

Carnegie Mellon University  
CARNEGIE INSTITUTE OF TECHNOLOGY

THESIS

SUBMITTED IN PARTIAL FULFILLMENT OF THE REQUIREMENTS  
FOR THE DEGREE OF  
Doctor of Philosophy

TITLE: Advanced Amine and Ammonia Systems for Greenhouse Gas Control at  
Fossil Fuel Power Plants

PRESENTED BY: Peter Versteeg

ACCEPTED BY THE DEPARTMENT OF: Engineering and Public Policy

\_\_\_\_\_  
ADVISOR, MAJOR PROFESSOR

\_\_\_\_\_  
DATE

\_\_\_\_\_  
DEPARTMENT HEAD

\_\_\_\_\_  
DATE

APPROVED BY THE COLLEGE COUNCIL:

\_\_\_\_\_  
DEAN

\_\_\_\_\_  
DATE

**Advanced Amine and Ammonia Systems for Greenhouse  
Gas Control at Fossil Fuel Power Plants**

Submitted in partial fulfillment of the requirements for

the degree of

Doctor of Philosophy

In

Engineering & Public Policy

Peter L. Versteeg

B.A.Sc., Engineering Science, University of Toronto

M.A.Sc., Chemical Engineering, University of Toronto

Carnegie Mellon University  
Pittsburgh, PA

May, 2012

## **Acknowledgments**

This work was supported by the U.S. Department of Energy's National Energy's Technology Laboratory (DOE/NETL) under Contract No.'s 24905.913.ER.1041723 and DE-AC26-04NT41917, and by the Natural Sciences and Engineering Research Council of Canada (NSERC).

First, I would like to thank my family. To my parents Bill and Judy Versteeg, you introduced me to science and engineering and encouraged me to read at a young age. For much of my youth I did not understand why this was so important, but I do now. This thesis is as much a result of your efforts as my own. To my fiancé Patricia Lai, this is the second thesis through which you have supported me through. You have encouraged me as I have followed my own path, despite its difficulties. I love you. To my sister Naomi Lomele and my brothers, Benjamin, and Michael Versteeg, you have been a constant source of incredible fun and happiness through these years. As I write this, I most look forward to spending time with you when this is all over.

Next, I would like to thank Professor Ed Rubin. Your guidance, the many opportunities and challenges you provided, and your never-ending support made all the difference during my time here in Pittsburgh. I learned a lot over the years under your mentorship, and I really appreciate the time and effort you put into advising me from day one. One day, I would like to have a great career which you have helped to get started. But much more so, I would like to have the kind of positive and long lasting influence on people's lives that you have, every day. I can see that the hard work never ends and that you embrace that and everything that goes with it. And I am very grateful for the opportunity to have learned this from you.

In our research group I would like to thank in particular Haibo, Karen, and Hari. I have learned a lot over our many conversations, and your feedback, troubleshooting, and help with this work greatly improved it. I would like to thank Justin and Kyle as well for letting me bounce ideas off of them, and providing a wealth of technical guidance on and common sense to my work here.

I owe many thanks to my great friends here in the Department of Engineering & Public Policy, including Anu, Kelly, Eric, Jon, Austin, Apurba, Dan, Stefan, Rob, Kim, Shantanu, Tim, Shira, Catherine, Brinda, and Sean. Great friends back home have also cheered me on including Tanya, Angela, Tooraj, Mirza, Tamima, Diana, Amanda, Liam, Jon, Ajay, Sarah, Vivien, Naseem, and Brad. And I would like to thank the members of the CMU Squash as well including Rick, Alvaro, Amy, Charlie, Rahul, Yousef, Jonathan, and Frank for a lot of great games over the last few years. For all of you, I hope our paths cross again often in the future.

Finally, I would also like to thank my committee chair Ed Rubin, committee members Granger Morgan, Jay Apt, and John Kitchin, the many other great professors and the amazing staff in the Department of Engineering & Public Policy. You made my time here an experience to remember.

## Abstract

In an effort to lower future CO<sub>2</sub> emissions, a wide range of technologies are being developed to scrub CO<sub>2</sub> from the flue gases of fossil fuel-based electric power and industrial plants. This thesis models two leading post-combustion CO<sub>2</sub> capture technologies, a chilled ammonia-based CO<sub>2</sub> capture process and an advanced amine-based CO<sub>2</sub> capture process, and presents performance and cost estimates of these systems on pulverized coal and natural gas combined cycle power plants.

The process modeling software package Aspen Plus® was used to develop performance and cost estimates for the chilled ammonia-based CO<sub>2</sub> capture technology and general response surface equations were created for the model. Assumptions about plant financing and utilization, as well as uncertainties in cooling costs and chemical reaction rates that affect absorber cost were found to produce a wide range of cost estimates for ammonia-based CO<sub>2</sub> capture systems. With uncertainties included, costs for a supercritical power plant with ammonia-based CO<sub>2</sub> capture ranged from \$80/MWh to \$160/MWh, with the 95% confidence interval ranging from \$95/MWh to \$143/MWh (with all costs in constant 2007 US dollars).

For the advanced amine-based CO<sub>2</sub> capture technology, an existing amine-based response surface model developed using Protreat® simulations was modified to match the performance and cost characteristics of a modern amine-based system. The response surface models of both technologies were incorporated into the Integrated Environmental Control Model for use in developing performance and cost estimates of pulverized coal and natural gas combined cycle power plants with these technologies. The baseline costs for a supercritical power plant with advanced amine-based CO<sub>2</sub> capture was \$105/MWh and for the natural gas combined cycle power plant with advanced amine-based CO<sub>2</sub> capture was \$85/MWh.

Both post-combustion CO<sub>2</sub> capture technologies are then compared in terms of performance and cost for different ranges of fuel type, fuel cost, plant size, and CO<sub>2</sub> capture system train size. A probabilistic cost difference analysis is also used to compare these technologies. The amine-based CO<sub>2</sub> capture system is found to have a higher revenue requirement in all the case studies and only a 2% chance of having a lower revenue requirement than the advanced amine system in the probabilistic cost difference. Combined, these results suggest that the advanced amine system will have a cost advantage over the ammonia system in most cases, in the absence of significant new improvements in the ammonia system design. Finally, the importance of these estimates for policy makers is discussed.

# Contents

Acknowledgments.....	1
Abstract.....	3
Contents .....	4
List of Figures.....	7
List of Tables.....	9
Chapter 1. The Role of Carbon Capture and Storage in Climate Change Mitigation .....	11
1.1. CO <sub>2</sub> Emissions and Climate Change.....	11
1.2. CO <sub>2</sub> Emissions and Electric Power.....	12
1.3. A Potential Solution: Carbon Capture and Storage.....	14
1.4. The Current State of Post-Combustion CO <sub>2</sub> Capture and Storage .....	15
1.4.1. Commercial Projects for CO <sub>2</sub> Capture and Storage.....	16
1.4.2. Full-Scale Demonstration Plants.....	20
1.5. Objectives and Scope of this Thesis .....	22
1.6. Organization of this Thesis .....	23
Chapter 2. Ammonia-Based Post-Combustion CO <sub>2</sub> Capture.....	24
2.1. Introduction.....	24
2.2. Baseline Assumptions and Baseline Power Plants (Case Studies 1-3).....	24
2.2.1. Baseline Assumptions .....	24
2.2.2. Case Study 1-3 Results: PC and NGCC Power Plants without CCS.....	27
2.3. The Potential Advantages of an Ammonia-Based CO <sub>2</sub> Capture Process.....	28
2.4. Historical Development of Ammonia-Based Capture Processes.....	29
2.4.1. The Powerspan ECO <sub>2</sub> Capture Process.....	30
2.4.2. The Alstom Chilled Ammonia Process (CAP).....	30
2.5. Past Performance and Cost Estimates of the Chilled Ammonia Process.....	33
2.6. Aspen Plus® for Process Modeling.....	36
2.6.1. The Ammonia-Based CO <sub>2</sub> Capture Performance Model .....	36
2.6.2. Initial Comparison for Absorber Solvent Flow Rates.....	41
2.6.1. Plant Performance Results for the LowNH <sub>3</sub> and the HighNH <sub>3</sub> Systems .	44
2.6.1. Plant Economic Results for the LowNH <sub>3</sub> and the HighNH <sub>3</sub> Systems.....	46
2.7. Sensitivity Analysis in Other Variables in the Chilled Ammonia Process.....	49
2.7.1. Lean Solvent NH <sub>3</sub> Concentration and NH <sub>3</sub> /CO <sub>2</sub> Sensitivity.....	49
2.7.2. Effects of Absorber Temperature .....	52
2.7.3. Water Wash Requirements for Ammonia Removal.....	53
2.8. Refined Plant Performance Results for Ammonia-Based CO <sub>2</sub> Capture .....	54
2.9. Uncertainty Analysis on Key CO <sub>2</sub> Capture System Performance Parameters.....	59
2.10. Discussion.....	63
2.11. Chapter Conclusion.....	64
Chapter 3. Developing a Chilled Ammonia Response Surface Model for the IECM.....	67
3.1. Motivation for Developing a General Power Plant and CO <sub>2</sub> Capture Model.....	67

3.2.	The Integrated Environmental Control Model (IECM).....	68
3.3.	Developing a Response Surface Model of the Chilled Ammonia System.....	70
3.3.1.	Fractional Factorial Experimental Design.....	71
3.4.	The Response Surface Models.....	76
3.4.1.	Response Model Calculation Order .....	76
3.4.2.	The Response Model Equations.....	78
3.5.	IECM Performance Estimates.....	92
3.6.	IECM Cost Estimates.....	106
3.7.	Layout of the IECM User Screens for the Chilled Ammonia System .....	128
3.7.1.	Introduction .....	128
3.7.2.	Inputs Screens for the IECM Chilled Ammonia Model.....	129
3.7.3.	Output Screens for the IECM Chilled Ammonia Model.....	133
3.8.	Case Studies of PC and NGCC Plants with Chilled Ammonia-Based Capture..	136
3.8.1.	Introduction to Case Studies .....	136
3.8.2.	Summary Results from Case Studies 4-6.....	136
3.9.	Conclusion .....	136
Chapter 4.	An Advanced Amine System Response Surface Model for the IECM.....	138
4.1.	Motivation.....	138
4.2.	Amine-Based Capture Processes .....	138
4.3.	Parameters for the Advanced Amine-Based CO <sub>2</sub> Capture System .....	140
4.4.	Model Adjustments for the Advanced Amine-Based CO <sub>2</sub> Capture System .....	143
4.5.	Layout of the IECM User Screens for the Amine System .....	148
4.6.	Comparison of Plants with Conventional vs. Advanced Amine-Based Capture	150
4.7.	Case Studies of PC and NGCC Plants with Advanced Amine-Based Capture...	151
4.7.1.	Introduction to the Case Studies .....	152
4.7.2.	Summary Results from Case Studies 7-9.....	152
4.8.	Chapter Conclusion.....	152
Chapter 5.	Comparing Chilled Ammonia and Advanced Amine-Based CO <sub>2</sub> Capture.....	154
5.1.	Introduction.....	154
5.2.	Baseline Model Comparisons .....	154
5.3.	Sensitivity Analysis Across Critical Parameters.....	157
5.3.1.	Sensitivity Analysis by Fuel Type and Cost.....	158
5.3.2.	Sensitivity Analysis by Plant Size and Train Size.....	161
5.4.	Probabilistic Cost Difference for Ammonia and Amine-Based Systems.....	165
5.4.1.	Introduction .....	165
5.4.2.	Selection of Uncertainty Distributions.....	166
5.5.	Chapter Conclusion.....	168
Chapter 6.	Policy Implications .....	170
6.1.	Introduction.....	170
6.2.	The Potential Effect of Performance and Cost Estimates on Policy Decisions..	171
6.2.1.	Information for EPA Guidelines.....	171

6.2.2.	Effect of CO <sub>2</sub> Prices .....	172
6.2.3.	The Effect of Site Specific Factors .....	173
6.3.	Potential Retrofit Costs .....	175
6.4.	Technological Maturity of Ammonia and Amine Technologies .....	177
6.5.	Chapter Conclusion.....	178
Chapter 7.	Summary and Conclusions .....	179
7.1.	Thesis Summary.....	179
7.2.	Main Results and Implications.....	181
Chapter 8.	References .....	183
Appendix A.	Case Study Generation .....	190
Appendix B.	Ammonia Based-CO <sub>2</sub> Capture at an NGCC Plant.....	207
Appendix C.	Aspen Plus Code.....	209

## List of Figures

Figure 1.1: Historical and projected global greenhouse gas emissions and CO <sub>2</sub> concentrations (left) and estimates of the associated average equilibrium temperature increase (right) (from IPCC, 2007).....	12
Figure 1.2: An amine-based CO <sub>2</sub> capture system used to purify natural gas at British Petroleum's In Salah Plant in Algeria. Photo courtesy of IEA Greenhouse Gas Programme.....	18
Figure 1.3: Amine-based post-combustion CO <sub>2</sub> capture systems treating a portion of the flue gas from a coal-fired power plant in Oklahoma, USA (left) and a natural gas combined cycle (NGCC) plant in Massachusetts, USA (right). Photos courtesy of ABB Lummus, Fluor Daniels and Chevron. ....	19
Figure 2.1: Conceptual diagram of primary ammonia system flows.....	31
Figure 2.2: The ammonia-based CO <sub>2</sub> capture performance model as built in Aspen Plus® V7.238	
Figure 2.3: CO <sub>2</sub> capture and absorber NH <sub>3</sub> slip for several lean solvent NH <sub>3</sub> /CO <sub>2</sub> ratios.....	42
Figure 2.4: CO <sub>2</sub> capture vs. NH <sub>3</sub> concentration.....	43
Figure 2.5: Absorber NH <sub>3</sub> slip vs. NH <sub>3</sub> concentration. Diamonds represent 90% CO <sub>2</sub> capture...	43
Figure 2.6: Lean solvent NH <sub>3</sub> concentration vs. solids flow rate. Diamonds represent 90% CO <sub>2</sub> capture. ....	44
Figure 2.7: CO <sub>2</sub> capture as a function of lean sorbent NH <sub>3</sub> concentration and NH <sub>3</sub> /CO <sub>2</sub> ratio. The labels in the figures represent the lean solvent NH <sub>3</sub> /CO <sub>2</sub> ratio.....	51
Figure 2.8: NH <sub>3</sub> slip (ppm) as a function of lean sorbent NH <sub>3</sub> concentration and NH <sub>3</sub> /CO <sub>2</sub> ratio. The labels in the figures represent the lean solvent NH <sub>3</sub> /CO <sub>2</sub> ratio. ....	51
Figure 2.9: Wt% solids in rich sorbent exiting the absorber as a function of lean sorbent NH <sub>3</sub> concentration and NH <sub>3</sub> /CO <sub>2</sub> ratio. The labels in the figures represent the lean solvent NH <sub>3</sub> /CO <sub>2</sub> ratio.....	52
Figure 2.10: Ammonia slip and absorber cooling requirements as a function of absorber temperature for three levels of CO <sub>2</sub> capture (80, 90 and 95%). The labels represent the % CO <sub>2</sub> captured. Reducing the temperature below 283°K results in a step change in the unit energy required for cooling. ....	53
Figure 2.11: Wash water flow rate and associated NH <sub>3</sub> stripper energy required to reduce the ammonia slip in the flue gas stream to 10 ppm, for various levels of ammonia slip exiting the absorber.....	54
Figure 2.12: Levelized cost of electricity and CO <sub>2</sub> avoidance cost as a function of CO <sub>2</sub> capture efficiency for the baseline PC power plant. All costs in constant 2007 US dollars. ....	59
Figure 2.13: A tornado graph indicating the change in the revenue required for a +/- 10% change in the input of ten important variables.....	60
Figure 2.14: Cumulative probability distribution of the levelized cost of electricity of the baseline PC plant with ammonia-based CO <sub>2</sub> capture .....	63
Figure 3.1: Configure Plant Screen for the IECM for a PC plant without environmental controls. ....	69

Figure 3.2: Configure Plant Screen for the IECM with environmental controls intended to meet requirements for a New Source Performance Standard.....	70
Figure 4.1: Modified CO <sub>2</sub> Capture / Config Menu .....	149
Figure 4.2: Modified CO <sub>2</sub> Capture / Capture Menu .....	150
Figure 4.3: Net plant efficiency by power plant and coal type for plants with conventional MEA and advanced amine CO <sub>2</sub> capture systems .....	151
Figure 5.1: Net plant efficiency by power plant and coal type.....	159
Figure 5.2: Revenue requirement by power plant and coal type .....	159
Figure 5.3: Revenue requirement by power plant and coal cost.....	160
Figure 5.4: Revenue requirement by power plant and natural gas cost.....	160
Figure 5.5: Revenue requirement by PC power plants with and without CO <sub>2</sub> capture as a function of plant size. The CO <sub>2</sub> capture systems use the default train sizes.....	162
Figure 5.6: Revenue requirement by PC power plants with CO <sub>2</sub> capture as a function of plant size. Each CO <sub>2</sub> capture system uses a small train size of 208.7 tonnes/hr and a large train size of 907.2 tonnes/hr.....	163
Figure 5.7: Revenue requirement by NGCC power plants with and without CO <sub>2</sub> capture system as a function of plant size. The CO <sub>2</sub> capture systems use the default train sizes. ....	164
Figure 5.8: Revenue requirement by NGCC power plants with and without CO <sub>2</sub> capture system as a function of plant size. Each CO <sub>2</sub> capture system uses a small train size of 208.7 tonnes/hr and a large train size of 907.2 tonnes/hr. ....	164
Figure 5.9: Probabilistic revenue required difference between the supercritical PC plant with the ammonia-based CO <sub>2</sub> capture system and the supercritical PC plant with the advanced amine-based CO <sub>2</sub> capture system. ....	168
Figure 6.1: Effect of a CO <sub>2</sub> price on the levelized cost of electricity for: (a) a supercritical power plant with an ammonia-based system that captures 90% CO <sub>2</sub> (Case 5); (b) the same plant with an ammonia-based system that captures 80% CO <sub>2</sub> (IECM); and (c) a supercritical power plant without a CCS System (Case 2). ....	173
Figure 6.2: Effect of available cooling on the plant efficiency and the revenue required for the supercritical power plant with an ammonia-based system that captures 90% CO <sub>2</sub> (Case 5). Shown in the graph are two cases, one with no additional capital cost, and one with 20% additional cost applied to the process facilities cost of the CO <sub>2</sub> capture system.....	175
Figure B.1: Ammonia concentration vs. the CO <sub>2</sub> capture system steam use and plant revenue requirement.....	208

## List of Tables

Table 1.1: Installed electrical generation capacity and energy production by type.....	13
Table 1.2: Commercial post-combustion capture processes at power plants and selected industrial facilities that capture, transport and sequester CO <sub>2</sub> <sup>1</sup> .....	17
Table 1.3: Planned demonstration projects at power plants with full-scale post-combustion CO <sub>2</sub> capture <sup>1</sup> .....	21
Table 2.1: Baseline plant assumptions.....	26
Table 2.2: Summary of case study results for power plants without CO <sub>2</sub> capture.....	28
Table 2.3: Comparison of plant derating estimates for amine and ammonia-based CO <sub>2</sub> capture systems.....	34
Table 2.4: Key variables for the power plant and ammonia-based CO <sub>2</sub> capture system.....	41
Table 2.5: Key process conditions and flows for the LowNH <sub>3</sub> and HighNH <sub>3</sub> system designs ....	45
Table 2.6: Power plant performance estimates. All values are in MWe equivalent. ....	46
Table 2.7: Power plant cost estimates with ammonia-based CO <sub>2</sub> capture, values are in 2007 \$Millions.....	48
Table 2.8: Power plant performance estimates. All values are in MWe equivalent.....	56
Table 2.9: Power plant cost estimates. All values are in 2007 \$Millions.....	57
Table 2.10: Nominal values and uncertainty parameters assessed in the ammonia-based CO <sub>2</sub> capture system.....	61
Table 3.1: Varied parameters in the model for coal and natural gas plants.....	72
Table 3.2: Partial structure of the fractional factorial design.....	72
Table 3.3: Response variables in the model.....	75
Table 3.4: Power plant performance estimates. All values are in MWe equivalent.....	93
Table 3.5: Power plant cost estimates. All values are in \$2007 constant dollars. ....	107
Table 3.6: Duplication of components when an additional train is added.....	126
Table 3.7: Inputs required from the performance model in the IECM.....	129
Table 3.8: Inputs required from the IECM user.....	130
Table 3.9: Additional inputs required from the IECM user when bypass is enabled.....	130
Table 3.10: Inputs required from the IECM user.....	130
Table 3.11: Inputs required from the IECM user.....	131
Table 3.12: Inputs required from the IECM user.....	131
Table 3.13: Inputs required from the IECM user.....	132
Table 3.14: Inputs required from the IECM user.....	132
Table 3.15: Inputs required from the IECM user.....	133
Table 3.16: Mass balance outputs required from the performance model in the IECM.....	133
Table 3.17: Energy usages of the ammonia system equipment.....	134
Table 3.18: Capital costs of the ammonia system equipment.....	135
Table 3.19: Operating and maintenance costs of the ammonia system equipment.....	135

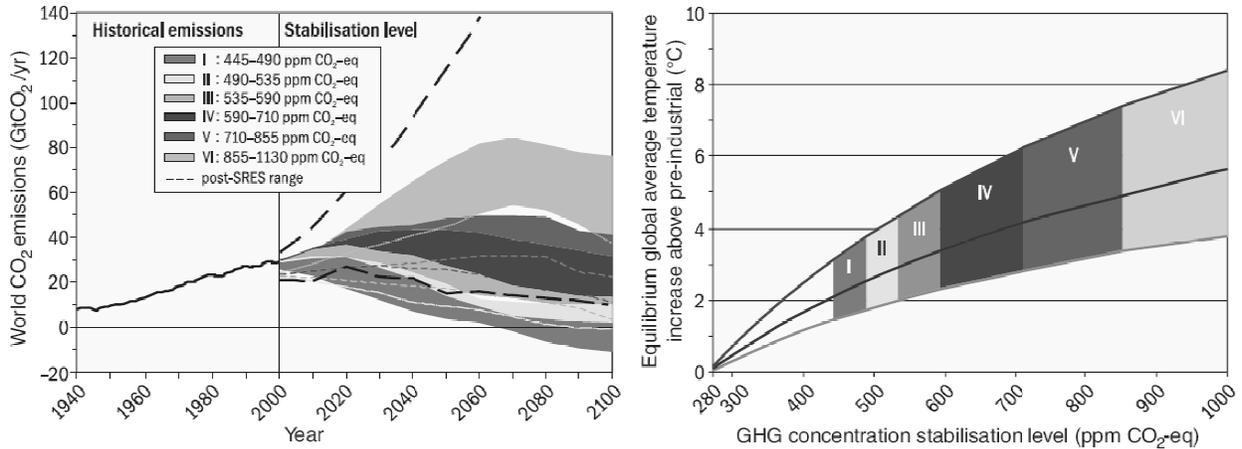
Table 3.20: Summary of case study results for power plants with ammonia-based CO <sub>2</sub> capture	136
Table 4.1: Set Parameters / CO <sub>2</sub> Capture / CO <sub>2</sub> Capture System Process / Config Menu.....	141
Table 4.2: Set Parameters / CO <sub>2</sub> Capture / CO <sub>2</sub> Capture System Process / Performance Menu	141
Table 4.3: Set Parameters / CO <sub>2</sub> Capture / CO <sub>2</sub> Capture System Process / Capture Menu .....	142
Table 4.4: Set Parameters / CO <sub>2</sub> Capture / CO <sub>2</sub> Capture System Process / CO <sub>2</sub> Storage Menu	142
Table 4.5: Set Parameters / CO <sub>2</sub> Capture / CO <sub>2</sub> Capture System Process / Retrofit Cost Menu	142
Table 4.6: Set Parameters CO <sub>2</sub> Capture / CO <sub>2</sub> Capture System Process / Capital Cost Menu..	143
Table 4.7: Set Parameters / CO <sub>2</sub> Capture / CO <sub>2</sub> Capture System Process / O&M Cost Menu..	143
Table 4.8: Comparison of power plants with conventional and advanced amine-based CO <sub>2</sub> capture systems.....	150
Table 4.9: Summary of case study results for power plants with advanced amine-based CO <sub>2</sub> capture .....	152
Table 5.1: Power plant performance estimates for all case studies. ....	155
Table 5.2: Power plant cost estimates for all case studies. All values are in 2007 \$Millions ....	156
Table 5.3: Nominal values and uncertainty parameters assessed in the CO <sub>2</sub> Capture systems ..	167
Table 6.1: Summary of case study results for the retrofit analysis.....	176
Table A.1: IECM parameters changed for Case 1: subcritical plants without CO <sub>2</sub> capture.....	190
Table A.2: IECM parameters changed for Case 2: supercritical plants without CO <sub>2</sub> capture....	192
Table A.3: IECM parameters changed for Case 3: NGCC plants without CO <sub>2</sub> capture .....	194
Table A.4: IECM parameters changed for Case 4: subcritical plants with ammonia-based CO <sub>2</sub> capture .....	195
Table A.5: IECM parameters changed for Case 5: supercritical plants with ammonia-based CO <sub>2</sub> capture .....	197
Table A.6: IECM parameters changed for Case 6: NGCC plants with ammonia-based CO <sub>2</sub> capture .....	199
Table A.7: IECM parameters changed for Case 7: subcritical plants with amine-based CO <sub>2</sub> capture .....	201
Table A.8: IECM parameters changed for Case 8: supercritical plants with amine-based CO <sub>2</sub> capture .....	203
Table A.9: IECM parameters changed for Case 9: NGCC plants with amine-based CO <sub>2</sub> capture .....	205

# **Chapter 1. The Role of Carbon Capture and Storage in Climate Change Mitigation**

## **1.1. CO<sub>2</sub> Emissions and Climate Change**

As anthropogenic greenhouse gas emissions increase atmospheric greenhouse gas concentrations, the resulting climate system warming poses an increasingly serious environmental challenge. According to the Intergovernmental Panel on Climate Change (IPCC), it is now very likely that these emissions are responsible for most of the recent measured global average surface temperature increase, about 0.65°C in total or 0.10-0.16°C per decade, in the 50 years between 1956 and 2005 (IPCC, 2007). As shown Figure 1.1, equilibrium global average temperatures depend on the concentration of atmospheric greenhouse gases, and these concentrations in turn depends on the rate of global greenhouse gas emissions (ibid). Rising global average temperatures increase the risk of extreme changes in the climate, and impacts natural systems supporting water availability, species habitat, and food production.

A significant global reduction in greenhouse gas emissions is required to limit these risks and impacts (IPCC, 2007; NAP, 2011a; NAP, 2011b). In particular, there is some consensus that this warming should be limited to 2°C above pre-industrial levels, requiring an 80%-85% decrease in global CO<sub>2</sub> emissions by mid-century (IPCC, 2007; Haszeldine, 2009). If these goals to reduce the risks associated with climate change are to be met, a large reduction of CO<sub>2</sub> emissions is needed.



**Figure 1.1: Historical and projected global greenhouse gas emissions and CO<sub>2</sub> concentrations (left) and estimates of the associated average equilibrium temperature increase (right) (from IPCC, 2007)**

## 1.2. CO<sub>2</sub> Emissions and Electric Power

Today, approximately 30 gigatonnes of CO<sub>2</sub> out of the 49 gigatonnes of CO<sub>2</sub>-equivalent that are emitted globally from anthropogenic sources each year are the result of burning fossil fuels (IPCC, 2007). Through power plant combustion processes, fossil fuels play a large role in powering the North American electric grid which covers the U.S., Canada, and parts of Baja California Norte in Mexico (through the Western Interconnection) (MIT, 2011). This grid and is one of the largest and most complex machines ever built (Grant, 2007). It is responsible for about 40% of U.S. CO<sub>2</sub> emissions and about 16% of Canadian greenhouse gas emissions (EIA, 2010a; Statistics Canada, 2009).

Table 1.1 describes the breakdown of power generation capacity in the North American electric grid for the U.S. and Canada. The U.S. is heavily invested in fossil fuels with more than 340 GW of coal and 450 GW of natural gas capacity (EIA, 2010b). Canada meanwhile has significant hydro electric generation capacity, but is also still dependent on coal and natural gas.

**Table 1.1: Installed electrical generation capacity and energy production by type**

Source	U.S. Installed Capacity, 2010 (GW) <sup>1</sup>	U.S. Generation, 2010 (BkWh/year) <sup>2</sup>	Canada Installed Capacity, 2007 (GW) <sup>3,4</sup>	Canada Generation, 2007 (BkWh/year) <sup>3</sup>
Hydro	78.2	260.2	73.4	364.1
Wind	39.5	94.7	1.6	2.9
Coal	342.3	1,847.3	16.2	94.3
Nuclear	106.7	807.0	13.3	88.2
Natural Gas	467.2	987.7	10.9	37.4
Other	104.6	128.1	8.8	30.6
Total	1138.5	4,125.0	124.2	617.5

<sup>1</sup>(EIA, 2010b). <sup>2</sup>(EIAa, 2010a). <sup>3</sup>For electric utilities and industry (Statistics Canada, 2009). <sup>4</sup>(National Energy Board, 2008)

Electricity production through fossil fuels at a single power plant can lead to a significant mass flow of CO<sub>2</sub> emissions into the environment. For example, a typical 550 MWe coal-fired power plant with a standard suite of environmental controls emits about 3-4 million tonnes of CO<sub>2</sub> per year or about 0.3-0.4% of a GtCO<sub>2</sub>/year (for a sense of this scale, refer to the y-axis of the left frame in Figure 1.1, shown in GtCO<sub>2</sub>/year). For larger plants the emissions can be significantly higher. For example, the large Nanticoke coal-fired power plant in southern Ontario was originally built with a nameplate capacity of 3960 MWe and it released almost 15 million tonnes of CO<sub>2</sub> per year into the atmosphere. In recent years the Nanticoke plant has reduced its capacity and it now releases approximately 6 million tonnes of CO<sub>2</sub> per year, but this still represents about 0.9% of Canada's current total CO<sub>2</sub> emissions (about 690 million tonnes of CO<sub>2</sub> per year) (Environment Canada, 2011a; Environment Canada, 2011b). Globally, coal is abundant and inexpensive, and its use in some parts of the world is growing. The US, China, and India in particular have large reserves of coal, and China is expanding its coal use significantly, building the equivalent of about two 500 MWe coal-fired power plants each week (MIT, 2007). In 2008, coal-fired electric power generation released 35% of U.S. CO<sub>2</sub> emissions from fossil fuel combustion processes, while natural gas based electric power generation released an additional 6% (EPA, 2010).

### **1.3. A Potential Solution: Carbon Capture and Storage**

One approach to CO<sub>2</sub> emission reductions from large point sources is carbon capture and storage (or sequestration) (CCS). CCS could significantly reduce CO<sub>2</sub> emissions from coal and natural gas-fired power plants, as well as from other large industrial sources. CCS involves two steps: 1) capturing the CO<sub>2</sub> normally emitted during combustion, and 2) transporting and injecting the CO<sub>2</sub> in an appropriate geologic formation. In a power plant equipped with CCS, the CO<sub>2</sub> in the combustion gases is selectively captured and purified using a CO<sub>2</sub> capture technology instead of being vented to the atmosphere. The CO<sub>2</sub> is then compressed to form a liquid-like supercritical fluid that is stored in an underground formation. The potential for CCS to reduce CO<sub>2</sub> emissions is significant because of the broad use of fossil fuels in electric power generation. Many organizations acknowledge the potential of CCS, recognizing it as a necessary component in lowering CO<sub>2</sub> emissions if fossil fuels are going to continue to be used in power generation (DOE/NETL, 2010a). CCS is also thought to be a key part of a least-cost climate strategy (IPCC, 2007; NAP, 2011b).

There are several different power plant designs that could be used for both power production and simultaneous CO<sub>2</sub> capture. These include post-combustion capture in which a CO<sub>2</sub> scrubbing system is added to the end of a power plant flue gas stream; pre-combustion capture in which fuel is converted into H<sub>2</sub> and CO<sub>2</sub> and these are separated so that the H<sub>2</sub> can be combusted while the CO<sub>2</sub> is sent to storage; and oxy-combustion capture in which air is separated into N<sub>2</sub> and O<sub>2</sub> and the fuel is combusted with O<sub>2</sub> resulting in a concentrated CO<sub>2</sub> stream which can be sent to storage. All of these technologies are thought to be technically possible at large scales, and each of these technologies has advantages and disadvantages under different circumstances. However, in their current state of development these technologies are expensive and have not yet been

proven at full commercial scale. A growing community of universities, government labs, and industrial partners are investigating less expensive CCS technologies and are moving these technologies through pilot and demonstration phases towards commercialization.

Both of the technologies investigated in this thesis are post-combustion CO<sub>2</sub> capture technologies. The plant configuration with post-combustion CO<sub>2</sub> capture is the most likely to be competitive as a retrofit option for existing power plants (Rochelle, 2009). This is an important consideration because a large portion of future power sector CO<sub>2</sub> emissions will occur at existing power plants, and post-combustion CO<sub>2</sub> capture may be a useful technology to reduce CO<sub>2</sub> emissions at these plants (ITFCCS, 2010). In addition, other industrial point sources account for a significant amount of CO<sub>2</sub> emissions, and post-combustion capture can also be applied to many of these sources including at oil and gas refineries, chemical plants, pulp and paper plants, and iron and steel plants (van Straelen, 2010; ITFCCS, 2010). The development of post-combustion CO<sub>2</sub> capture technologies is being done in this context.

#### **1.4. The Current State of Post-Combustion CO<sub>2</sub> Capture and Storage**

In recent years, carbon capture research and development programs have expanded rapidly throughout the world; thus, any summary of “current” activities and projects is soon out of date. For this reason, it is difficult to be comprehensive in covering all post-combustion capture-related activities. Rather, this section attempts to synthesize key findings from the literature and from online databases which track and report on the status of CO<sub>2</sub> capture technology developments. Excellent publicly available databases and CCS project status reports are maintained by organizations including the U.S. Department of Energy’s National Energy Technology Laboratory (DOE/NETL), the International Energy Agency’s Greenhouse Gas Control Programme (IEAGHG), the Massachusetts Institute of Technology (MIT) Carbon

Sequestration Program and the Global Carbon Capture and Storage Institute (GCCSI). In many cases, the information from public databases and reports presented below has also been supplemented and checked by additional data from companies and research groups involved in capture technology development and testing. In each of the sections below, the objective is to summarize the current status of post-combustion capture technology developments (as of December 2011).

