.

Specific enthalpy of the flue gas exiting the CAP island will be calculated based on the mass balance composition assuming 90% CO₂ removal, 100% SO₂ removal, and water saturation; no significant removal of other gases is anticipated in the absorber. Though the exhaust temperature from the CAP island is not known, Alstom includes a second direct contact heat exchanger to recover some of the heat removed in preconditioning (DOE/NETL, 2007). The temperature of the gas exiting the CAP island is assumed to be the wet bulb temperature of the flue gas entering the CAP island.

CO₂ exiting the stripper reflux drum and leaving the CAP island is estimated to be pure CO₂ saturated with water. The temperature of the CO₂ exiting the stripper reflux drum in the MHI design was not known but is assumed to be 21 °C as found in the MEA island analysis.

Steam is primarily used in the reboiler to regenerate solvent and produce concentrated CO₂. Steam use is estimated by the net heat required for regeneration, based on the amount of CO₂ recovered, 267 Btu/lb CO₂.

The heat balance in the water streams is difficult to estimate with certainty because the amount of fresh makeup water added to scrub the absorber outlet gas is not known for the MHI design. It is not clear whether the scrubbing water is derived from the direct contact cooler or fresh makeup water. For estimation purposes, the water used for scrubbing the absorber outlet gas is assumed to be derived entirely from the direct contact cooler. The net amount of water condensed is therefore the difference in the water in the flue gas entering the CAP island and the water in the gases leaving the CAP island in the stack gas and CO₂ gas streams. The specific enthalpy of the net condensed water is estimated at 5 °C warmer than the wet bulb temperature of the flue gas entering the CAP island.

Work is transferred to the flue gas and working fluids through the action of blowers and pumps. All this work is assumed to be powered by electricity. The electric motors driving this equipment are assumed to be 95% efficient; 5% of the electric power used is assumed lost to the ambient environment and does not contribute to the energy balance around the CAP island.

D.9 CO₂ SI CONTROL SYSTEM COST ALGORITHM DEVELOPMENT

Sorbent-based CO₂ capture can be developed in a variety of configurations to conform to sorbent properties and market constraints. For this estimate, sorbent-based CO₂ capture is assumed to utilize an internally cooled moving bed reactor for CO₂ sorption. Sorbent regeneration is assumed to require indirect steam in a separate moving bed reactor. Parallel to MEA costs, sorbent-based CO₂ capture costs will be estimated with two islands: a sorbent island and a compressor island. Sorbent island costs will require inputs specific to the

sorbent system. As there is no base plant for comparison, the capital cost for the sorbent island will be estimated based on major components and modified to a comparable cost by MEA island. The costs for compressor island will be modeled in the same way as described in the MEA process.

The sorbent island consists of three major subsystems: preconditioning, sorption, and regeneration.

D.9.1 Preconditioning

Preconditioning is assumed to be required for most sorbent applications to prevent condensation on the sorbent. If the absorption temperature is less than 5 °C above the flue gas supply temperature, a DCC is assumed to be required to cool the gas and remove moisture. For absorption temperatures greater than 5 °C above the island inlet temperature, it is assumed that preconditioning is not required.

During pretreatment, water is circulated through a direct contact cooler resulting in condensation of water from the flue gas. The condensed water is recycled through a heat exchanger to reduce the water temperature and is then sprayed back into the direct contact cooler. A slip stream of condensed water is purged from the direct contact cooler prior to the heat exchanger. The heat duty of the heat exchanger is estimated as the difference between the enthalpy of the flue gas entering the direct contact cooler and the enthalpy of the gas exiting the direct contact cooler plus the enthalpy difference of the moisture condensed in the direct contact cooler. Enthalpy of the flue gas entering the direct contact cooler is calculated based on mass flows and temperatures exiting previous unit operation. Enthalpy of the gas flow exiting the direct contact cooler is based on the mass flow exiting the direct contact cooler assumed to be saturated with water at 5 °C less than the absorber temperature; a 35 °C absorber temperature would require a 30 °C DCC exhaust temperature. Enthalpy of the moisture condensed from the flue gas is based on the mass flow of moisture condensed and a temperature 5 °C lower than the wet bulb temperature of the flue gas entering the DCC. For applications after a wet FGD, the gas entering the DCC is essentially saturated and the inlet temperature is equal to the wet bulb temperature. Though the pretreatment may also involve SO₂ polishing, this heat duty is expected to be inconsequential in comparison with the heat duty of condensing water vapor from the flue gas.

Direct Contact Cooler

Water recirculating within the DCC is assumed to be cooled in a counter-current shell and tube heat exchanger. Cooling water is assumed to be available at 16 °C and is discharged from the heat exchanger at 27 °C. Condensed water enters the DCC 5 °C cooler than the inlet flue gas wet bulb temperature (T_{i-5}) and discharges 10 °C cooler than the absorber temperature (T_a-10). Using an overall heat transfer coefficient of 1200 J/(m² s °C), the surface area of the heat exchanger can be estimated as:

$$A(m^2) = \frac{\dot{Q}}{1200} \cdot \frac{\ln[(T_a - 10 - 16)/(T_i - 5 - 27)]}{(T_a - 10 - 16) - (T_i - 5 - 27)} \quad (\text{Eq. D-19})$$

\dot{Q} = heat duty in J/s

T_a = absorber temperature in °C

T_i = inlet flue gas wet bulb temperature in °C

The maximum surface area in a single unit is assumed to be 1000 m². The number of heat exchangers is estimated by dividing the total surface area by 1000 m² and rounding up to the next largest integer. The bare module cost of these heat exchangers can then be estimated using a power law:

$$\text{Cost}(2002\$) = n \cdot 80,000 \left(\frac{A/n}{1000} \right)^{0.86} \quad (\text{Eq. D-20})$$

where

A= surface area, m²

n=number of exchangers.

DCC Recirculation Pump

The amount of water recirculating through the heat exchangers and to the DCC is calculated based on the heat duty of the DCC and the temperature change of the water across the DCC.

$$\dot{q}(m^3 / s) = \frac{\dot{Q} \cdot (T_i - T_a + 5)}{4,187,000} \quad (\text{Eq. D-21})$$

where

T_a= temperature gas at the inlet of absorber, C

T_i=temperature of gas at the inlet of the direct contact cooler, C

Q= heat duty removed by the direct contact cooler, J/Q.

At this level of estimate, a single centrifugal pump is assumed to be associated with each heat exchanger.

$$(2002\$) = n \cdot 19200 \cdot \left(\frac{\dot{q}}{n} \right)^{0.40} \quad (\text{Eq. D-22})$$

Power consumption will depend greatly on the pressure drop through the nozzles and the type of pump selected. For costing purposes, power consumption will assume an 85% pump efficiency and a 95% drive efficiency and a 350 kPa (51 psi) pressure drop.

$$P(kW) = n \cdot \frac{\dot{q} \cdot 350}{0.85 \cdot 0.95} \quad (\text{Eq. D-23})$$

where

P=power consumption, kW.

D.9.2 Absorber

Though the absorber in a dry sorbent system may be engineered in a variety of configurations, at this time a countercurrent moving bed is assumed. Hot sorbent from the regenerator is assumed to be added to the top of the moving bed at the regeneration temperature and loaded sorbent removed from the bottom of the moving bed at the sorption temperature. Conditioned flue gas is assumed to enter the bottom of the reactor heated slightly above DCC exhaust due to the compression of a blower and to exhaust the absorber at the regeneration temperature. The absorber is assumed to be cooled with non-contact cooling water to cool the sorbent from regeneration temperature to absorption temperature and maintain the sorbent at absorption temperature during carbon capture.

Absorber Feed Conveyor

Due to the anticipated conveying capacity, 3-belt conveyors are assumed to collect, raise, and distribute the absorber feed for each conveying system. Each conveyor system is assumed to be limited to 0.66 m³/s and a 20° incline. Estimating the cost of the absorber feed conveyor requires estimation of the volume of sorbent to be fed and the conveying distance and height. The estimated volume of sorbent to be fed depends on estimated CO₂ loading and sorbent bulk density which demands the input by the user.

$$\dot{V}[m^3/s] = CO_2 \cdot L / \rho \quad (\text{Eq. D-24})$$

where

V=estimated volume of sorbent to be fed, m³/s

CO₂ = removal rate (kg CO₂/s)

L = sorbent loading (kg CO₂/kg sorbent)

ρ = bulk density (kg sorbent/m³).