#### **1.4.1. Commercial Projects for CO<sub>2</sub> Capture and Storage**

Post-combustion CO<sub>2</sub> capture systems have been in use commercially for many decades, mainly in industrial processes for purifying natural gas streams, though they have also been used on combustion-based flue gas streams. The use of amines to capture CO<sub>2</sub> was first patented eighty years ago (Bottoms, 1930) and since that time amine-based systems have been used to meet CO<sub>2</sub> product specifications in industries ranging from natural gas production to the food and beverage industry (DOE/NETL, 2010b; Rochelle, 2009; EPA 2006). A number of vendors currently offer commercial amine-based processes, including the Fluor Daniel Econamine FG Plus process, the Mitsubishi Heavy Industries KM-CDR process, the Lummus Kerr-McGee process, the Aker Clean Carbon Just Catch process, the Cansolv CO<sub>2</sub> capture system and the HTC Pureenergy Process (CATF, 2009; GCCSI, 2009).

There are hundreds of commercial aqueous amine systems in operation. Most systems currently are used for removing acid gases from a product stream, and the captured CO<sub>2</sub> is typically vented to the atmosphere. Table 1.2 lists several of these projects, including three recent projects at natural gas treatment plants (two in Norway, one in Algeria), in which the captured CO<sub>2</sub> is sequestered in deep geological formations to prevent its release to the atmosphere. One of these projects, the Statoil natural gas production facility at Sleipner in the North Sea, has been

operating since 1996. This is the longest-running commercial CCS project. Figure 1.2 shows a photograph of the amine-based CO<sub>2</sub> capture unit installed more recently at a natural gas treatment plant in Algeria. That unit is part of an integrated CCS system that includes CO<sub>2</sub> capture, pipeline transport and storage in a nearby geological formation.

**Table 1.2: Commercial post-combustion capture processes at power plants and selected industrial facilities that capture, transport and sequester CO<sub>2</sub><sup>1</sup>**

Project Name and Location	Plant and Fuel Type	Year of Startup	Approx. Capture Plant Capacity	Capture System Type (Vendor)	CO <sub>2</sub> Captured (10 <sup>6</sup> tonnes/yr)
Projects in the U.S.					
IMC Global Inc. Soda Ash Plant (Trona, CA)	Coal and petroleum coke boilers	1978	43 MW	Amine (Lummus)	0.29
AES Shady Point Power Plant (Panama City, OK)	Coal-fired power plant	1991	9 MW	Amine (Lummus)	0.06
Bellingham Cogeneration Facility (Bellingham, MA)	Natural gas-fired power plant	1991	17 MW	Amine (Fluor)	0.11
Warrior Run Power Plant (Cumberland, MD)	Coal-fired power plant	2000	8 MW	Amine (Lummus)	0.05
Projects outside the U.S.					
Soda Ash Botswana Sua Pan Plant (Botswana)	Coal-fired power plant	1991	17 MW	Amine (Lummus)	0.11
Sumitomo Chemicals Plant (Japan)	Gas and coal boilers	1994	8 MW	Amine (Fluor)	0.05
Statoil Sleipner West Gas Field (North Sea, Norway) <sup>2</sup>	Natural gas separation	1996	N/A	Amine (Aker)	1.0
Petronas Gas Processing Plant (Kuala Lumpur, Malaysia)	Natural gas-fired power plant	1999	10 MW	Amine (MHI)	0.07
BP Gas Processing Plant (In Salah, Algeria) <sup>2</sup>	Natural gas separation	2004	N/A	Amine (Multiple)	1.0
Mitsubishi Chemical Kurosaki Plant (Kurosaki, Japan)	Natural gas-fired power plant	2005	18 MW	Amine (MHI)	0.12
Snøhvit Field LNG and CO <sub>2</sub> Storage Project (North Sea, Norway) <sup>2</sup>	Natural gas separation	2008	N/A	Amine (Aker)	0.7
Huaneng Co-Generation Power Plant (Beijing, China)	Coal-fired power plant	2008	0.5 MW	Amine (Huaneng)	0.003

<sup>1</sup>(DOE/NETL, 2010b; IEAGHG, 2010; MIT, 2010; GCCSI, 2009). <sup>2</sup>Plants with an Integrated CCS System.



**Figure 1.2: An amine-based CO<sub>2</sub> capture system used to purify natural gas at British Petroleum's In Salah Plant in Algeria. Photo courtesy of IEA Greenhouse Gas Programme.**

As shown in Table 1.2, CO<sub>2</sub> is also captured at several coal-fired and natural gas-fired power plants where a portion of the flue gas stream is fitted with a CO<sub>2</sub> capture system. Figure 1.3 shows the amine systems installed at two U.S. power plants, one burning coal, and the other burning natural gas. At these plants, the captured CO<sub>2</sub> is sold to nearby food processing facilities, which use it to make dry ice or carbonated beverages. The oldest and largest commercial CO<sub>2</sub> capture system operating on flue gases is the IMC Global soda ash plant in California. Here, the mineral trona is mined locally and combined with CO<sub>2</sub> to produce sodium carbonate (soda ash), a widely-used industrial chemical (IEAGHG, 2010). In all cases, these products soon release the CO<sub>2</sub> to the atmosphere (e.g., from carbonated beverages).



**Figure 1.3: Amine-based post-combustion CO<sub>2</sub> capture systems treating a portion of the flue gas from a coal-fired power plant in Oklahoma, USA (left) and a natural gas combined cycle (NGCC) plant in Massachusetts, USA (right). Photos courtesy of ABB Lummus, Fluor Daniels and Chevron.**

Only ABB Lummus (now CB&I Lummus) currently has commercial flue gas CO<sub>2</sub> capture units operating at coal-fired power plants, while both Fluor Daniel and Mitsubishi Heavy Industries have commercial installations at gas-fired plants (see Table 1.2). Both Fluor and Mitsubishi Heavy Industries now also offer commercial guarantees for post-combustion capture at coal-fired power plants.

These vendors (and others) use amine-based solvents for CO<sub>2</sub> capture. In most cases the exact composition of the solvent is proprietary. The currently operating Lummus systems employ a solution of about 20 weight percent monoethanolamine (MEA) in water, while the Fluor systems use a solvent that has a 30 weight percent MEA concentration (Feron, 2009; Jensen, 2006). Higher amine concentrations can be beneficial in reducing the energy requirements of a CO<sub>2</sub> capture process since there is less water in the solution that requires heating in the regeneration process. Capital cost also is reduced since higher amine concentrations lead to smaller equipment sizes. On the other hand, amines such as MEA are highly corrosive and higher amine

concentrations require chemical additives or more costly construction materials in order to prevent corrosion. Tradeoffs among these factors underlie some of the differences in capture system designs offered by different vendors. The systems and solvents currently offered commercially by Fluor (Econamine FG Plus) and Mitsubishi Heavy Industries (KS-1) are advertised by these vendors as reducing the capture energy requirements relative to older MEA-based system designs by roughly 25 percent, which lowers the overall cost.

#### **1.4.2. Full-Scale Demonstration Plants**

Although several CO<sub>2</sub> capture systems have operated commercially for nearly two decades on a portion of power plant flue gases, no capture units have yet been applied to the full flue gas stream of a modern coal-fired or natural gas-fired power plant. Thus, demonstrating post-combustion CO<sub>2</sub> capture at full scale is widely regarded as crucial for gaining the acceptance of this technology by electric utility companies and by the institutions that finance and regulate power plant construction and operation. Several years ago, for example, the European Union called for twelve such demonstrations in Europe, while in the United States there have been calls for at least six to ten full-scale projects (ZEP, 2008; MIT, 2007; Pew Center, 2007).

To date however, no such demonstrations have yet occurred, nor (to the best of the author's knowledge) has full financing yet been guaranteed for any of the full-scale demonstration projects that have been announced. One reason is the high cost of each project, estimated at roughly one billion dollars for CO<sub>2</sub> capture at a 400 MW unit operating for five years (Pew Center, 2007). Several previously announced demonstrations of full-scale power plant capture and storage systems have been canceled or delayed due to sharp escalations in construction costs and the evolving U.S. political environment. Even more recently, a 160 MW demonstration project in the U.S. was canceled not long after being announced (Sourcewatch, 2010).

Nevertheless, it appears reasonable to assume that at least some of the large-scale projects currently planned for post-combustion CO<sub>2</sub> capture in the U.S. and other countries will materialize over the next several years, with costs shared between the public and private sectors.

Table 1.3 lists the features and locations of the major post-combustion capture demonstration projects planned at power plants in the United States and other countries as of December 2011. Most of these CO<sub>2</sub> capture systems would be installed at existing coal-fired plants, with the captured CO<sub>2</sub> transported via pipeline to a geological storage site, often in conjunction with enhanced oil recovery to reduce project costs. Most of the planned demonstration projects have expected startup dates of 2014 or later. This means that such projects are currently in the early stages of detailed design and that final commitments of full funding for construction and operation have not yet been made.

**Table 1.3: Planned demonstration projects at power plants with full-scale post-combustion CO<sub>2</sub> capture<sup>1</sup>**

Project Name and Location	Plant and Fuel Type	Year of Startup	Approx. Capture Plant Capacity	Capture System Type (Vendor)	CO <sub>2</sub> Captured (10 <sup>6</sup> tonnes)
<b>Projects in the U.S.</b>					
Tenaska Trailblazer Energy Center (Sweetwater, TX)	Coal-fired power plant	2014	600 MW	Amine (Fluor)	4.3
NRG Energy WA Parish Plant (Houston, TX)	Coal-fired power plant	2015	240 MW	Amine (Fluor)	1.5
<b>Projects outside the U.S.</b>					
SaskPower Boundary Dam Polygon (Estevan, Canada)	Coal-fired power plant	2014	115 MW	Amine (Cansolv)	1.0
TransAlta Project Pioneer Keephills 3 Power Plant (Wabamun, Canada)	Coal-fired power plant	2015	200 MW	Chilled Ammonia (Alstom)	1.0
Vattenfall Janschwalde (Janschwalde, Germany)	Coal-fired power plant	2015	125 MW	Amine (TBD)	N/A <sup>3</sup>
PGE Bechatów Power Station (Bechatów, Poland)	Coal-fired power plant	2015	360 MW	Amine (Alstom, Dow Chemical)	1.0
Porto Tolle (Rovigo, Italy)	Coal-fired power plant	2015	200 MW <sup>2</sup>	Amine (TBD)	1.0
SSE Peterhead Power Station (Peterhead, UK)	Gas-fired power plant	2015	385 MW <sup>2</sup>	N/A <sup>3</sup>	1.0

<sup>1</sup>(DOE/NETL, 2010b; IEAGHG, 2010; MIT, 2010; GCCSI, 2009).<sup>2</sup>Estimated from other reported data. <sup>3</sup>Currently not available

While many technologies are being pursued for post-combustion CO<sub>2</sub> mitigation, at present only two technologies have emerged that are being incorporated into the designs of these planned large-scale post-combustion demonstration plants: amine and ammonia-based post-combustion CO<sub>2</sub> capture technologies. Both processes are described in more detail in later chapters of this thesis. Plans for scale-up to a demonstration project are predicated on the successful operation of smaller scale pilot plants as well as funding, and public and legislative support for larger facilities.

### **1.5. Objectives and Scope of this Thesis**

The expected startup dates for the first large scale post-combustion CO<sub>2</sub> capture technologies shown in Table 1.3 suggest that these processes are still in the early stages of technological development. Yet, it is not difficult to find strongly divided opinions on the use of CCS. Here are two examples at the extremes (Stephens and Jiusto, 2010):

“Carbon capture and storage technologies are essential to allow the continued use of coal to generate electricity while we substantially reduce emissions of green house gases to combat global warming. While the technologies are complex, the overall value of introducing them into the U.S. and global economies is undeniable” (from the Center for American Progress, 2009).

“Carbon capture and storage is a scam... Governments and businesses need to reduce their emissions, not search for excuses to keep burning coal” (from Greenpeace, 2009).

Since CCS in power plants has not yet been realized at a commercial scale there are significant uncertainties in the technology. Therefore, included in the above opinions are large unstated assumptions about costs, risks, other technological options, and judgments about what the future should look like (Stephens and Jiusto, 2010).

This is a difficult situation for policy makers who face questions about the performance, cost, and the effectiveness of these technologies as they assess CO<sub>2</sub> emission mitigation options (Schaeffer et al., 2008). The purpose of this thesis is to present a transparent, engineering-based, technical and economic assessment of the two leading post-combustion CO<sub>2</sub> capture technologies: amine-based systems using MEA, and Chilled Ammonia-based systems. The hope is that this will help create a more fact-based dialogue about CO<sub>2</sub> capture technologies, which in turn will better inform the discourse and contribute to better policy making.

### **1.6. Organization of this Thesis**

This chapter has briefly introduced the connection between CO<sub>2</sub> emissions and climate change, and has established that post-combustion based CCS could be a useful technology for significantly reducing CO<sub>2</sub> emissions at large point sources. This chapter has also limited the scope of this thesis to an investigation of the performance and economics of the two leading post-combustion CO<sub>2</sub> capture technologies: ammonia and amine-based systems. Chapter 2 of this thesis first provides an initial assessment of ammonia-based CO<sub>2</sub> capture. Chapter 3 then describes the development of a general ammonia-based CO<sub>2</sub> capture model for use in the Integrated Environmental Control Model (IECM) that allows one to estimate the performance and costs of power plants with this system. In Chapter 4, a model of an existing amine system in the IECM is modified to reflect more modern designs of amine-based CO<sub>2</sub> systems. In Chapter 5, the ammonia and amine-based systems are compared. Chapter 6 then describes the policy implications of this comparison. Finally, in Chapter 7, the results are summarized and general conclusions drawn on these two major approaches to CO<sub>2</sub> capture.

## **Chapter 2. Ammonia-Based Post-Combustion CO<sub>2</sub> Capture**

### **2.1. Introduction**

This chapter will present a performance assessment and preliminary cost analysis of a supercritical coal-fired power plant with an ammonia-based post-combustion CO<sub>2</sub> capture system. First, this chapter sets the context by defining a set of baseline power plants at which the CO<sub>2</sub> capture systems in this thesis will be implemented. Next, additional background on the development of ammonia-based CO<sub>2</sub> post-combustion CO<sub>2</sub> capture systems is provided. Then, the development of a performance and cost model of an ammonia-based CO<sub>2</sub> capture system is described, along with results for application on the supercritical coal-fired power plant. The technical details provided are intended to be a starting point for estimating the performance and costs for ammonia-based CO<sub>2</sub> capture in a general process, and these details are used in later chapters.

### **2.2. Baseline Assumptions and Baseline Power Plants (Case Studies 1-3)**

This thesis is concerned with the performance, costs, and emissions of power plants with post-combustion CO<sub>2</sub> capture technologies. Before these details are presented it is useful to establish a set of baseline assumptions used throughout the rest of the chapters as well as baseline performance, costs, and CO<sub>2</sub> emissions for power generation technologies without CO<sub>2</sub> capture.

#### **2.2.1. Baseline Assumptions**

The baseline plant configurations, performance, and financial assumptions used for the power plants throughout this thesis are based on a widely-used set of “baseline” plant characteristics specified by the U.S. Department of Energy (DOE/NETL, 2007a) as shown in Table 2.1. As shown in Table 2.1, the pulverized coal power plants in this thesis are either subcritical or

supercritical units that burn Illinois No.6 bituminous coal. The natural gas combined cycle plants burn natural gas with two 7FB turbines. The fixed charge factor reflects whether these plants are financially considered to be relatively low risk (plants without CCS) or relatively high risk (plants with CCS), with higher fixed charge factors used for plants with CCS owing to the increased complexity of these systems and to the lack of commercially proven CO<sub>2</sub> capture technology at large scales. All costs are evaluated in \$2007 constant dollars, and in this analysis inflation is not considered in calculating the costs of electricity. The wide varieties of owner's costs that are sometimes included in other financial analysis are not included here in this analysis for simplicity and clarity. For more detailed estimates for any power plant, owner's costs specific to the project would typically be included.

**Table 2.1: Baseline plant assumptions**

	Pulverized Coal Power Plant	Natural Gas Combined Cycle Plants
<b>Power Plant Specifications</b>		
Nominal Net Plant Power Output (MWe)	550	Constrained by Turbine Size <sup>2</sup>
Steam Cycle	Subcritical <sup>3</sup> or Supercritical <sup>3</sup>	Subcritical <sup>3</sup>
Environmental Controls	SCR, Fabric Filter, Wet FGD	None
Cooling	Wet Cooling Tower	Wet Cooling Tower
Fuel Type	Illinois No. 6	Natural Gas (93% CH <sub>4</sub> )
Fuel Heating Value, As Received (HHV, kJ/kg)	27,113	52,970
Fuel Cost (2007\$/GJ)	\$1.7	\$6.4
Fuel Cost Nominal Escalation (real escalation plus inflation, %)	0%	0%
Capacity Factor (%)	75%	75%
<b>Financial Assumptions</b>		
Cost Year and Type	2007 Constant Dollars	2007 Constant Dollars
Annual Inflation Rate	N/A	N/A
Real Escalation Rate	0%	0%
Weighted Cost of Capital Before Taxes <sup>4</sup> without CCS (with CCS)	8.25% (9.075%)	8.25% (9.075%)
Percentage Debt without CCS (with CCS)	50% (45%)	50% (45%)
Real Bond Interest Rate without CCS (with CCS)	4.5% (5.5%)	4.5% (5.5%)
Percentage Equity without CCS (with CCS)	50% (55%)	50% (55%)
Real Stock Return	12%	12%
Fixed Charge Factor without CCS (with CCS) <sup>f</sup>	0.113 (0.143)	0.113 (0.143)
Years of Construction	5	3
Plant Book Life	30 Years	30 Years
Federal Tax Rate	36%	36%
State Tax Rate	6%	6%
Property Tax Rate	0%	0%
<b>CO<sub>2</sub> Capture System Specifications (When Applicable)</b>		
Flue Gas CO <sub>2</sub> Capture Requirement (%)	90%	90%
Flue Gas CO <sub>2</sub> Concentration into CO <sub>2</sub> Capture System (%)	13.5%	4.2% <sup>1</sup>
CO <sub>2</sub> Product Pressure	15.3MPa	15.3MPa
CO <sub>2</sub> Transportation Distance (km)	80 km	80 km
CO <sub>2</sub> Transport Storage & Maintenance Costs (\$/tonne CO <sub>2</sub> )	\$3.75/tonne CO <sub>2</sub>	\$3.75/tonne CO <sub>2</sub>
Project Contingency	16.4%	16.4%
Process Contingency	4.7%	4.7%

<sup>1</sup>The flue gas CO<sub>2</sub> concentration for the NGCC plant was adjusted by changing the excess air. This is consistent with the DOE/NETL, 2007a report. <sup>2</sup>The NGCC plant uses 2 GE 7FB combustion turbines, which each produce 425.5 MWg. <sup>3</sup>The thermal efficiency of the steam cycle is based on DOE/NETL, 2007a Case 11 and 12 for the PC plant, and Case 14 for the NGCC plant. The subcritical steam cycle is a 16.5MPa/566°C/566°C, and the supercritical cycle is 24.1MPa/593°C/593°C. <sup>4</sup>Also called the discount rate. This is calculated as  $i = (1+c)(1+e_t) - 1$ , where  $i$  is the discount rate,  $c$  is the constant dollar cost of capital (in the absence of inflation), and  $e_t$  is the annual inflation rate. The constant dollar cost of capital is calculated as  $c = \text{Percentage Debt} \times \text{Real Bond Interest Rate} + \text{Percentage Equity} \times \text{Real Stock Return}$ , all in constant dollars.

<sup>f</sup>The fixed charge factor is calculated based on year by year carrying charges and a present worth factor according to the equation:  $FCF = [CC_1 \times (1+i)^{-1} + CC_2 \times (1+i)^{-2} + \dots + CC_n \times (1+i)^{-n}] / a_n$  where  $n$  is the book life of the plant,  $i$  is the interest rate,  $CC$  is the year by year carrying charges of the plant, and  $a_n$  is the present value worth factor for a uniform series. The year by year carrying charges are the sum of: (the return on debt, the return on equity, the payable income taxes, book depreciation, property taxes, and insurance)/the total plant cost (TPC). The value of  $a_n$  is calculated according to the following equation:  $a_n = [(1+i)^n - 1] / [i \times (1+i)^n]$  (EPRI, 1993).

Changes in the assumptions in Table 2.1 can significantly change the overall results, as outlined in the uncertainty analysis included in the following chapters. Due to the large impact that assumptions can have on final costs, the costs for power plants produced in this thesis are most usefully compared to other plants within this thesis, and are not to be compared directly with costs presented in other studies that use different assumptions.

### **2.2.2. Case Study 1-3 Results: PC and NGCC Power Plants without CCS**

The first three case study results presented below in Table 2.2 show representative costs, performance, and CO<sub>2</sub> emissions of a subcritical coal-fired power plant (Case 1), a supercritical coal-fired power plant (Case 2), and a natural gas combined cycle plant (Case 3), all without CCS, using the baseline assumptions in Table 2.1. These results were generated using the baseline assumptions and the existing models in the Integrated Environmental Control Model (IECM V7.0) developed by Carnegie Mellon University. The IECM is freely downloadable online at: <http://www.cmu.edu/epp/iecm/>, and readers are welcome to reproduce the results for Cases 1-3, as well as results presented throughout this thesis, by downloading and using the model for themselves. Qualitatively, the case studies are comparable with the 2007 NETL Baseline report (DOE/NETL, 2007a) though the baseline cost assumptions and therefore the overall costs have been updated to reflect changes in NETL assumptions (DOE/NETL, 2010c). The results from later chapters will be compared against these baseline results. For reproducibility, a more detailed outlines of the models developed for Cases 1-3 can be found in Appendix A.

**Table 2.2: Summary of case study results for power plants without CO<sub>2</sub> capture**

Cases	Gross Plant Output (MWe)	Net Plant Output (MWe)	Net Plant Efficiency (%)	CO <sub>2</sub> Emissions (million tonnes/year)	Capital Cost (\$/kW -net) <sup>1</sup>	Revenue Required (\$/MWh) <sup>1</sup>
Case 1: PC Subcritical	582.8	550.0	36.8%	3.1	1545	53.7
Case 2: PC Supercritical	581.3	550.0	39.1%	2.9	1608	53.7
Case 3: NGCC	546.1 <sup>2</sup>	531.6 <sup>2</sup>	50.0%	1.3	565.1	58.5

<sup>1</sup>All costs are in \$U.S. 2007 constant dollars. <sup>2</sup>The output of the NGCC plant is fixed by the turbine size.

There is an important caveat associated with the capital cost numbers described in Table 2.2. These numbers, as per the 2007 NETL Baseline report (DOE/NETL, 2007a), do not include owner's costs. These additional costs are included in some studies and can add significantly to the capital cost estimates and resulting revenue requirement of these plants.

### 2.3. The Potential Advantages of an Ammonia-Based CO<sub>2</sub> Capture Process

The context for the development of new a post-combustion CO<sub>2</sub> capture process is the current technological limitations and costs of amine-based CO<sub>2</sub> capture systems. Amine-based CO<sub>2</sub> capture processes are the most understood of the post-combustion CO<sub>2</sub> capture technologies that are being designed for large-scale power plants (Rochelle, 2009). However, current amine-based processes are expected to impose large energy penalties and substantial costs when applied to power plant designs. These penalties and costs will be explored in more detail in Chapter 4.

Ammonia-based post-combustion CO<sub>2</sub> capture is being developed to lower CO<sub>2</sub> emissions in the electric power and industrial sectors in parallel with amine-base post-combustion CO<sub>2</sub> capture systems. As of April 2012, this technology is being advanced primarily by Alstom Power in the form of the Chilled Ammonia Process (CAP). In the CAP, an ammonia-based solution is used to selectively capture CO<sub>2</sub> from flue gas streams in a reduced temperature absorber.

The proposed advantages of ammonia-based systems include the high CO<sub>2</sub> carrying capacity of ammoniated solutions when solids are allowed to precipitate in the process; the potential for low

reboiler regeneration energy because of the higher CO<sub>2</sub> carrying capacity; the potential for reduced CO<sub>2</sub> compressor power because CO<sub>2</sub> can be regenerated above atmospheric pressure; and reduced solvent cost because ammonia is relatively inexpensive. Overall, post-combustion CO<sub>2</sub> capture based on ammonia is more complicated and less understood but may be an improvement over amine-based technologies particularly because the process may have a lower energy penalty compared to an amine-based process (Hilton, 2009).

#### **2.4. Historical Development of Ammonia-Based Capture Processes**

A 2005 study found that post-combustion CO<sub>2</sub> capture using ammonia appeared to be a promising avenue to reducing the costs of CO<sub>2</sub> capture over that of amine processes (Ciferno et al., 2005). The preliminary study results found that since ammonia potentially could capture multiple pollutants simultaneously (including CO<sub>2</sub>, SO<sub>2</sub>, NO<sub>x</sub> and Hg), the overall plant cost could be reduced significantly by removing the need for additional environmental control equipment. It was also found that an ammonia-based CO<sub>2</sub> capture process would use considerably less steam from the steam cycle than a process based on amines because the regenerator heat duty would be lower. In addition, substantially less compressor power would be required since the CO<sub>2</sub> could be regenerated at higher pressure than with an amine system. It was noted in this study that ammonia is known to have a higher volatility than MEA and thus is more easily released into the flue gas stream during the absorption step (a process called “ammonia slip”). However, the study suggested that if this and other engineering challenges could be overcome, the overall cost of an ammonia-based system for CO<sub>2</sub> capture would be substantially less than for an amine-based system.

Controlling ammonia slip to acceptable levels turned out to be a major engineering challenge and meeting that challenge is expected to add considerably to the cost of the system (Figueroa et al.,

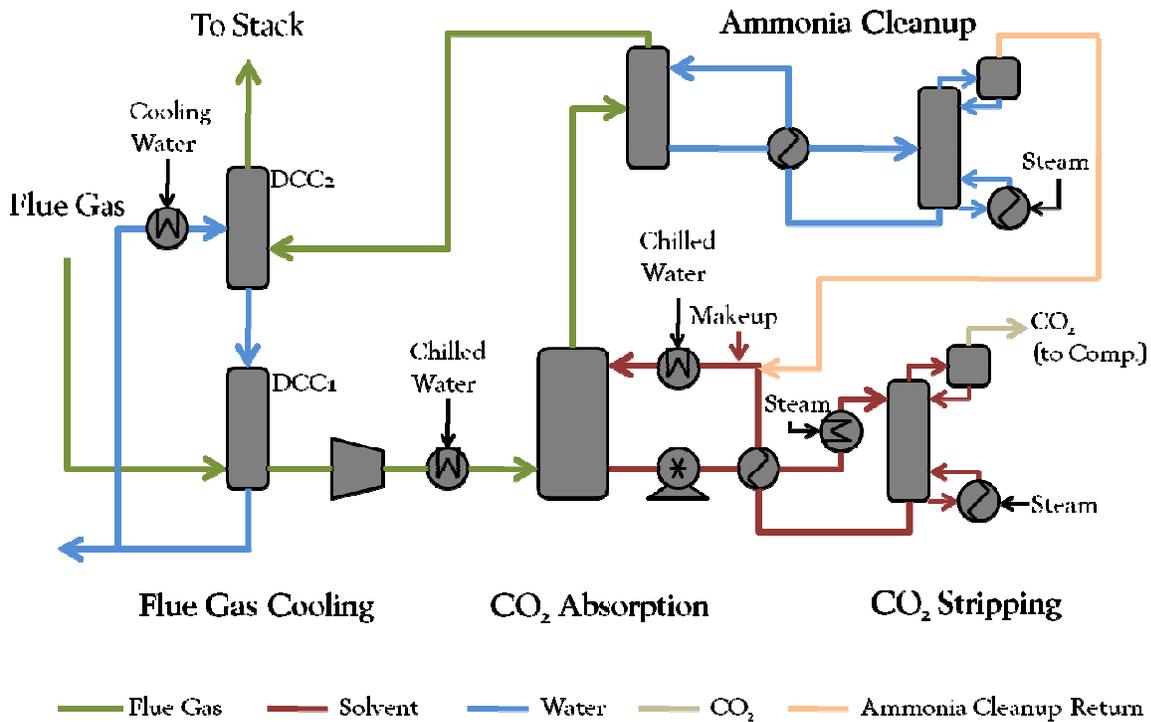
2008). The two major companies that have been involved in developing ammonia-based CO<sub>2</sub> capture, Powerspan and Alstom Power, have attempted to solve this challenge in different ways as outlined below.

#### **2.4.1. The Powerspan ECO<sub>2</sub> Capture Process**

Powerspan was founded in 1994 and has since developed their business model around the Electro-Catalytic Oxidation (ECO) process for NO<sub>x</sub>, SO<sub>2</sub>, particulate, and Hg capture as well as the ECO<sub>2</sub> process for additional CO<sub>2</sub> capture (Powerspan, 2012). These processes use ammonia to capture acid gases from power plant flue gas streams in a combined system, in lieu of separate units such as a wet flue gas desulfurization and selective catalytic reduction systems. The ECO<sub>2</sub> process uses ammonium carbonate to capture CO<sub>2</sub> and operates with a moderate temperature (as opposed to a low temperature) absorption process. Ammonium sulfate from the SO<sub>2</sub> capture portion of the process is used to control ammonia slip in the process. Thus, while amine-based systems must severely limit exposure of the amine solvent to acid gases like SO<sub>2</sub> and NO<sub>2</sub> to prevent solvent loss and degradation, ammonia does not degrade in the presence of these gases; instead, it forms ammonium sulfate and nitrate, which have value as fertilizer by-products (CATF, 2009). Powerspan had tested its ECO<sub>2</sub> process for CO<sub>2</sub> capture at a 1 MW pilot plant at First Energy's R.E. Burger plant (ibid), but has not announced plans to develop the process at larger scales. There is also only very limited public information available on the details of the ECO<sub>2</sub> process. As such, this process is not considered further in this thesis.

#### **2.4.2. The Alstom Chilled Ammonia Process (CAP)**

In contrast, the Chilled Ammonia Process (CAP) being developed by Alstom Power is designed as a stand-alone post-combustion CO<sub>2</sub> capture system. The system diagram depicting the main process details and flows is shown in Figure 2.1 below.



**Figure 2.1: Conceptual diagram of primary ammonia system flows**

In the CAP, the flue gas stream and the CO<sub>2</sub> absorber are cooled to about 283°K (10°C), a temperature that reduces the vapor pressure of ammonia low enough such that excessive amounts of ammonia slip do not exit the absorber with the cleaned flue gas stream. To do this, power plant flue gasses are initially cooled using circulating water and a direct contact cooler, and most of the water in the gases is condensed out. The flue gases are further cooled in a cross flow heat exchanger using chilled water from a vapor compressor. The chilled flue gasses feed into a CO<sub>2</sub> absorption column, where the gases are contacted with a lean solvent mixture. The lean solvent contains ammonia, carbon dioxide, and water, at a specified lean loading NH<sub>3</sub>/CO<sub>2</sub> ratio. Higher ratios result in increased ammonia slip over the absorber and therefore increased flue gas cleaning demands, while lower ratios require additional solvent flow or higher ammonia concentrations to capture high levels of CO<sub>2</sub>. The CO<sub>2</sub> rich stream leaves the bottom of the absorber and is compressed to about 3.0 MPa by a high pressure pump. The rich solution then

flows through a cross flow heat exchanger where it is heated by the hot lean solution coming off the reboiler, and if any solids remain in the stream a heater is used to dissolve them. As with amine system designs, the CO<sub>2</sub> “rich” solution stream is then sent to a stripper column where heat is added (using steam extracted from the power plant steam turbine) to strip CO<sub>2</sub> from the solution. The CO<sub>2</sub> stripper is operated at high pressure and regenerates the CO<sub>2</sub> at about 2.8 MPa. This leaves a nearly pure CO<sub>2</sub> stream that can be cleaned, dried and compressed for transport to a geological storage site. The regenerated solvent is then re-circulated back to the absorber by passing first through a cross flow heat exchanger where its temperature is lowered through heating of the rich solvent, and then it is further cooled before entering the absorber in another heat exchanger with a flow of chilled water. The absorber gasses are cleaned of ammonia in a water wash and are then heated in a second direct contact cooler, before being released through the stack. The precipitation of both ammonium bisulfate (NH<sub>4</sub>HSO<sub>3</sub>) and ammonium chloride (NH<sub>4</sub>Cl) can cause the formation of a visible plume out of the stack (Kohl and Neilson, 1997), and the second direct contact cooler removes these ammonium salts from the flue gas and reduces plume formation through heating of the gas. Finally, distillate from the second stripper containing ammonia, carbon dioxide, and water is fed back to the CO<sub>2</sub> capture system.

As mentioned previously, the energy required in the stripper to regenerate the ammonia-based solvent is believed to be smaller than the energy required for amine systems. However, there is also an important tradeoff between the energy required to cool the process and the additional equipment and energy costs of reducing ammonia slip to acceptable levels. Thus, the overall process design must be optimized to achieve good performance at minimum cost.

Alstom has operated two pilot plants using their chilled ammonia process—one in the United States and one in Norway. The most recent is the pilot plant at the American Electric Power

(AEP) Mountaineer power station in West Virginia, a 1300 MW coal-fired plant, where a flue gas slip stream equivalent to about 20 MW has been fitted with the Alstom Chilled Ammonia Process. This was the first successful integration of CO<sub>2</sub> capture, transport and geological sequestration at a coal-fired power plant. Data from this pilot plant was to provide the basis for the proposed Pioneer Keephills 3 demonstration plant listed in Table 1.3. Estimating the performance and cost of the Chilled Ammonia Process at a large power plant is a primary goal of this thesis.

### 2.5. Past Performance and Cost Estimates of the Chilled Ammonia Process

For estimating performance and costs for post-combustion CO<sub>2</sub> capture technologies in general, this thesis uses the plant derating of CO<sub>2</sub> capture on the power plant as a measure of performance, and the levelized revenue required (also called the levelized cost of electricity or LCOE) as a measures of cost, as shown in equation 1 and equation 2.

Plant Derating (%)

$$= \frac{\text{Plant Efficiency without Capture} - \text{Plant Efficiency with Capture}}{\text{Plant Efficiency without Capture}} \quad (\text{eq. 1})$$

Revenue Required (\$/MWh)

$$= \frac{\text{Total Plant Costs} * \text{Fixed Charge Factor} + \text{O\&M Costs}}{8760 * \text{Capacity Factor} * \text{MW of Net Plant Capacity}} \quad (\text{eq. 2})$$

The plant derating for CO<sub>2</sub> capture is expressed as the percentage reduction in net plant output for a constant energy input and is occasionally reported as an “energy penalty”, though the term “energy penalty” has multiple conflicting definitions in the literature. The revenue requirement is the income the plant would have to receive from the sale of electricity to pay for the capital and operating costs of the plant over the plant lifetime. The revenue required calculation has two

components, the total plant costs multiplied by a fixed charge factor, and operating costs. The fixed charge factor is calculated in the footnotes of Table 2.1.

Plant derating estimates of amine and ammonia-based CO<sub>2</sub> capture systems from previous studies are shown in Table 2.3, and as it can be seen from the table, most plant derating estimates for ammonia systems are lower than for amine systems.

**Table 2.3: Comparison of plant derating estimates for amine and ammonia-based CO<sub>2</sub> capture systems**

Study Type	Plant Derating	Authors, Publication, Affiliation & Year	Notes
<b>Amine</b>			
Performance and Cost Estimate	28%	Buchanan et al., EPRI 2000	Calculated from data
Performance and Cost Estimate	30%	Woods et al., DOE/NETL 2007a	Calculated from data
Vendor Estimate	23%	Kishimoto et al., MHI 2009	Estimated from data
Expert Elicitation	25-28%	Chung et al., 2009	Power plant retrofit in 2030
<b>Ammonia</b>			
Performance and Cost Estimate	17%	Ciferno et al., NETL 2005	Calculated in study
Performance and Cost Estimate	10-14%	Gal, EPRI 2006	Calculated from data
Vendor Estimate	9%	Peltier, Powermag 2008	Calculated from data from Alstom
Performance Estimate	13%	Valenti et al. 2008	Calculated from data
Performance Estimate	28%	Mathias et al., Fluor 2009	Estimated from data
Vendor Estimate	17%	McLarnon et al., Powerspan 2008	Calculated from data
Vendor Estimate	20%	Hilton et al., Alstom 2009	Estimated by Alstom
Expert Elicitation	17-20%	Chung et al. 2009	Power Plant Retrofit in 2030

Early investigations into the use of ammonia-based post-combustion CO<sub>2</sub> capture thus indicated significant improvements in performance over traditional amine technologies, with consequent expected benefits for lower cost (Bai and Yeh, 1997; Ciferno et al., 2005; Gal, 2006). However, early plant derating estimates of ammonia-based CCS (about half that of a conventional amine system) are lower than later estimates because initial studies assumed that an ammonia-based capture system could be designed around a low energy reaction between ammonium carbonate ((NH<sub>4</sub>)<sub>2</sub>CO<sub>3</sub>) and ammonium bicarbonate (NH<sub>4</sub>HCO<sub>3</sub>). In contrast, further studies indicated that due to unwanted side reactions and problems with high levels of ammonia becoming entrained in the flue gas exiting the absorber, many of the advantages of ammonia-based capture over amines

would be offset by higher auxiliary loads and increased equipment costs elsewhere in the process (Mathias et al., 2009). Problems with slow rates of reaction within the absorber have further called into question the overall potential economic benefits of ammonia-based CO<sub>2</sub> capture due to the potential for large absorber sizes (Derks et al., 2009; Qin et al., 2010).

Unfortunately, while authors have often written about the economic implications of different benefits and drawbacks of CO<sub>2</sub> capture systems, detailed calculations describing these economics are largely absent from the literature and this holds true in particular for ammonia-based CO<sub>2</sub> capture. What is available in terms of economic calculations is not detailed and avoids modeling side reactions in the chemistry (e.g. Gal, 2006; Ciferno et al., 2005). As well, for most of the plant derating estimates supplied in Table 2.3, detailed and transparent information leading to the numbers behind these estimates are also unavailable.

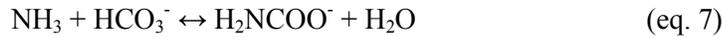
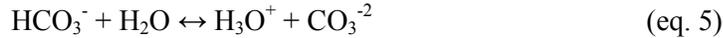
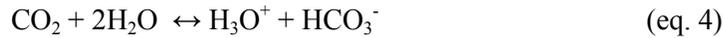
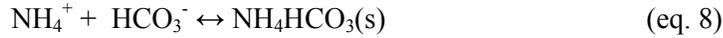
What is needed to fill this gap is a detailed systematic analysis of the potential performance and costs of CO<sub>2</sub> capture using ammonia from a systems perspective, as well as an assessment of how uncertainties in key performance and cost variables affect overall system costs. The intent of this chapter is to fill that gap. The following sections describe: (1) a process model developed in Aspen Plus® V7.2 and used to estimate the performance of a CO<sub>2</sub> capture system using an ammoniated solution applied to the flue gas stream of a coal-fired power plant. (2) Initial performance and cost results of this model for low and high solvent flow rates and ammonia concentrations. (3) More detailed analysis in an attempt to optimize performance and cost results where the effects of different NH<sub>3</sub>/CO<sub>2</sub> ratios, solids concentrations, absorber temperature and water wash conditions are investigated. In several stages of the analysis uncertainty is included to quantify the effects on system cost of uncertainty or variability in key system parameters.

## 2.6. Aspen Plus® for Process Modeling

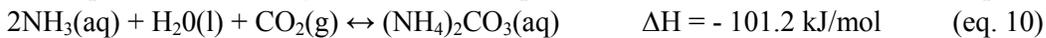
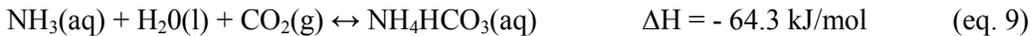
To simulate CO<sub>2</sub> capture using ammonia, Aspen Plus® was chosen. Aspen Plus® contains thermodynamic packages which can describe equilibrium behavior under a wide range of conditions. It is used in modeling thermodynamics within established process equipment, it is familiar in both industry and academia, and it has been successful in simulating processes including CO<sub>2</sub> capture technologies.

### 2.6.1. The Ammonia-Based CO<sub>2</sub> Capture Performance Model

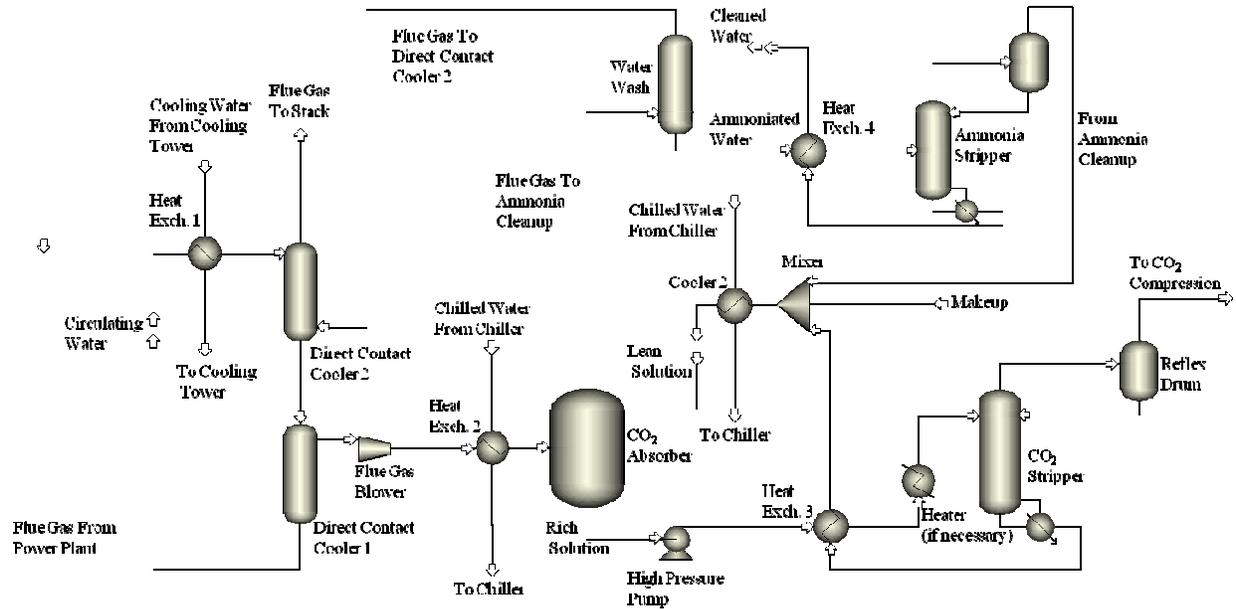
Ammonia-based CO<sub>2</sub> capture was simulated in the Aspen Plus® V7.2 framework using an electrolyte model intended for the study of CO<sub>2</sub> capture by ammonia under equilibrium conditions (Aspentech, 2010). Aspen Plus® V7.2 is used to represent the thermodynamics of the system using the electrolyte non-random two liquid property package. To represent non-ideal behavior in the NH<sub>3</sub>-CO<sub>2</sub>-H<sub>2</sub>O system the model uses the Redlich-Kwong equation of state for the vapor phase (a modification of the ideal gas law) and the electrolyte non-random two liquid activity coefficient model for the liquid phase. In the liquid phase, CO<sub>2</sub> in solution exists as dissolved molecular CO<sub>2</sub> as well as bicarbonate (HCO<sub>3</sub><sup>-</sup>), carbonate (CO<sub>3</sub><sup>2-</sup>), and carbamate (H<sub>2</sub>NCOO<sup>-</sup>) ions while ammonia in solution exists as dissolved molecular NH<sub>3</sub>, ammonium (NH<sub>4</sub><sup>+</sup>) and carbamate (H<sub>2</sub>NCOO<sup>-</sup>) ions. The activity coefficient model represents the NH<sub>3</sub>-CO<sub>2</sub>-H<sub>2</sub>O system as a series of ionic reactions describing the interaction between these species as shown in equations 3-7, as well as the reaction describing the precipitation of ammonium bicarbonate as shown in equation 8. When precipitation of solids occurs, it can increase the CO<sub>2</sub> carrying capacity of the solution. These reactions occur and compete simultaneously, with the concentration of any species depending strongly on the concentrations of the other components as well as on state variables.

**Ionic Reactions****Salt Precipitation Reaction**

In the literature the absorption of  $\text{CO}_2$  is also represented by a series of molecular reactions as shown by equations 9 to 13 (PCP, 2011; Qin et al., 2011), with ammonium bicarbonate ( $\text{NH}_4\text{HCO}_3$ ), ammonium carbonate ( $(\text{NH}_4)_2\text{CO}_3$ ), and ammonium carbamate ( $\text{NH}_2\text{COONH}_4$ ) being formed. Salt precipitation lowers the concentration of aqueous ammonium bicarbonate allowing additional  $\text{CO}_2$  to be absorbed, and increasing the carrying capacity of the solution. In theory, multiple ammonia-based  $\text{CO}_2$  capture systems can be designed by taking advantage of different aspects of the chemistry.

**Vapor-Liquid Reactions****Salt Precipitation Reaction**

The basic process steps and conditions for ammonia-based CCS are taken from a patent application (Gal, 2008), a presentation by Alstom (Hilton et al., 2009), and past modeling efforts (Mathias et al., 2009), and the major features are modeled in Aspen Plus® (V7.2) as shown in Figure 2.2.



**Figure 2.2: The ammonia-based CO<sub>2</sub> capture performance model as built in Aspen Plus<sup>®</sup> V7.2**

The capture of CO<sub>2</sub> is done at low temperatures (273-293°K or 0-20°C) because this favors the reaction between aqueous ammonium carbonate and ammonium bicarbonate (eq. 11) and reduces the vapor pressure of ammonia above the absorber, effectively lowering the ammonia slip. Overall, low temperature absorption is accomplished by cooling and then chilling the flue gas, chilling the lean solution before it enters the absorber, and by removing the exothermic heat of reaction released within the absorber from the capture of CO<sub>2</sub> and the precipitation of ammonium bicarbonate.

Many of the specific details of the Alstom process (temperatures, pressures, and flows) currently are proprietary. In this analysis, the author has made every attempt to provide reasonable values for these numbers based on the ranges provided by Alstom, data provided by other models, and values in the literature. In the base design for the ammonia system, the flue gas from the power plant at 330°K (57°C) enters a direct contact cooler and its temperature is lowered with cooling water to 301°K (28°C). In this step, most of the water vapor in the gas condenses. The gas then

passes through a blower to compensate for the pressure drop through the CO<sub>2</sub> capture system wherein compression increases the gas temperature to 323°K (50°C). The gas temperature is then again lowered to 279°K (6°C) as it passes through a chilled water heat exchanger before entering the absorber. The lean solution from the CO<sub>2</sub> regenerator passes through a cross flow heat exchanger and then is further cooled to 281°K (8°C) in another chilled water heat exchanger before it enters the absorber. Within the absorber the lean solution contacts the flue gas and selectively captures CO<sub>2</sub>. The resulting rich solution may contain significant quantities of precipitated solids. Heat released within the absorber is removed by chilling the absorber internally. That solution leaves at the bottom of the absorber while the decarbonized flue gas leaves at the top.

While not explicitly described in the public literature by Alstom, it is assumed that streams chilled to temperatures below 283°K (10°C) require electrical power of 0.55 kW/ton refrigeration, while streams chilled to between 283°K (10°C) and 302°K (29°C) require 0.47 kW/ton refrigeration (DOE/NETL, 2004; Platts, 2004). It is also assumed that water from the cooling tower at 294°K (21°C) provides cooling for streams or components cooled to 302°K (29°C) or above, such as the CO<sub>2</sub> regenerator condenser.

The rich solution exiting the absorber passes through a high pressure pump which increases the solution pressure to 3.0 MPa. The solution then goes through a heat exchanger with a cold-inlet hot-outlet temperature approach of 5.5°C. If solids in the rich solution are not dissolved entirely in the heat exchanger, a separate heater is used to dissolve these solids before the rich solution enters the CO<sub>2</sub> stripper. In the high pressure (2.8MPa) CO<sub>2</sub> stripper the lean solution is regenerated and returned to the absorber while the concentrated CO<sub>2</sub> stream goes to further compression.

As mentioned previously, a significant amount of ammonia may enter the vapor phase and exit the absorber with the flue gas. This ammonia slip is removed in a water wash system. Some ammonia also may be captured in the circulating water used by the direct contact coolers, but this analysis assumes that the water wash is the primary method for reducing ammonia slip. An ammonia stripper is used to clean the washing water, which is then recycled for contacting with the flue gas. The water wash stripper distillate, including ammonia, carbon dioxide and water, is returned to the capture process. Once the absorber gasses are cleaned of ammonia they are heated in a second direct contact cooler before being released through the stack. Several components were modeled separately in Excel<sup>®</sup>, including the CO<sub>2</sub> compressor equipment and the water chillers that supply cooling loads.

A summary of the important process variables for the power plant used in this analysis and the ammonia-based CO<sub>2</sub> capture system are shown in Table 2.4. Where specific values of process variables were unavailable in the open literature, values were selected within the ranges specified by the CAP patent or were estimated by the author. The characteristics of the power plant used in this part of the study are derived from the DOE/NETL Bituminous Baseline report (DOE/NETL, 2007a). It feeds flue gas into the CO<sub>2</sub> capture system and is similar to the supercritical baseline power plant already introduced in this thesis (Case 2), except it larger so as to account for the substantial steam and electricity loads of the capture system.

**Table 2.4: Key variables for the power plant and ammonia-based CO<sub>2</sub> capture system**

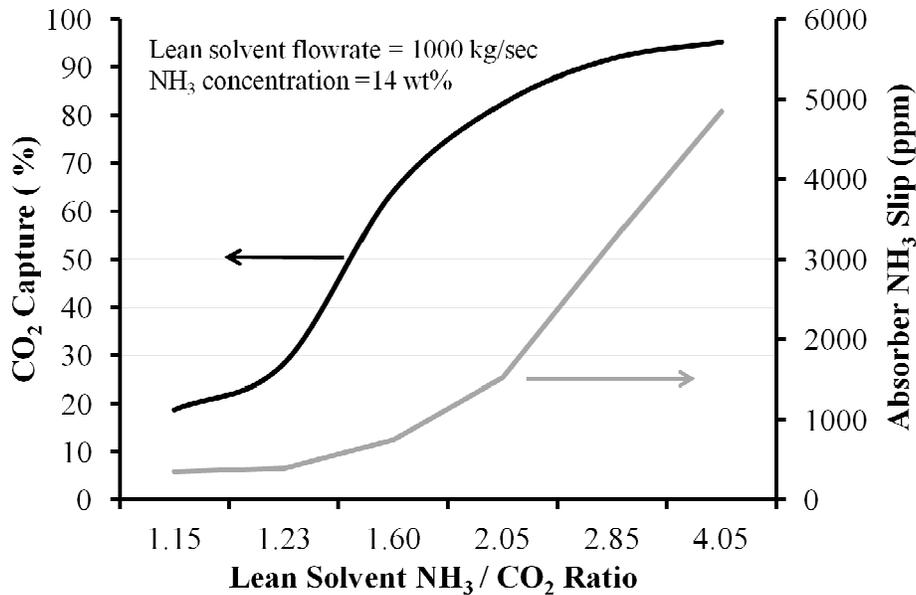
Parameter	Units	Value
<b>Base Power Plant<sup>1</sup></b>		
Coal Flow Rate, Illinois No.6	kg/hr	266,089
Coal Higher Heating Value (As Received)	kJ/kg	27,113
Coal Cost	2007\$/GJ	1.7
Supercritical Steam Cycle	MPa/°C/°C	24.1/593/593
Gross Plant Power Output with Amine-Based CO <sub>2</sub> Capture	MWe	663.4
SCR NO <sub>x</sub> Removal Efficiency	%	86
Wet FGD System SO <sub>2</sub> Removal Efficiency	%	98
Flue Gas Flow Rate into CO <sub>2</sub> Capture System	kg/hr	3,099,560
Flue Gas CO <sub>2</sub> Mole Fraction into CO <sub>2</sub> Capture System	%	13.3
Flue Gas Temperature into CO <sub>2</sub> Capture System	°K	330.4
Flue Gas Pressure into CO <sub>2</sub> Capture System	kPa	104.8
Flue Gas SO <sub>2</sub> into CO <sub>2</sub> Capture System <sup>2</sup>	ppm	38
Flue Gas NO <sub>x</sub> into CO <sub>2</sub> Capture System <sup>2</sup>	ppm	42
<b>Ammonia-Based CO<sub>2</sub> Capture System</b>		
Water Flow Rate into Direct Contact Cooler 2	kg/sec	1452
HeatX1 Cooling Water Flow Rate	kg/sec @ 299.8°K	5000
HeatX2 Chilling Water Flow Rate	kg/sec @ 275.9°K	1000
HeatX3 Temperature Approach	°C	5.5
Lean Solvent Flow Rate	kg/sec	1000
Absorber Temperature	°K	283.1
Lean Solvent NH <sub>3</sub> Wt%	%	0-30
Lean Solvent NH <sub>3</sub> /CO <sub>2</sub> Ratio	mol/mol	1.5-4.0
Allowable Ammonia Slip after Water Wash	ppm	10

<sup>1</sup>The power plant parameters are based the supercritical coal-fired power plant in the 2007 NETL Bituminous Baseline Report (DOE/NETL, 2007a). <sup>2</sup>This analysis assumes that the SO<sub>2</sub> and NO<sub>x</sub> entering the CO<sub>2</sub> capture system is removed by Direct Contact Cooler 1 and therefore has a negligible impact on the CO<sub>2</sub> capture process.

### 2.6.2. Initial Comparison for Absorber Solvent Flow Rates

Using the model, an initial comparison was done to assess the impacts of low and high lean solvent flow rates on the absorption of CO<sub>2</sub>. Therefore, in this initial comparison the primary focus of the model is on the CO<sub>2</sub> absorber. The chilled flue gasses feed into a CO<sub>2</sub> absorber and are contacted with a lean solvent mixture which contains ammonia, carbon dioxide, and water. After a series of preliminary trial runs, the lean solvent flow rate and NH<sub>3</sub> concentration were given initial values of 1000 kg/sec and 14 wt% respectively as these values allowed a wide range of CO<sub>2</sub> capture for several different ratios of NH<sub>3</sub>/CO<sub>2</sub> in the lean solvent mixture. Results for CO<sub>2</sub> capture and subsequent absorber ammonia slip for these initial values for a range of NH<sub>3</sub>/CO<sub>2</sub> ratios is shown in Figure 2.3. As this figure shows, higher ratios result in increased

ammonia slip over the absorber and therefore increased flue gas cleaning demands, while lower ratios required additional solvent flow or higher ammonia concentrations to capture 90% of the CO<sub>2</sub>. The ratio of 2.85 was chosen as a compromise between these two issues for the initial comparisons that follow.



**Figure 2.3: CO<sub>2</sub> capture and absorber NH<sub>3</sub> slip for several lean solvent NH<sub>3</sub>/CO<sub>2</sub> ratios**

The ammonia concentrations and solvent flow rates in the CO<sub>2</sub> capture process affect performance and equipment cost. The impact of varied ammonia concentrations over a consistent design is absent in the literature and so this section presents a sensitivity analysis of the effect of ammonia concentration and solvent flow rate on CO<sub>2</sub> absorption. The results for percent CO<sub>2</sub> capture in the absorber, absorber ammonia slip, and solids content out of the bottom of the absorber are shown in Figure 2.4, Figure 2.5, and Figure 2.6 respectively.

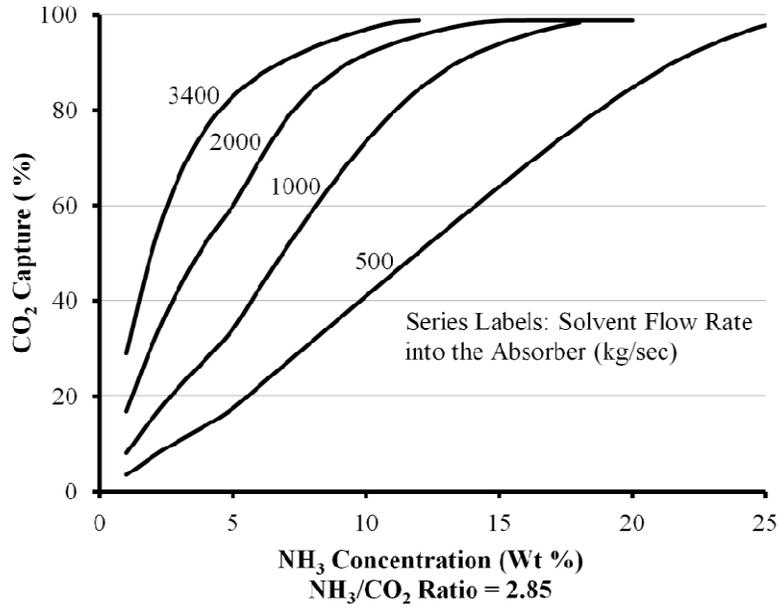


Figure 2.4:  $\text{CO}_2$  capture vs.  $\text{NH}_3$  concentration

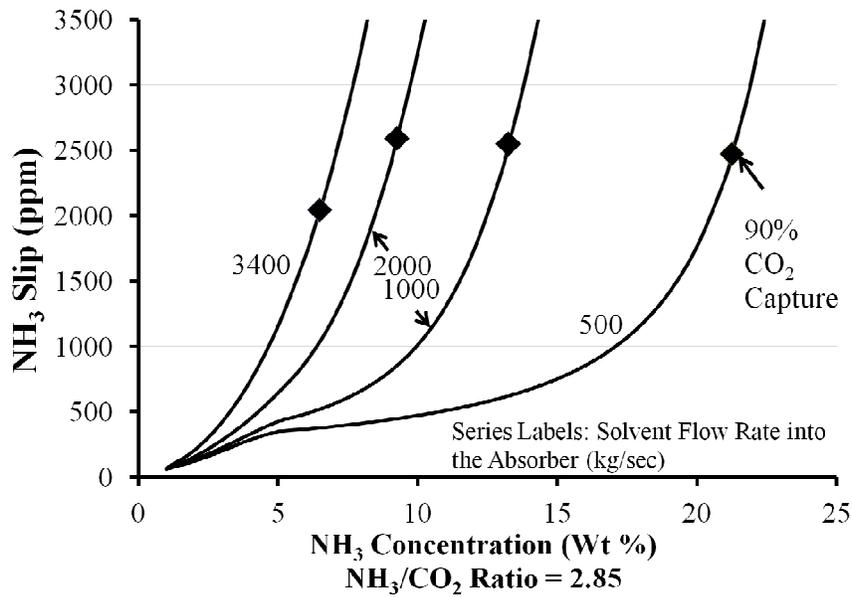
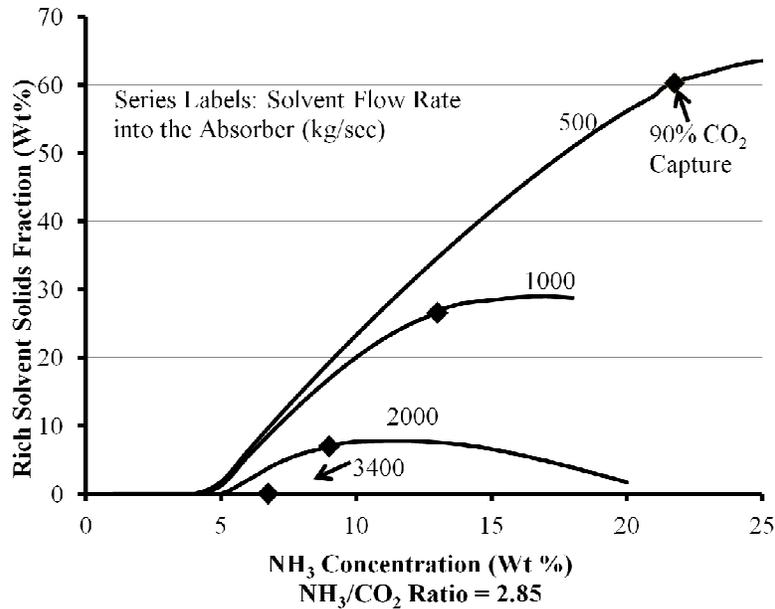


Figure 2.5: Absorber  $\text{NH}_3$  slip vs.  $\text{NH}_3$  concentration. Diamonds represent 90%  $\text{CO}_2$  capture.



**Figure 2.6: Lean solvent NH<sub>3</sub> concentration vs. solids flow rate. Diamonds represent 90% CO<sub>2</sub> capture.**

As shown in the above figures, increases in CO<sub>2</sub> capture results in higher ammonia slip, while the use of low flow rates or high ammonia concentrations cause ammonia salts to precipitate in large quantities.

### 2.6.1. Plant Performance Results for the LowNH<sub>3</sub> and the HighNH<sub>3</sub> Systems

From the results of the sensitivity analysis of the absorber two CO<sub>2</sub> capture system designs were first considered, a low concentration ammonia system operating without solids (LowNH<sub>3</sub>) and a high concentration system operating with a rich solvent absorber outlet stream of 60 wt% solids (HighNH<sub>3</sub>). Additional lean solvent ammonia would result in salt precipitation for LowNH<sub>3</sub>, while HighNH<sub>3</sub> has a rich solvent solids content equivalent to the highest solids wt % for ammonia-based CCS found in the literature (Mathias et al., 2009). The designs for both systems include flue gas cooling, absorber cooling, and ammonia cleanup of the flue gas and the full model shown in Figure 2.2 was used. For both LowNH<sub>3</sub> and HighNH<sub>3</sub>, the key design variables, process conditions, and predicted flows are shown in Table 2.5. The minimum temperature

approach for the heat exchangers in both systems is 5.5°C, with the exception of Heat Exch. 3 for LowNH<sub>3</sub>, which has a temperature approach of 20°C to control for costs for this piece of equipment.

**Table 2.5: Key process conditions and flows for the LowNH<sub>3</sub> and HighNH<sub>3</sub> system designs**

Parameter	LowNH <sub>3</sub>	HighNH <sub>3</sub>
Flue Gas Flow Rate into the System (kg/sec)	860	860
Flue Gas CO <sub>2</sub> Mole Fraction	13.3	13.3
Flue Gas Circulating Water Flow Rate (kg/sec)	1452	1452
Flue Gas Water Wash Cleaning Water Flow Rate (kg/sec)	32	36
Flue Gas HeatX 2 Chilling Load (10 <sup>3</sup> tons cooling/ hr @ 3°C) <sup>1</sup>	10	10
CO <sub>2</sub> Absorber Solvent Flow Rate (kg/sec)	3,400	500
CO <sub>2</sub> Absorber Lean Solvent NH <sub>3</sub> Concentration (wt%)	6.75%	21.5%
CO <sub>2</sub> Absorber Chilling Load (10 <sup>3</sup> tons cooling/ hr @ 3°C) <sup>1</sup>	70	107
CO <sub>2</sub> Absorber CO <sub>2</sub> Removal Efficiency (%)	90%	90%
CO <sub>2</sub> Absorber Ammonia Slip (ppm)	2112	2488
CO <sub>2</sub> Absorber Rich Stream Solids Content (wt%)	No Solids Occur	60%
Lean Solvent Chilling Load (10 <sup>3</sup> tons cooling/ hr @ 3°C) <sup>1</sup>	88	5
Overall Ammonia Slip (ppm)	<1	<1
Overall CO <sub>2</sub> Capture (%)	90%	90%
Overall CO <sub>2</sub> Product Purity (vol%)	99.8%	99.8%

<sup>1</sup>Ton cooling (or ton refrigeration) is a standard unit of energy used for large refrigeration units and is equal to 12,000 Btu.

The power usage of both LowNH<sub>3</sub> and HighNH<sub>3</sub>, including the electrical equivalent in steam drawn off after the intermediate turbine, is shown in Table 2.6 along with the resulting estimated performance characteristics of the power plant. The performance estimates are based primarily on performance data from Aspen Plus<sup>®</sup> as well as data scaled from the Integrated Environmental Control Model (IECM V6.2). The plant derating of the HighNH<sub>3</sub> system was found to be 28.5% while the plant derating of the LowNH<sub>3</sub> system was found to be 39.4%. This suggests that the LowNH<sub>3</sub> system likely will not be competitive since it has a higher plant derating and requires significantly larger equipment. HighNH<sub>3</sub> performs relatively well due to the energy benefits associated with higher CO<sub>2</sub> loading and reduced heating, cooling, and transportation energy requirements.

**Table 2.6: Power plant performance estimates. All values are in MWe equivalent.**

	No CO <sub>2</sub> Capture (Case 2)	LowNH <sub>3</sub>	HighNH <sub>3</sub>	Notes and Primary Data Sources for Calculation
Potential Power Available	581.3	827.6	827.6	Based on coal flow rate
Auxiliary Steam Load				
Heater			14.8	Aspen Plus
CO <sub>2</sub> Stripper		163.0	87.0	Aspen Plus
NH <sub>3</sub> Stripper		3.5	3.6	Aspen Plus
Steam Turbine Power	581.3	661.1	722.2	Based on aux. steam load
Auxiliary Electrical Load				
Flue Gas Blower		18.9	18.9	ΔP=3 psi, scaled IECM data
Gas Cooling Water Pumps		3.2	3.2	Aspen Plus, scaled IECM data
Chiller for Heat Exch 2		5.7	5.7	Aspen Plus, RDC 2003, Platts 2004
Chiller for Absorber Cooling		38.4	58.9	Aspen Plus, RDC 2003, Platts 2004
Chiller for Solvent Cooling		48.5	3.0	Aspen Plus, RDC 2003, Platts 2004
Absorber Cooling Pumps		3.5	5.4	Aspen Plus, scaled IECM data
Solvent Circulation Pumps		1.5	0.2	Aspen Plus, scaled IECM data
Econamine FG Plus System				
CO <sub>2</sub> Compression		17.0	17.0	Aspen Plus, scaled IECM data, DOE/NETL 2007a
Balance of Plant	31.3	49.0	49.0	Scaled IECM data
Plant Net Power	550.0	475.2	560.8	
Plant Efficiency (% HHV)	39.1%	23.7%	28.0%	
Plant Derating of CO <sub>2</sub> Capture (%)		39.4%	28.5%	eq. 1

### 2.6.1. Plant Economic Results for the LowNH<sub>3</sub> and the HighNH<sub>3</sub> Systems

Preliminary cost results for the two ammonia-based CO<sub>2</sub> capture system designs in 2007 constant dollars are shown in Table 2.7. The cost estimates are based on data from the IECM, sources from the literature, and Aspen Icarus®. The costs for a number of components are scaled according to methodologies described elsewhere (Rao, 2002). The plants in Table 2.7 are assumed to have a levelized capacity factor of 75% over their lifetimes as assumed in the baseline assumptions of Table 2.1. The capital costs for both the HighNH<sub>3</sub> and the LowNH<sub>3</sub> systems are significantly higher than for the plant without a CO<sub>2</sub> capture system. Revenue required estimates for the plants with HighNH<sub>3</sub> and LowNH<sub>3</sub> systems are \$U.S.103MWh and \$U.S.133/MWh respectively. The plant with the HighNH<sub>3</sub> system has a higher efficiency, which

leads to a significant cost advantage compared to the plant with the LowNH<sub>3</sub> system. In comparison with LowNH<sub>3</sub>, HighNH<sub>3</sub> also benefits from smaller equipment sizes associated with higher loading and reduced heating, cooling, and transportation requirements. This analysis suggests that ammonia-based CO<sub>2</sub> capture systems with higher concentrations that allow solids precipitation will be more cost effective than if solids precipitation was not allowed to occur, and therefore systems with a low ammonia-concentration will not be considered further.

The absorber in the HighNH<sub>3</sub> system is required to handle significant amounts of solids and was therefore considered a spray tower with capital costs similar to that of a wet flue gas desulfurization system. This cost estimate may be optimistic because modeling is based on equilibrium assumptions which may not apply in all cases. Given that a very large absorber would be required for a close approach to equilibrium, and that previous investigations on the kinetics of ammonia based CO<sub>2</sub> capture have shown that absorption may be slower than for amine-based CO<sub>2</sub> capture (Qin et al., 2010), the absorber for this system may be significantly more expensive than the estimate provided here. This issue is considered further in the uncertainty analysis near the end of this chapter.