Given the user supplied height [m], the length of the lifting conveyor is:

$$D_l = \frac{\Delta Z}{\sin(20^\circ)} \quad (\text{Eq. D-25})$$

where

ΔZ=supplied height, m

D_l=length of the lifting conveyor, m.

For estimation purposes, the collection and distribution conveyors are assumed to run the length of the absorber. The length of the absorber is estimated with the user supplied face velocity of the inlet flue gas and an assumed width of 6.1 m (20 ft).

$$D_c = D_d = \frac{V_a}{n \cdot F \cdot 6.1} \quad (\text{Eq. D-26})$$

where

D_c = length of collection, m

D_d = length of distribution, m

V_a = gas volumetric flow into absorber (m³/s)

n = number of conveyor systems

F = face velocity of gas (m/s).

Cost of the entire conveyor system, excluding drive, is estimated as:

$$(2002\$) = n \cdot [25000 + 2200 \cdot (D_c + D_l + D_d)] \quad (\text{Eq. D-27})$$

Conveyor Drives

Each conveyor is assumed to be driven by an electric motor. The power for the conveyor is estimated with the length, lift, and loading of the conveyor; conveyor speed is assumed constant for this estimate. A constant 80% efficiency is assumed for each drive. The power required for collection and distribution is assumed equal.

$$P_c = P_d = \left[3.91 + 0.07245 \cdot D_c + 0.0295 \cdot \left(0.4 + \frac{D_c}{91.42} \right) \cdot \dot{V} \cdot \rho \right] \quad (\text{Eq. D-28})$$

where

P_c = power for the collection conveyor, kW

P_d = power for the distribution conveyor, kW.

Power required for lifting

$$P_l = \left[3.91 + 0.07245 \cdot D_l + 0.0295 \cdot \left(0.4 + \frac{D_l}{91.45} \right) \cdot \dot{V} \cdot \rho + 0.00969 \cdot \Delta Z \cdot \dot{V} \cdot \rho \right] \quad (\text{Eq. D-29})$$

where

P_l = power of lifting conveyor, kW.

The total cost for drives for conveyers is estimated as:

$$(2002\$) = n \cdot (P_c + P_l + P_d) \cdot 100 \quad (\text{Eq. D-30})$$

Absorber

At this stage of estimation, the moving bed costs will be based on the cost of heat exchangers plus the cost of a shell. The cost of the shell will be estimated based on volume.

$$(2002\$) = 11300 \cdot \left(\frac{V_a \cdot \Delta Z}{F} \right)^{0.55} \quad (\text{Eq. D-31})$$

where

V_a = gas volumetric flow into absorber (m^3/s)

ΔZ = height of the absorber (m)

F = face velocity of gas (m/s).

Absorber costs are assumed to be driven by the heat removal requirement. Heat is removed with non-contact cooling water. The heat duty of the absorber is approximated by the heat of adsorption and the sensible heat of the sorbent less the sensible heat of the flue gas.

$$Q = Q_a + Q_s - Q_f \quad (\text{Eq. D-32})$$

where

Q = total heat, kJ

Q_a = adsorption heat, kJ

Q_s = sensible heat of sorbent, kJ

Q_f = sensible heat of the flue gas, kJ.

The heat of sorption is estimated base on the required CO_2 removal and the user supplied specific heat of sorption.

$$Q_a = \text{CO}_2 \cdot \Delta H \quad (\text{Eq. D-33})$$

where

Q_a = adsorption heat, kJ

CO_2 = CO_2 removal (kg/s)

ΔH = specific heat of sorption ($\text{J}/\text{kg CO}_2$).

The sensible heat of the sorbent is estimated using the user supplied sorbent heat capacity

$$Q_s = C_p \cdot \dot{V} \cdot \rho \cdot (T_r - T_a) \quad (\text{Eq. D-34})$$

where

Q_s = sensible heat of sorbent, kJ

C_p = specific heat capacity, kJ/kg

V = volume of sorbent per hour

T_r = Temperature at the outlet of regenerator, °C

T_a = Temperature at the outlet of the absorber, °C.

Flue gas enters the absorber at a temperature close to the absorption temperature and exits the absorber at regeneration temperature; there is heat removal associated with the exhausting flue gas. For estimating purposes, the sensible heat of the exhaust gas rising from the absorber inlet temperature to the regeneration temperature will be used to estimate this heat removal. The sensible heat should be summed for each component of the flue gas exhausting the absorber. The heat associated with SO_x, NO_x, and CO is expected to be minimal and/or unaffected by CO₂. Water condensation is likely in many sorption schemes. The condensed water is then evaporated during sorbent regeneration, which is, in turn, condensed in the reflux to produce nearly pure CO₂. The heat required and released from the absorber and regenerator can be canceled out, leaving only heat removals across only the reflux.

$$Q_f = \sum (C_p)_i \cdot \dot{m}_i \cdot (T_r - T_i) \quad (\text{Eq. D-35})$$

where

Q_f = sensible heat of the flue gas, kJ

C_p = the specific heat capacity of flue gas component., kJ/kg. °C

m = the mass of the gas component, kg

T_r = temperature of the flue gas out of the regenerator, °C

T_i = temperature of flue gas after the direct contact cooler, °C.

The heat exchanger surface areas are estimated from the heat removal requirement and the temperature driving force. The overall heat transfer coefficient is likely a function of the sorbent and the gas velocity. The user-specified heat transfer coefficient is assumed constant across the absorber for this estimate; a default of 250 J/(m²s °C) will be assumed. Potential for condensation on heat exchange tubes will be ignored for this estimate.

$$A = \frac{Q_a}{U} \cdot \frac{\ln[(T_a - 16)/(T_a - 27)]}{(27 - 16)} + \frac{(Q_s - Q_f)}{U} \cdot \frac{\ln[(T_r - 27)/(T_a - 16)]}{(T_r - 27) - (T_a - 16)} \quad (\text{Eq. D-36})$$

where

Q_a = adsorption heat, kJ

U = heat transfer coefficient, default at 250 J/(m² s °C)

Q_s = sensible heat of sorbent, kJ

Q_f = sensible heat of the flue gas, kJ

T_r = temperature of the flue gas out of the regenerator, °C

T_a = temperature of the flue gas out of the absorber, °C.

For costing purposes, the heat exchange is assumed to be performed in U-tube heat exchangers with a maximum surface area of 1000 m².

$$(2002\$) = n \cdot 360 \cdot \left(\frac{A}{n} \right)^{0.74} \quad (\text{Eq. D-37})$$

where

n = number of exchangers

A = total surface area of exchangers, m^2

(2002\$) = purchase price of the material in 2002 dollars.

D.9.3 Blower/ID Fan

At this stage of costing, a blower is specified based on the anticipated pressure drop across the absorber bed without additional consideration of bypassing or system failure. The pressure drop across the absorber, ΔP , will be estimated employing user provided head loss (Pa/m) across the sorbent and the user provided sorbent height (m) used in the absorber pricing. The blower is assumed to be adiabatic. At this stage, a constant heat capacity ratio of 1.4 will be used. For cases where the direct contact cooler is used to cool the gas to 5 °C below the absorber temperature, the fluid power is estimated:

$$W_f = \dot{m} \cdot R \cdot (T_a - 5 + 273) \cdot 3.5 \cdot \left[\left(\frac{P_i + \Delta P}{P_i} \right)^{0.286} - 1 \right] \quad (\text{Eq. D-38})$$

where

W_f = work performed by the fan, kW

\dot{m} = flow rate of the flue gas, kg

R = gas constant, $8.314, \text{JK}^{-1}\text{mol}^{-1}$

P_i = flue gas pressure before the absorber, pa/m

ΔP = head loss across the absorber, pa/m.

Purchase price of the blower is then estimated as:

$$(2002\$) = 3170 \cdot W_f^{0.60} \quad (\text{Eq. D-39})$$

The enthalpy (ΔH) increase of the gas is estimated using a constant efficiency of 80%.

$$\Delta \dot{H} = \frac{W_f}{0.80} \quad (\text{Eq. D-40})$$

The power consumption (P) will include drive efficiency, assumed to be 95%

$$P = \frac{\Delta\dot{H}}{0.95} \quad (\text{Eq. D-41})$$

D.9.4 Regenerator

Carbon dioxide will be recovered in a regenerator, analogous to a stripper for the MEA process. The gas recovered from the regenerator is expected to consist of nominally the CO₂ and water removed by the absorber without significant contamination. At this point, the regenerator is expected to rely on temperature swing with a regeneration temperature significantly higher than the sorption temperature. The regenerator, expected to be large and capable of accepting solid feed, is expected to operate at near atmospheric pressure. The loaded sorbent is expected to enter the top of the regenerator at the absorber temperature and be heated to near the regeneration temperature. For this estimate, exhaust gases are assumed to be withdrawn from below the top of the sorbent bed at regenerator temperature to avoid potential condensation issues. Lean sorbents are removed from the bottom of the regenerator at regeneration temperatures.

Regenerator Feed Conveyor

At this stage, the regenerator is assumed to be the same size and shape as the absorber. For cost estimates, the conveyors are assumed to cost the same as the absorber feed conveyors.

Conveyor Drives

The cost of conveyor drives is estimated in the same manner as the absorber conveyor drives. The sizes of the conveyor drives are expected to be slightly larger than estimated due to the mass of CO₂ and water absorbed on the sorbent. The volume of the sorbent is assumed to remain unchanged while the density increases. Therefore the product of volumetric flow and density, the mass flow, is equal to the mass flow of absorbent feed plus the CO₂ and water absorbed.

Regenerator

The regenerator, comprised of shell and heat exchanger, is priced in the same way as the absorber. The shell price is estimated to be the same as the absorber shell since they are assumed to be the same dimensions.

$$(2002\$) = 11300 \cdot \left(\frac{V_a \cdot \Delta Z}{F} \right)^{0.55} \quad (\text{Eq. D-42})$$

where

V_a = gas volumetric flow into regenerator, m³/s

ΔZ = height of the regenerator, m

F = face velocity of gas, m/s.

The heat exchanger required to reverse the heat of absorption is calculated as

$$Q_r = Q_a = CO_2 \cdot \Delta H \quad (\text{Eq. D-43})$$

where

CO_2 = CO_2 removal, kg/s

ΔH = specific heat of sorption, J/kg CO_2

Q_r = heat required for sorbent regeneration, kJ

Q_a = heat released for sorbent absorption, kJ.

The heat exchangers are required to (1) heat the sorbent and absorbed CO_2 and water (as gases) to regeneration temperatures and (2) to reverse the adsorption process.

$$Q_s = [C_p \cdot \dot{V} \cdot \rho + \sum (C_p)_i \cdot \dot{m}_i] \cdot (T_r - T_a) \quad (\text{Eq. D-44})$$

where

C_p = specific heat capacity, kJ/kg. °C

V = gas volumetric flow into regenerator, m^3/s

ρ = density of the gas, kg/m^3

m = mass of the sorbent

T_r = temperature of gas at outlet of the regenerator, °C

T_a = temperature of gas at outlet of the absorber, °C.

At this stage of estimation, heat is assumed to be provided from saturated condensing steam at 150 °C. The overall heat transfer coefficient is assumed to be equal to the overall heat transfer coefficient used in the absorber.

$$A = \frac{Q_r}{U \cdot (150 - T_r)} + \frac{Q_s \cdot \ln[(150 - T_a)/(150 - T_r)]}{U \cdot (T_r - T_a)} \quad (\text{Eq. D-45})$$

where

A = surface area of the exchangers, m^2

T_r = temperature of gas at outlet of the regenerator, °C

T_a = temperature of gas at outlet of the absorber, °C

Q_r = heat required for sorbent regeneration, kJ

Q_a = heat released for sorbent absorption, kJ

U = heat transfer coefficient, default at 250 $J/(m^2 \text{ s } ^\circ C)$.

For costing purposes, the heat exchange is assumed to be performed in U-tube heat exchangers with a maximum surface area of 1000 m^2 .

$$(2002\$) = n \cdot 360 \cdot \left(\frac{A}{n} \right)^{0.74} \quad (\text{Eq. D-46})$$

Gas Cooler

A cooler/partial condenser will bring the CO₂ and steam down to a low temperature ahead of compression. For this estimate, the gas exit temperature is assumed to be 21 °C. The amount of heat removed is equal to the sensible heat of cooling the CO₂ and water vapor plus the latent heat of condensing the water, 2.541 10⁶ J/kg.

$$Q = 2.541 \cdot 10^6 \cdot \dot{m}_{water} + \sum (C_p)_i \cdot \dot{m}_i \cdot (T_r - 21) \quad (\text{Eq. D-47})$$

m_i = mass of the sorbent

T_r = temperature of gas at outlet of the regenerator, °C

Q = heat removed by the exchanger, kJ/s

Cooling water is assumed to enter the gas cooler at 16 °C and exit at 27 °C. For this estimate, an overall heat transfer coefficient of 250 J/(m²s°C) is assumed. The heat exchange surface area is estimated assuming countercurrent flow

$$A = \frac{Q \cdot \ln[(T_r - 27)/5]}{U \cdot (T_r - 32)} \quad (\text{Eq. D-48})$$

where

A = surface area of the exchangers, m²

T_r = temperature of gas at outlet of the regenerator, °C

Q = heat removed by the exchanger, kJ/s

U = heat transfer coefficient, default at 250 J/(m² s °C).

The maximum surface area in a single unit is assumed to be 1000 m². The number of heat exchangers is estimated by dividing the total surface area of a unit by 1000 m² and rounding up to the next largest integer. The bare module cost of these heat exchangers can then be estimated using a power law:

$$\text{Cost}(2002\$) = n \cdot 80,000 \left(\frac{A/n}{1000} \right)^{0.86} \quad (\text{Eq. D-49})$$

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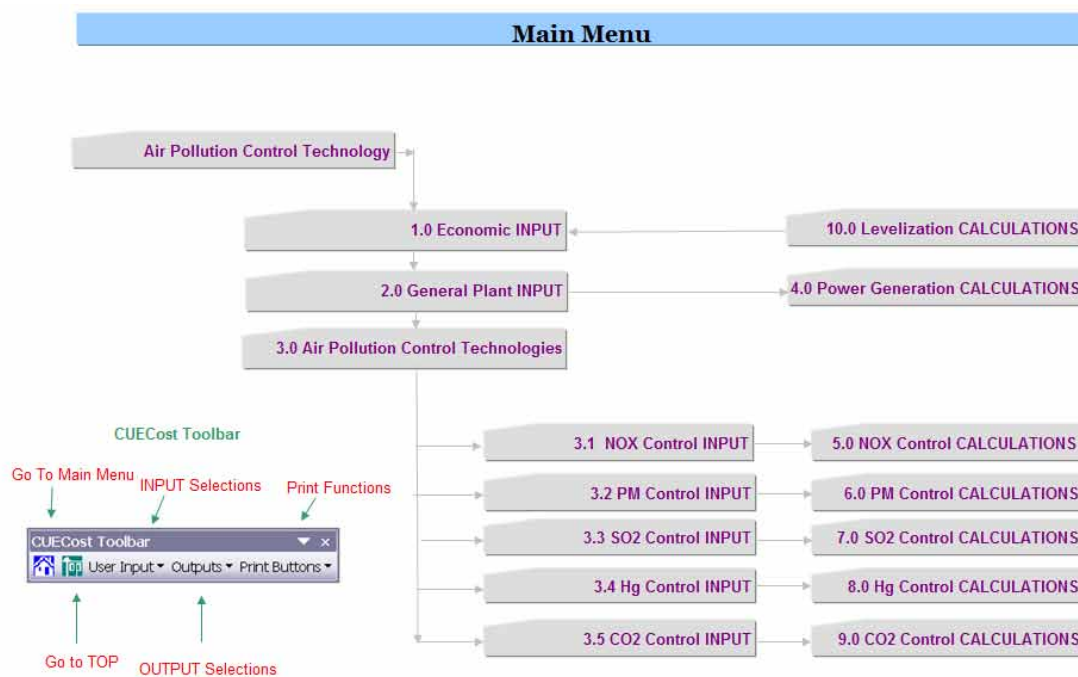
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APPENDIX E INPUT WORKSHEET SCREENS

E.1 GETTING STARTED

After downloading the workbook to the hard drive, the first thing to do is to create a copy of the workbook and save it under a different name. Once the workbook has been saved to the hard drive, it can be opened using Microsoft Excel 5.0 or a newer version of Excel.