**Table 2.7: Power plant cost estimates with ammonia-based CO<sub>2</sub> capture, values are in 2007**

	<b>\$Millions</b>			
	No CO <sub>2</sub> Capture (Case 2)	LowNH <sub>3</sub>	HighNH <sub>3</sub>	Notes and Primary Data Sources for Calculation
<b>CO<sub>2</sub> Capture Process Area Costs</b>				
DCC #1		30.9	30.9	Aspen Plus, scaled IECM data, Rao 2002
DCC #2		23.3	23.3	Aspen Plus, scaled IECM data, Rao 2002
Flue Gas Blower		6.4	6.4	Aspen Plus, scaled IECM data, Rao 2002
Heat Exch. 1		6.7	6.7	Aspen Icarus
Heat Exch. 2		2.9	2.9	Aspen Icarus
Heat Exch. 1 Pumps		1.4	1.4	Aspen Plus, scaled IECM data, Rao 2002
Heat Exch. 2 Pumps		0.5	0.5	Aspen Plus, scaled IECM data, Rao 2002
Cooling Water Circ Pumps		0.7	0.7	Aspen Plus, scaled IECM data, Rao 2002
Chiller System		74.3	54.3	Aspen Plus, Platts 2004
Absorber		74.4	105.1	Aspen Plus, scaled IECM data
Absorber Pumps		1.9	2.5	Aspen Plus, scaled IECM data, Rao 2002
Heat Exch. 3		74.7	19.2	Aspen Icarus
Solvent Circulation Pumps		16.5	5.2	Aspen Plus, scaled IECM data, Rao 2002
Solvent Heater		0.0	2.5	Aspen Icarus
Solvent Cooler		37.5	2.3	Aspen Icarus
CO <sub>2</sub> Stripper		66.5	21.0	Aspen Plus, scaled IECM data, Rao 2002
CO <sub>2</sub> Stripper Reboiler		33.2	7.2	Aspen Plus, scaled IECM data, Rao 2002
Water Wash		2.2	2.2	Aspen Icarus
Heat Exch. 4		0.1	0.1	Aspen Icarus
NH <sub>3</sub> Stripper		1.5	1.5	Aspen Icarus
NH <sub>3</sub> Cleanup Pumps		1.0	1.1	Aspen Plus, scaled IECM data, Rao 2002
Steam Extractor		3.3	3.3	Scaled IECM data
Sorbent Processing		1.1	1.1	Scaled IECM data
Drying and Compress Unit		18.3	18.3	Aspen Plus, scaled IECM data
General Facilities Capital		7.5	5.0	1.57 % PFC, DOE/NETL 2007a
Eng. & Home Office Fees		45.0	30.1	9.37 % PFC, DOE/NETL 2007a
Project Contingency Cost		78.5	52.6	16.38 % PFC, DOE/NETL 2007a
Process Contingency Cost		22.4	15.0	4.67 % PFC, DOE/NETL 2007a
CO <sub>2</sub> System (TCR)		633.2	424.0	Based on Area Costs
Base Plant (TCR) <sup>1</sup>	670.8	865.4	884.1	Scaled IECM data
Cooling Tower (TCR)	35.8	62.7	62.7	Scaled IECM data
NO <sub>x</sub> Control (TCR)	25.0	33.7	33.7	Scaled IECM data
TSP Control (TCR)	37.4	49.8	49.8	Scaled IECM data
SO <sub>2</sub> Control (TCR)	112.1	138.7	138.7	Scaled IECM data
CO <sub>2</sub> System and TS&M		21.6	21.6	Scaled IECM data
O&M/Year				
Balance of Plant O&M/Year	103.1	128.9	128.9	Scaled IECM data
Plant Total Capital Requirement	881.1	1783.5	1593.0	Based on TCR Costs
Total O&M Costs/Year	103.1	150.6	150.6	Total O&M
Capital Required (\$/kW-net)	1601.0	3753.3	2840.5	Based on Performance
Revenue Required (\$/MWh)	53.7	132.9	103.0	eq. 2

<sup>1</sup>The base plant cost is reduced for the low concentration ammonia-based CO<sub>2</sub> capture system design because a smaller steam turbine is required.

## **2.7. Sensitivity Analysis in Other Variables in the Chilled Ammonia Process**

Having established high concentration ammonia-based CO<sub>2</sub> capture processes as likely to be less expensive than low concentration ammonia-based CO<sub>2</sub> capture, the preliminary analysis for the HighNH<sub>3</sub> system from the previous section was used as a basis for exploring the effect of other critical variables on the performance and costs of the Chilled Ammonia Process. These are outlined in the next sections.

### **2.7.1. Lean Solvent NH<sub>3</sub> Concentration and NH<sub>3</sub>/CO<sub>2</sub> Sensitivity**

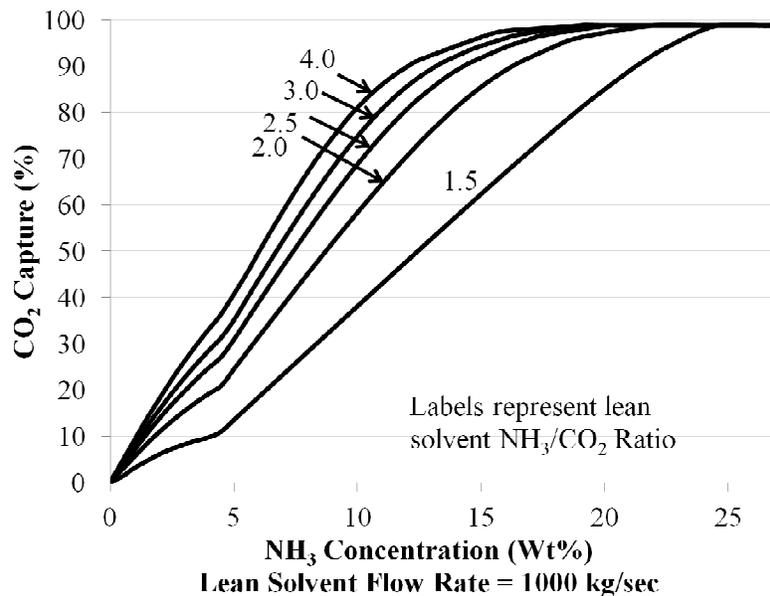
To reduce energy demand, the process attempts to take advantage of eq. 11, the low energy reaction between ammonium carbonate and ammonium bicarbonate. Ideally if eq. 11 were the only reaction pathway, the lean solution would consist primarily of ammonium carbonate and water, with an NH<sub>3</sub>/CO<sub>2</sub> ratio near 2.0. However several researchers have noted that other reactions occur leading to the formation of unwanted species (Mathias et al., 2009), and under many circumstances where the NH<sub>3</sub>/CO<sub>2</sub> ratio is 2.0, ammonium carbonate may not be even the primary constituent in the lean solution (Qin et al., 2011). This leads to questions regarding the best lean solvent NH<sub>3</sub>/CO<sub>2</sub> ratio for the process. The patent cited above for the CAP (Gal, 2008) recommends using a range between 1.0 and 4.0. In this section, the impact of variations in lean solution NH<sub>3</sub>/CO<sub>2</sub> ratio and NH<sub>3</sub> concentration on absorber CO<sub>2</sub> capture, NH<sub>3</sub> slip, and solids formation is investigated over this range.

As noted in the previous section, while lean solutions with high ammonia concentrations and relatively low solvent flow rates had the potential for a high CO<sub>2</sub> capture efficiency, one consequence was high solids precipitation in the rich solvent, as shown in Figure 2.6. At a pilot plant operated by We Energies and Alstom in Pleasant Prairie, Wisconsin, it was found that solids formation for chilled ammonia posed an operational challenge due to blockages in process

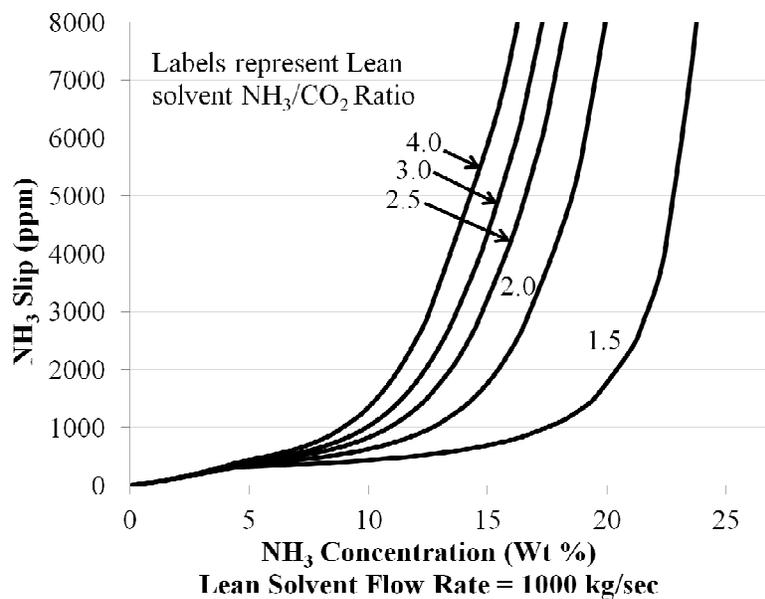
equipment (Bollinger et al., 2010). Due to this concern over operating with high solids in the rich solution, the present analysis employs a baseline lean solvent flow rate of 1000 kg/sec to reduce solids formation in the rich solution to approximately 30 to 40 wt% in the baseline design (vs. 60 wt% at the lower sorbent flow of 500 kg/sec rate in the previous section). This is similar to wet limestone flue gas desulfurization systems which can operate with an absorber slurry density of 15 to 20 wt% suspended solids and 35 to 45 wt% suspended solids after the first dewatering stage (Kohl and Nielsen, 1997; Stultz and Kitto, 2005).

The absorber temperature was held constant at 283.1°K (10°C) and the NH<sub>3</sub> concentration and NH<sub>3</sub>/CO<sub>2</sub> ratio of the lean solution were varied parametrically to explore system behavior under these conditions. The resulting CO<sub>2</sub> capture efficiency, NH<sub>3</sub> slip and solids precipitation are shown in Figure 2.7, Figure 2.8, and Figure 2.9.

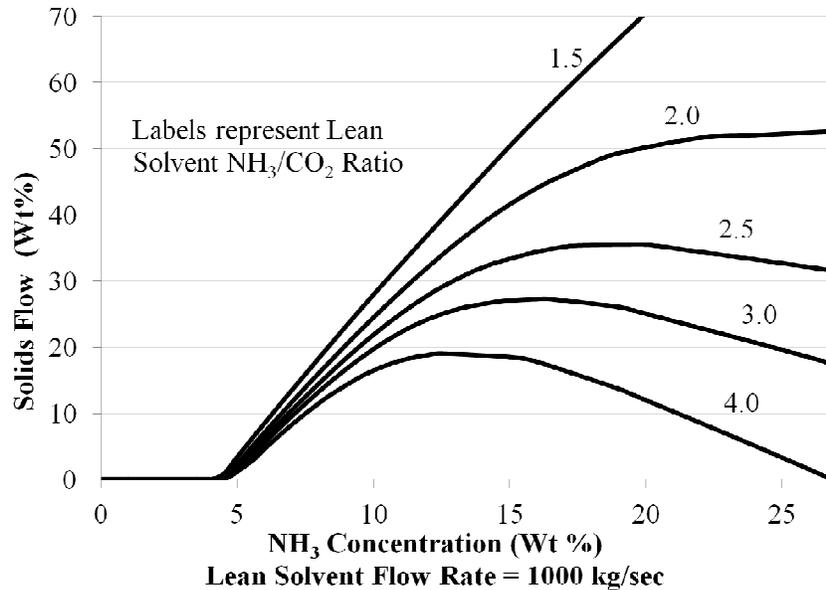
These results indicate that both the ammonia concentration and the NH<sub>3</sub>/CO<sub>2</sub> ratio significantly affect CO<sub>2</sub> capture, ammonia slip, and solids precipitation in the absorber. Increases in either the NH<sub>3</sub> concentration or the NH<sub>3</sub>/CO<sub>2</sub> ratio increase the percent CO<sub>2</sub> captured as well as the NH<sub>3</sub> slip, while increases in NH<sub>3</sub> concentration and decreases in NH<sub>3</sub>/CO<sub>2</sub> ratio increase the wt% solids. Figure 2.7 further indicates that a decrease in the NH<sub>3</sub>/CO<sub>2</sub> ratio below 2.0 results in a significant reduction in the fraction of CO<sub>2</sub> that can be captured as well as a large increase in the amount of solids precipitated. NH<sub>3</sub>/CO<sub>2</sub> ratios above 3.0 result in lower levels of solids precipitation but high levels of ammonia slip and only marginal improvements in CO<sub>2</sub> capture efficiency. Based on these results, an NH<sub>3</sub>/CO<sub>2</sub> ratio of 2.5 was chosen for use in the remainder of this analysis as this provided a balance between reduced solids handling requirements—to aid in absorber and process design at higher levels of CO<sub>2</sub> capture— and reduced ammonia slip.



**Figure 2.7:**  $\text{CO}_2$  capture as a function of lean sorbent  $\text{NH}_3$  concentration and  $\text{NH}_3/\text{CO}_2$  ratio. The labels in the figures represent the lean solvent  $\text{NH}_3/\text{CO}_2$  ratio.



**Figure 2.8:**  $\text{NH}_3$  slip (ppm) as a function of lean sorbent  $\text{NH}_3$  concentration and  $\text{NH}_3/\text{CO}_2$  ratio. The labels in the figures represent the lean solvent  $\text{NH}_3/\text{CO}_2$  ratio.

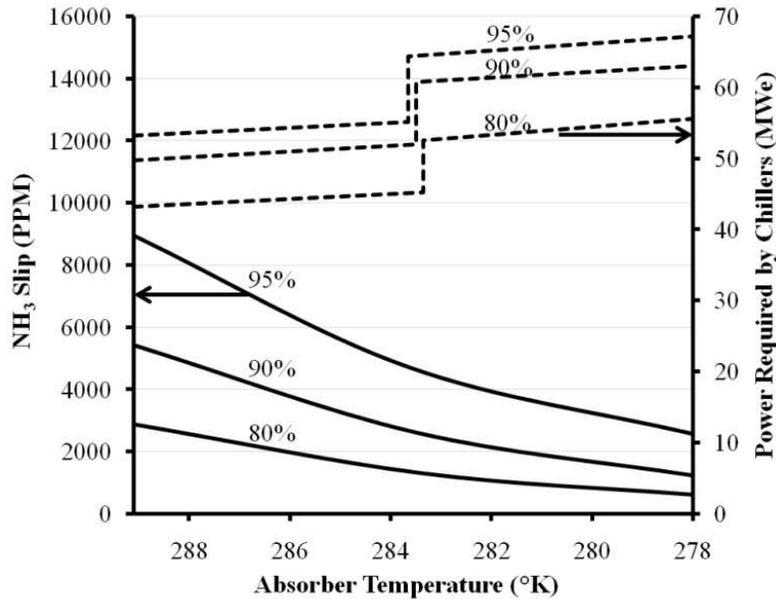


**Figure 2.9: Wt% solids in rich sorbent exiting the absorber as a function of lean sorbent NH<sub>3</sub> concentration and NH<sub>3</sub>/CO<sub>2</sub> ratio. The labels in the figures represent the lean solvent NH<sub>3</sub>/CO<sub>2</sub> ratio.**

### 2.7.2. Effects of Absorber Temperature

Lowering the temperature of the absorption process can help control ammonia slip, but the trade-off is that lower temperatures increase the chilling energy requirements of the process and are expected to have a negative effect on the reaction kinetics, increasing the absorber size. In this section, the lean solvent flow rate and NH<sub>3</sub>/CO<sub>2</sub> ratio were held constant at 1000 kg/sec and 2.5, respectively, while the temperature of the absorber was varied between 277.6 and 288.8°K (4.5 and 15.6°C). The resulting ammonia slip and absorber cooling power requirements are shown in Figure 2.10 for several different levels of CO<sub>2</sub> capture, which is adjusted by changing the NH<sub>3</sub> concentration in the lean solvent. The temperature of the flue gas and the lean solvent entering the absorber were held constant by Heat Exchanger 2 and Cooler 2, respectively, shown in Figure 2.2. The power required by the chillers to cool the absorber assumes 0.55 kW/per ton of refrigeration for chilling below 283°K (10°C), and 0.47 kW/ton refrigeration for chilling above 283°K (DOE, 2004; Platts, 2004). The chiller power requirement increases with increasing

percentages of CO<sub>2</sub> captured (as more exothermic heat of reaction is removed) and also with reductions in the absorber temperature. Ammonia slip also increases for higher fractions of CO<sub>2</sub> capture, but decreases at lower absorber temperatures. Most of the absorber cooling energy is for removal of the exothermic heat released by the capture of CO<sub>2</sub> and the precipitation of solids.

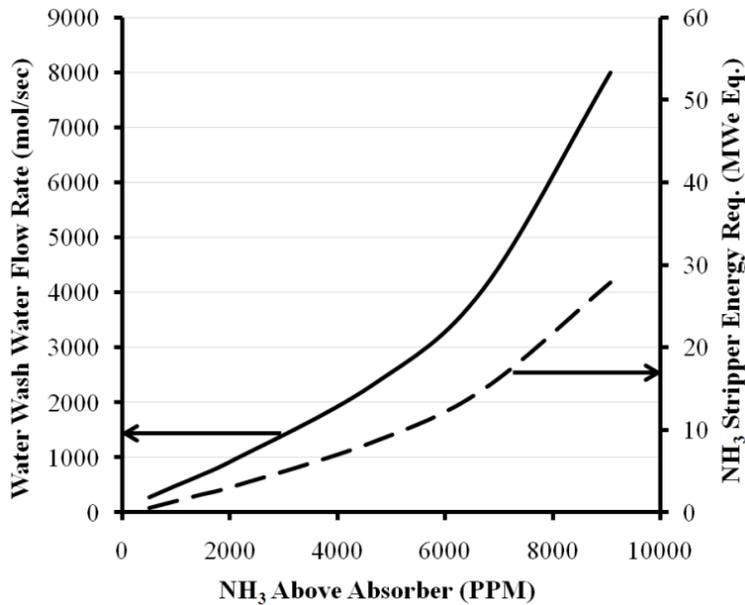


**Figure 2.10: Ammonia slip and absorber cooling requirements as a function of absorber temperature for three levels of CO<sub>2</sub> capture (80, 90 and 95%). The labels represent the % CO<sub>2</sub> captured. Reducing the temperature below 283°K results in a step change in the unit energy required for cooling.**

### 2.7.3. Water Wash Requirements for Ammonia Removal

The water wash above the absorber removes ammonia from the flue gas by contacting the gas stream with water in a column. The cleaned flue gas exits the top of the column with a low ammonia concentration while the resulting sour water is sent through a heat exchanger to a sour water stripper. Vapor from the sour water stripper containing NH<sub>3</sub>, CO<sub>2</sub>, and H<sub>2</sub>O is returned to the CO<sub>2</sub> capture process. Figure 2.11 shows the water required to reduce the NH<sub>3</sub> concentration in the flue gas stream to 10 parts per million (ppm). The corresponding steam energy requirement also is shown as electrical energy equivalent, assuming the steam enthalpy is 3276

kJ/kg, the condensate enthalpy is 749 kJ/kg, and the heat-to-electricity conversion efficiency used to penalize the power plant for the loss of low-pressure steam use is 0.22 (IECM V6.2). In this analysis the lean solvent flow rate, the  $\text{NH}_3/\text{CO}_2$  ratio, and the absorber temperature were held constant at 1000 kg/sec, 2.5, and 283.1°K (10°C) respectively, and the ammonia slip exiting the absorber was varied by adjusting the lean solvent wt%  $\text{NH}_3$ . As shown in Figure 2.11, both the wash water requirement and the  $\text{NH}_3$  stripper energy requirement increase as the ammonia level entrained in the flue gas stream increases.



**Figure 2.11: Wash water flow rate and associated  $\text{NH}_3$  stripper energy required to reduce the ammonia slip in the flue gas stream to 10 ppm, for various levels of ammonia slip exiting the absorber.**

## 2.8. Refined Plant Performance Results for Ammonia-Based $\text{CO}_2$ Capture

Estimates of the performance and cost of the coal-fired power plant with ammonia-based  $\text{CO}_2$  capture are shown in Table 2.8 and Table 2.9. These results are based on performance data from the Aspen Plus® model of the ammonia capture system and associated cost data from Aspen

Icarus<sup>®</sup>, together with scaled data from the literature and from the Integrated Environmental Control Model power plant simulator (IECM V6.2).

For this baseline estimate, the CO<sub>2</sub> capture efficiency was 90%, the NH<sub>3</sub> concentration was 14.4 wt % at a lean solvent flow rate of 1000 kg/sec, the NH<sub>3</sub>/CO<sub>2</sub> ratio was 2.5, the absorber temperature was 283.1°K (10°C), ammonia slip after the water wash was limited to 10 ppm, and the solids content in the rich solution was 33 wt%. Note that a more stringent emissions limit for NH<sub>3</sub> slip (e.g. 2 ppm, as found in some SCR facilities) would further increase the cost of the ammonia-based system.

Where not available directly from the models, equipment power consumption was scaled linearly from reference loads and process flow conditions, and equipment capital costs were scaled from reference costs and process flows and conditions. This scaling methodology for specific equipment has been described previously (Rao, 2002). All costs are reported in constant 2007 dollars, and were scaled using the Marshall & Swift Equipment Cost Index (“Marshall & Swift”, 2009). The Revenue Required (or levelized cost of electricity, LCOE) also is calculated in real terms excluding the effects of inflation, and including transport, storage, and maintenance costs of \$3.75/tonne CO<sub>2</sub> (DOE/NETL, 2007a).

**Table 2.8: Power plant performance estimates. All values are in MWe equivalent**

	No CO <sub>2</sub> Capture (Case 2)	Ammonia Capture System	Notes and Primary Data Sources for Ammonia System
Potential Power Available	581.3	827.6	Based on coal flow rate
Auxiliary Steam Load <sup>1</sup>			
Heater		5.1	Aspen Plus
CO <sub>2</sub> Stripper		103.4	Aspen Plus
NH <sub>3</sub> Stripper		3.5	Aspen Plus
Steam Turbine Power	581.3	715.6	Based on aux. steam load
Auxiliary Electrical Load			
Flue Gas Blower		18.9	ΔP=3 psi, scaled IECM data
Heat Exchanger 1 Pumps		2.2	Aspen Plus, scaled IECM data
Heat Exchanger 2 Pumps		0.4	Aspen Plus, scaled IECM data
Gas Cooling Water Pumps		0.6	Aspen Plus, scaled IECM data
Chiller for Heat Exch 2		5.7	Aspen Plus, RDC 2003, Platts 2004
Chiller for Absorber Cooling		48.2	Aspen Plus, RDC 2003, Platts 2004
Chiller for Solvent Cooling		6.0	Aspen Plus, RDC 2003, Platts 2004
Absorber Cooling Pumps		5.1	Aspen Plus, scaled IECM data
Solvent Circulation Pumps		3.5	Aspen Plus, scaled IECM data
Econamine FG Plus System			
CO <sub>2</sub> Compression		16.9	Aspen Plus, scaled IECM data, DOE/NETL 2007a
Balance of Plant	31.3	49.0	Scaled IECM data
Plant Net Power	550.0	558.7	
Plant Efficiency (% HHV)	39.1%	27.9%	
Plant Derating of CO <sub>2</sub> Capture (%)		28.6%	

<sup>1</sup>The auxiliary steam load is shown as electrical energy equivalent, assuming the steam has an enthalpy of 3276 kJ/kg and a pressure of 902 kPa, the water condensate has an enthalpy of 749 kJ/kg, and the heat-to-electricity conversion efficiency used to penalize the power plant for the loss of low-pressure steam use is 0.22 (IECM, V6.2).

**Table 2.9: Power plant cost estimates. All values are in 2007 \$Millions**

	No CO <sub>2</sub> Capture (Case 2) <sup>1</sup>	Ammonia Capture System <sup>1</sup>	Notes and Primary Data Sources
<b>CO<sub>2</sub> Capture Process Area Costs</b>			
DCC #1		30.9	Aspen Plus®, scaled IECM data, Rao, 2002
DCC #2		23.3	Aspen Plus®, scaled IECM data, Rao, 2002
Flue Gas Blower		6.3	Aspen Plus®, scaled IECM data, Rao, 2002
Heat Exch. 1		6.7	Aspen Icarus®
Heat Exch. 2		2.9	Aspen Icarus®
Heat Exch. 1 Pumps		1.4	Aspen Plus®, scaled IECM data, Rao 2002
Heat Exch. 2 Pumps		0.5	Aspen Plus®, scaled IECM data, Rao 2002
Cooling Water Circ Pumps		0.7	Aspen Plus®, scaled IECM data, Rao 2002
Chiller System		54.6	Aspen Plus®, DOE, 2004; Platts, 2004
Absorber		105.1	Aspen Plus®, scaled IECM data <sup>2</sup>
Absorber Pumps		2.4	Aspen Plus®, scaled IECM data, Rao 2002
Heat Exch. 3		41.6	Scaled Aspen Icarus® data
Solvent Circulation Pumps		7.9	Aspen Plus®, scaled IECM data, Rao 2002
Solvent Heater		2.2	Aspen Icarus®
Solvent Cooler		2.2	Aspen Icarus®
CO <sub>2</sub> Stripper		35.1	Aspen Plus®, scaled IECM data, Rao 2002
CO <sub>2</sub> Stripper Reboiler		13.4	Aspen Plus®, scaled IECM data, Rao 2002
Water Wash		2.2	Aspen Icarus®
Heat Exch. 4		0.1	Aspen Icarus®
NH <sub>3</sub> Stripper		1.5	Aspen Icarus®
NH <sub>3</sub> Cleanup Pumps		0.8	Aspen Plus®, scaled IECM data, Rao 2002
Steam Extractor		3.3	Scaled IECM data
Sorbent Reclaimer		1.1	Scaled IECM data
Sorbent Processing		1.1	Scaled IECM data
Drying and Compress Unit		18.6	Aspen Plus®, scaled IECM data
CO <sub>2</sub> System (PFC)		365.9	
General Facilities Capital		5.7	1.57 % PFC from Woods et al. 2007
Eng. & Home Office Fees		34.3	9.37 % PFC from Woods et al. 2007
Project Contingency Cost		59.9	16.38 % PFC from Woods et al. 2007
Process Contingency Cost		17.1	4.67 % PFC from Woods et al. 2007
CO <sub>2</sub> System (TCR)		483.0	Based on Area Costs
Base Plant (TCR)	670.8	884.1	Scaled IECM data
Cooling Tower (TCR)	35.8	62.7	Scaled IECM data
NO <sub>x</sub> Control (TCR)	25.0	33.7	Scaled IECM data
TSP Control (TCR)	37.4	49.8	Scaled IECM data
SO <sub>2</sub> Control (TCR)	112.1	138.7	Scaled IECM data
CO <sub>2</sub> System and TS&M O&M/Year <sup>3</sup>		22.3	Scaled IECM data
Balance of Plant O&M/Year	103.1	128.9	Scaled IECM data
Plant Total Capital Requirement	881.1	1652.0	Based on TCR Costs
Total O&M Costs/Year	103.1	151.3	Total O&M
Capital Required (\$/kW-net)	1601.0	2956.8	Based on Performance
Revenue Required (\$/MWh)	53.7	105.4	
CO <sub>2</sub> Avoidance Cost (\$/tonne avoided)		73.2	

<sup>1</sup>These plants were modeled in the IECM with a 75% capacity factor and a fixed charge factor of 0.143 for the plant with CCS (higher risk) and 0.113 for the plant without CCS (lower risk) as in Rubin and Zhai, 2011, and as outlined in the baseline assumptions of Table 2.1. A risk premium for CCS also is assumed by DOE/NETL in recent cost studies. <sup>2</sup>The reference IECM cost was a wet FGD scrubber. <sup>3</sup>O&M costs for the plant with CCS includes (1) CO<sub>2</sub> transport, storage, and maintenance costs of \$3.75/tonne CO<sub>2</sub>, (2) fixed costs of \$8.0M/year, and (3) variable costs related to solvent losses. In the ammonia system, flue gas

contaminants act as nucleation sites for the condensation of water in DCC1 and are mostly removed (Gal, 2008) resulting in few operational problems (Hilton, 2009). Absorber slip losses at 10 ppm are approximately 0.02 kg/tonne CO<sub>2</sub> at a unit cost of \$135/tonne 28% aqueous NH<sub>3</sub> (\$483/tonne NH<sub>3</sub>) (DOE/NETL, 2007a), which equals to \$0.01/tonne CO<sub>2</sub> or \$0.04M/year. If all the SO<sub>2</sub> entering the capture system formed heat stable salts and an activated carbon bed (\$0.6M/year) was required, NH<sub>3</sub> losses would be approximately 0.13 kg/tonne CO<sub>2</sub> or \$0.3M/year, raising the LCOE only slightly to \$105.7/MWh.

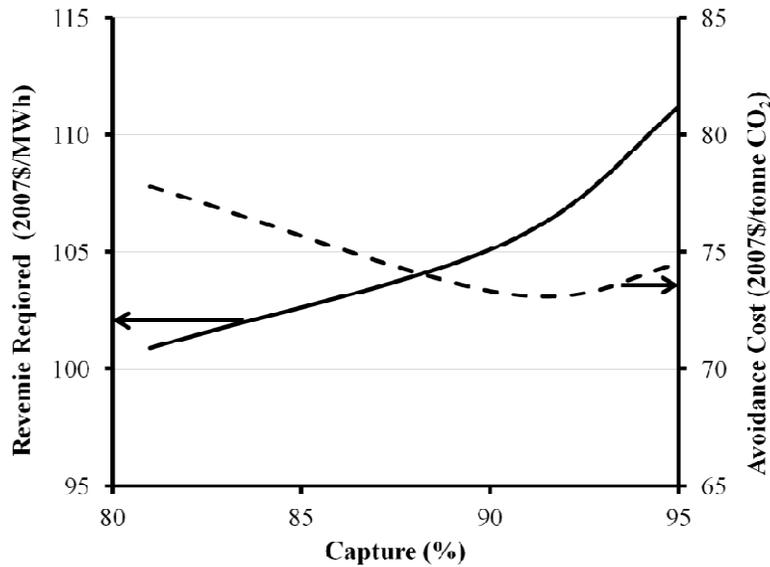
As shown in Table 2.9, the revenue required for the ammonia-based CO<sub>2</sub> capture system is \$105/MWh. The ammonia-based CO<sub>2</sub> system benefits from low steam loads and reduced compressor power requirements, but the chilling loads and associated costs offset these benefits.

By varying the lean solvent NH<sub>3</sub> concentration for the same process conditions and cost estimating methodology, the revenue requirement and CO<sub>2</sub> avoidance cost were calculated as a function of the CO<sub>2</sub> capture efficiency. The avoidance cost is calculated as shown in eq. 14.

Avoided Cost

$$= \frac{[\text{Revenue Required with CCS} - \text{Revenue Required without CCS}] \{ \$/\text{MWh} \}}{[\text{CO}_2 \text{ Emissions without CCS} - \text{CO}_2 \text{ Emissions with CCS}] \{ \text{tons}/\text{MWh} \}} \quad (\text{eq. 14})$$

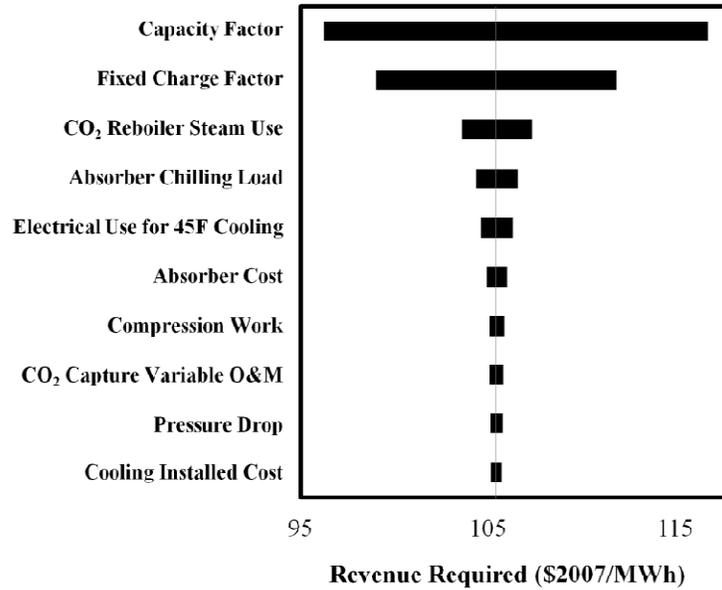
The results are shown in Figure 2.12. At capture efficiencies above approximately 90% the revenue required rises at an increasing rate due to the increasing steam demands for CO<sub>2</sub> regeneration and NH<sub>3</sub> cleanup, and the increased chilling requirements of the CO<sub>2</sub> capture system. There is a minimum in the CO<sub>2</sub> avoidance cost between approximately 90% and 94% capture. Lower levels of CO<sub>2</sub> capture lead to higher avoidance costs due to the high capital requirements of the process, while avoidance costs rise above 95% CO<sub>2</sub> capture due to the rapidly increasing costs of NH<sub>3</sub> cleanup.



**Figure 2.12: Revenue Required and CO<sub>2</sub> avoidance cost as a function of CO<sub>2</sub> capture efficiency for the baseline PC power plant. All costs in constant 2007 US dollars.**

### **2.9. Uncertainty Analysis on Key CO<sub>2</sub> Capture System Performance Parameters**

In this section we explore the effects of uncertainty on the results presented thus far using two methods of analysis. Figure 2.13 first shows the effect on the revenue requirement of a uniform  $\pm 10\%$  change from the baseline value of several key variables for the plant with ammonia-based capture. The tornado diagram shows that plant utilization and financing assumptions dominate the revenue requirement calculation for equal changes in all variables. However, significant revenue requirement changes are also driven by the CO<sub>2</sub> stripper energy requirements, chilling requirements and absorber capital cost.



**Figure 2.13: A tornado graph indicating the change in the revenue required for a +/- 10% change in the input of ten important variables.**

In practice, of course, some parameters are more uncertain or variable than others. Thus, a probabilistic analysis was undertaken to more realistically characterize the impact on total plant cost of uncertainties or variability in key system parameters relative to the results presented above for the plant with ammonia capture. The parameter distribution functions for this analysis are shown in Table 2.10. Uncertainty distributions were compiled following the methodology outlined in Frey and Rubin (1991), with distributions inferred either from the literature or estimated by the author. Uncertain variables were grouped into three categories: capture system performance parameters, capture system cost parameters, and plant financing and utilization parameters.