The workbook will originally open to the "1.0 General Input Sheet". This is the worksheet where all of the necessary inputs are entered. However, a main menu is created for the user where all of the sheets are linked with buttons. The screen the user will encounter is:



E.2 INPUTS

As the user proceeds down following the menu, (s)he will encounter the following worksheets:

- Power Generation Technique Choices
- Air Pollution Control (APC) Technology Choices
- General Plant Technical Inputs
- Economic Inputs
- Limestone Forced Oxidation (LSFO) Inputs
- Lime Spray Dryer (LSD) Inputs
- Particulate Control Inputs
- NO_x Control Inputs
- Hg Control Technology Inputs
- CO₂ Control Amine Technology Input

E.2.1 Economic Inputs

This is the area of the worksheet where the economic factors are input. These factors are used in developing the capital and O&M costs for the control technologies.

Item/Description	Units	Range	Default	Case 1	Case 2	Case 3	Case 4
Economic Factor							
Cost Basis -Year Dollars	year		2006	D	D	D	D
Service Life (Levelization Period)	years		30	D	D	D	D
Sales Tax	%		6%	D	D	D	D
Escalation/Inflation Adjustment (GDP or Chem Index)		C or D	GDP	D	D	D	D
Construction Labor Rate	\$/h		\$35	D	D	D	D
Prime Contractor's Markup	%		3%	D	D	D	D
Inflation Rate	%		2%	D	D	D	D
Escalation Rate	%		3%	D	D	D	D
	%		9%	D	D	D	D
Capital Carrying Charges							
First-year Carrying Charge (Current \$'s)	%	Calculator	16%	16%	16%	16%	16%
Levelized Carrying Charge (Constant \$'s)	%		8%	8%	8%	8%	8%
Non-Carrying Expense (O&M)							
Levelizing Factor (L30) (Constant \$'s)			1.49	1.488	1.488	1.488	1.488
Variable Cost Factors							
Operating Labor Rate (include benefit)	\$/h		\$25.0	D	D	D	D
Power Cost	Mills/kWh		60	D	D	D	D
Steam Cost	\$/1000 lbs		3.5	D	D	D	D
Demineralized Water	\$/lb		\$0.0030	D	D	D	D
MakeUp Water	\$/1000 lb		\$0.05	D	D	D	D

User needs to click the icon of calculator, a quick worksheet pops up to facilitate calculating carrying charges and levelization factors.

INPUT

Factors during Operation (for Carrying Charge and levelization)										
Return of Debt	%	5%	5%	5%	5%	5%	5%	5%	5%	5%
Ratio of Debt	%	50%	50%	50%	50%	50%	50%	50%	50%	50%
Return of Equity	%	10%	10%	10%	10%	10%	10%	10%	10%	10%
Ratio of Equity	%	50%	50%	50%	50%	50%	50%	50%	50%	50%
Property Taxes and Insurance	%	2%	2%	2%	2%	2%	2%	2%	2%	2%
Income Tax	%	50%	50%	50%	50%	50%	50%	50%	50%	50%
Investment Tax Credit	%	10%	10%	10%	10%	10%	10%	10%	10%	10%
Book life of the Plant	Year	30	30	30	30	30	30	30	30	30
Average (price) Inflation rate (Long-Term)	%	3%	3%	3%	3%	3%	3%	3%	3%	3%
Depreciation method (1= 30 Yr Straight Line; 2=30 Yr Straight Line;3=20 Yr ARCS Schedule)		3	3	3	3	3	3	9	3	3
Consumables (O&M) Escalation	%	3%	3%	3%	3%	3%	3%	3%	3%	3%

OUTPUT

Carrying Charges										
First-year Carrying Charge (current \$'s)*	%	15.9%	15.9%	15.9%	15.9%	15.9%	15.9%	15.9%	15.9%	15.9%
Levelized Carrying Charge (current \$'s)	%	11.4%	11.4%	11.4%	11.4%	11.4%	11.4%	11.4%	11.4%	11.4%
First-year Carrying Charge (constant \$'s)	%	11.7%	11.7%	11.7%	11.7%	11.7%	11.7%	11.7%	11.7%	11.7%
Levelized Carrying Charge (constant \$'s)*	%	8.3%	8.3%	8.3%	8.3%	8.3%	8.3%	8.3%	8.3%	8.3%
Non-Carrying Expense										
Levelizing Factor (L30) (Current \$'s) (for O&M)		2.08	2.08	2.08	2.08	2.08	2.08	2.08	2.08	2.08
Levelizing Factor (L30) (Constant \$'s) (for O&M)*		1.49	1.49	1.49	1.49	1.49	1.49	1.49	1.49	1.49

E.2.2 Power Generation Technique Choices

This is the area of the worksheet where the user can choose what power generation technique will be evaluated. The following screen shows how this area looks and what options are available.

2.0 Power Generation Technology

Item/Description	Units	Range	Default	Case 1	Case 2	Case 3
Plant Information						
Cost Basis -Year (For Power Generation Estimation only)			2005	D	D	D
Location - State		All States	PA	D	D	D
Power Generation Technologies			1	3	2	2
				(1=Subcritical; 2=SuperCritical; 3=Ultra-SuperCritical)		
General Plant Factors						
Gross Plant output	MW	500-800	580	580	580	580
Net Plant Output	MW	500-750	500	D	D	D
Plant Heat Rate	Btu/kWh		10,500	D	D	D
Plant Capacity Factor	%	40-90%	65%	D	D	D
Coal Type		Goto Coal Data	5	2	2	2
				See Coal Types		
Price of Coal	\$/MMBtu		2.05	D	D	D
Other Operating Information						
Percent Excess Air in Boiler	%		120%	D	D	D
Uncontrolled NOx from Boiler			algorithm	D	D	D
Air Heater Inleakage	%		12%	D	D	D
Air Heater Outlet Gas Temperature	°F		300	D	D	D
Inlet Air Temperature	°F		80	D	D	D
Ambient Absolute Pressure	in. Hg		29.4	D	D	D
Pressure After Air Heater	in. H ₂ O		-12	D	D	D
Moisture in Air	lb/lb dry air		0.013	D	D	D
Ash Split				D	D	D
Fly Ash	%		80%	D	D	D
Seismic Zone	integer	1-5	1	D	D	D
Conversion of SO ₂ to SO ₃			10%	D	D	D

E.2.3 APC Technology Choices

This is the area of the worksheet where the user can choose what control technologies are needed. The following screen shows how this area looks and what options are available.

APC Technology Choices							
NO _x Control							
Combustion Control (0=no additional control, 1= combustion control)	integer	0, 1	1	1	1	1	1
Post Combustion Control (0=none, 1= SCR, 2= SNCR, 3 = NGF)	integer	0, 1, 2, 3	1	1	1	1	1
Particulate Control (0= none, 1= Fabric Filter, 2 = ESPc, 3=ESPh)	integer	0, 1, 2, 3	2	1	1	1	1
SO ₂ Control (0=none, 1= LSF0, 2 = LSD)	integer	0, 1, 2	1	1	1	1	1
Additional Mercury Control with Sorbent Injection? (0=no, 1=yes)	integer	0 or 1	1	1	1	1	1
If Sorbent Injection, add downstream PJFF? (0=no, 1=PJFF)	integer	0 or 1	1	1	1	1	1
CO ₂ Control (0=none, 1=CAP, 2=MEA, 3=Sorbent)	Integer	0, 1, 2, 3	1	1	3	1	1

E-2.4 NO_x Control Inputs

Data necessary for sizing and costing the NO_x control processes are input in the worksheet below. This information is used with the combustion calculations to size one of the four processes.