**Table 2.10: Nominal values and uncertainty parameters assessed in the ammonia-based CO<sub>2</sub> capture system**

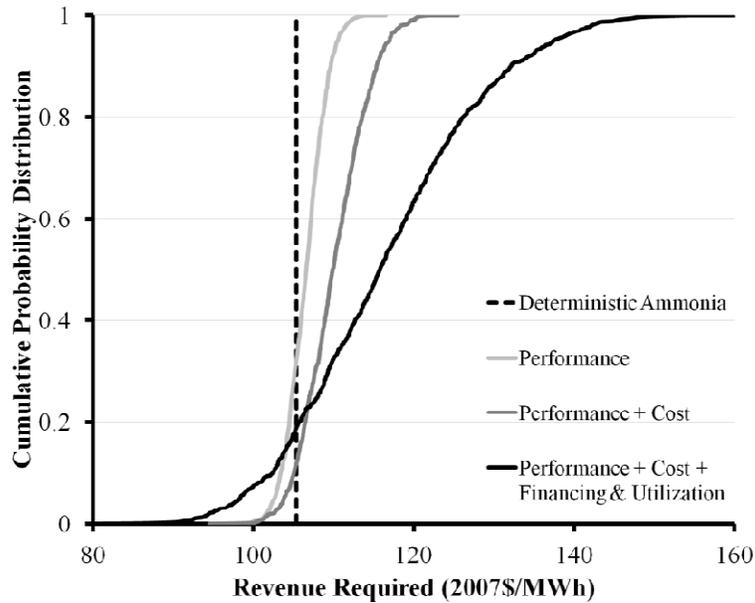
Parameter	Units	Nominal (x)	Values (or $\sigma$ as % of x)	References
<b>Capture System Performance</b>				
Chilling Loads Required at <283°K	Tons Refrigeration	103,000	Normal(x, 10%)	Author Estimate
Chilling Loads Required at 283°K to 302°K	Tons Refrigeration	21,333	Normal(x, 10%)	Author Estimate
Chiller Electrical Use, 276°K Water Product	kW/Ton Refrigeration	0.55	Triangular(0.47, 0.55, 0.60)	Platts 2004
Chiller Electrical Use, 280°K Water Product	kW/Ton Refrigeration	0.47	Triangular(0.47, 0.47, 0.55)	Platts 2004
CO <sub>2</sub> Regeneration Heat Requirement	kJ/kg CO <sub>2</sub>	2293	Normal(x, 10%)	Author Estimate
Pumping Head	kPa	207	Triangular(150,207,250)	Rao 2002
Pump Efficiency	%	75	Uniform(70,75)	Rao 2002
$\Delta$ P Across CO <sub>2</sub> Capture System	kPa	20.7	Triangular(14,26,30)	Rao 2002
Blower Efficiency	%	75	Uniform(70,75)	Rao 2002
CO <sub>2</sub> Compression, 27.5 bar to 152.7 bar	kWh/kg CO <sub>2</sub>	0.03	Triangular(0.028, 0.03, 0.032)	
<b>Capture System Cost</b>				
Reference Chilling Equipment Costs (PFC)	\$2007/Ton Refrigeration	441	Uniform(0.7x-1.3x)	Author Estimate
Reference IECM Costs (PFC)	\$2007	251.9	Uniform(0.7x-1.3x)	Author Estimate
Reference Aspen Icarus® Costs (PFC)	\$2007	59.4	Uniform(0.7x-1.4x)	Author Estimate
CO <sub>2</sub> Absorber Costs (PFC)	\$2007	105.1	Uniform(0.7x-2.5x)	Zhuang 2011
General Facilities Capital	% of PFC	1.57 <sup>a</sup>	Normal(x, 10%)	Berkenpas et al. 1999
Eng. & Home Office Fees	% of PFC	9.37 <sup>a</sup>	Triangular(0.7x, 1x, 1.5x)	Berkenpas et al. 1999
Project Contingency Cost	% of PFC	16.38 <sup>a</sup>	Normal(x, 20%)	Berkenpas et al. 1999
Process Contingency Cost	% of PFC	4.67 <sup>a</sup>	Normal(x, 30%)	Berkenpas et al. 1999
CO <sub>2</sub> System Fixed O&M/Year	\$2007 Million/year	8.0	Uniform(0.7x-1.3x)	Author Estimate
CO <sub>2</sub> System Variable O&M/Year + TS&M	\$2007 Million/year	14.0	Uniform(0.7x-1.3x)	Author Estimate
<b>Plant Financing &amp; Utilization</b>				
Power Plant Fixed Charge Factor	fraction	0.143	Uniform(0.130, 0.180)	Rubin and Zhai, 2011
Power Plant Levelized Capacity Factor	--	0.75	Uniform(0.65, 0.85)	Rubin and Zhai, 2011

One important source of uncertainty is the reaction rate for CO<sub>2</sub> capture in the absorber, which is dependent on species concentration and temperature. Details of these reaction rates are not yet well understood and for this reason the Aspen Plus® model used here assumes (for simplicity and transparency) the bounding case of equilibrium conditions. Experimental work, however,

suggests that rates for ammonia based-CO<sub>2</sub> capture could be 3 to 10 times slower than for MEA (Darde et al., 2011, Qin et al., 2010). Under these conditions, absorber sizing estimates indicate that the absorber for an ammonia-based process would be 2 to 3 three times larger, and thus more costly, than for an amine-based process (Chang, 2009; Zhuang, 2011). This cost uncertainty stemming from uncertain reaction rates is reflected in the distribution of the CO<sub>2</sub> absorber cost in Table 2.10. The skewed distribution reflects a likelihood of much higher cost relative to the baseline case of an equilibrium reactor, and is an important source of uncertainty in the process.

The results of the uncertainty analysis are shown in Figure 2.14. Using only the performance parameter uncertainty distributions, the analysis indicates that the probability that the revenue required will be equal to or lower than the deterministic case is about 30%. However, when the uncertainties in cost parameters are also included, the cumulative distribution function shifts predominately to the right (i.e., higher costs), primarily due to the effects of higher CO<sub>2</sub> absorber cost related to uncertain reaction rates, as discussed earlier. This distribution function shows only a 10% chance that the revenue required will be equal to or less than the deterministic case. This result highlights the importance of the absorber cost and suggests that further research into how reaction rates affects absorber size for specific absorber designs is needed. Finally, when uncertainties in plant financing and utilization parameters are also included, the probability that the revenue required will be less than or equal to the deterministic value rises to 20%. The cumulative probability distribution also widens significantly. Costs now range from \$80/MWh to \$160/MWh, with the 95% confidence interval ranging from \$95/MWh to \$143/MWh. This broad range indicates the importance of financial and plant utilization assumptions on the overall cost of the plant, as seen earlier in Figure 2.13. With all uncertainties included the ammonia-based system has a median (50% probability) revenue required of \$115.8/MWh. The average revenue

required is only slightly higher at \$116.5/MWh, reflecting the asymmetric distributions for several parameters. These values are approximately 10% higher than the deterministic revenue required reported earlier.



**Figure 2.14: Cumulative probability distribution of the levelized cost of electricity of the baseline PC plant with ammonia-based CO<sub>2</sub> capture**

The uncertainty distributions in Figure 2.14 also affect the CO<sub>2</sub> avoidance cost. In contrast to the deterministic estimate of \$73/tonne CO<sub>2</sub> for the ammonia-based system with 90% removal, the median revenue required in Figure 2.14 corresponds to an avoidance cost of \$88/tonne CO<sub>2</sub> while the 95% confidence interval for revenue required corresponds to avoidance costs of \$60 to \$127/tonne CO<sub>2</sub> avoided. This represents the likely range of carbon price or tax required in a market-based regime to make ammonia-based CCS more economical than a similar PC plant with no CCS.

## 2.10. Discussion

The uncertainty analysis presented above explores possible departures from the baseline performance and cost assumptions and produces a broader range of potential cost estimates for

the ammonia-based CO<sub>2</sub> capture system assessed in this chapter. Those costs are generally higher than the deterministic estimate for the baseline system. However, since ammonia-based CO<sub>2</sub> capture is not as mature as other technologies such as amine-based CO<sub>2</sub> capture there may be considerable room for improvement in the technology (Chung et al., 2009). For example, though slow reaction rates in the absorber have the potential to significantly increase system costs, additives are currently being investigated to speed up reaction rates (Chang, 2009, Lee, 2007). And although the analysis presented in this section limited the NH<sub>3</sub> concentration and lean solvent flow rate to avoid aggressive solids formation, a lower solvent flow rate coupled with higher NH<sub>3</sub> concentration could lead to reduced energy requirements and lower cost if the high solids formation can be managed in a way that does not compromise system reliability. There is some precedent for this as large-scale solids handling has been successfully accomplished in wet limestone-based flue gas desulfurization systems applied to coal-fired power plants (Stultz and Kitto, 2005). Future process improvements may also take advantage of the potential for heat integration of the CO<sub>2</sub> regeneration and NH<sub>3</sub> cleanup systems where significant amounts of heat are rejected by the CO<sub>2</sub> compressors and chillers—an option not investigated in this analysis

## **2.11. Chapter Conclusion**

In this chapter the performance and cost of two ammonia-based post-combustion CO<sub>2</sub> capture systems operating at a new supercritical coal-fired power plant were modeled. This assessment showed that for a fixed coal input, the plant derating of a CO<sub>2</sub> capture system operating with high ammonia concentrations (HighNH<sub>3</sub>) was found to be 28.5%. The plant derating of a CO<sub>2</sub> capture system operating with low ammonia concentrations (LowNH<sub>3</sub>) was substantially higher at 39.4%. Preliminary estimates of the revenue requirement of the plants with HighNH<sub>3</sub> and LowNH<sub>3</sub> systems are \$U.S. 133/MWh and \$U.S. 103/MWh respectively. The results from this

performance assessment and preliminary cost analysis suggest that the LowNH<sub>3</sub> system will not be competitive.

For the HighNH<sub>3</sub> ammonia system the absorber CO<sub>2</sub> capture efficiency, NH<sub>3</sub> slip, and solids precipitation were evaluated for changes in lean solution NH<sub>3</sub> concentration, NH<sub>3</sub>/CO<sub>2</sub> ratio, and absorber temperature. Reductions in NH<sub>3</sub> slip were also assessed for changes in absorber temperature and water wash flow rate. For 90% CO<sub>2</sub> capture the revenue required for the final plant design was estimated at \$US 105/MWh. The revenue required for this system was found to depend strongly on the fraction of CO<sub>2</sub> captured as well as on key process design parameters such as lean solution NH<sub>3</sub> concentration. Uncertainties in system performance and cost also were estimated probabilistically. Assumptions about plant financing and utilization, as well as uncertainties in cooling costs and reaction rates that affect absorber cost were found in particular to produce a wide range of cost estimates for ammonia-based CO<sub>2</sub> capture systems.

The intent of this chapter was to provide reasonable preliminary performance and cost estimates of ammonia-based CO<sub>2</sub> capture system designs. In the future the ammonia-based CO<sub>2</sub> capture performance and cost models could benefit from improved thermodynamic models, more detailed simulations of individual pieces of equipment including in particular the absorber, rate-based as opposed to equilibrium modelling, and cost estimates by vendors.



## **Chapter 3. Developing a Chilled Ammonia Response Surface Model for the IECM**

### **3.1. Motivation for Developing a General Power Plant and CO<sub>2</sub> Capture Model**

Performance and cost estimates for power and industrial plants with CO<sub>2</sub> capture technologies are valuable to policy makers as they inform potential policies regarding CO<sub>2</sub> mitigation. Unfortunately, performance and cost estimates can also be difficult to find or expensive and time consuming to produce, and existing estimates can be challenging to modify if different financial, technological, or environmental conditions are assumed. For example, in 2007 the U.S. National Energy Technology Laboratory estimated that on a new supercritical coal-fired power plant with amine-based CO<sub>2</sub> capture and long term storage, CCS would lower the plant efficiency by 11.9 percentage points and would increase the cost of electricity by \$52/MWh compared to a plant without CCS (DOE/NETL, 2007a). This is a valuable baseline but it is also based on a large number of particular financial, technological and site specific assumptions, and it can be difficult to extend these results to alternate scenarios. The development of accessible analytical tools grounded in scientific principles that can provide flexible estimates, tools such as the Integrated Environmental Control Model (IECM), are important resources for policy makers and can help put the development of other technologies into context.

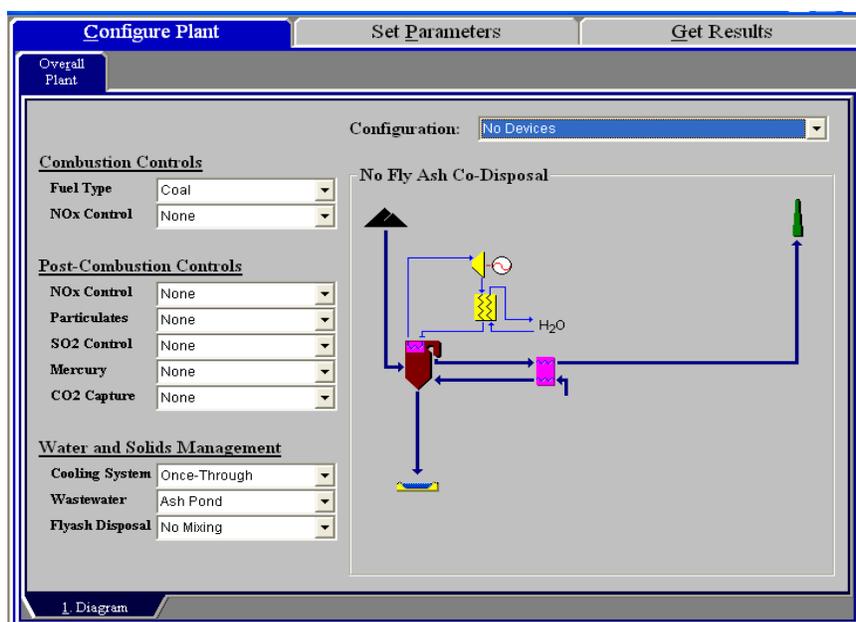
This chapter outlines work done to integrate an ammonia-based post-combustion CO<sub>2</sub> capture module into the IECM. The model created in Aspen Plus<sup>®</sup> V 7.2 and described in Chapter 2 was simulated under a variety of process conditions, and the results were used to develop an ammonia-based CO<sub>2</sub> capture module in the IECM. This allowed for a systematic evaluation of ammonia-based CO<sub>2</sub> capture systems on hypothetical plants with various financial assumptions and technology and fuel configurations. Following the integration of the model into the IECM,

the model was used to do an analysis of plants with ammonia-based CO<sub>2</sub> capture over a range of conditions and on different types of power plants. The updated version of the IECM is intended as a starting point for policy makers, researchers, engineers, and others to do their own preliminary performance and cost analysis for pulverized or natural gas power plants using this leading post-combustion CO<sub>2</sub> capture technology. The model is technical but every effort will be made to keep the model accessible to as wide an audience as possible.

### **3.2. The Integrated Environmental Control Model (IECM)**

The integrated environmental control model is a freely available tool used by researchers, engineers, policy makers, and others for developing preliminary cost estimates for power plants burning coal and natural gas under a wide variety of plant configurations. A user can select from several different types of power plants including coal-fired power plants, natural-gas combined cycle plants, and integrated gasification combined cycle (IGCC) plants, and get preliminary performance and cost information for any of the types.

In the model, a user can configure the overall design of the plant, set key parameters for specific pieces of equipment, and look at the performance and cost results of the power plant. As the user changes the plant configuration to reflect their preferences, a simple schematic diagram of the power plant is shown. This schematic is illustrated for a coal-fired power plant in Figure 3.1.

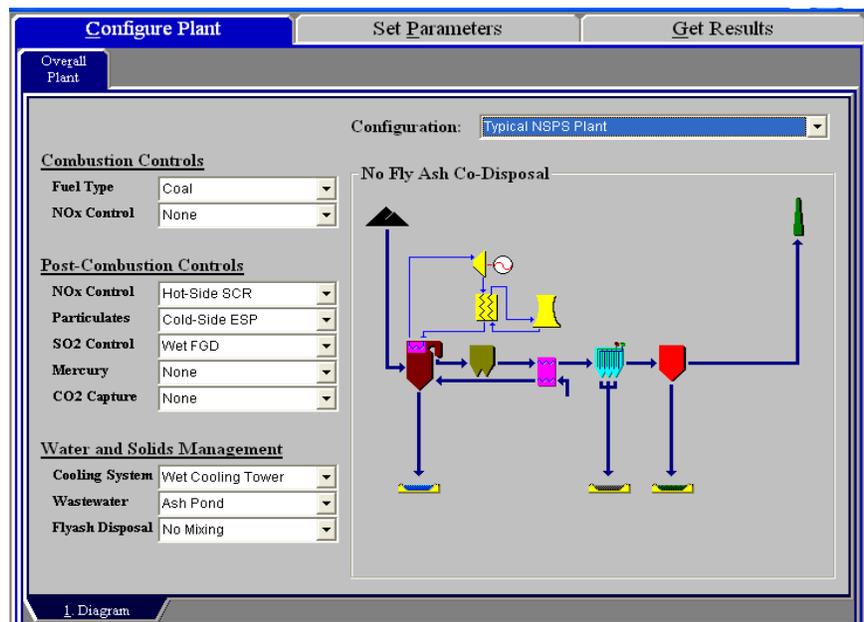


**Figure 3.1: Configure Plant Screen for the IECM for a PC plant without environmental controls.**

As shown in Figure 3.1, coal from a coal pile is fed into a boiler, steam is produced which is used to generate electricity, the bottom ash is sent for disposal, and the flue gases from combustion are sent to the stack and then the atmosphere. This is a plant without any environmental controls and is not representative of a plant that would typically be built in the United States today. Most plants would be built to meet a certain level of environmental emissions constraints, and technologies can be added onto the base plant configuration to meet those constraints.

Those alternate configurations are represented in the model through the options on the left of Figure 3.1. For the pulverized coal power plant, there are combustion controls for controlling  $\text{NO}_x$ , and there are also post-combustion controls for  $\text{NO}_x$ ,  $\text{SO}_2$ , Mercury, and  $\text{CO}_2$  Capture. There are also options for waste and solids management, including managing plant water use, wastewater, and flyash disposal. When any of these alternative options are selected, the schematic for the plant configuration can change. For example, a user can select a wet FGD for  $\text{SO}_2$  removal as well as a  $\text{CO}_2$  capture system for  $\text{CO}_2$  removal, and the plant configuration

changes accordingly. The IECM already has several default configurations that a user might be interested in, and a user can select them from the upper right corner. For the pulverized coal power plant, one can choose a New Source Performance Standard (NSPS) plant, which sets the plant configuration to meet a standard set of current federal regulations for environmental emissions. This plant is shown in Figure 3.2.



**Figure 3.2: Configure Plant Screen for the IECM with environmental controls intended to meet requirements for a New Source Performance Standard**

A user can go into considerable detail in configuring the base plant and the environmental controls that they wish to select, and can also see an extensive variety of results for the plant. The next sections describe the development of a general response surface model of the ammonia-based CO<sub>2</sub> capture system to be incorporated as an environmental control module in the IECM.

### **3.3. Developing a Response Surface Model of the Chilled Ammonia System**

This section describes the methodology for developing the response surface model of the chilled ammonia system in the IECM. A response surface model is a mathematical model that describes and approximates the behavior of a large system under a wide variety of conditions. In the case

of the ammonia-based CO<sub>2</sub> capture system, data from a wide variety of model runs in Aspen Plus® V7.2 was used to develop the response surface model.

### **3.3.1. Fractional Factorial Experimental Design**

In Chapter 2, the performance and cost results for the ammonia system were dependent on a number of factors (cooling water temperature, absorber temperature, NH<sub>3</sub>/CO<sub>2</sub> lean solvent loading ratio, solvent NH<sub>3</sub> concentration, solvent flowrate, percent CO<sub>2</sub> capture etc.). An understanding of the impact of all of these factors is critical in the development of an ammonia-based CO<sub>2</sub> capture module for the IECM, but the impact of these factors has not yet been fully explored in a systematic way. This section will explain how these factors were systematically explored.

Initially, runs for the ammonia-based CO<sub>2</sub> capture model built in Aspen Plus® V7.2 and shown in Figure 2.2 were tried over a range of conditions and for variations in several of the important system variables. Exploring the model space in a full factorial design (varying each of the input variables systematically to evenly cover the model space) proved challenging because for this approach a large number of model runs were required and many of these runs experienced time consuming convergence difficulties in Aspen Plus®.

To cover the full model space in an efficient way, the input variables were changed systematically in a fractional factorial design to lower the model runs needed. A fractional factorial design takes advantage of the fact that most systems are dominated by lower order effects. For example, for the key input variables in Table 3.1, the number of model runs can be reduced from a full factorial design exploring all the variables at both the upper and lower bounds in the range ( $2^9$  runs for coal plants plus  $2^9$  runs for natural gas plants = 1024 total runs)

to a much more compact design by using the generators: A, B, C, D, E, F, G, ABCDEFG, BCDEFG ( $2^7$  runs for coal plants plus  $2^7$  runs for natural gas plants = 256 total runs, see variable labels in Table 3.1). A portion of the structure of this design is shown in Table 3.2. .

**Table 3.1: Varied parameters in the model for coal and natural gas plants**

Label	Parameter	Type	Units	Coal Lower Bound (-)	Coal Upper Bound (+)	Nat. Gas Lower Bound (-)	Nat. Gas Upper Bound (+)
A	Input Flue Gas Temperature	Input	°C	37	66		
B	Input Flue Gas CO <sub>2</sub> Content	Input	mole/sec	2,800	4,200	1,000	1,600
C	Input Flue Gas N <sub>2</sub> Content	Input	mole/sec	16,000	24,000	20,000	26,000
D	Input Flue Gas H <sub>2</sub> O Content	Input	mole/sec	4,000	6,000	2,000	3,000
E	DCC Cooling Water Temperature	Input	°C	18	30	21	32
F	DCC Circulating Water Flow Rate	Input	mole/sec	65,000	95,000	65,000	95,000
G	Absorber Lean in NH <sub>3</sub> Mole Flow <sup>1</sup>	Input	mol/sec	4,000	7,000	2,300	3,000
H	Absorber Lean in H <sub>2</sub> O Mole Flow	Input	mol/sec	13,000	19,000	12,000	15,000
I	Water Wash Water Flow Rate	Input	mol/sec	1,000	10,000	2,000	8,000

<sup>1</sup>In this analysis the Absorber Lean CO<sub>2</sub> Molar Flow Rate (mol/sec) was fixed at 0.4\*the Absorber Lean NH<sub>3</sub> Molar Flow Rate (mol/sec) to keep the NH<sub>3</sub>/CO<sub>2</sub> Ratio at 2.5, according to the analysis described in Chapter 2.

**Table 3.2: Partial structure of the fractional factorial design.**

A	B	C	D	E	F	G	H = ABCDEFG	I = BCDEFG
-	-	-	-	-	-	-	-	+
-	-	-	-	-	-	+	+	-
-	-	-	-	-	+	-	+	-
-	-	-	-	-	+	+	-	+
-	-	-	-	+	-	-	+	-
...	...	...	...	...	...	...	...	...
+	+	+	+	+	+	+	+	+

\*Negative (-) represents the lower bound, positive (+) represents the upper bound of the variable from Table 3.1

The variable ranges in Table 3.1 aim to cover large variations around flue gas flows and concentrations that may be found in typical coal and natural gas power plants. A set of additional model runs were also done in regions of interest in the model space between the upper and lower bounds. This was done specifically for the CO<sub>2</sub> concentrations of flue gasses found at coal and natural gas-fired power plants in the literature (i.e. DOE/NETL, 2007a) and for lean solvent flow rates and ammonia concentrations that led to 90% CO<sub>2</sub> capture in the absorber, to ensure appropriate coverage. These additional model runs were only completed on a section of the

model consisting of the direct contact coolers, absorber, and water wash owing to convergence difficulties related to the loops around the CO<sub>2</sub> and NH<sub>3</sub> strippers.

The combination of exploring the model space using a fractional factorial design as well as targeted runs aimed at capturing the details near regions of particular interest led to broad coverage of the model space. The resulting data for each successful model run was merged as a complete dataset.

Several other model variables were calculated by Aspen Plus® as a consequence of the varied input parameters in the model. For example, the flue gas flow, gas concentration, lean solvent flow rate, and ammonia concentration in the lean solvent together determines the amount of cooling the absorber requires, the high pressure pump power requirements, and the amount of CO<sub>2</sub> that must be regenerated in the CO<sub>2</sub> stripper to complete the material and energy balances in the model. This in turn helps determine chiller power requirements, CO<sub>2</sub> heater stripper steam requirements, and eventually leads to cost estimates.

Multivariate linear regression equations were created based on the dataset and these equations were incorporated as a module in the IECM. The minimum dataset sample size required for any regression was calculated according to a multivariate sample size calculator (Soper, 2010), and each regression met this minimum requirement. For example, if all 9 predictors in Table 3.1 were used at least 113 data points would be required for statistical significance at the 0.05 probability level for an anticipated effect size of 0.15 and a statistical power level of 0.8 according to the statistics calculator.

The response variables for the model are shown in Table 3.3. These response variables from the performance model are required to estimate the performance and costs of the components in the ammonia-based CO<sub>2</sub> capture system.

**Table 3.3: Response variables in the model**

Parameter	Type	Units
Aspen Flue Flow Rate into DCC1	Response	cum/sec
Aspen Flue Flow Rate into DCC2	Response	cum/sec
Circulating Water Flow Rate into HeatX1	Response	kg/sec
Gas Flow Rate Into Absorber	Response	cum/sec
Lean Solvent Flow Rate	Response	kg/sec
Rich Solvent Flow	Response	kg/sec
Absorber Ammonia Slip	Response	Ppm
Water Wash Water Flow Rate	Response	kg/sec
Solids Fraction in the Rich Solution	Response	%
Nitrogen (N <sub>2</sub> ) to Stack	Response	cum/sec
Oxygen (O <sub>2</sub> ) to Stack	Response	cum/sec
Carbon Dioxide (CO <sub>2</sub> ) to Stack	Response	cum/sec
Carbon Monoxide (CO) to Stack	Response	cum/sec
Hydrochloric Acid (HCl) to Stack	Response	cum/sec
Sulfur Dioxide (SO <sub>2</sub> ) to Stack	Response	cum/sec
Sulfuric Acid (equivalent SO <sub>3</sub> ) to Stack	Response	cum/sec
Nitric Oxide (NO) to Stack	Response	cum/sec
Nitrogen Dioxide (NO <sub>2</sub> ) to Stack	Response	cum/sec
Ammonia (NH <sub>3</sub> ) to Stack	Response	cum/sec
Argon (Ar) to Stack	Response	cum/sec
Particulate (Flyash) to Stack	Response	kg/sec
Water Vapor (H <sub>2</sub> O) to Stack	Response	cum/sec
HTX2 Chilling Water Flow	Response	kg/sec
Absorber Chilling Water Flow Rate	Response	kg/sec
Lean Solution Cooler Chilling Water Flow	Response	kg/sec
CO <sub>2</sub> Capture System Cooling Water Required	Response	kg/sec
CO <sub>2</sub> Compressor Cooling Water Flow	Response	kg/sec
Water Bleed	Response	mol/sec
Cooler Chilling Load	Response	Btu/hr
Absorber Chilling Load	Response	Btu/hr
HeatX2 Chilling Load	Response	Btu/hr
Steam Flow to CO <sub>2</sub> Stripper & Heater	Response	Btu/hr
Steam Flow to NH <sub>3</sub> Stripper	Response	Btu/hr
High Pressure Pump Power Usage	Response	MWe
Aspen Heat Exchanger 1 Surface Area	Response	sqm
Aspen Heat Exchanger 2 Surface Area	Response	sqm
Aspen Heat Exchanger 1 Liquid Flow Rate	Response	kg/sec
Aspen Heat Exchanger 2 Liquid Flow Rate	Response	kg/sec
Aspen Heat Exchanger 3 Surface Area	Response	sqm
Solution Cooler Surface Area	Response	sqm
CO <sub>2</sub> to Compressor	Response	mol/sec
Reclaimer Waste	Response	kg/sec
Ammonia Makeup	Response	mol/sec
Water Makeup	Response	mol/sec

### **3.4. The Response Surface Models**

This section introduces the key response equations that comprise the response model used by the ammonia-based CO<sub>2</sub> capture system module in the IECM. The response model is built in stages and the resulting performance and cost estimates were tested against the Aspen Plus® model runs and existing cost calculations. Since in some cases the ability to calculate later equations depends on earlier equations, these calculations are completed sequentially by the IECM. The predictors for each independent variable were chosen using a manual approach. Multiple predictors were initially selected for each independent variable based on the author's knowledge of the process. These predictor variables were chosen because they were likely to impact the independent variable. For example, it was expected by the author that the most important variables impacting the lean solvent flow rate required in the absorber would be the flue gas flow rate and composition, the CO<sub>2</sub> capture required by the IECM user, and the NH<sub>3</sub> concentration of the lean solution. Plots of the dependent versus independent variables were created and each regression was fit with the dependent variables where significant dependencies were seen.

#### **3.4.1. Response Model Calculation Order**

Several of the equations in the overall response model are dependent of one another, and so they are calculated in a particular order that generally follows the process flow of the system. For example, the rich solvent flow is dependent on the lean in solvent flow rate calculated in the regression before it. The order of the overall calculation is outlined below.

- 1) Direct Contact Cooler (DCC) Flow Rates
  - a. Aspen Flue Flow Rate into DCC1
  - b. Aspen Flue Flow Rate into DCC2
  - c. Circulating Water Flow Rate into HeatX1
  
- 2) CO<sub>2</sub> Capture and Ammonia Cleanup Variables
  - a. Gas Flow Rate Into Absorber

- b. Lean Solvent Flow Rate
    - i. For <91% CO<sub>2</sub> Capture
    - ii. For => 91% CO<sub>2</sub> Capture
  - c. Rich Solvent Flow
  - d. Absorber Ammonia Slip
  - e. Water Wash Water Flow Rate
  - f. Solids Fraction in the Rich Solution
- 3) Flue Gas to Stack Variables
- a. Nitrogen (N<sub>2</sub>)
  - b. Oxygen (O<sub>2</sub>)
  - c. Carbon Dioxide (CO<sub>2</sub>)
  - d. Carbon Monoxide (CO)
  - e. Hydrochloric Acid (HCl)
  - f. Sulfur Dioxide (SO<sub>2</sub>)
  - g. Sulfuric Acid (equivalent SO<sub>3</sub>)
  - h. Nitric Oxide (NO)
  - i. Nitrogen Dioxide (NO<sub>2</sub>)
  - j. Ammonia (NH<sub>3</sub>)
  - k. Argon (Ar)
  - l. Particulate (Flyash)
  - m. Water Vapor (H<sub>2</sub>O)
  - n. Stack Temperature
- 4) Water Balance Variables
- a. Chilled Water
    - i. HTX2 Chilling Water Flow
    - ii. Absorber Chilling Water Flow Rate
    - iii. Lean Solution Cooler Chilling Water Flow
  - b. Cooling Water
    - i. CO<sub>2</sub> Capture System Cooling Water Required
    - ii. CO<sub>2</sub> Compressor Cooling Water Flow
  - c. Waste Water
    - i. Water Bleed
- 5) Refrigeration Loads
- a. Cooler Chilling Load
  - b. Absorber Chilling Load
  - c. HeatX2 Chilling Load
- 6) Steam Flow Requirements
- a. Steam Flow to CO<sub>2</sub> Stripper & Heater
  - b. Steam Flow to NH<sub>3</sub> Stripper
- 7) Aspen Calculated Power Usage
- a. High Pressure Pump Power Usage

- 8) Heat Exchanger Variables
  - a. Aspen Heat Exchanger 1 Surface Area
  - b. Aspen Heat Exchanger 2 Surface Area
  - c. Aspen Heat Exchanger 1 Liquid Flow Rate
  - d. Aspen Heat Exchanger 2 Liquid Flow Rate
  - e. Aspen Heat Exchanger 3 Surface Area
  - f. Solution Cooler Surface Area

9) CO<sub>2</sub> to Compressor

10) Reclaimer Waste

11) Makeup

- a. Ammonia Makeup
- b. Water Makeup

### 3.4.2. The Response Model Equations

The individual response model equations are outlined below in the same order as above. Given initial values of the independent variables, a user should be able to calculate the dependent variable in each case by hand if necessary. In a number of cases the response model equations are nonlinear to take into account the extensive nonlinearities inherent in some parts of the process. In other cases, it was found that the data was modeled more accurately as two individual regressions separated by a boundary level. Each of the response model equations should be read from left to right on sequential lines for the headings below. The regression equations are given in mixed units, and a user can convert the coefficients to different units as needed.

- 1) Direct Contact Cooler (DCC) Flow Rates
  - a. Aspen Flue Flow Rate into DCC1 ( $R^2 = 0.98$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Aspen Flue Flow Rate in DCC1			cum/sec	=
	-282.8621			+
	0.025923	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.030188	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.012649	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
	2.210346	Flue Gas Temperature	°F	

b. Aspen Flue Flow Rate into DCC2 ( $R^2 = 0.99$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Aspen Flue Flow Rate in DCC2			cum/sec	=
	82.63099			+
	-0.7653	CO <sub>2</sub> Capture	%	+
	0.0076455	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.0236881	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.0001557	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
	0.0866089	Water Wash Flow Rate	kg/sec	+
	-0.1246694	Cooling Water Temperature	°F	

c. Circulating Water Flow Rate into HeatX1

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Circulating Water Flow Rate			kg/sec	=
	1.783			*
		Aspen Flue Flow Rate in DCC1	cum/sec	

2) CO<sub>2</sub> Capture and Ammonia Cleanup Variables

a. Gas Flow Rate Into Absorber ( $R^2 = 0.99$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Gas Flow Rate Into Absorber			cum/sec	=
	-0.418934			+
	0.0263549	CO <sub>2</sub> Capture	%	+
	0.0202873	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.0201861	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.0017601	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
	-0.0165	Circulating Water Flow Rate	kg/sec	+
	0.0143333	Water Wash Flow Rate	kg/sec	+
	0.1143204	Flue Gas Temperature	°F	+
	0.1018737	Cooling Water Temperature	°F	

The Lean Solvent Flow Rate regression is composed of two regressions separated by a boundary at 91% CO<sub>2</sub> capture. It was found that two regressions more accurately described the dataset than one alone. In the below regressions, a jump discontinuity is avoided at the 91% boundary by defining the regressions as equal at the boundary.

b. Lean Solvent Flow Rate

i. For  $\leq 91\%$  CO<sub>2</sub> Capture ( $R^2 = 0.98$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
ln(Lean Sorbent Flow)				=
	7.401809			+
	0.020091	CO <sub>2</sub> Capture	%	+
	-1.27139	ln(NH <sub>3</sub> Wt%)	%	+
	0.000227	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	6.06E-06	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	1.30E-06	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
Lean Sorbent Flow			kg/sec	=
		exp(ln(Lean Sorbent Flow))		

ii. For  $> 91\%$  CO<sub>2</sub> Capture ( $R^2 = 0.84$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
ln(Lean Sorbent Flow)				=
	7.401809			+
	0.020091*91			+
	0.05474001	(CO <sub>2</sub> Capture-91)	%	+
	-1.27139	ln(NH <sub>3</sub> Wt%)	%	+
	0.000227	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	6.06E-06	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	1.30E-06	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
Lean Sorbent Flow			kg/sec	=
		exp(ln(Lean Sorbent Flow))		

The equivalent molar flow rates of individual components (NH<sub>3</sub>, CO<sub>2</sub>, H<sub>2</sub>O) of the lean solvent determine the lean solvent flow rate, the NH<sub>3</sub>/CO<sub>2</sub> mole ratio, and the NH<sub>3</sub> Wt%. These relationships are described below and were used in Aspen Plus® but are not available in detail on the IECM user screens. However, a user could calculate these details by using data in the IECM and solving for the three unknowns in the three equations below.

- Lean In Flow (kg/sec) = 17/1000 \* Equivalent Lean In NH<sub>3</sub> Flow Rate (mol/sec) + 40/1000 \* Equivalent Lean In CO<sub>2</sub> Flow Rate (mol/sec) + 18/1000 \* Equivalent Lean In H<sub>2</sub>O Flow Rate (mol/sec)

2.  $\text{NH}_3/\text{CO}_2$  Mole Ratio = Equivalent  $\text{NH}_3$  Flow Rate (mol/sec)/ Equivalent  $\text{CO}_2$  Flow Rate (mol/sec)
3.  $\text{NH}_3$  Wt% =  $(17 * \text{Equivalent NH}_3 \text{ Flow Rate (mol/sec)}) / (17 * \text{Equivalent NH}_3 \text{ Flow Rate (mol/sec)} + 40 * \text{Equivalent CO}_2 \text{ Flow Rate (mol/sec)} + 18 * \text{Equivalent H}_2\text{O Flow Rate (mol/sec)})$

c. Rich Solvent Flow ( $R^2 = 0.99$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Rich Solvent Flow			kg/sec	=
	-117.4991			+
	1.715576	$\text{CO}_2$ Capture	%	+
	-.6278055	$\text{NH}_3$ Wt%	%	+
	0.9850088	Lean in Solvent Flow Rate	kg/sec	+
	0.0316657	$\text{CO}_2$ in the Incoming Flue Gas	mol/sec	+
	0.0010072	$\text{N}_2$ in the Incoming Flue Gas	mol/sec	+
	0.0013217	$\text{H}_2\text{O}$ in the Incoming Flue Gas	mol/sec	

d. Absorber Ammonia Slip ( $R^2 = 0.98$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
$\ln(\text{Absorber NH}_3 \text{ Slip})$				=
	6.254814			+
	2.71e-08	$\text{CO}_2$ Capture <sup>4</sup>	%	+
	0.007752	$\text{NH}_3$ Wt%	%	+
	0.0837074	$\ln(\text{Lean in Solvent Flow Rate})$	kg/sec	+
	-1.029565	$\ln(\text{CO}_2 \text{ in the Incoming Flue Gas})$	mol/sec	+
	0.7730576	$\ln(\text{N}_2 \text{ in the Incoming Flue Gas})$	mol/sec	+
	0.0075508	$\ln(\text{H}_2\text{O in the Incoming Flue Gas})$	mol/sec	+
Absorber $\text{NH}_3$ Slip			ppm	=
		$\exp(\ln(\text{Absorber NH}_3 \text{ Slip}))$		

e. Water Wash Water Flow Rate ( $R^2 = 0.99$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
ln(Water Wash Flow Rate)				=
	-3.459428			+
	-4.122067	ln(CO <sub>2</sub> Capture)	%	+
	2.949985	ln(NH <sub>3</sub> Wt%)	%	+
	1.976051	ln(Lean in Solvent Flow Rate)	kg/sec	+
	1.225447	ln(Absorber Ammonia Slip)	ppm	+
	-0.0570676	ln(Water Wash Ammonia Slip)	ppm	+
	-1.316468	ln(CO <sub>2</sub> in the Incoming Flue Gas)	mol/sec	+
	0.5253992	ln(N <sub>2</sub> in the Incoming Flue Gas)	mol/sec	+
	-0.0312865	ln(H <sub>2</sub> O in the Incoming Flue Gas)	mol/sec	+
Water Wash Flow Rate			kg/sec	=
		exp(ln(Water Wash Flow Rate))		

f. Solids Fraction in the Rich Solution ( $R^2 = 0.96$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Solids Fraction			wt%	=
	-11.05579			+
	-0.1659079	CO <sub>2</sub> Capture	%	+
	3.806864	NH <sub>3</sub> Wt%	%	+
	0.0028427	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	-0.0002865	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	-0.0001919	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+

3) Flue Gas to Stack Variables

a. Nitrogen (N<sub>2</sub>)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Nitrogen (N <sub>2</sub> ) To Stack			mol/sec	=
		Nitrogen (N <sub>2</sub> ) Into Model	mol/sec	

b. Oxygen (O<sub>2</sub>)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Oxygen (O <sub>2</sub> ) To Stack			mol/sec	=
		Oxygen (O <sub>2</sub> ) Into Model	mol/sec	

c. Carbon Dioxide (CO<sub>2</sub>)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
CO <sub>2</sub> To Stack			mol/sec	=
		CO <sub>2</sub> Into Model	mol/sec	-
		CO <sub>2</sub> Capture	mol/sec	*
		CO <sub>2</sub> Into Model	mol/sec	

d. Carbon Monoxide (CO)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
CO To Stack			mol/sec	=
		CO Into Model	mol/sec	-
		CO Removal Efficiency in IECM	mol/sec	*
		CO Into Model	mol/sec	

e. Hydrochloric Acid (HCl)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
HCl To Stack			mol/sec	=
		HCl Into Model	mol/sec	-
		HCl Removal Efficiency in IECM	mol/sec	*
		HCl Into Model	mol/sec	

f. Sulfur Dioxide (SO<sub>2</sub>)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
SO <sub>2</sub> To Stack			mol/sec	=
		SO <sub>2</sub> Into Model	mol/sec	-
		SO <sub>2</sub> Removal Efficiency in IECM	mol/sec	*
		SO <sub>2</sub> Into Model	mol/sec	

g. Sulfuric Acid (equivalent SO<sub>3</sub>)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
SO <sub>3</sub> To Stack			mol/sec	=
		SO <sub>3</sub> Into Model	mol/sec	-
		SO <sub>3</sub> Removal Efficiency in IECM	mol/sec	*
		SO <sub>3</sub> Into Model	mol/sec	

h. Nitric Oxide (NO)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
NO To Stack			mol/sec	=
		NO Into Model	mol/sec	-
		NO Removal Efficiency in IECM	mol/sec	*
		NO Into Model	mol/sec	

i. Nitrogen Dioxide (NO<sub>2</sub>)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
NO <sub>2</sub> To Stack			mol/sec	=
		NO <sub>2</sub> Into Model	mol/sec	-
		NO <sub>2</sub> Removal Efficiency in IECM	mol/sec	*
		NO <sub>2</sub> Into Model	mol/sec	

j. Ammonia (NH<sub>3</sub>)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
NH <sub>3</sub> to Stack			mol/sec	=
		[Nitrogen (N <sub>2</sub> ) to stack	mol/sec	+
		Oxygen (O <sub>2</sub> ) to stack	mol/sec	+
		Water Vapor (H <sub>2</sub> O) to stack	mol/sec	+
		Carbon Dioxide (CO <sub>2</sub> ) to stack	mol/sec	+
		Carbon Monoxide (CO) to stack	mol/sec	+
		Hydrochloric Acid (HCl) to stack	mol/sec	+
		Sulfur Dioxide (SO <sub>2</sub> ) to stack	mol/sec	+
		Sulfuric Acid to stack	mol/sec	+
		Nitric Oxide (NO) to stack	mol/sec	+
		Nitrogen Dioxide (NO <sub>2</sub> ) to stack	mol/sec	+
		Argon (Ar) to stack]	mol/sec	*
		Ammonia Slip Above Water Wash	ppm	/
		1E6		+
		Ammonia (NH <sub>3</sub> ) Into Model	mol/sec	

Since the user specifies the ammonia slip above the water wash, the ammonia to stack is calculated based on this slip and the flow rates of the other gases to the stack, as well as the ammonia coming into the capture system from the upstream environmental controls. The circulating water may remove some ammonia from the flue gas, but since this water goes to waste water treatment the ammonia removed by the circulating water is assumed to reach the environment eventually, and the model treats the ammonia above the water wash as ammonia that leaves out the stack. This quantity for mass balance purposes is calculated as:

$$\text{NH}_3 \text{ Flow Out the Stack } \left[ \frac{\text{mol}}{\text{sec}} \right] = \frac{\text{Ammonia Slip Above Water Wash [ppm]} * \sum \text{Mol Flowrates of All Gasses To Stack } \left[ \frac{\text{mol}}{\text{sec}} \right]}{1\text{E}6[\text{ppm}]}$$

k. Argon (Ar)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Argon (Ar) To Stack			Mole/sec	=
		Argon (Ar) Into Model	Mole/sec	

l. Particulate (Flyash)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Particulate (Flyash)To Stack			mol/sec	=
		Particulate Into Model	mol/sec	-
		Removal Efficiency in IECM	mol/sec	*
		Particulate Into Model	mol/sec	

Particulate is a separate mass flow within the flue gas in the IECM framework. It is not counted as any of the above flue gas species.

m. Water Vapor (H<sub>2</sub>O) (R<sup>2</sup> = 0.99)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Water Vapor To Stack			mol/sec	=
	-2388.283			+
	-1.518133	CO <sub>2</sub> Capture	%	+
	-1.822638	NH <sub>3</sub> Wt%	%	+
	0.0178865	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.048015	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.0004356	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
	32.39481	Cooling Water Temperature	°F	

n. Stack Temperature

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Stack Temperature			°F	=
		Cooling Water Temperature	°F	+
		10	°F	

The temperature approach on Heat Exchanger 1 is 5.5°C, and therefore the stack temperature is calculated as 5.5°C more than the cooling water temperature.

4) Water Balance Variables

a. Chilled Water

i. HTX2 Chilling Water Flow

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
HTX2 Chilled Water Flow			kg/sec	=
	1000		kg/sec	+

The HTX2 Chilled Water flow is fixed at 1000 kg/sec. For changes in system and power plant size, the HTX2 surface area and therefore the cost of the heat exchanger changes with larger or smaller flue gas flow rates to maintain a constant flue gas temperature entering the absorber of

5.6°C. Alternative options to lower the flue gas temperature to 5.6°C would have included changing the chilled water flow rate, changing the temperature of the chilled water, changing the temperature approach of the heat exchanger, or doing a combination of these options. Changing the heat exchanger surface area and fixing everything else as constant was the option that led to the most straightforward cost calculations.

ii. Absorber Chilling Water Flow Rate ( $R^2 = 0.96$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Absorber Chilling Water Flow Rate			kg/sec	=
	-9112.466			+
	72.1217	CO <sub>2</sub> Capture	%	+
	175.0502	NH <sub>3</sub> Wt%	%	+
	2.1176	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.0711	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.2883	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	

iii. Lean Solution Cooler Chilling Water Flow

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Lean Solution Cooler Chilling Water Flow			kg/sec	=
	18		kg/sec	

The Lean Solution Cooler Chilling Water Flow is fixed at 18 kg/sec. For changes in power plant size, the cooler surface area changes for larger or smaller lean solvent flow rates, for similar reasons as above.

b. Cooling Water

i. CO<sub>2</sub> Capture System Cooling Water Required (R<sup>2</sup> = 0.81)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
CO <sub>2</sub> Capture System Cooling Water Required			kg/sec	=
	7637.364			+
	4.7476	CO <sub>2</sub> Capture	%	+
	-32.3213	NH <sub>3</sub> Wt%	%	+
	-0.3061	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	-0.0549	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	-0.1733	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
	10.9584	Water Wash Flow Rate	kg/sec	+
	3.9532	Rich Solvent Flow	kg/sec	

Cooling water within the CO<sub>2</sub> capture system consists of cooling water for HTX1, for the CO<sub>2</sub> Stripper Condenser, for the NH<sub>3</sub> Stripper Condenser.

ii. CO<sub>2</sub> Compressor Cooling Water Flow

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
CO <sub>2</sub> Compressor Cooling Water Flow			[kg/sec]	=
	2.07		tonne H <sub>2</sub> O/tonne CO <sub>2</sub> to compression	

Power consumption for the CO<sub>2</sub> compressor from 1 to 152.7 bar in the IECM equals 0.099 kWh/kg CO<sub>2</sub>, while water consumption equals  $3123 \text{ {tonne H}_2\text{O}}/461.1 \text{ {tonne CO}_2} = 6.77 \text{ {tonne H}_2\text{O/tonne CO}_2}$ . The ammonia-based CO<sub>2</sub> capture system regenerates CO<sub>2</sub> at 27.5 bar, and therefore only requires 0.0304 {kWh/kg CO<sub>2</sub>} to compress to 152.7 bar. Therefore, the CO<sub>2</sub> compression cooling water required for the ammonia system is estimated as  $6.77 \text{ {tonne H}_2\text{O}}/\text{{tonne CO}_2} * (0.0304 \text{ {kwh/kg CO}_2}/0.099\text{{kwh/kg CO}_2}) = 2.07 \text{ {tonne H}_2\text{O}}/\text{{tonne CO}_2}$  to compression}.

c. Waste Water

i. Water Bleed ( $R^2 = 0.97$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Water Bleed			mol/sec	=
	5377.593			+
	-0.2095	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	-0.0859	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.7305	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
	0.6079	Circulating Water Flow Rate	kg/sec	+
	-0.8363	Flue Gas Temperature	°F	+
	-40.0557	Cooling Water Temperature	°F	

5) Refrigeration Loads (an energy measurement, with units in either Btu/hr or tons refrigeration/hr)

a. Cooler Chilling Load ( $R^2 = 0.82$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Cooler Chilling Load			tons refrigeration /hr	=
	-4291.751			+
	199.5634	CO <sub>2</sub> Capture	%	+
	-675.2407	NH <sub>3</sub> Wt%	%	+
	0.8383	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.2256	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	

b. Absorber Chilling Load ( $R^2 = 0.92$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Absorber Chilling Load			Tons refrigeration /hr	=
	-85704.43			+
	899.0939	CO <sub>2</sub> Capture	%	+
	1208.371	NH <sub>3</sub> Wt%	%	+
	17.28144	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.5816745	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.9546541	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	

c. HeatX2 Chilling Load ( $R^2 = 0.86$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
HeatX2 Chilling Load			tons refrigeration/hr	=
	-7863.238			+
	0.6541	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.4418	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	1.0217	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
	-2.7794	Circulating Water Flow Rate	kg/sec	+
	29.6179	Flue Gas Temperature	°F	+
	21.4856	Cooling Water Temperature	°F	

6) Steam Flow Requirements

a. Steam Flow to CO<sub>2</sub> Stripper & Heater (R<sup>2</sup> = 0.98)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Steam Flow to CO <sub>2</sub> Stripper			Btu/hr	=
	-7.19e+08			+
	5065583	CO <sub>2</sub> Capture	%	+
	2.48e+07	NH <sub>3</sub> Wt%	%	+
	184594.8	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	368130.9	Rich Solvent Flow	kg/sec	

b. Steam Flow to NH<sub>3</sub> Stripper (R<sup>2</sup> = 0.83)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Steam Flow to NH <sub>3</sub> Stripper			Btu/hr	=
	3.82E7			+
	-619	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	389857	Water Wash Flow Rate	kg/sec	+
	7287	Absorber NH <sub>3</sub> Slip	ppm	

The regression equations below estimate the power consumption of components calculated by Aspen Plus®. The pump power usage is a simple function of the rich solvent flow rate, as indicated in the table below.

7) Aspen Calculated Power Usage.

a. High Pressure Pump Power Usage (also called CO<sub>2</sub> Capture System Circulation Pumps or Solvent Circulation Pump) (R<sup>2</sup> = 0.98)

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
High Pressure Pump Power Usage			MW	=
	0.0081			+
	0.0031	Rich Solvent Flow	kg/sec	

8) Heat Exchanger Variables (these are used primarily for calculating the surface area of the key heat exchangers).

a. Aspen Heat Exchanger 1 ( $R^2 = 0.99$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Aspen Heat Exchanger 1			sqm	=
	10663.24			+
	10.7825	Circulating Water Flow Rate	kg/sec	+
	-152.1726	Cooling Water Temperature	°F	

b. Aspen Heat Exchanger 2 ( $R^2 = 0.96$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Aspen Heat Exchanger 2			sqm	=
	-379.8505			+
	0.1773122	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.1413364	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.0490353	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
	-0.4017035	Circulating Water Flow Rate	kg/sec	+
	2.05628	Flue Gas Temperature	°F	

c. Aspen Heat Exchanger 1 Liquid Flow Rate

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Aspen Heat Exchanger 1 Liquid Flow Rate			kg/sec	=
	5000			+

The Aspen Heat Exchanger 1 Liquid Flow Rate (the cooling water from the cooling tower) is fixed in this version of the model. As outlined for other major heat exchangers above, there are other options for determining the specifications of this heat exchanger for changes in plant size and flue gas flow rate, but fixing this liquid flow rate was done for reasons of simplicity.

d. Aspen Heat Exchanger 2 Liquid Flow Rate

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Aspen Heat Exchanger 2 Liquid Flow Rate			kg/sec	=
	1000			+

The Aspen Heat Exchanger 2 Liquid Flow Rate (the chilled water from the chillers) is fixed in this version of the model. As outlined for other major heat exchangers above, there are other

options for determining the specifications of this heat exchanger for changes in plant size and flue gas flow rate, but fixing this liquid flow rate was done for simplicity.

e. Aspen Heat Exchanger 3 ( $R^2 = 0.95$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Aspen Heat Exchanger 3			sqm	=
	29882.73			+
	-377.6261	CO <sub>2</sub> Capture	%	+
	-138.8811	NH <sub>3</sub> Wt%	%	+
	-9.4264	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.8638	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
	68.0621	Rich Solvent Flow	kg/sec	

f. Solution Cooler Surface Area ( $R^2 = 0.88$ )

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Solution Cooler Surface Area			sqm	=
	-428.6294			+
	19.7596	CO <sub>2</sub> Capture	%	+
	-2.0463	CO <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.2135	N <sub>2</sub> in the Incoming Flue Gas	mol/sec	+
	0.3976	H <sub>2</sub> O in the Incoming Flue Gas	mol/sec	+
	15.4230	Rich Solvent Flow	kg/sec	

9) CO<sub>2</sub> to Compressor

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
CO <sub>2</sub> To Compressor			mol/sec	=
		CO <sub>2</sub> Into Model	mol/sec	*
		CO <sub>2</sub> Capture	mol/sec	

10) Reclaimer Waste/CO<sub>2</sub> Product

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Reclaimer Waste/CO <sub>2</sub> Product			kg/tonne CO <sub>2</sub>	=
	6.0		kg/tonne CO <sub>2</sub>	

The reclaimer waste for the ammonia-based CO<sub>2</sub> capture system is estimated based on the amine system in the IECM. The reclaimer waste in the amine system is 3020 [kg/hr] for a CO<sub>2</sub> product flow rate of 500.7 [tonne/hr], meaning the reclaimer waste is 6.0 [kg/tonne CO<sub>2</sub>]. This is the same value used for the ammonia system.

11) Makeup

a. Ammonia Makeup

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Ammonia Makeup			kg/hr	=
	0.5	Reclaimer Waste	kg/hr	+
	1	NH <sub>3</sub> to Stack	mol/sec	*
	3600		sec/hr	*
	17		g NH <sub>3</sub> /mol	*
	0.001		kg/g	

Ammonia losses are assumed to constitute 50% of the reclaimer waste, plus ammonia lost out the stack and with the CO<sub>2</sub> product. In this analysis, the ammonia loss with the CO<sub>2</sub> product is assumed to be negligible.

b. Water Makeup

Dependent Variable	Coeff.	Independent Variable	Unit	Calc.
Water Makeup			tonne/hr	=
	0		tonne/hr	

There is actually a small requirement to purge water in the system to prevent a buildup of water. Depending on the system configuration, this could be large (~100-150 tonnes per hour), but is typically much less (~10-20 tonnes/hr). It is assumed that the sorbent reclaimer handles this purge, and therefore no makeup is required.

### 3.5. IECM Performance Estimates

The following performance and cost calculations have been programmed into the IECM and are completed for each model run. Some of these calculations are used directly for the ammonia-based CO<sub>2</sub> capture system while others are required to calculate the overall performance or cost of the power plant. The reference equipment and associated performance estimates listed in the following sections are taken from the literature, Aspen Plus®, and from the existing amine models in the IECM. When used, the amine model in the IECM is called the Amine Reference Model in the following sections.

Table 3.4 shows an overview and set of results of performance calculations for a plant with an ammonia-based CO<sub>2</sub> capture system, based on work from Chapter 2. The power consumption of the major components of the ammonia-based CO<sub>2</sub> capture system are listed, as well as the overall power consumption from other major plant systems (base plant, environmental controls for removing NO<sub>x</sub>, SO<sub>2</sub>, particulate matter etc.). Calculations for the overall net plant power and plant efficiency are also included. The calculation methods and associated equations for each of these performance based calculations are provided below and the overall structure of the calculations follows that of the table.

**Table 3.4: Power plant performance estimates. All values are in MWe equivalent.**

	Ammonia Capture System	Notes
Potential Power Available	827.6	Based on Coal Flow Rate
Auxiliary Steam Load		
CO <sub>2</sub> Stripper & Heater	108.5	
NH <sub>3</sub> Stripper	3.5	
Steam Turbine Power	715.6	Based on Aux. Steam Load
Auxiliary Electrical Load		
Flue Gas Blower	18.9	
Heat Exchanger 1 Pumps	2.2	
Heat Exchanger 2 Pumps	0.4	
Gas Cooling Water Pumps	0.6	
Chiller for Heat Exch 2	5.7	
Chiller for Absorber Cooling	48.2	
Chiller for Solvent Cooling	6.0	
Absorber Cooling Pumps	5.1	
Solvent Circulation Pumps (See Regression No.7)	3.5	
CO <sub>2</sub> Compression	16.9	
Balance of Plant	49.0	Based on Other IECM Models
Plant Net Power	558.7	Based on All IECM Models
Plant Efficiency (% HHV)	27.9%	Based on All IECM Models

Generally, the performance requirements for each piece of equipment is dependent on the user specifications and configurations in the IECM, the response model regression equations, data from the literature, and assumptions made in the original modeling effort.

### **Potential Power (Equivalent Plant Size without CCS)**

The potential power available as calculated in Table 3.4 is a parameter used for comparing plants, and is based on the coal flow rate of a baseline supercritical coal-fired power plant without CCS (DOE/NETL, 2007a). This is a simplified calculation, and is not used by the IECM (which relies on a large number parameters, material, and energy balances solved in an iterative manner). The calculation is, however, illustratively useful for comparing two similar power plants, one with and one without CO<sub>2</sub> capture.

$$\text{Potential Power \{MW\}} = \left[ \frac{\text{Coal Flow Rate With CCS \{lb/hr\}}}{\left[ \frac{\text{Coal Flow Rate w/o CCS \{lb/hr\}}}{\text{Gross Plant Size w/o CCS \{MW\}} \right]} \right]$$

### **Auxiliary Steam Loads (MWe Equivalent)**

The auxiliary steam load calculates the MWe Equivalent steam use of the CO<sub>2</sub> stripper & solution heater and the NH<sub>3</sub> stripper. This value is calculated because as steam is diverted for each of these pieces of equipment less steam is available to generate power in the steam turbine and the lost power is counted against that equipment. The electrical equivalent loss (MWe) due to the steam requirements of this equipment is calculated in several steps. First, the steam flow rate required is calculated directly, and then this value is used along with the Heat-to-Electricity Conversion Efficiency to calculate the electrical equivalent loss. The full calculation steps are as follows:

First, the steam mass flow rate required for each component is calculated as:

$$\dot{M}_{\text{Steam Flow}}\{\text{lb/hr}\} = \frac{\text{Heat Energy}\{\text{Btu/hr}\}}{\text{Enthalpy}_{\text{Steam Inlet}}\left\{\frac{\text{Btu}}{\text{lb Steam}}\right\} - \text{Enthalpy}_{\text{Steam Condensate}}\left\{\frac{\text{Btu}}{\text{lb Steam}}\right\}}$$

Where:

Heat Energy is estimated by Heat Energy Regressions (Steam Flow to CO<sub>2</sub> Stripper & Heater, Steam Flow to NH<sub>3</sub> Stripper).

$$\text{Enthalpy}_{\text{Steam Inlet}} = 1397.7 \{\text{Btu/lb}\} \text{ (DOE/NETL, 2007a)}$$

$$\text{Enthalpy}_{\text{Steam Condensate}} = 319.5 \{\text{Btu/lb}\} \text{ (DOE/NETL, 2007a)}$$

Then, the equivalent electrical loss (MWe Equivalent) due to steam drawn off by these system components is calculated as:

$$\text{MW}_{\text{Eq.}} = (\text{Heat} - \text{to} - \text{Electricity Efficiency}) * \dot{M}_{\text{Steam Flow}} * \text{Enthalpy}_{\text{Steam Inlet}} * 2.97\text{E}^{-7}$$

Where:

The Heat – to – Electricity Efficiency represents the energy conversion efficiency of the plant for converting steam to electricity, and is dimensionless. The default value for the Heat – to – Electricity Efficiency is 0.22 but it can be defined by the user in the IECM for specific applications.

$$2.97\text{E}^{-7} = \text{A Conversion Factor for Btu to MW}\{\text{MW}/[\text{Btu/hr}]\}$$

The overall equations for calculating the MWe Equivalent auxiliary steam loads are as follows:

CO2 Stripper & Heater Steam Use {MWeq}

= [Heat – to – Electricity Efficiency]

$$* \left[ \frac{\text{Steam Flow to CO2 Stripper \& Heater \{Btu/hr\}}}{\text{Enthalpy}_{\text{Steam Inlet}} \left\{ \frac{\text{Btu}}{\text{lb Steam}} \right\} - \text{Enthalpy}_{\text{Steam Condensate}} \left\{ \frac{\text{Btu}}{\text{lb Steam}} \right\}} \right]$$

$$* \left[ \text{Enthalpy}_{\text{Steam Inlet}} \left\{ \frac{\text{Btu}}{\text{lb Steam}} \right\} * 2.97\text{E}^{-7} \right]$$

NH3 Stripper Steam Use {MWeq}

= [Heat – to – Electricity Efficiency]

$$* \left[ \frac{\text{Steam Flow to NH3 Stripper \{Btu/hr\}}}{\text{Enthalpy}_{\text{Steam Inlet}} \left\{ \frac{\text{Btu}}{\text{lb Steam}} \right\} - \text{Enthalpy}_{\text{Steam Condensate}} \left\{ \frac{\text{Btu}}{\text{lb Steam}} \right\}} \right]$$

$$* \left[ \text{Enthalpy}_{\text{Steam Inlet}} \left\{ \frac{\text{Btu}}{\text{lb Steam}} \right\} * 2.97\text{E}^{-7} \right]$$

### Steam Turbine Power

The steam turbine power is calculated as the Potential Power {MW} minus the Auxiliary Steam Loads of the CO<sub>2</sub> stripper & heater, and the NH<sub>3</sub> stripper in MWe equivalent.

### Auxiliary Electrical Load - Flue Gas Blower

The auxiliary electrical load required by the Flue Gas Blower is calculated according to the equation:

Flue Gas Blower {MWe}

$$\begin{aligned} &= \text{Flue Gas Blower}_{\text{Ref}} \{\text{MWe}\} \\ &* \left[ \frac{\Delta\text{P Across DCC1, CO}_2 \text{ Absorber, Water Wash, \& DCC2} \{\text{psi}\}}{\Delta\text{P Across Reference System} \{\text{psi}\}} \right] \\ &* \left[ \frac{\text{Aspen Flue Flow Rate in DCC1} \{\text{cum/sec}\}}{\text{IECM Reference Flue Flow Rate} \{\text{cum/sec}\}} \right] \end{aligned}$$

Where:

Flue Gas Blower<sub>Ref</sub> {MWe} is from the reference amine system in the IECM = 6.3 MWe

$\Delta\text{P Across DCC1, CO}_2 \text{ Absorber, Water Wash, \& DCC2} \{\text{psi}\}$  is the gas phase pressure drop through these pieces of equipment. The default value for  $\Delta\text{P}$  is 3 psi but this value can be defined in the user interface screen in the IECM for specific applications.

$\Delta\text{P Across Reference System} \{\text{psi}\}$  is from the reference amine system in the IECM and equals 1 psi.

Aspen Flue Flow Rate in DCC1 {cum/sec} is calculated as a function of the flue gas molecular flow rates as well as the temperature of the flue gas flowing into DCC1. This variable is estimated from the Aspen Flue Flow Rate in DCC1 Regression Equation

IECM Reference Flue Flow Rate {cum/sec} is calculated in the IECM Amine Reference Model and is equal to 884.50 {cum/sec}

In the model, the default blower efficiency is assumed to be 75%, but this can be changed by the user and the power requirements will change as well accordingly.

### **Auxiliary Electrical Load - Heat Exchanger 1 Pumps**

The auxiliary electrical load required by the Heat Exchanger 1 Pumps is calculated according to the equation:

$$\begin{aligned} &\text{Heat Exchanger 1 Pumps \{MWe\}} \\ &= \text{Pumps Usage}_{\text{Ref}} [\text{MWe}] \\ &\quad * \left[ \frac{\text{Liquid Flow Rate \{tonne/hr\}}}{\text{Liquid Flow Rate in Reference System \{tonne/hr\}}} \right] \end{aligned}$$

Where:

Pump Usage<sub>Ref</sub> {MWe} is the pump power usage used to circulate solvent in the reference amine system in the IECM = 1.046 MWe

Liquid Flow Rate {tonne/hr} is cooling water from the cooling tower at a temperature of 80F and a flow rate of 18000 tonnes/hr.

Liquid Flow Rate in Reference System {tonne/hr} is the solvent circulation rate from the reference amine system in the IECM = 8660 tonne/hr.

In the model, the default pump efficiency is assumed to be 75%, but this number can be changed by the user.

### **Auxiliary Electrical Load - Heat Exchanger 2 Pumps**

The auxiliary electrical load required by the Heat Exchanger 2 Pumps is calculated according to the equation:

Heat Exchanger 2 Pumps {MWe}

= Pumps Usage<sub>Ref</sub> {MWe}

$$* \left\{ \frac{\text{Liquid Flow Rate \{tonne/hr\}}}{\text{Liquid Flow Rate in Reference System \{tonne/hr\}}} \right\}$$

Where:

Pump Usage<sub>Ref</sub> {MWe} is the pump power usage used to circulate solvent in the reference amine system in the IECM = 1.046 MWe

Liquid Flow Rate {tonne/hr} is chilling water from the chillers at a temperature of 37°F and a flow rate of 3600 tonnes/hr.

Liquid Flow Rate in Reference System {tonne/hr} is the solvent circulation rate from the reference amine system in the IECM = 8660 tonne/hr.

In the model, the default pump efficiency is assumed to be 75%, but this number can be changed by the user.

### **Auxiliary Electrical Load - Gas Cooling Water Pumps**

The auxiliary electrical load required by the Gas Cooling Water Pumps is calculated according to the equation:

Gas Cooling Water Pumps {MWe}

= Pumps Usage<sub>Ref</sub> {MWe}

$$* \left[ \frac{\text{Liquid Flow Rate \{tonne/hr\}}}{\text{Liquid Flow Rate in Reference System \{tonne/hr\}}} \right]$$

Where:

Pump Usage<sub>Ref</sub> {MWe} is the pump power usage used to circulate solvent in the reference amine system in the IECM = 1.046 MWe

Liquid Flow Rate {tonne/hr} is from the Circulating Water Flow Rate from the IECM ammonia user interface screen.

Liquid Flow Rate in Reference System {tonne/hr} is the solvent circulation rate from the reference amine system in the IECM = 8660 tonne/hr.

In the model, the default pump efficiency is assumed to be 75%, but this number can be changed by the user.

### **Auxiliary Electrical Load - Chiller for Heat Exch 2**

The Chiller for Heat Exch 2 supplies chilled water in order to cool the flue gas before it enters the absorber system. The chilling load required is dependent on the temperature of the chilled water, which is fixed here at 37°F, and the cooling duty of the heat exchanger. The auxiliary electrical load required by the Chiller for Heat Exch 2 is calculated according to the equation:

Chiller for Heat Exch 2 {MWe}

$$= \text{Chilling Electrical Usage} \left\{ \frac{\text{kWe}}{\text{Ton Cooling}} \right\} * \frac{1}{1000} \left\{ \frac{\text{MWe}}{\text{kWe}} \right\} \\ * \left[ \frac{\text{Chiller for Heat Exch 2 Chilling Load} \left\{ \frac{\text{Btu}}{\text{hr}} \right\}}{12000 \left\{ \frac{\text{BTU}}{\text{Ton Cooling}} \right\}} \right]$$

Where:

Chilling Electrical Usage{kWe/Ton Cooling} is the power usage required to cool water from 80°F to 37 °F, and equals 0.55 (Platts, 2004).

Chiller for Heat Exch 2 Chilling Load [Btu/hr] is from the Chiller for Heat Exch 2 Chilling Load Regression Equation.

### **Auxiliary Electrical Load - Chiller for Absorber Cooling**

The Chiller for Absorber Cooling supplies chilled water in order to cool the flue gas before it enters the absorber system. The chilling load required is dependent on the temperature of the chilled water, which is fixed here at 37°F, and the cooling duty of the heat exchanger within the absorber. The auxiliary electrical load required by the Chiller for Absorber Cooling is calculated according to the equation:

$$\begin{aligned} & \text{Chiller for Absorber Cooling \{MWe\}} \\ & = \text{Chilling Electrical Usage} \left\{ \frac{\text{kWe}}{\text{Ton Cooling}} \right\} * \frac{1}{1000} \left\{ \frac{\text{MWe}}{\text{kWe}} \right\} \\ & * \left[ \frac{\text{Chiller for Absorber Cooling Chilling Load} \left\{ \left[ \frac{\text{Btu}}{\text{hr}} \right] \right\}}{12000 \left\{ \frac{\text{BTU}}{\text{Ton Cooling}} \right\}} \right] \end{aligned}$$

Where:

Chilling Electrical Usage{kWe/Ton Cooling} is the power usage required to cool water from 80°F to 37°F, and equals 0.55 (Platts, 2004).