3.1 NO_x Control Technology

Item/Description	Units	Range	Default	Case 1	Case 2	Case 3
Combustion Technology Selected?				Selected	Selected	Selected
Uncontrolled NO _x level	lb/MMBtu		calculated	D	D	D
Boiler Type	W:Wall		T	D	D	D
Burner Type	2=LNB and OFA		1	D	D	D
Retrofit Difficulty Factor	number		13	D	D	D
General Facilities	percent		5.0%	D	D	D
Engineering	percent		10.0%	D	D	D
Contingency	percent		15.0%	D	D	D
Duration of Project	years		1	D	D	D
SCR Technology Selected?				Selected	Selected	Selected
Inlet NO _x level	lb/MMBtu		calculated	D	D	D
NH ₃ /NO _x Stoichiometric Ratio	NH ₃ /NO _x	0.7-1.0	0.9	D	D	D
NO _x Reduction Efficiency	Fraction	0.60-0.90	0.90	D	D	D
Space Velocity (Calculated if zero)	1/h		0	D	D	D
Time to First Catalyst Replenishment	years	2-5	3	D	D	D
Ammonia Cost	\$/ton		400	D	D	D
Catalyst Cost	\$/m ³		5000	D	D	D
Solid Waste Disposal Cost	\$/ton		11.48	D	D	D
Retrofit Difficulty Factor			15	D	D	D
General Facilities (% of Installed Cost)	%		5%	D	D	D
Engineering Fees (% of Installed Cost)	%		10%	D	D	D
Contingency (% of Installed Cost)	%		15%	D	D	D
Duration of Project	years		2	D	D	D
Maintenance (% of Installed Cost)	%		0.66%	D	D	D
Co-benefit Application						
Mercury Oxidation Rate - bituminous coal	%		90.0%	D	D	D
Mercury Oxidation Rate - subbituminous coal	%		0.0%	D	D	D
SNCR Technology Selected?				Not Selected	Not Selected	Not Selected
Inlet NO _x level	lb/MMBtu		calculated	D	D	D
Reagent	integer	Urea 2:Ammoni	1	D	D	D

E.2.5 Particulate Control Inputs

Data necessary for sizing and costing the particulate control equipment are input into the worksheet below. This information is used with the combustion calculations to size either an ESP or FF.

3.2 PM Control Technology

Item/Description	Units	Range	Default	Case 1	Case 2	Case 3
Outlet Particulate Emission Limit	lbs/MMBtu		0.03	D	D	D
Fabric Filter Selected				Selected	Selected	Selected
Pressure Drop	in. H ₂ O		6	D	D	D
Type (1 = Reverse Gas, 2 = Pulse Jet)	Integer		2	D	D	D
Gas-to-Cloth Ratio	ACFM/ft ²		3.5	D	D	D
Bag Material (RGFF fiberglass only) (1 = Fiberglass, 2 = Nomex, 3 = Ryton)	Integer		2	D	D	D
Bag Diameter	inches		6	D	D	D
Bag Length	feet		20	D	D	D
Bag Reach			3	D	D	D
Compartments out of Service	%		10%	D	D	D
Bag Life	Years		5	D	D	D
Retrofit Factor			1.3	D	D	D
Contingency (% of installed cost)	%		15%	D	D	D
General Facilities (% of installed cost)	%		5%	D	D	D
Engineering Fees (% of installed cost)	%		10%	D	D	D
Project Duration	years		2	D	D	D
Maintenance (% of installed cost)	%		5%	D	D	D
ESP Selected				Not Selected	Not Selected	Not Selected
Strength of the electric field in the ESP = E	kV/cm		10.0	D	D	D
Plate Spacing	in.		12	D	D	D
Plate Height	ft		36	D	D	D
Pressure Drop	in. H ₂ O		3	D	D	D
Retrofit Factor			1.3	D	D	D
Contingency (% of Installed Cost)	%		15%	D	D	D
General Facilities (% of Installed Cost)	%		5%	D	D	D
Engineering Fees (% of Installed cost)	%		10%	D	D	D
Project Duration	years		2	D	D	D

E.2.6 SO₂ Control Inputs

Data necessary for sizing and costing an SO₂ control system are input into the worksheet below. This information is used with the combustion calculations to design the system.

3.3 SO₂ Control Technology

Item/Description	Units	Range	Default	Case 1	Case 2	Case 3
Lime Spray Dryer Selected?				Not Selected	Not Selected	Not Selected
SO ₂ Removal Required	%	70-95%	90%	D	D	D
Is SDA being retrofit upstream of existing ESP? (0 = no, 1 = yes)	integer	0, 1	0	D	D	D
Adiabatic Saturation Temperature	*F	100-170	127	D	D	D
Flue Gas Approach to Saturation	*F	10-50	20	D	D	D
Recycle Slurry Solids Concentration	Wt. %	10-50	35%	D	D	D
Number of Absorbers (Max. Capacity = 300 MW per spray dryer)	integer	1-7	1	D	D	D
Absorber Material (1 = Alloy, 2 = RLCS)	integer	1 or 2	2	D	D	D
Spray Cooler Pressure Drop	in. H ₂ O		1	D	D	D
Reagent Bulk Storage	days		30	D	D	D
Reagent Cost (delivered)	\$/ton		\$65	D	D	D
Dry Waste Disposal Cost	\$/ton		\$30	D	D	D
Retrofit Factor			1.3	D	D	D
Contingency (% of Installed Cost)	%		15%	D	D	D
General Facilities (% of Installed Cost)	%		5%	D	D	D
Engineering Fees (% of Installed Cost)	%		10%	D	D	D
Project Duration	years		2	D	D	D
Maintenance Factor (% of TPC)	%		2%	D	D	D
LSFO Selected?				Selected	Selected	Selected
Year Equipment Placed in Service	year		2004	D	D	D
SO ₂ Removal Required	%	90-98%	95%	D	D	D
L/G Ratio	gal / 1000 acf	95-160	125	D	D	D
Design Scrubber with Dibasic Acid Addition? (1 = yes, 2 = no)	integer	1 or 2	1	D	D	D
Adiabatic Saturation Temperature	*F	100-170	127	D	D	D
Reagent Feed Ratio	Factor	1.0-2.0	1.05	D	D	D

E.2.7 Mercury Control Inputs

This is where the data necessary for sizing and costing the mercury control processes are input. This information is used with the combustion calculations to size powdered activated carbon (PAC) and pulse-jet fabric filter (PJFF) processes.

3.4 Mercury Control Technology

Item/Description	Units	Range	Default	Case 1	Case 2	Case 3
Sorbent Injection Technology Selected?						
Hg CEMS (0=no, 1=yes)	integer	0 or 1	1	D	D	-
Hg Reduction Required from Coal	percent		0.8	D	D	D
Sorbent Type, 1=EPAC, 2=PAC, 3=other			2	D	D	D
Maximum Temperature before Spray Cooling	°F	≥325	325	D	D	D
Sorbent Recycle Used?	Yes/No		No	D	D	D
Spray Cooling Desired?	Yes/No		No	D	D	D
EPAC Cost (Delivered Cost of Treated PAC)	\$/ton		1500	D	D	D
PAC Cost (Delivered)	\$/ton		1000	D	D	D
Other Sorbent Cost (Delivered)	\$/ton		1000	D	D	D
Before Sorbent Injection, Fly Ash Sold (1) or Disposed of (2)	1 or 2		1	D	D	D
Does Sorbent Adversely Impact Fly Ash Sales? (0=no, 1=yes)	integer	0 or 1	0	D	D	D
Revenue from Fly Ash Sales	\$/ton	0 to 35	6	D	D	D
Dry Waste Disposal Cost	\$/ton	1 to 25	6	D	D	D
Retrofit Factor			13	D	D	D
Process Contingency, % of Process Capital	%		5%	D	D	D
General Facilities (% of Installed Cost)	%		5%	D	D	D
Engineering Fees (% of Installed Cost)	%		10%	D	D	D
Project Contingency	%		15%	D	D	D
Duration of Project	years		1	D	D	D
Maintenance Factors (% of Installed Cost)	%		5%	D	D	D
PJFF Downstream of PAC Selected?						
PJFF to COHPAC (ie, TOXCON), 0=no, 1=yes	0 or 1		1	D	D	D
Cost of Bags, Installed (\$/bag)	\$/bag		80	D	D	D
Estimated #Bags/MW	integer		20	D	D	D
Average Bag Life	years		5	D	D	D
Pressure Drop	in. H ₂ O		8.0	D	D	D
Outlet Emissions	lb/MWbtu		0.012	D	D	D
Retrofit Difficulty Factor			13	D	D	D
Process Contingency, % of Process Capital	%		5%	D	D	D

E.2.8 CO₂ Control Inputs

This is where the data necessary for sizing and costing the CO₂ control processes are input. This information is used with the combustion calculations to size amine-based CO₂ control processes.