Chiller for Absorber Cooling Chilling Load {Btu/hr} is from the Chiller for Absorber Cooling Chilling Load Regression Equation

### **Auxiliary Electrical Load - Chiller for Solvent Cooling**

The Chiller for Solvent Cooling supplies chilled water in order to cool the solvent before it enters the absorber. The chilling load required is dependent on the temperature of the chilled water which is fixed here at 37°F, the temperature of the solvent, the flow rate of the solvent, the composition of the solvent, and the temperature approach of the heat exchanger cooling the solvent. The auxiliary electrical load required by the Chiller for Solvent Cooling is calculated according to the equation:

$$\begin{aligned} & \text{Chiller for Solvent Cooling \{MWe\}} \\ & = \text{Chilling Electrical Usage} \left\{ \frac{\text{kWe}}{\text{Ton Cooling}} \right\} * \frac{1}{1000} \left\{ \frac{\text{MWe}}{\text{kWe}} \right\} \\ & * \left[ \frac{\text{Chiller for Solvent Cooling Chilling Load} \left\{ \frac{\text{Btu}}{\text{hr}} \right\}}{12000 \left\{ \frac{\text{BTU}}{\text{Ton Cooling}} \right\}} \right] \end{aligned}$$

Where:

Chilling Electrical Usage{kWe/Ton Cooling} is the power usage required to cool water from 80°F to 37°F, and equals 0.55 (Platts, 2004).

Chiller for Solvent Cooling Chilling Load {Btu/hr} is from the Chiller for Solvent Cooling Chilling Load Regression Equation.

### **Auxiliary Electrical Load – Absorber Cooling Pumps**

The auxiliary electrical load required by the Absorber Cooling Pumps is calculated according to the equation:

$$\begin{aligned} & \text{Absorber Cooling Pumps \{MWe\}} \\ & = \text{Pumps Usage}_{\text{Ref}} \{MWe\} \\ & \quad * \left[ \frac{\text{Absorber Cooling Water Flow Rate \{tonne/hr\}}}{\text{Liquid Flow Rate in Reference System \{tonne/hr\}}} \right] \end{aligned}$$

Where:

Pump Usage<sub>Ref</sub> {MWe} is the pump power usage used to circulate Absorber in the reference amine system in the IECM = 1.046 MWe.

Liquid Flow Rate {tonne/hr} is a utility measurement from the Aspen Plus® Model, and is from the Absorber Cooling Water Flow Rate Regression Equation.

Liquid Flow Rate in Reference System {tonne/hr} is the Absorber circulation rate from the reference amine system in the IECM = 8660 tonne/hr.

In the model, the default pump efficiency is assumed to be 75%, but this number can be changed by the user.

### **Auxiliary Electrical Load – Solvent Circulation Pumps**

The solvent circulation pump energy consists of the energy required from the high pressure pump. The high pressure pump pressurizes the solvent slurry coming off the bottom of the absorber from 0.1 MPa 3.0 MPa. The auxiliary electrical load required by the Solvent Circulation Pumps is calculated according to the equation:

Solvent Circulation Pumps {MWe} = High Pressure Pump Power Usage {MWe}

Where:

High Pressure Pump Power Usage {MWe} is a utility measurement from the Aspen Plus® Model, and is from the High Pressure Pump Power Usage Regression Equation

### **Auxiliary Electrical Load – NH<sub>3</sub> Cleanup Pumps**

The auxiliary electrical load required by the NH<sub>3</sub> Cleanup Pumps is calculated according to the equation:

$$\begin{aligned} \text{NH}_3 \text{ Cleanup Pumps \{MWe\}} \\ &= \text{Pumps Usage}_{\text{Ref}} \{MWe\} \\ &\quad * \left[ \frac{\text{NH}_3 \text{ Cleanup Water Flow Rate \{kg/sec\}}}{\text{Liquid Flow Rate in Reference System \{tonne/hr\}}} \right] \end{aligned}$$

Where:

Pump Usage<sub>Ref</sub> {MWe} is the pump power usage used to circulate Absorber in the reference amine system in the IECM = 1.046 MW.

NH<sub>3</sub> Cleanup Water Flow Rate {kg/sec} is from the NH<sub>3</sub> Cleanup Water Flow Rate Regression Equation.

Liquid Flow Rate in Reference System {kg/sec} is the Absorber circulation rate from the reference amine system in the IECM = 2405.5 kg/sec.

In the model, the default pump efficiency is assumed to be 75%, but this number can be changed by the user. The value of the NH<sub>3</sub> Cleanup Pumps is small relative to the power usage of the

other system components. Typically, it will be on the order of ~ 0.01 MWe, and so while it is calculated by the IECM, is not listed as one of the main auxiliary electrical consumers in the output screens.

### **Auxiliary Electrical Load – CO<sub>2</sub> Compression**

The total compression work and associated electrical use required is dependent on the amount of CO<sub>2</sub> that goes to the compressor, as well as the initial and final CO<sub>2</sub> stream pressures. The auxiliary electrical load required by the CO<sub>2</sub> Compressors is calculated according to the equation:

$$\begin{aligned} \text{CO}_2 \text{ Compressor \{MWe\}} \\ &= \text{CO}_2 \text{ Flow Rate to Compressor \{kg/hr\}} \\ &\quad * \text{Energy Required to Compress from 28 bar to final pressure in bar } \left\{ \frac{\text{kWh}}{\text{kg CO}_2} \right\} \end{aligned}$$

Where:

CO<sub>2</sub> Flow Rate to Compressor {kg/hr} is calculated in the model as the CO<sub>2</sub> To Compressor.

Energy Required to Compress from 28 bar to final pressure in bar  $\left\{ \frac{\text{kWh}}{\text{kg CO}_2} \right\}$  is derived from the CO<sub>2</sub> compression model within the IECM. For compression between 28 bar and 152 bar, this value equals 0.03.

### **Auxiliary Electrical Load – Additional Cooling Tower Pump Requirements**

The large cooling demands from the ammonia-based CO<sub>2</sub> capture system affect the water requirements in the plant and change the water flow to the cooling tower, which affects pumping

requirements. The cooling tower pumping requirements for the IECM are detailed in the Wet Cooling Tower documentation for the IECM.

### **3.6. IECM Cost Estimates**

Table 3.5 below shows an overview and set of results of cost calculations for a plant with an ammonia-based CO<sub>2</sub> capture system, based on work in Chapter 2. The cost of major components of the ammonia-based CO<sub>2</sub> capture system are listed, as well as the overall costs from other major plant systems (base plant, environmental controls for removing NO<sub>x</sub>, SO<sub>2</sub>, particulate matter etc.). Calculations for the overall revenue required are also included. The calculation methods and associated equations for each of these cost calculations are provided in the following sections. Typically, where costs in the IECM or the literature were not available, cost estimates for equipment were determined through the equipment sizing and costing functions of Aspen Icarus®. The costs in Aspen Icarus® were given in first quarter 2008 dollars and are scaled in the IECM to the appropriate dollar year as specified by the user using the Marshal & Swift Index or a similar index. These costs as well as the parameters used to determine these costs are shown below. Not all cost details are directly available in the IECM results screens in the interest of clarity, and space, and costs for some pieces of equipment are aggregated and presented as costs for functional areas (for example, all the heat exchanger costs below have been aggregated as one total heat exchanger cost on the IECM result screen). However, using the regression equations, the data supplied by the IECM, and the equations below, a user should be able to reproduce the costs for individual pieces of equipment.

**Table 3.5: Power plant cost estimates. All values are in \$2007 constant dollars.**

	Ammonia Capture System	Notes
CO2 Capture Process Area Costs		
DCC #1	30.9	
DCC #2	23.3	
Flue Gas Blower	6.3	
Heat Exch. 1	6.7	
Heat Exch. 2	2.9	
Heat Exch. 1 Pumps	1.4	
Heat Exch. 2 Pumps	0.5	
Cooling Water Circulation Pumps	0.7	
Chiller System	54.6	
Absorber	105.1	
Absorber Pumps	2.4	
Heat Exch. 3	41.6	
Solvent Circulation Pumps	7.9	
Solvent Heater 1	2.2	
Solvent Cooler	2.2	
CO <sub>2</sub> Stripper	35.1	
CO <sub>2</sub> Stripper Reboiler	13.4	
Water Wash	2.2	
Heat Exch. 4	0.1	
NH <sub>3</sub> Stripper	1.5	
NH <sub>3</sub> Cleanup Pumps	0.8	
Steam Extractor	3.3	
Sorbent Reclaimer	1.1	
Sorbent Processing	1.1	
Drying and Compress Unit	18.6	
General Facilities Capital	5.7	Based on IECM Data
Eng. & Home Office Fees	34.3	Based on IECM Data
Project Contingency Cost	59.9	Based on IECM Data
Process Contingency Cost	17.1	Based on IECM Data
CO2 System (TCR)	483.0	Based on Area Costs
Base Plant (TCR)	884.1	Based on IECM Data
Cooling Tower (TCR)	62.7	Based on IECM Data
NO <sub>x</sub> Control (TCR)	33.7	Based on IECM Data
TSP Control (TCR)	49.8	Based on IECM Data
SO <sub>2</sub> Control (TCR)	138.7	Based on IECM Data
CO2 Transport & Storage O&M	22.3	Based on IECM Data
Balance of Plant O&M	128.9	Based on IECM Data
Plant Total Capital Requirement	1652.0	Based on TCR Costs
Total O&M Costs	151.3	Total O&M
Capital Required (\$/kW-net)	2956.8	Based on Performance
Revenue Required (\$/MWh)	105.4	eq. 2

The scaling of most equipment is non-linear because the value to be scaled is assumed to benefit from economies of scale as the size of the equipment increases. The form of the equation for much of the cost scaling is similar to the one below.

$$X = X_{\text{Ref}} * \left(\frac{Y}{Y_{\text{Ref}}}\right)^{0.6}$$

Where

X = the cost of the piece of equipment as estimated by the IECM.

X<sub>Ref</sub> = a reference cost of a similar piece of reference equipment that may be larger or smaller, or may process more or less of a key component of the system.

Y = A process parameter of the piece of equipment in which costs are to be estimated (material flow, energy requirements).

Y<sub>Ref</sub> = a reference process parameter of the reference equipment.

The reference equipment and associated costs are taken from the literature, Aspen Icarus®, and in many cases, and from the existing amine models in the IECM. When used, the amine model in the IECM is called the Amine Reference Model in the following sections.

### Cost Calculations – DCC1

The total cost for DCC1 is based on the volumetric flow rate through the direct contact cooler and is calculated according to the equation:

$$\text{DCC1 Cost \{\$\}} = \text{Reference System Cost \{\$\}} * \left[ \frac{\text{Aspen Flue Flow Rate in DCC1 \{cum/sec\}}}{\text{IECM Reference Flue Flow Rate\{cum/sec\}}} \right]^{0.6}$$

Where:

Reference System Cost {\\$} is calculated in the IECM Amine Reference Model and is equal to 32.48 {\\$M}.

Aspen Flue Flow Rate in DCC1 {cum/sec} is calculated as a function of the flue gas molecular flow rates as well as the temperature of the flue gas flowing into DCC1. This variable is estimated from the Aspen Flue Flow Rate in DCC1 Regression Equation.

IECM Reference Flue Flow Rate {cum/sec} calculated in the IECM Amine Reference Model and is equal to 884.50 {cum/sec}.

### **Cost Calculations – DCC2**

The total cost for DCC2 is based on the volumetric flowrate through the direct contact cooler and is calculated according to the equation:

$$\text{DCC2 Cost \{\$\}} = \text{Reference System Cost \{\$\}} * \left[ \frac{\text{Aspen Flue Flow Rate in DCC2 \{cum/sec\}}}{\text{IECM Reference Flue Flow Rate \{cum/sec\}}} \right]^{0.6}$$

Where:

Reference System Cost {\\$} is calculated in the IECM Amine Reference Model and is equal to 32.48 {\\$M}.

Aspen Flue Flow Rate in DCC2 {cum/sec} is calculated is a function of the flue gas molecular flow rates as well as the temperature of the flue gas flowing into DCC2. This variable is estimated from the Aspen Flue Flow Rate in DCC2 Regression Equation.

IECM Reference Flue Flow Rate {cum/sec} calculated in the IECM Amine Reference Model and is equal to 884.50 {cum/sec}.

### Cost Calculations – Flue Gas Blower

The total cost for the Flue Gas Blower is based on the volumetric flow rate through DCC1 (and therefore also the Flue Gas Blower) and is calculated according to the equation:

Flue Gas Blower Cost { $\$$ }

$$= \text{Reference System Cost } \{\$\} * \left[ \frac{\text{Aspen Flue Flow Rate in DCC1 } \{\text{cum/sec}\}}{\text{IECM Reference Flue Flow Rate } \{\text{cum/sec}\}} \right]^{0.6}$$

Where:

Reference System Cost { $\$$ } is calculated in the IECM Amine Reference Model and is equal to 6.639 { $\$M$ }.

Aspen Flue Flow Rate in DCC1 {cum/sec} is calculated is a function of the flue gas molecular flow rates as well as the temperature of the flue gas flowing into DCC1. This variable is estimated from the Aspen Flue Flow Rate in DCC1 Regression Equation.

IECM Reference Flue Flow Rate {cum/sec} calculated in the IECM Amine Reference Model and is equal to 884.50 {cum/sec}.

The default blower efficiency is assumed to be 75%, but this value can be changed by the user in the model.

### Cost Calculations – Heat Exchanger 1

The total cost for Heat Exchanger 1 is based on the heat exchanger surface area and is calculated according to the equation:

$$\begin{aligned} & \text{Heat Exchanger 1 Installed Cost \{\$\}} \\ & = \text{Reference System Cost \{\$\}} \\ & * \left[ \frac{\text{Heat Exchanger 1 Surface Area \{sqm\}}}{\text{Reference Heat Exchanger Surface Area \{sqm\}}} \right]^{0.6} \end{aligned}$$

Where:

Reference System Cost {\\$} is calculated by the Ammonia System Aspen Icarus® Reference Model and is equal to 6.671 {\\$M}.

Reference Heat Exchanger Surface Area {sqm} is calculated by the Ammonia System Aspen Plus® Reference Model and is equal to 14155.3 {sqm}.

Heat Exchanger 1 Surface Area {sqm} is calculated as a function of the temperature approach of the heat exchange (default 5.5°C between the cold side inlet and the hot side outlet), the flow rates, flow compositions, flow rate heat capacity, chemical reactions, and the temperature of the flows into and out of the heat exchanger. This variable is estimated from the Aspen Heat Exchanger 1 Regression Equation.

### **Cost Calculations – Heat Exchanger 2**

The total cost for Heat Exchanger 2 is based on the heat exchanger surface area and is calculated according to the equation:

$$\begin{aligned} & \text{Heat Exchanger 2 Installed Cost \{\$\}} \\ & = \text{Reference System Cost \{\$\}} \\ & * \left[ \frac{\text{Heat Exchanger 2 Surface Area \{sqm\}}}{\text{Reference Heat Exchanger Surface Area \{sqm\}}} \right]^{0.6} \end{aligned}$$

Where:

Reference System Cost {\$} is calculated by the Ammonia System Aspen Icarus® Reference Model and is equal to 2.871 {\$M}.

Reference Heat Exchanger Surface Area {sqm} is calculated by the Ammonia System Aspen Plus® Reference Model and is equal to 3102.62 {sqm}.

Heat Exchanger 2 Surface Area {sqm} is calculated as a function of the temperature approach of the heat exchange (default 5.5°C between the cold side inlet and the hot side outlet), the flow rates, flow compositions, flow rate heat capacity, chemical reactions, and the temperature of the flows into and out of the heat exchanger. This variable is estimated from the Aspen Heat Exchanger 2 Regression Equation.

### Cost Calculations – Heat Exchanger 1 Pumps

The total cost for Heat Exchanger 1 Pumps is calculated according to the equation:

$$\begin{aligned} \text{Heat Exchanger 1 Pumps Installed Cost \{\$\}} \\ &= \text{Reference System Cost \{\$\}} \\ &\quad * \left[ \frac{\text{Heat Exchanger 1 Liquid Flow Rate \{tonne/hr\}}}{\text{Reference Liquid Flow Rate \{tonne/hr\}}} \right]^{0.6} \end{aligned}$$

Where:

Reference System Cost {\$} is based on the circulating water pumps cost in an NETL Reference Study and equals 2.065 {\$M} (DOE/NETL, 2007a).

Reference Liquid Flow Rate {tonne/hr} is based on the circulating water pumps flow rate in an NETL Reference Study and equals 32706 {tonne/hr} (DOE/NETL 2007a).

Heat Exchanger 1 Liquid Flow Rate {tonne/hr} is estimated from the Aspen Heat Exchanger 1 Liquid Flow Rate Regression Equation.

### **Cost Calculations – Heat Exchanger 2 Pumps**

The total cost for Heat Exchanger 2 Pumps is calculated according to the equation:

$$\begin{aligned} &\text{Heat Exchanger 2 Pumps Installed Cost \{\$\}} \\ &= \text{Reference System Cost \{\$\}} \\ &\quad * \left[ \frac{\text{Heat Exchanger 2 Liquid Flow Rate \{tonne/hr\}}}{\text{Reference Liquid Flow Rate \{tonne/hr\}}} \right]^{0.6} \end{aligned}$$

Where:

Reference System Cost {\\$} is based on the circulating water pumps cost in an NETL Reference Study and equals 2.065 {\\$M} (DOE/NETL, 2007a).

Reference Liquid Flow Rate {tonne/hr} is based on the circulating water pumps flow rate in an NETL Reference Study and equals 32706 {tonne/hr} (DOE/NETL, 2007a).

Heat Exchanger 2 Liquid Flow Rate {tonne/hr} is estimated from the Aspen Heat Exchanger 2 Liquid Flow Rate Regression Equation.

### **Cost Calculations – Cooling Water Circulation Pumps**

The total cost for Cooling Water Circulation Pumps is calculated according to the equation:

Cooling Water Circulation Pumps Installed Cost {\$}

= Reference System Cost {\$}

$$* \left[ \frac{\text{Cooling Water Circulation Liquid Flow Rate \{tonne/hr\}}}{\text{Reference Liquid Flow Rate \{tonne/hr\}}} \right]^{0.6}$$

Where:

Reference System Cost {\$} is based on the circulating water pumps cost in an NETL Reference Study and equals 2.065 {\$M} (DOE/NETL, 2007a).

Reference Liquid Flow Rate {tonne/hr} is based on the circulating water pumps flow rate in an NETL Reference Study and equals 32706 {tonne/hr} (DOE/NETL, 2007a).

Cooling Water Circulation Liquid Flow Rate {tonne/hr} is estimated from the Circulating Water Flow Rate IECM Interface Screen.

### Cost Calculations – Chiller System

The total cost for Chiller System is based on the chilling loads required by the ammonia-based CO<sub>2</sub> capture system. The total cost is calculated according to the equation:

Chiller Installed Cost {\$}

$$= \text{Chilling System Installed Cost} \left\{ \frac{\$}{\text{ton chilling}} \right\}$$

$$* \frac{[\text{DCC2 Chilling Load \{Btu\}} + \text{Absorber Chilling Load \{Btu\}} + \text{Lean Solution Chilling Load \{Btu\}}]}{12000 \left\{ \frac{\text{Btu}}{\text{ton chilling}} \right\}}$$

Where:

Chilling System Installed Cost {\$/ton chilling} is equal to \$441.245 in \$2007, scaled from \$350 in \$2000, from a reference in the literature (RDC, 2003) using the Marshal & Swift Index.

DCC2 Chilling Load {Btu}, Absorber Chilling Load {Btu}, and Lean Solution Chilling Load {Btu} are estimated from the regression equations associated with each of these variables.

### Cost Calculations – Absorber

Currently, the most appropriate absorber design for the ammonia-based CO<sub>2</sub> capture system is a spray tower absorber designed to handle a significant amount of solids precipitation. Spray tower absorber equipment is typically used in wet flue gas desulfurization (FGD) processes, and is considerably different than the traditional packed or trayed columns found in amine scrubbers. Therefore the base costs for the ammonia-based CO<sub>2</sub> capture system absorber are taken from the wet FGD model rather than the existing amine system model in the IECM. The total cost of the spray tower absorber and associated equipment is calculated according to the equation:

Absorber Installed Cost { \$ }

$$= \text{Reference System Cost } \{ \$ \} * \left[ \frac{\text{Gas Flow Rate into Absorber } \{ \text{cum/sec} \}}{\text{Reference Gas Flow Rate } \{ \text{cum/sec} \}} \right]^{0.6}$$

Where:

Reference Installed Cost { \$ } is equal to 105.1 { \$M } in \$2007, based on the wet FGD system cost in the IECM reference amine model. It is appropriate to use the cost of the wet FGD system of a plant with an amine system, because the wet FGD system will have to

process approximately the same amount of flue gas as the absorber in the CO<sub>2</sub> capture system. A plant without a CO<sub>2</sub> capture system that produces the same amount of net electricity will have to process considerably less flue gas all other things being equal.

Gas Flow Rate into Absorber {cum/sec} is from the Gas Flow Rate Into Absorber regression equation.

Reference Gas Flow Rate {cum/sec} is based on the gas flow rate in the Aspen Plus® Reference Study and equals 497.3 {cum/sec}.

### **Cost Calculations – Absorber Pumps**

The total cost for Absorber Pumps is calculated according to the equation:

Absorber Pumps Installed Cost {\$}

$$= \text{Reference System Cost } \{\$\} * \left[ \frac{\text{Absorber Liquid Flow Rate } \{\text{tonne/hr}\}}{\text{Reference Liquid Flow Rate } \{\text{tonne/hr}\}} \right]^{0.6}$$

Where:

Reference System Cost {\$} is based on the circulating water pumps cost in an NETL Reference Study and equals 2.065 {\$M} (DOE/NETL, 2007a).

Reference Liquid Flow Rate {tonne/hr} is based on the circulating water pumps flow rate in an NETL Reference Study and equals 32706 {tonne/hr} (DOE/NETL, 2007a).

Absorber Liquid Flow Rate {tonne/hr} is estimated from the Absorber Cooling Water Flow Rate Regression Equation.

### Cost Calculations – Heat Exchanger 3 (The Rich/Lean Heat Exchanger)

The total cost for Heat Exchanger 3 is based on the heat exchanger surface area and is calculated according to the equation:

$$\begin{aligned} &\text{Heat Exchanger 3 Installed Cost \{\$\}} \\ &= \text{Reference System Cost \{\$\}} \\ &\quad * \left[ \frac{\text{Heat Exchanger 3 Surface Area \{sqm\}}}{\text{Reference Heat Exchanger Surface Area \{sqm\}}} \right]^{0.6} \end{aligned}$$

Where:

Reference System Cost {\\$} is calculated by the Ammonia System Aspen Icarus® Reference Model and is equal to 19.343 {\\$M}.

Reference Heat Exchanger Surface Area {sqm} is calculated by the Ammonia System Aspen Plus Reference Model and is equal to 10078.5 {sqm}.

Heat Exchanger 3 Surface Area {sqm} is calculated as a function of the temperature approach of the heat exchange (default 5.5°C between the cold side inlet and the hot side outlet), the flow rates, flow compositions, flow rate heat capacity, chemical reactions, and the temperature of the flows into and out of the heat exchanger. This variable is estimated from the Aspen Heat Exchanger 3 Regression Equation.

### Cost Calculations – Solvent Circulation Pumps (The High Pressure Pump)

The total cost for Solvent Circulation Pump is calculated according to the equation:

Solvent Circulation Pumps Installed Cost { \$ }

= Reference System Cost { \$ }

$$* \left[ \frac{\text{Solvent Circulation Liquid Flow Rate \{tonne/hr\}}}{\text{Reference Liquid Flow Rate \{tonne/hr\}}} \right]^{0.6}$$

Where:

Reference System Cost { \$ } is based on the circulating water pumps cost in the IECM amine reference model and is equal to 13.38 { \$M }.

Reference Liquid Flow Rate { tonne/hr } is based on the circulating water pumps flow rate in the IECM amine reference model and is equal to 2405.5 { kg/sec }.

Solvent Circulation Liquid Flow Rate { tonne/hr } is estimated from Rich Solvent Flow Regression Equation.

### Cost Calculations – Solution Heater 1

The total cost for Solution Heater is calculated according to both the solvent flow rate through the heater as well as the heat transferred, according to the equation:

Solution Heater Installed Cost { \$ }

= Reference System Cost { \$ }

$$* \left[ \frac{\text{Solution Heater Liquid Flow Rate \{tonne/hr\}}}{\text{Reference Liquid Flow Rate \{tonne/hr\}}} \right]$$

$$* \left[ \frac{\text{Steam Flow to Heater \{Btu/hr\}}}{\text{Reference Steam Use \{Btu/hr\}}} \right]^{0.6}$$

Where:

Reference System Cost {\$} is based on Aspen Icarus® costing in the ammonia system reference model and is equal to 26.38 {\$M}.

Reference Liquid Flow Rate {tonne/hr} is the Absorber circulation rate from the reference amine system in the IECM and is equal to 8660 {tonne/hr}.

Solvent Circulation Liquid Flow Rate {tonne/hr} is estimated from Rich Solvent Flow Regression Equation.

Reference Steam Use {Btu/hr} is from the steam use in the CO<sub>2</sub> stripper reboiler in the IECM amine reference model, and is equal to 1.88E9 {Btu/hr}.

Steam Flow to Heater {Btu/hr} is from the Steam Flow to Heater Regression Equation.

### Cost Calculations – Solution Cooler

The total cost for the Solution Cooler is based on the heat exchanger surface area and is calculated according to the equation:

$$\begin{aligned} & \text{Solution Cooler Installed Cost \{ \$ \}} \\ & = \text{Reference System Cost \{ \$ \}} \\ & \quad * \left[ \frac{\text{Solution Cooler Surface Area \{ sqm \}}}{\text{Reference Heat Exchanger Surface Area \{ sqm \}}} \right]^{0.6} \end{aligned}$$

Where:

Reference System Cost {\$} is calculated by the Ammonia System Aspen Icarus® Reference Model and is equal to 2.315 {\$M}.

Reference Heat Exchanger Surface Area {sqm} is calculated by the Ammonia System Aspen Plus® Reference Model and is equal to 17138.89 {sqm}.

Solution Cooler Surface Area {sqm} is calculated as a function of the temperature approach of the heat exchange (default 5.5°C between the cold side inlet and the hot side outlet), the flow rates, flow compositions, flow rate heat capacity, chemical reactions, and the temperature of the flows into and out of the heat exchanger. This variable is estimated from the Solution Cooler Surface Area Regression Equation.

### Cost Calculations – CO<sub>2</sub> Stripper

The total cost for Solvent Circulation Pump is calculated according to the equation:

CO<sub>2</sub> Stripper Installed Cost {\$}

$$= \text{Reference System Cost \{\$\}} * \left[ \frac{\text{CO}_2 \text{ Stripper Liquid Flow Rate \{tonne/hr\}}}{\text{Reference Liquid Flow Rate \{tonne/hr\}}} \right]^{0.6}$$

Where:

Reference System Cost {\$} is based on the CO<sub>2</sub> Stripper cost in the IECM amine reference model and is equal to 54.07 {\$M}.

Reference Liquid Flow Rate {tonne/hr} is the Absorber circulation rate from the reference amine system in the IECM and is equal to 8660 {tonne/hr}.

Solvent Circulation Liquid Flow Rate {tonne/hr} is estimated from the Rich Solvent Flow Regression Equation.

## Cost Calculations – CO<sub>2</sub> Stripper Reboiler

The total cost for the CO<sub>2</sub> Stripper Reboiler is calculated according to both the solvent flow rate through the reboiler as well as the heat transferred, according to the equation:

$$\begin{aligned} &\text{CO}_2 \text{ Stripper Reboiler Installed Cost \{\$\}} \\ &= \text{Reference System Cost \{\$\}} \\ &\quad * \left[ \frac{\text{CO}_2 \text{ Stripper Reboiler Liquid Flow Rate \{tonne/hr\}}}{\text{Reference Liquid Flow Rate \{tonne/hr\}}} \right. \\ &\quad \left. * \frac{\text{Steam Flow to Reboiler \{Btu/hr\}}}{\text{Reference Steam Use \{Btu/hr\}}} \right]^{0.6} \end{aligned}$$

Where:

Reference System Cost {\\$} is based on the cost of the reboiler in the IECM amine system reference model and is equal to 26.38 {\\$M}.

Reference Liquid Flow Rate {tonne/hr} is the Absorber circulation rate from the reference amine system in the IECM and is equal 8660 {tonne/hr}.

Solvent Circulation Liquid Flow Rate {tonne/hr} is estimated from the Rich Solvent Flow Regression Equation.

Reference Steam Use {Btu/hr} is from the steam use in the CO<sub>2</sub> stripper reboiler in the IECM amine reference model, and is equal to 1.88E9 {Btu/hr}.

Steam Flow to Reboiler {Btu/hr} is from the Steam Flow to Reboiler Regression Equation.

### **Cost Calculations – Water Wash Absorber**

The total cost for CO<sub>2</sub> water wash absorber was sized according to Aspen Icarus®, and is assumed to be constant and equal to 2.2 {\$M} in \$2007.

### **Cost Calculations – Heat Exch. 4**

The total cost for Heat Exch. 4 was sized according to Aspen Icarus®, and is assumed to be constant and equal to 0.1 {\$M} in \$2007. The calculated cost of this heat exchanger is considerably lower than the other heat exchangers in the ammonia-based CO<sub>2</sub> capture system primarily both because the liquid flow rates are typically much smaller, and because the hot and cold streams are already closer in temperature than elsewhere in the process. These two factors significantly reduce the heat transfer requirement and therefore the surface area and corresponding cost. Due to the small cost of this heat exchanger relative to the other equipment in the process, the cost of this exchanger is held constant in the model through all power plant sizes and model conditions by default. For significantly different applications, the user can adjust the total cost of the heat exchangers in the IECM by adjusting the appropriate retrofit parameters in the user screen.

### **Cost Calculations – NH<sub>3</sub> Stripper**

The total cost for the NH<sub>3</sub> Stripper was sized according to Aspen Icarus®, and is assumed to be constant and equal to 1.51 {\$M} in \$2007. The calculated cost of this stripper is considerably lower than most of the other components in the ammonia-based CO<sub>2</sub> capture system primarily because it is smaller and has a lower liquid flow rate and heat transfer surface area in the reboiler. Due to the small cost of the NH<sub>3</sub> stripper relative to the other equipment in the process, the cost of this exchanger is held constant in the model through all power plant sizes and model conditions by default. For significantly different applications, the user can adjust the total cost of

the overall water wash system in the IECM by adjusting the appropriate retrofit parameters in the user screen.

### **Cost Calculations – NH<sub>3</sub> Cleanup Pumps**

The total cost for NH<sub>3</sub> Cleanup Pump (the water wash pumps) is calculated according to the equation:

NH<sub>3</sub> Cleanup Pumps Installed Cost {\$}

$$= \text{Reference System Cost \{\$\}} * \left[ \frac{\text{NH}_3 \text{ Cleanup Liquid Flow Rate \{tonne/hr\}}}{\text{Reference Liquid Flow Rate \{tonne/hr\}}} \right]^{0.6}$$

Where:

Reference System Cost {\$} is based on the circulating water pumps cost in the IECM amine reference model and is equal to 13.38 {\$M}.

Reference Liquid Flow Rate {tonne/hr} is based on the circulating water pumps flow rate in the IECM amine reference model and is equal to 2405.5 {kg/sec}.

Solvent Circulation Liquid Flow Rate {tonne/hr} is estimated from the Water Wash Flow Rate Regression Equation.

### **Cost Calculations – Steam Extractor**

The total cost for the Steam Extractor is taken from the IECM amine system reference model, and is assumed to be constant and equal to 3.3 {\$M} in \$2007. This value is assumed constant regardless of the size of the plant. For specific applications, the user can adjust the total cost of the steam extractor in the IECM by adjusting the appropriate retrofit parameters in the user screen.

### **Cost Calculations – Sorbent Reclaimer**

The total cost for the Sorbent Reclaimer is taken from the IECM amine system reference model, and is assumed to be constant and equal to 1.12 {\$M} in \$2007. This value is assumed constant regardless of the size of the plant. For specific applications, the user can adjust the total cost of the steam extractor in the IECM by adjusting the appropriate retrofit parameters in the user screen.

### **Cost Calculations – Sorbent Processing**

The total cost for Sorbent Processing is taken from the IECM amine system reference model, and is assumed to be constant and equal to 1.13 {\$M} in \$2007. This value is assumed constant regardless of the size of the plant. For specific applications, the user can adjust the total cost of the steam extractor in the IECM by adjusting the appropriate retrofit parameters in the user screen.

### **Cost Calculations – Drying and Compression Unit**

The total compression work and therefore electrical use required is dependent on the amount of CO<sub>2</sub> that goes to the compressor, as well as the initial and final stream pressures. The costs of the compressors are assumed to be proportional to the MWe usage required. Therefore, the total cost for the Drying and Compression Unit is calculated according to the equation:

$$\begin{aligned} & \text{Drying and Compression Unit Installed Cost \{\$\}} \\ & = \text{Reference System Cost \{\$\}} \\ & * \left[ \frac{\text{Auxiliary Electrical Load for CO}_2 \text{ Compressor \{MWe\}}}{\text{Reference Electrical Use for CO}_2 \text{ Compressor \{MWe\}}} \right] \end{aligned}$$

Where:

Reference System Cost {\$} is based on the Drying and Compression Unit cost in the IECM amine reference model and is equal to 57.3 {\$M}.

Reference Electrical Use for CO<sub>2</sub> Compressor is based on the IECM amine reference model and is equal to 52.89 [MWe].

The Auxiliary Electrical Load for CO<sub>2</sub> Compressor {MWe} is calculated in the auxiliary electrical load section above.

### **Selection of Default Parameters**

There are many adjustable parameters in the ammonia-based CO<sub>2</sub> capture module, and a set of default values were selected for these parameters, primarily based analysis presented in Chapter 2. These parameters represent the best currently available information on this technology and may be updated in future versions of the IECM as new information comes out or as specific applications of the technology are developed.

### **Ammonia System Train Size and Spares**

The default maximum train size for the ammonia-based CO<sub>2</sub> capture module is set at 1000 tons CO<sub>2</sub>/hr (907.2 tonnes CO<sub>2</sub>/hr). This is significantly larger than public designs for amine-based CO<sub>2</sub> capture systems, but train sizes have been increasing through the years. In addition, the absorber in the ammonia-based CO<sub>2</sub> capture system is envisioned as a spray tower, and spray towers for wet FGD systems have been designed at large scales. For example, a 62 foot diameter General Electric Environmental Services Inc. absorber has been built that is part of a forced oxidation limestone FGD system processing flue gas from a 700 MWe boiler at a power plant

(Kohl, 1997). The train size influences costs in a significant way, especially because each train in an ammonia system is very capital intensive. The user has the option of setting the maximum train size for the ammonia-based CO<sub>2</sub> capture module on the user input screen.