3.5 Carbon Dioxide Control Technology

Item/Description	Units	Range	Default	Case 1	Case 2	Case 3
Chilled Ammonia Process Selected?						
				Selected	Not Selected	Selected
Efficiency of CO ₂ removal	%	≥90	90%	D	D	D
Flue gas temperature out of direct contact cooler	°F	35	35	D	D	D
Flue gas temperature entering the absorber	°F	30-32	32	D	D	D
Flue gas temp exiting the absorber	°F	32-50	35	D	D	D
CO ₂ temperature exiting the stripper reflux	°F	65-72	70	D	D	D
Cycle of Concentration for Cooling Water		2-10	5	D	D	D
Reagent of Ammonia						
Concentration of Ammonia	%	28-30%	28%	D	D	D
Price of Ammonia (28%)	\$/ton	100-200	150	D	D	D
Ammonia slip to flue gas	ppm	2-10	5	D	D	D
Regenerator Pressure	Psi	300-600	400	D	D	D
Compressor						
CO ₂ Product Pressure	psi	500-2500	2200	D	D	D
CO ₂ compressor stage		3	3	D	D	D
Number of operator		6-8	8	D	D	D
Retrofit Factor			1.0	D	D	D
Maintenance Factor (% of TPC)	%		3%	D	D	D
Contingency (% of Installed Cost)	%		15%	D	D	D
General Facilities (% of Installed Cost)	%		10%	D	D	D
Engineering Fees (% of Installed Cost)	%		7%	D	D	D
Time for Retrofit	years		2	D	D	D
MEA Process Selected?						
				Not Selected	Not Selected	Not Selected
Efficiency of CO ₂ removal	%	90%	90%	D	D	D
KS-1 or Other MEA	1=KS-1, 2= MEA		1	1	2	1
Cycle of Concentration for Cooling Water		2-10	5	D	D	D
Reagen Price						
Price of MEA	\$/ton		2142	D	D	D
Price of NaOH (20% solution)	\$/ton		413	D	D	D

APPENDIX F PROGRAMS FOR ECONOMIC PARAMETERS

The programs included in this appendix are for calculations of economic parameters, including the TEC, TPI, current \$ carrying charge, constant \$ carrying charge, first year current \$ carrying charges, first year constant \$ carrying charges, and levelization factors. Comments were added in each program for clarity.

'This module contains functions to calculate the Economic parameters

'This function calculates TCE value

'Arguments: inflation rate, float; escalation, float; period, float

Function TCE (inflation, escalation, period)

$$EA = (1 + \text{inflation}) \times (1 + \text{escalation}) - 1$$

$$TCE = PV(EA, \text{period}, -1) \times (1 + EA) / \text{period}$$

End Function

'This function calculates TPI value, the parameters

'Inside the parenthesis already show what you should input

Function TPI(inflation, escalation, interest, period)

$$EA = (1 + \text{inflation}) \times (1 + \text{escalation}) - 1$$

$$Z = (1 + \text{interest}) / (1 + EA)$$

$$TPI = (Z ^ \text{period} - 1)$$

$$TPI = TPI / \text{period}$$

$$TPI = TPI / (Z - 1)$$

End Function

'This function calculates Current Carrying Charge

'Arguments: rd (cost of debt), float; wd(ratio of debt), float; re(cost of equity), float;

'we (ratio of equity), float; PTI (property tax and insurance), float; T (tax), float;

'ITC (investment tax credit), float; BL(book life), float; depreMethod (depreciation method), integer

Function CurrentCC(ByVal rd, ByVal wd, ByVal re, ByVal we, ByVal PTI, ByVal T, ByVal ITC, ByVal BL, ByVal depreMethod)

rd = rd

wd = wd

re = re

we = we

PTI = PTI

T = T

ITC = ITC

Dim Arcs(21) As Double

'Define depreciation rate for each year

Dim Depr(31) As Double

'Loop variable

Dim i As Integer

'Calculate the weighted average cost of capital (WACC)

Dim EFI As Double

'Straight line depreciation

Dim sltxdp As Double

sltxdp = 1 / BL

'Cumulative CC × V in the EPRI formula

'Detail of the original formula can be seen in TAG-Technical Assessment Guide, Vol. 3,

'Fundamentals and Methods, Supply-1986, EPRI P-4463-SR

Dim SP As Double

SP = 0

'V current value factor in the EPRI formula

Dim V As Double

'Deferred income tax

Dim DT As Double

'tax paid

Dim TP As Double

'tax depreciation rate

Dim TD As Double

'Carrying charge

Dim cc As Double

'An annuity factor in EPRI formula

Dim A As Double

'temp variable of rd

Dim rdTemp As Double

'temp variable of re

Dim reTemp As Double

'Book depreciation rate net of ITC

Dim BD As Double

BD = (1 - ITC) / BL

'Remaining book value

Dim SB As Double

SB = 1 - ITC

'Assign value to ACRS schedule

Arcs(0) = 0

Arcs(1) = 7.5

Arcs(2) = 6.9

Arcs(3) = 6.4

Arcs(4) = 5.9

Arcs(5) = 5.5

Arcs(6) = 5.1

Arcs(7) = 4.7

Arcs(8) = 4.5

Arcs(9) = 4.5

Arcs(10) = 4.5

Arcs(11) = 4.5

Arcs(12) = 4.5

Arcs(13) = 4.5

Arcs(14) = 4.5

Arcs(15) = 4.5

Arcs(16) = 4.4

Arcs(17) = 4.4

Arcs(18) = 4.4

Arcs(19) = 4.4

Arcs(20) = 4.4

'Assign value 0 to initialize the array

A = 0#

'Compute return rate used to determine carrying charges

EFI = (wd × rd + we × re)

For i = LBound(Depr) To UBound(Depr)

Depr(i) = 0

Next i

'Calculate tax depreciation with 3 cases, we assume that the book life is greater than the depreciation life

'Calculate straight line tax depreciation

If (depreMethod = 1) Then

'For straight line tax depreciation over book life

For i = 1 To BL

Depr(i) = 1 / BL

Next i

'For straight line tax depreciation over ACRS

Elseif (depreMethod = 2) Then

For i = 1 To 20

Depr(i) = 1 / 20

```

Next i
  'For ACRS depreciation schedule
Else
  For i = 1 To 20
    Depr(i) = Arcs(i) / 100
  Next i
End If

'Start loop over the book life
For i = 1 To BL
  'Present value factor
  V = (1 + EFI) ^ (-i)
  If (EFI <> 0) Then
    A = (1 - V) / (EFI)
  Else
    A = i
  End If
  'tax depreciation rate
  TD = Depr(i)
  'return on equity
  reTemp = SB × re × we
  'return on debt
  rdTemp = SB × rd × wd
  'deferred income tax
  DT = (TD - sltxdp) × T
  'tax paid.
  TP = T / (1 - T) × (BD - TD + DT + reTemp)
  ' year by year carrying charges
  cc = BD + DT + rdTemp + reTemp + TP + PTI
  'cum. present value of carrying charge
  SP = SP + cc × V
  'Depreciation book value, net def. tax
  SB = SB - BD - DT
Next i
'carryCharges = SP / A
CurrentCC = SP / A
End Function
'This function calculates first year Current Carrying Charge
'Arguments: rd (cost of debt), float; wd(ratio of debt), float; re(cost of equity), float;
'we (ratio of equity), float;PTI (property tax and insurance), float; T ( tax), float;
'ITC (investment tax credit), float; BL(book life), float; depreMethod (depreciation method),
integer

```

Function firstYearCurrentCC(ByVal rd, ByVal wd, ByVal re, ByVal we, ByVal PTI, ByVal T, ByVal ITC, ByVal BL, ByVal depreMethod)

rd = rd
wd = wd
re = re
we = we
PTI = PTI
T = T
ITC = ITC

Dim Arcs(21) As Double
'Define depreciation rate for each year
Dim Depr(31) As Double
'loop variable
Dim i As Integer
'Calculate the weighted average cost of capital (WACC)
Dim EFI As Double
'straight line depreciation
Dim sltxdp As Double
sltxdp = 1 / BL
'cumulative CC × V in the EPRI formula
'Detail of the original formula can be seen in TAG-Technical Assessment Guide, Vol. 3, Fundamentals and Methods, Supply-1986, EPRI P-4463-SR
Dim SP As Double
SP = 0
'V current value factor in the EPRI formula
Dim V As Double
'deferred income tax
Dim DT As Double
'tax paid.
Dim TP As Double
'tax depreciation rate
Dim TD As Double
'Carrying charge
Dim cc As Double
'A annuity factor in EPRI formula
Dim A As Double
'temp variable of rd
Dim rdTemp As Double
'temp variable of re
Dim reTemp As Double
'book depreciation rate net of ITC
Dim BD As Double

BD = (1 - ITC) / BL

'remaining book value

Dim SB As Double

SB = 1 - ITC

'assign value to ACRS schedule

Arcs(0) = 0

Arcs(1) = 7.5

Arcs(2) = 6.9

Arcs(3) = 6.4

Arcs(4) = 5.9

Arcs(5) = 5.5

Arcs(6) = 5.1

Arcs(7) = 4.7

Arcs(8) = 4.5

Arcs(9) = 4.5

Arcs(10) = 4.5

Arcs(11) = 4.5

Arcs(12) = 4.5

Arcs(13) = 4.5

Arcs(14) = 4.5

Arcs(15) = 4.5

Arcs(16) = 4.4

Arcs(17) = 4.4

Arcs(18) = 4.4

Arcs(19) = 4.4

Arcs(20) = 4.4

'assign value 0 to initialize the array

A = 0#

'compute return rate used to determine carrying charges

EFI = (wd × rd + we × re)

For i = LBound(Depr) To UBound(Depr)

Depr(i) = 0

Next i

'calculate tax depreciation with 3 cases, we assume that the book life is greater than the depreciation life

'calculate straight line tax depreciation

If (depreMethod = 1) Then

'for straight line tax depreciation over booklife

For i = 1 To BL

Depr(i) = 1 / BL

Next i

'for straight line tax depreciation over ACRS

Elseif (depreMethod = 2) Then

For i = 1 To 20

```
    Depr(i) = 1 / 20
    Next i
    'for ACRS depreciation schedule
Else
    For i = 1 To 20
        Depr(i) = Arcs(i) / 100
    Next i
End If
For i = 1 To 1
    'Present value factor
    V = (1 + EFI) ^ (-i)
    If (EFI <> 0) Then
        A = (1 - V) / (EFI)
    Else
        A = i
    End If
    'tax depreciation rate
    TD = Depr(i)
    'Return on equity
    reTemp = SB × re × we
    'Return on debt
    rdTemp = SB × rd × wd
    'Deferred income tax
    DT = (TD - sltxdp) × T
    'tax paid.
    TP = T / (1 - T) × (BD - TD + DT + reTemp)
    ' Year by year carrying charges
    cc = BD + DT + rdTemp + reTemp + TP + PTI
    'cum. present value of carrying charge
    SP = SP + cc × V
    'Depreciation book value, net def. tax
    SB = SB - BD - DT
Next i
'carryCharges = SP / A
firstYearCurrentCC = SP / A
End Function
```

'This function calculates constant Carrying Charge

'Arguments: rd (cost of debt), float; wd(ratio of debt), float; re(cost of equity), float;

'we (ratio of equity), float; PTI (property tax and insurance), float; T (tax), float;

'ITC (investment tax credit), float; BL(book life), float; inflation, float; depreMethod (depreciation method), integer

Function ConstantCC(ByVal rd, ByVal wd, ByVal re, ByVal we, ByVal PTI, ByVal T, ByVal ITC, ByVal BL, ByVal inflation, ByVal depreMethod)

rd = rd

wd = wd

inflation = inflation

re = re

we = we

PTI = PTI

T = T

ITC = ITC

Dim Arcs(21) As Double

'Define depreciation rate for each year

Dim Depr(31) As Double

'loop variable

Dim i As Integer

'Calculate the weighted average cost of capital (WACC)

Dim EFI As Double

'Straight line depreciation

Dim sltxdp As Double

sltxdp = 1 / BL

'Cumulative CC × V in the EPRI formula

'Detail of the original formula can be seen in TAG-Technical Assessment Guide, Vol. 3,

'Fundamentals and Methods, Supply-1986, EPRI P-4463-SR

Dim SP As Double

SP = 0

'V current value factor in the EPRI formula

Dim V As Double

'Deferred income tax

Dim DT As Double

'tax paid.

Dim TP As Double

'tax depreciation rate

Dim TD As Double

'Carrying charge

Dim cc As Double

'An annuity factor in EPRI formula


```
Dim A As Double
'temp variable of rd
Dim rdTemp As Double
'temp variable of re
Dim reTemp As Double
'book depreciation rate net of ITC
Dim BD As Double
BD = (1 - ITC) / BL
'Remaining book value
Dim SB As Double
SB = 1 - ITC
'Assign value to ACRS schedule
Arcs(0) = 0
Arcs(1) = 7.5
Arcs(2) = 6.9
Arcs(3) = 6.4
Arcs(4) = 5.9
Arcs(5) = 5.5
Arcs(6) = 5.1
Arcs(7) = 4.7
Arcs(8) = 4.5
Arcs(9) = 4.5
Arcs(10) = 4.5
Arcs(11) = 4.5
Arcs(12) = 4.5
Arcs(13) = 4.5
Arcs(14) = 4.5
Arcs(15) = 4.5
Arcs(16) = 4.4
Arcs(17) = 4.4
Arcs(18) = 4.4
Arcs(19) = 4.4
Arcs(20) = 4.4

'Assign value 0 to annuity
A = 0#
'Compute return rate used to determine carrying charges
'Calculate rd and re without inflation
rd = (1 + rd) / (1 + inflation) - 1
re = (1 + re) / (1 + inflation) - 1
EFI = wd × rd + we × re
'Initialize depreciation value in the depreciation array.
For i = LBound(Depr) To UBound(Depr)
    Depr(i) = 0
```

Next i

'Calculate tax depreciation with 3 cases, we assume that the book life is greater than the depreciation life

'Calculate straight line tax depreciation

If (depreMethod = 1) Then

'For straight line tax depreciation over book life

For i = 1 To BL

Depr(i) = 1 / BL

Next i

'For straight line tax depreciation over ACRS

Elseif (depreMethod = 2) Then

For i = 1 To 20

Depr(i) = 1 / 20

Next i

'for ACRS depreciation schedule

Else

For i = 1 To 20

Depr(i) = Arcs(i) / 100

Next i

End If

'start the loop over the book life

For i = 1 To BL

'Present value factor

$V = (1 + \text{EFI})^{-i}$

If (EFI <> 0) Then

$A = (1 - V) / \text{EFI}$

Else

A = i

End If

'Tax depreciation rate

TD = Depr(i)

'return on equity

reTemp = SB × re × we

'return on debt

rdTemp = SB × rd × wd

'Deferred income tax

DT = (TD - sltxdp) × T

'tax paid.

TP = T / (1 - T) × (BD - TD + DT + reTemp)

'Year by year carrying charges

cc = BD + DT + rdTemp + reTemp + TP + PTI

'Cum. presents value of carrying charge

SP = SP + cc × V

'Depreciation book value, net def. tax

```

    SB = SB - BD - DT
Next i
'carryCharges = SP / A
    ConstantCC = SP / A
End Function
'This function calculates first year Current Carrying Charge
'Arguments: rd (cost of debt), float; wd(ratio of debt), float; re(cost of equity), float;
'we (ratio of equity), float;PTI (property tax and insurance), float; T ( tax), float;
'ITC (investment tax credit), float; BL(book life), float; inflation, float; depreMethod
(depreciation method), integer
Function firstYearConstantCC(ByVal rd, ByVal wd, ByVal re, ByVal we, ByVal PTI, ByVal T,
ByVal ITC, ByVal BL, ByVal inflation, ByVal depreMethod)

    rd = rd
    wd = wd
    inflation = inflation
    re = re
    we = we
    PTI = PTI
    T = T
    ITC = ITC

    Dim Arcs(21) As Double
    'Define depreciation rate for each year
    Dim Depr(31) As Double
    'Loop variable
    Dim i As Integer
    'Calculate the weighted average cost of capital (WACC)
    Dim EFI As Double
    'Straight line depreciation
    Dim sltxdp As Double
    sltxdp = 1 / BL
    'Cumulative CC × V in the EPRI formula
    'Detail of the original formula can be seen in TAG-Technical Assessment Guide, Vol. 3,
    'Fundamentals and Methods, Supply-1986, EPRI P-4463-SR
    Dim SP As Double
    SP = 0
    'V current value factor in the EPRI formula
    Dim V As Double
    'Deferred income tax
    Dim DT As Double
    'tax paid.
    Dim TP As Double
    'tax depreciation rate

```

Dim TD As Double

'Carrying charge

Dim cc As Double

'An annuity factor in EPRI formula

Dim A As Double

'temp variable of rd

Dim rdTemp As Double

'temp variable of re

Dim reTemp As Double

'book depreciation rate net of ITC

Dim BD As Double

$BD = (1 - ITC) / BL$

'Remaining book value

Dim SB As Double

$SB = 1 - ITC$

'Assign value to ACRS schedule

$Arcs(0) = 0$

$Arcs(1) = 7.5$

$Arcs(2) = 6.9$

$Arcs(3) = 6.4$

$Arcs(4) = 5.9$

$Arcs(5) = 5.5$

$Arcs(6) = 5.1$

$Arcs(7) = 4.7$

$Arcs(8) = 4.5$

$Arcs(9) = 4.5$

$Arcs(10) = 4.5$

$Arcs(11) = 4.5$

$Arcs(12) = 4.5$

$Arcs(13) = 4.5$

$Arcs(14) = 4.5$

$Arcs(15) = 4.5$

$Arcs(16) = 4.4$

$Arcs(17) = 4.4$

$Arcs(18) = 4.4$

$Arcs(19) = 4.4$

$Arcs(20) = 4.4$

'Assign value 0 to annuity

$A = 0\#$

'Compute return rate used to determine carrying charges

'Calculate rd and re without inflation

$rd = (1 + rd) / (1 + inflation) - 1$

$re = (1 + re) / (1 + inflation) - 1$

```
EFI = wd × rd + we × re
'Initialize depreciation value in the depreciation array.
For i = LBound(Depr) To UBound(Depr)
  Depr(i) = 0
Next i
'Calculate tax depreciation with 3 cases, we assume that the book life is greater than the
depreciation life
'Calculate straight line tax depreciation
If (depreMethod = 1) Then
  'For straight line tax depreciation over book life
  For i = 1 To BL
    Depr(i) = 1 / BL
  Next i
  'For straight line tax depreciation over ACRS
Elseif (depreMethod = 2) Then
  For i = 1 To 20
    Depr(i) = 1 / 20
  Next i
  'For ACRS depreciation schedule
Else
  For i = 1 To 20
    Depr(i) = Arcs(i) / 100
  Next i
End If
'Start the loop over the book life
For i = 1 To 1
  'present value factor
  V = (1 + EFI) ^ (-i)
  If (EFI <> 0) Then
    A = (1 - V) / (EFI)
  Else
    A = i
  End If
  'Tax depreciation rate
  TD = Depr(i)
  'Return on equity
  reTemp = SB × re × we
  'Return on debt
  rdTemp = SB × rd × wd
  'Deferred income tax
  DT = (TD - sltxdp) × T
  'tax paid.
  TP = T / (1 - T) × (BD - TD + DT + reTemp)
  ' Year by year carrying charges
```

$cc = BD + DT + rdTemp + reTemp + TP + PTI$

'Cum. present value of carrying charge

$SP = SP + cc \times V$

'Depreciation book value, net def. tax

$SB = SB - BD - DT$

Next i

'carry Charges = SP / A

$firstYearConstantCC = SP / A$

End Function

'This function calculate current levelization for O&M cost

'Arguments: rd (cost of debt), float; wd(ratio of debt), float; re(cost of equity), float;

'we (ratio of equity), float; BL(book life), float; inflation, float; escalation, float

Function currentLL(rd, wd, re, we, BL, inflation, escalation)

inflation = inflation

escalation = escalation

rd = rd

wd = wd

re = re

we = we

'Define discount (weighted average cost of capital)

Dim discount

$discount = wd \times rd + we \times re$

Dim EA

$EA = (1 + inflation) \times (1 + escalation) - 1$

Dim k

$k = (1 + EA) / (1 + discount)$

Dim An

$An = ((1 + discount) ^ BL - 1) / (discount \times (1 + discount) ^ BL)$

Dim Ln

$Ln = (k \times (1 - k ^ BL)) / (An \times (1 - k))$

currentLL = Ln

End Function

Function constantLL(rd, wd, re, we, BL, inflation, escalation)

inflation = inflation

escalation = escalation

rd = rd

wd = wd

re = re

we = we

$$rd = (1 + rd) / (1 + inflation) - 1$$

$$re = (1 + re) / (1 + inflation) - 1$$

Dim discount

$$discount = wd \times rd + we \times re$$

Dim EA

$$EA = (1 + escalation) - 1$$

Dim k

$$k = (1 + EA) / (1 + discount)$$

Dim An

$$An = ((1 + discount) ^ BL - 1) / (discount \times (1 + discount) ^ BL)$$

Dim Ln

$$Ln = (k \times (1 - k ^ BL)) / (An \times (1 - k))$$

$$constantLL = Ln$$

$$'constantLL = re$$

End Function

The function below is a function to calculate the compressibility of compressors across the compressor island in CO₂ compression

'Define a function

Function CompressionPower(initialPressure, finalPressure, numOfStage, CompressionTemp)

Dim pressureRatio As Double

Dim pressureOfEachStage(10) As Double

Dim Z(10) As Double

Dim i As Integer

Dim averageZ

'initialize pressure

pressureOfEachStage(0) = initialPressure

'initialize Z

If pressureOfEachStage(0) > 800 Then

Z(0) = 0.5

Else

Z(0) = 1 - 0.4 / 800 × pressureOfEachStage(0)

End If

pressureRatio = ((finalPressure + 1.5 × numOfStage) / initialPressure) ^ (1# / numOfStage)

'calculate Z factors

CompressionPower = 0#

For i = 1 To numOfStage

```
pressureOfEachStage(i) = pressureOfEachStage(i - 1) × pressureRatio
'calculate outlet Z
Z(i) = 1 - 0.5 / 2000 × pressureOfEachStage(i)
If Z(i) < 0.5 Then
    Z(i) = 0.5
End If

'calculate inlet Z
If pressureOfEachStage(i - 1) > 800 Then
    Z(i - 1) = 0.5
Else
    Z(i - 1) = 1 - 0.4 / 800 × pressureOfEachStage(i - 1)
End If
'Calculate average Z
averageZ = (Z(i) + Z(i - 1)) / 2
CompressionPower = CompressionPower + 188.9 × (CompressionTemp + 273.15) ×
1.28 × averageZ / (1.28 - 1) × (pressureRatio ^ (0.28 / 1.28) - 1) / 0.8
Next
End Function
```