When an additional train is added not all process components related to the CO<sub>2</sub> capture system necessarily need to be duplicated as some equipment can be sized to handle the requirements of multiple trains. Table 3.6 below describes which process components are duplicated when an additional train is added. Due to the limited operating experience of this type of process at scale, the selection components that are duplicated when additional trains are added may not represent designs that are actually implemented. For specific applications, the user can adjust the train size and then also adjust the individual cost for specific components as needed in the retrofit screen for the ammonia-based CO<sub>2</sub> capture system.

**Table 3.6: Duplication of components when an additional train is added**

Process Area	Multiple Trains
DCC	Share among trains
Flue Gas Blower	Split into trains if necessary
Chiller System	Share among trains
CO <sub>2</sub> Absorber Vessel	Split into trains if necessary
Heat Exchangers	Split into trains if necessary
Circulation Pumps	Split into trains if necessary
Sorbent Regeneration	Split into trains if necessary
Ammonia Water Wash	Split into trains if necessary
Steam Extractor	Share among trains
Sorbent Processing and Reclaimer	Share among trains
Drying and Compression Unit	Split into trains if necessary
NH <sub>3</sub> Stripping	Split into trains if necessary
Auxiliary Natural Gas Boiler	Share among trains
Auxiliary Steam Turbine	Share among trains

When more than one train is required, the cost of components that are duplicated (split into trains) is adjusted according to the following equation:

Total Installed Cost {\$}

= Calculated Installed Cost For One Train {\$}

$$* \frac{(\text{No of Trains Required} + \text{No. of Spares Required})}{\text{No. Of Trains Required}^{0.6}}$$

For example, consider the costs of the CO<sub>2</sub> absorber. Ignoring the impact of multiple trains and spares, the cost of the absorber is calculated as in the previous section:

Absorber Installed Cost {\$}

$$= \text{Reference System Cost} \{ \$ \} * \left[ \frac{\text{Gas Flow Rate into Absorber} \{ \text{cum/sec} \}}{\text{Reference Gas Flow Rate} \{ \text{cum/sec} \}} \right]^{0.6}$$

If three trains and one spare absorber are required, however, the total gas flow rate is processed by three absorbers, the spare absorber is unused during regular operation, and each operating absorber processes one third of the flue gas. The total cost of the three absorbers and the spare is calculated as:

Total Absorber Installed Cost {\$}

= (3 + No. of Spares) \* Reference System Cost {\$}

$$* \left[ \frac{\frac{1}{3} \text{ Total Gas Flow Rate into Absorbers} \{ \text{cum/sec} \}}{\text{Reference Gas Flow Rate} \{ \text{cum/sec} \}} \right]^{0.6}$$

which equals:

Total Absorber Installed Cost { $\$$ }

$$= \text{Reference System Cost } \{\$\} * \left[ \frac{\text{Total Gas Flow Rate into Absorbers } \left\{ \frac{\text{cum}}{\text{sec}} \right\}}{\text{Reference Gas Flow Rate } \left\{ \frac{\text{cum}}{\text{sec}} \right\}} \right]^{0.6} \\ * \left[ \frac{3 + \text{No. of Spares}}{3^{0.6}} \right]$$

which equals:

Total Absorber Installed Cost { $\$$ }

= Calculated Installed Cost For One Train { $\$$ }

$$* \left[ \frac{\text{No of Trains Required} + \text{No. of Spares Required}}{\text{No. Of Trains Required}^{0.6}} \right]$$

This concludes the section outlining the equations for estimating performance and cost for the ammonia-based CO<sub>2</sub> capture system. These details run behind the scenes in the model and are not typically accessible to the user. The next section will outline the design of the interface that allows a user to configure the ammonia-based CO<sub>2</sub> capture module in the IECM.

### **3.7. Layout of the IECM User Screens for the Chilled Ammonia System**

#### **3.7.1. Introduction**

The capture screens in the IECM have been designed to provide the user with options for specifying the parameters of the ammonia-based CO<sub>2</sub> capture system, and the outputs provide a concise summary of the performance and costs of the components in the system. The following sections describe the inputs required from the IECM to run the model, the inputs required from the user (though there are default values preset for all the key inputs), and the resulting output

screens after the IECM has calculated all the values in the performance and costs response model.

When the Ammonia System is selected from the CO<sub>2</sub> Capture Menu in the Configure Plant / Overall Plant / Diagram Screen, the ammonia model is used as the CO<sub>2</sub> capture system. When the ammonia-based CO<sub>2</sub> capture system is selected, the ammonia model requires data from the IECM, including data from the IECM input screens that are described in the next sections.

### 3.7.2. Inputs Screens for the IECM Chilled Ammonia Model

Key inputs from the IECM Performance Model for the ammonia based CO<sub>2</sub> capture system are listed in Table 3.7 below. The units that are listed in these and subsequent sections are in mixed unit systems and are directly compatible with the regression equations. A user can select among alternative unit systems in the IECM as necessary.

**Table 3.7: Inputs required from the performance model in the IECM**

Parameter	Type	Units
Nitrogen (N <sub>2</sub> ) Into Model	IECM Module Output	mole/sec
Oxygen (O <sub>2</sub> ) Into Model	IECM Module Output	mole/sec
Water Vapor (H <sub>2</sub> O) Into Model	IECM Module Output	mole/sec
Carbon Dioxide (CO <sub>2</sub> ) Into Model	IECM Module Output	mole/sec
Carbon Monoxide (CO) Into Model	IECM Module Output	mole/sec
Hydrochloric Acid (HCl) Into Model	IECM Module Output	mole/sec
Sulfur Dioxide (SO <sub>2</sub> ) Into Model	IECM Module Output	mole/sec
Sulfuric Acid (equivalent SO <sub>3</sub> ) Into Model	IECM Module Output	mole/sec
Nitric Oxide (NO) Into Model	IECM Module Output	mole/sec
Nitrogen Dioxide (NO <sub>2</sub> ) Into Model	IECM Module Output	mole/sec
Ammonia (NH <sub>3</sub> ) Into Model	IECM Module Output	mole/sec
Argon (Ar) Into Model	IECM Module Output	mole/sec
Flue Gas Volumetric Flow Into Model	IECM Module Output	cum/sec
Flue Gas Temperature Into Model	IECM Module Output	°F
Available Cooling Water Temperature	IECM Module Output	°F
Particulate Into Model	IECM Module Output	kg/sec

### Inputs from the User on the Set Parameters / CO<sub>2</sub> Capture / Config Screen

Key inputs from the user on this screen are listed in the Table 3.8 below. Enabling Bypass on this menu leads to additional line items appearing, as shown in Table 3.9.

**Table 3.8: Inputs required from the IECM user**

Parameter	Type	Units
Sorbent Used	IECM Interface Screen	N/A
Auxiliary Natural Gas Boiler?	IECM Interface Screen	N/A
CO <sub>2</sub> Product Compressor Used?	IECM Interface Screen	N/A
Flue Gas Bypass Control	IECM Interface Screen	N/A

**Table 3.9: Additional inputs required from the IECM user when bypass is enabled**

Parameter	Type	Units
Maximum CO <sub>2</sub> Removal Efficiency	IECM Interface Screen	%
Overall CO <sub>2</sub> Removal Efficiency	IECM Interface Screen	%
Absorber CO <sub>2</sub> Removal Efficiency	IECM Interface Screen	%
Minimum Bypass	IECM Interface Screen	%
Allowable Bypass	IECM Interface Screen	%
Actual Bypass	IECM Interface Screen	%

### Inputs from the User on the Set Parameters / CO<sub>2</sub> Capture / Performance

Key inputs from the user on this screen are listed in Table 3.10 below.

**Table 3.10: Inputs required from the IECM user**

Parameter	Type	Units
Maximum CO <sub>2</sub> Removal Efficiency	IECM Interface Screen	%
Scrubber CO <sub>2</sub> Removal Efficiency	IECM Interface Screen	%
SO <sub>2</sub> Removal Efficiency	IECM Interface Screen	%
SO <sub>3</sub> Removal Efficiency	IECM Interface Screen	%
NO <sub>2</sub> Removal Efficiency	IECM Interface Screen	%
HCl Removal Efficiency	IECM Interface Screen	%
Particulate Removal Efficiency	IECM Interface Screen	%
Mercury Removal from CO <sub>2</sub> Absorber	IECM Interface Screen	%
Maximum Train CO <sub>2</sub> Capacity	IECM Interface Screen	tons/hr
Number of Operating Absorbers	IECM Interface Screen	integer
Number of Spare Absorbers	IECM Interface Screen	integer
Max CO <sub>2</sub> Compressor Capacity	IECM Interface Screen	tons/hr
No. of Operating CO <sub>2</sub> Compressors	IECM Interface Screen	integer
No. of Spare CO <sub>2</sub> Compressors	IECM Interface Screen	integer
NH <sub>3</sub> Scrubber Power Requirement	IECM Interface Screen	% MWg

### Inputs from the User on the Set Parameters / CO<sub>2</sub> Capture / Capture Screen

Key inputs from the user on this screen are listed in Table 3.11 below.

**Table 3.11: Inputs required from the IECM user**

Parameter	Type	Units
Ammonia Concentration	IECM Interface Screen	wt %
Overall Ammonia Slip	IECM Interface Screen	ppmv
Absorber NH <sub>3</sub> Slip	IECM Interface Screen	ppmv
Circulating Water Flow Rate	IECM Interface Screen	lb/sec
Gas Phase Pressure Drop	IECM Interface Screen	Psia
ID Fan Efficiency	IECM Interface Screen	%
Capture System Cooling Duty	IECM Interface Screen	t H <sub>2</sub> O/t CO <sub>2</sub>
Percent Cooling Supply by Chillers	IECM Interface Screen	%
Power Requirement by Chillers	IECM Interface Screen	kW/ton refrig.
Regen. Heat Requirement	IECM Interface Screen	Btu/lb CO <sub>2</sub>
Regen. Steam Heat Content	IECM Interface Screen	Btu/lb steam
Heat-to-Electricity Efficiency	IECM Interface Screen	%
Pump Efficiency	IECM Interface Screen	%
Percent Solids in Reclaimer Waste	IECM Interface Screen	%

### Inputs from the User on the Set Parameters / CO<sub>2</sub> Capture / CO<sub>2</sub> Storage Screen

Key inputs from the user on this screen are listed in Table 3.12 below.

**Table 3.12: Inputs required from the IECM user**

Parameter	Type	Units
CO <sub>2</sub> Product Pressure	IECM Interface Screen	Psig
Captured CO <sub>2</sub> Purity	IECM Interface Screen	vol %
H <sub>2</sub> O Content	IECM Interface Screen	vol %
Other Content	IECM Interface Screen	vol %
CO <sub>2</sub> Compressor Efficiency	IECM Interface Screen	%
CO <sub>2</sub> Unit Compression Energy	IECM Interface Screen	kWh/ton CO <sub>2</sub>
CO <sub>2</sub> Storage Method:	IECM Interface Screen	N/A

### Inputs from the User on the Set Parameters / CO<sub>2</sub> Capture / Retrofit Cost

Key inputs from the user on this screen are listed in the Table 3.13 below.

**Table 3.13: Inputs required from the IECM user**

Parameter	Type	Units
Direct Contact Coolers	IECM Interface Screen	
Flue Gas Blower	IECM Interface Screen	retro \$/new \$
Chiller System	IECM Interface Screen	retro \$/new \$
CO <sub>2</sub> Absorber Vessel	IECM Interface Screen	retro \$/new \$
Heat Exchangers	IECM Interface Screen	retro \$/new \$
Circulation Pumps	IECM Interface Screen	retro \$/new \$
Sorbent Regeneration	IECM Interface Screen	retro \$/new \$
Ammonia Water Wash	IECM Interface Screen	retro \$/new \$
Steam Extractor	IECM Interface Screen	retro \$/new \$
Sorbent Processing and Reclaimer	IECM Interface Screen	retro \$/new \$
CO <sub>2</sub> Drying and Compression Unit	IECM Interface Screen	retro \$/new \$
NH <sub>3</sub> Stripping	IECM Interface Screen	retro \$/new \$
Auxiliary Natural Gas Boiler	IECM Interface Screen	retro \$/new \$
Auxiliary Steam Turbine	IECM Interface Screen	retro \$/new \$

### Inputs from the User on the Set Parameters / CO<sub>2</sub> Capture / Capital Cost Screen

Key inputs from the user on this screen are listed in the Table 3.14 below.

**Table 3.14: Inputs required from the IECM user**

Parameter	Type	Units
Construction Time	IECM User Input	Years
General Facilities Capital	IECM User Input	%PFC
Engineering & Home Office Fees	IECM User Input	%PFC
Project Contingency Cost	IECM User Input	%PFC
Process Contingency Cost	IECM User Input	%PFC
Royalty Fees	IECM User Input	%PFC
Months of Fixed O&M	IECM User Input	Months
Months of Variable O&M	IECM User Input	Months
Misc. Capital Cost	IECM User Input	%TPI
Inventory Capital	IECM User Input	%TPC
TCR Recovery Factor	IECM User Input	%

### Inputs from the User on the Set Parameters / CO<sub>2</sub> Capture / O&M Cost Screen

Key inputs from the user on this screen are listed in the Table 3.15 below.

**Table 3.15: Inputs required from the IECM user**

Parameter	Type	Units
Ammonia Cost	IECM User Input	\$/ton
Water Cost	IECM User Input	\$/kgal
Auxiliary CCS Cooling Cost	IECM User Input	\$/ton cool H <sub>2</sub> O
Reclaimer Waste Disposal Cost	IECM User Input	\$/ton
Electricity Price (Base Plant)	IECM User Input	\$/MWh
Number of Operating Jobs	IECM User Input	jobs/shift
Number of Operating Shifts	IECM User Input	shifts/day
Operating Labor Rate	IECM User Input	\$/hr
Total Maintenance Cost	IECM User Input	%TPC
Maint. Cost Allocated to Labor	IECM User Input	% total
Administrative & Support Cost	IECM User Input	% total labor
CO <sub>2</sub> Transportation Cost	IECM User Input	\$/ton
CO <sub>2</sub> Storage Cost	IECM User Input	\$/ton

### 3.7.3. Output Screens for the IECM Chilled Ammonia Model

#### Mass Balance Outputs from the Ammonia Model

The mass balance outputs from the ammonia model are described below in Table 3.16.

**Table 3.16: Mass balance outputs required from the performance model in the IECM**

Parameter	Type	Units
Nitrogen (N <sub>2</sub> ) To Stack	NH <sub>3</sub> Module Output	mole/sec
Oxygen (O <sub>2</sub> ) To Stack	NH <sub>3</sub> Module Output	mole/sec
Water Vapor (H <sub>2</sub> O) To Stack	NH <sub>3</sub> Module Output	mole/sec
Carbon Dioxide (CO <sub>2</sub> ) To Stack	NH <sub>3</sub> Module Output	mole/sec
Carbon Monoxide (CO) To Stack	NH <sub>3</sub> Module Output	mole/sec
Hydrochloric Acid (HCl) To Stack	NH <sub>3</sub> Module Output	mole/sec
Sulfur Dioxide (SO <sub>2</sub> ) To Stack	NH <sub>3</sub> Module Output	mole/sec
Sulfuric Acid (equivalent SO <sub>3</sub> ) To Stack	NH <sub>3</sub> Module Output	mole/sec
Nitric Oxide (NO) To Stack	NH <sub>3</sub> Module Output	mole/sec
Nitrogen Dioxide (NO <sub>2</sub> ) To Stack	NH <sub>3</sub> Module Output	mole/sec
Ammonia (NH <sub>3</sub> ) To Stack	NH <sub>3</sub> Module Output	mole/sec
Argon (Ar) To Stack	NH <sub>3</sub> Module Output	mole/sec
Particulate To Stack	NH <sub>3</sub> Module Output	ton/hr
Water Bleed from Flue Gas Cooling System	NH <sub>3</sub> Module Output	kg/sec
CO <sub>2</sub> Flow to Compressor	NH <sub>3</sub> Module Output	mole/sec
Cooling Water Flow	NH <sub>3</sub> Module Output	kg/sec
Chilled Water Flow	NH <sub>3</sub> Module Output	kg/sec
Steam Flow to Heater (if necessary)	NH <sub>3</sub> Module Output	kg/sec
Steam Flow to NH <sub>3</sub> Cleanup System	NH <sub>3</sub> Module Output	kg/sec
Steam Flow to CO <sub>2</sub> Regeneration System	NH <sub>3</sub> Module Output	kg/sec

Note: Nitrogen (N<sub>2</sub>), Oxygen (O<sub>2</sub>), and Argon (Ar) pass through the CO<sub>2</sub> capture system unchanged in this version of the model.

### Energy Usages of the Ammonia System Equipment

The energy balance outputs from the ammonia model are described below in Table 3.17. The IECM interface calculates all these parameters, though in some cases the user sees an aggregate of several values.

**Table 3.17: Energy usages of the ammonia system equipment**

Parameter	Type	Units
Flue Gas Blower	NH <sub>3</sub> Module Output	MWe
DCC1 Pumps	NH <sub>3</sub> Module Output	MWe
DCC2 Pumps	NH <sub>3</sub> Module Output	MWe
Flue Gas Cooler Water Circulation Pumps	NH <sub>3</sub> Module Output	MWe
Absorber Cooling Pumps	NH <sub>3</sub> Module Output	MWe
Ammonia Cleanup Pumps	NH <sub>3</sub> Module Output	MWe
Drying and Compress Unit	NH <sub>3</sub> Module Output	MWe
High Pressure Pump	NH <sub>3</sub> Module Output	MWe
Abs Cooling (water)	NH <sub>3</sub> Module Output	Btu/hr
Lean Solution Cooling (water)	NH <sub>3</sub> Module Output	Btu/hr
DCC2 Chilling Load (water)	NH <sub>3</sub> Module Output	Btu/hr
CO <sub>2</sub> Flash Cooling (water)	NH <sub>3</sub> Module Output	Btu/hr
Heater (steam)	NH <sub>3</sub> Module Output	Btu/hr
CO <sub>2</sub> Stripper (steam)	NH <sub>3</sub> Module Output	Btu/hr
NH <sub>3</sub> Stripper (steam)	NH <sub>3</sub> Module Output	Btu/hr

### Capital Costs of the Ammonia System Equipment

The resulting capital cost outputs from the ammonia model are described below in Table 3.18. The IECM interface calculates all these parameters, though in some cases the user sees an aggregate of several values, for example all the heat exchanger costs are combined on the IECM Results screen.

**Table 3.18: Capital costs of the ammonia system equipment**

Parameter	Type	Units
Direct Contact Cooler 1	NH <sub>3</sub> Module Output	\$
Flue Gas Blower	NH <sub>3</sub> Module Output	\$
Heat Exchanger 1	NH <sub>3</sub> Module Output	\$
Heat Exchanger 2	NH <sub>3</sub> Module Output	\$
Direct Contact Cooler 2 Equipment	NH <sub>3</sub> Module Output	\$
DCC1 Pumps (Also Called Heat X1)	NH <sub>3</sub> Module Output	\$
DCC2 Pumps (Also Called Heat X2)	NH <sub>3</sub> Module Output	\$
Flue Gas Cooler Water Circulation Pumps	NH <sub>3</sub> Module Output	\$
Flue Gas Cooling - Chiller System	NH <sub>3</sub> Module Output	\$
CO <sub>2</sub> Spray Tower Absorber Vessel	NH <sub>3</sub> Module Output	\$
Absorber Cooling Pumps	NH <sub>3</sub> Module Output	\$
Rich/Lean Heat Exchanger	NH <sub>3</sub> Module Output	\$
Solution Heater 1	NH <sub>3</sub> Module Output	\$
Solution Cooler	NH <sub>3</sub> Module Output	\$
High Pressure Pump	NH <sub>3</sub> Module Output	\$
CO <sub>2</sub> Capture Packed Bed Stripper	NH <sub>3</sub> Module Output	\$
Reboiler	NH <sub>3</sub> Module Output	\$
Ammonia Cleanup Water Wash Unit	NH <sub>3</sub> Module Output	\$
Ammonia Cleanup Heat Exchanger	NH <sub>3</sub> Module Output	\$
Ammonia Cleanup Stripper	NH <sub>3</sub> Module Output	\$
Ammonia Cleanup Pumps	NH <sub>3</sub> Module Output	\$
Steam Extractor	NH <sub>3</sub> Module Output	\$
Sorbent Reclaimer	NH <sub>3</sub> Module Output	\$
Sorbent Processing	NH <sub>3</sub> Module Output	\$
Drying and Compress Unit	NH <sub>3</sub> Module Output	\$

### O&M Costs of the Ammonia System Equipment

The resulting O&M cost outputs from the ammonia model are described below in Table 3.19.

Fixed costs are also calculated as a fraction of other costs, as specified by the user.

**Table 3.19: Operating and maintenance costs of the ammonia system equipment**

Parameter	Type	Units
Ammonia	NH <sub>3</sub> Module Output	\$/Year
Natural Gas	NH <sub>3</sub> Module Output	\$/Year
Reclaimer Waste Disposal	NH <sub>3</sub> Module Output	\$/Year
Electricity	NH <sub>3</sub> Module Output	\$/Year
Auxiliary Power Credit	NH <sub>3</sub> Module Output	\$/Year
Water	NH <sub>3</sub> Module Output	\$/Year
CO <sub>2</sub> Transport	NH <sub>3</sub> Module Output	\$/Year
CO <sub>2</sub> Storage	NH <sub>3</sub> Module Output	\$/Year

### 3.8. Case Studies of PC and NGCC Plants with Chilled Ammonia-Based Capture

#### 3.8.1. Introduction to Case Studies

This section presents three case studies for power plants with ammonia-based post-combustion CO<sub>2</sub> capture systems: a subcritical PC plant (Case 4), a supercritical PC plant (Case 5), and an NGCC plant (Case 6). For these case studies, the IECM models for the ammonia-based post-combustion CO<sub>2</sub> capture system that were developed in this chapter were used along with the baseline plant assumptions in Table 2.1 to develop performance and cost estimates for these plants. Complete details for reproducing these results using the IECM are given in Appendix A.

#### 3.8.2. Summary Results from Case Studies 4-6

A summary of the performance results for the plants is given below in Table 3.20. The NGCC plant has the highest net efficiency, lowest CO<sub>2</sub> emissions, and lowest revenue required of the three plants with ammonia-based CO<sub>2</sub> capture. Of the pulverized coal plants, the supercritical plant has a higher net plant efficiency, lower CO<sub>2</sub> emissions, and a lower revenue requirement than the subcritical plant.

**Table 3.20: Summary of case study results for power plants with ammonia-based CO<sub>2</sub> capture**

Cases	Gross Plant Output (MWe)	Net Plant Output (MWe)	Net Plant Efficiency (%)	CO <sub>2</sub> Emissions (million tonnes/year)	Capital Cost (\$/kW -net)	Revenue Required (\$/MWh)
Case 4: PC Subcritical	723.9	550.0	25.5%	0.46	3049	109.9
Case 5: PC Supercritical	711.2	550.0	27.7%	0.41	3058	107.6
Case 6: NGCC	486.9	407.1	38.3%	0.13	1785	109.3

### 3.9. Conclusion

This chapter has outlined work done to integrate an ammonia-based post-combustion CO<sub>2</sub> capture into the IECM with updated modules for this technology. A general response surface model of an ammonia-based CO<sub>2</sub> capture system was presented using a dataset created using the

Aspen Plus® model developed in Chapter 2. The response surface model consisted of a large number of equations which were implemented as a module in the IECM. The resulting model in the IECM can use a large number of particular financial, technological and site specific assumptions, and using those assumptions the model will estimate the performance and costs for power plants with ammonia-based CO<sub>2</sub> capture. The resulting tool can provide flexible power plant performance and cost estimates, and could be an important resource for policy makers.

## **Chapter 4. An Advanced Amine System Response Surface Model for the IECM**

### **4.1. Motivation**

The existing amine system in the IECM was originally developed in 2000-2002 (Rao, 2002), and since that time results from studies that use amine-based CO<sub>2</sub> capture have shown significant improvements in the technology. This chapter develops an advanced amine system model in the IECM to replace the older model and to reflect these process improvements.

### **4.2. Amine-Based Capture Processes**

The class of solvents called amines (more properly, alkanolamines) comprise a family of organic compounds that are derivatives of alkanols (commonly called the alcohols group) that contain an “amino” group in the chemical structure such as RNH<sub>2</sub>, R<sub>2</sub>NH or R<sub>3</sub>N. There are multiple classifications of amines, each of which has different characteristics relevant to CO<sub>2</sub> capture (Kohl and Nielsen, 1997). In response to growing interest in large scale carbon dioxide (CO<sub>2</sub>) capture, Fluor and Mitsubishi Heavy Industries (MHI) have developed commercially available advanced CO<sub>2</sub> capture systems based on solutions of aqueous amines (EPRI, 2008).

One of the most popular amines for CO<sub>2</sub> capture is monoethanolamine (MEA). Several of the most mature amine-based CO<sub>2</sub> capture systems today are either based on MEA or are often compared to MEA-based processes. MEA reacts strongly with acid gases like CO<sub>2</sub>, has a fast reaction time, and has an ability to remove high percentages of CO<sub>2</sub> from a gas stream even at the low CO<sub>2</sub> concentrations found in power plant flue gasses. Other properties of MEA are undesirable for CO<sub>2</sub> capture systems, such as its significant cost, issues with corrosivity in process equipment, and high regeneration energy requirement.

Various research groups are involved in synthesizing and testing a variety of alternate amine mixtures and designer amines to achieve a more desirable set of overall properties for use in CO<sub>2</sub> capture systems. One major focus is on lowering the energy required for solvent regeneration, which has a major impact on process costs. Often there are tradeoffs to consider however. For example, the energy required for regeneration is typically related to the driving forces for achieving high capture capacities. Therefore reducing the regeneration energy can lower the driving force and increase the amount of solvent and size of absorber needed to capture a given amount of CO<sub>2</sub>, leading to an increase in capital cost. The potential high costs of manufacturing a new solvent also may detract from its benefits. Pilot plant projects are acquiring the data needed to assess such tradeoffs for new amine formulations.

Improvements in existing MEA-based CO<sub>2</sub> capture processes are also possible. These have focused at lowering solvent losses, providing improved heat integration, and research into additives that inhibit corrosion allowing for the use of carbon steel instead of more expensive stainless steel (Roberts et al., 2004). Fluor's most recent offering is the Econamine® FG+ process, which uses an aqueous mixture of monoethanolamine (MEA) and proprietary corrosion inhibitors (DOE/NETL, 2007a).

While many of the improvements in MEA-based processes are proprietary, the MEA-based CO<sub>2</sub> capture system incorporated into the IECM can be modified to match the information publicly available on the performance and cost of these systems. The original IECM capture model is based on an MEA capture process that was representative of the state of the technology in 2002. In the sections that follow, a new CO<sub>2</sub> capture model based on the Fluor's Econamine® FG+ process is developed, and this model has been integrated into the IECM. For this new model, the performance of the original MEA model was adjusted to reflect process improvements in amine-

based CO<sub>2</sub> capture, and the corresponding costs have also been updated to reflect technology improvements and the current pricing environment.

The sections that follow describes parameter updates made to the original MEA based CO<sub>2</sub> capture system that were applied to create this new advanced amine-based CO<sub>2</sub> capture system. These sections include specific details on model adjustments made for the advanced amine-based CO<sub>2</sub> capture system, menu updates for the capture models in the IECM in general, and a comparison between the original MEA and the advanced amine-based CO<sub>2</sub> capture system.

### **4.3. Parameters for the Advanced Amine-Based CO<sub>2</sub> Capture System**

This section describes default parameter modifications to the 2002 IECM MEA-based CO<sub>2</sub> capture model (Rao, 2002) that were made to represent the advanced MEA-based CO<sub>2</sub> capture processes in the DOE/NETL August 2007 Baseline Report, which is based on Fluor's Econamine® FG+ CO<sub>2</sub> capture system (DOE/NETL, 2007a). In the cases where the DOE/NETL August 2007 Baseline Report does not specify values for needed parameters, information was used from similar reports which included advanced amine-based CO<sub>2</sub> capture processes, or was transferred from the original MEA-based CO<sub>2</sub> capture model. The original and updated performance and cost parameters are shown in the Table 4.1 to Table 4.7 below, with the original parameters listed under IECM 5.2, and the updated parameters listed under IECM 6.2. These defaults are specific to when a new supercritical PC plant is chosen that has been configured with a Hot-Side SCR, Cost-Side ESP, Wet FGD (the default configuration for Case 2 in this thesis) and an Amine System with Advanced Amine (FG+) selected as the solvent. The defaults for other plant configurations are similar, with minor variations. For example, when an NGCC plant is chosen the IECM 6.2 default parameter for "SO<sub>2</sub> polisher used?" in Table 4.1 is not available

because SO<sub>2</sub> is not a major component of the flue gas streams for NGCC plants. Each of the tables below corresponds to a specific IECM CO<sub>2</sub> capture screen, listed as the title of the table.

**Table 4.1: Set Parameters / CO<sub>2</sub> Capture / CO<sub>2</sub> Capture System Process / Config Menu**

Parameter	IECM 5.2 Default	IECM 6.2 Default	References
Sorbent Used	MEA	Adv. Amine	
Auxiliary Natural Gas Boiler	None	None	
CO <sub>2</sub> Product Compressor Used?	Yes	Yes	
Flue Gas Bypass Control	No Bypass	No Bypass	
Direct Contact Cooler (DCC) Used?	Yes	Yes	
SO <sub>2</sub> Polisher Used?	N/A	Yes	DOE/NETL, 2007a
SO <sub>2</sub> Outlet Concentration (ppmv)	N/A	10	DOE/NETL, 2007a
Temperature Exiting DDC (°C)	50	45	IECM <sup>1</sup>

<sup>1</sup>The IECM calculates updated values based on other values supplied as inputs. These values may change with different plant configurations.

**Table 4.2: Set Parameters / CO<sub>2</sub> Capture / CO<sub>2</sub> Capture System Process / Performance Menu**

Parameter	IECM 5.2 Default	IECM 6.2 Default	References
Maximum CO <sub>2</sub> Removal Efficiency (%)	90%	90%	
Scrubber Removal Efficiency (%)	90%	90%	DOE/NETL, 2007a
SO <sub>2</sub> Removal Efficiency (%)	99.50%	100%	DOE/NETL, 2007a
SO <sub>3</sub> Removal Efficiency (%)	99.50%	99.50%	
NO <sub>2</sub> Removal Efficiency (%)	25%	25%	
HCl Removal Efficiency (%)	95%	95%	
PM Removal Efficiency (%)	50%	50%	
Max Train CO <sub>2</sub> Capacity (tons/hr)	230	230	
# Absorbers	2	3	IECM <sup>1</sup>
Spare Absorbers	0	0	
Max CO <sub>2</sub> Compressor Capacity (tons/hr)	330	330	
No. of Operating CO <sub>2</sub> Compressors	2	2	
No. of Spare CO <sub>2</sub> Compressors	0	0	
Scrubber Power Requirement (%MWg)	14.00%	9.187%	IECM <sup>1</sup>

<sup>1</sup>The IECM calculates updated values based on other values supplied as inputs. These values may change with different plant configurations.

**Table 4.3: Set Parameters / CO<sub>2</sub> Capture / CO<sub>2</sub> Capture System Process / Capture Menu**

Parameter	IECM 5.2 Default	IECM 6.2 Default	References
Sorbent concentration (%)	30%	30%	DOE/NETL, 2007b
Lean CO <sub>2</sub> Loading (mol CO <sub>2</sub> /mol sorb)	0.2	0.19	DOE/NETL, 2007b
Nominal Sorbent Loss (lb / ton CO <sub>2</sub> )	3	0.6 <sup>2</sup>	DOE/NETL, 2007a
Liq/Gas Ratio (Ratio)	2.876	3.072	IECM <sup>1</sup>
Ammonia Generation (mol NH <sub>3</sub> / mol sorb.)	1	1	
Gas Phase Pressure Drop (psia)	2	1	DOE/NETL, 2007b
ID Fan Efficiency (%)	75%	75%	
Makeup Water for Wash Section (% raw flue gas)	N/A	0.8	IECM <sup>1</sup>
Regenerator Heat Requirement (Btu / lb CO <sub>2</sub> )	1975	1516	DOE/NETL, 2007a
Steam Ht. Cont (Btu/lb Steam)	860.4	1397	DOE/NETL, 2007a
Heat to Electricity Efficiency	14%	22% <sup>3</sup>	DOE/NETL, 2007a
Solvent Pumping Head	30	30	
Pump Efficiency	75%	75%	
% Solids in Reclaimer Waste	40%	40%	
Capture System Cooling Duty (ton H <sub>2</sub> O/ton CO <sub>2</sub> )	N/A	46.19	IECM <sup>1</sup>

<sup>1</sup>The IECM calculates updated values based on other values supplied as inputs. These values may change with different plant configurations. <sup>2</sup>For a more thorough explanation and for the calculation of Nominal Sorbent Loss, see below. <sup>3</sup> For the calculation of the updated Heat-to-Electricity Efficiency, see below.

**Table 4.4: Set Parameters / CO<sub>2</sub> Capture / CO<sub>2</sub> Capture System Process / CO<sub>2</sub> Storage Menu**

Parameter	IECM 5.2 Default	IECM 6.2 Default	References
CO <sub>2</sub> Product Pressure (psig)	2000	2000	
CO <sub>2</sub> Compressor Efficiency (%)	80%	80%	
CO <sub>2</sub> Unit Compression Energy (kWh/ton CO <sub>2</sub> )	107	107	

**Table 4.5: Set Parameters / CO<sub>2</sub> Capture / CO<sub>2</sub> Capture System Process / Retrofit Cost Menu**

Parameter	IECM 5.2 Default <sup>1</sup>	IECM 6.2 Default <sup>1</sup>	References
SO <sub>2</sub> Polisher/ Direct Contact Cooler	1	1	
Flue Gas Blower	1	1	
CO <sub>2</sub> Absorber Vessel	1	1	
Heat Exchangers	1	1	
Circulation Pumps	1	1	
Sorbent Regenerator	1	1	
Reboiler	1	1	
Steam Extractor	1	1	
Sorbent Reclaimer	1	1	
Sorbent Processing	1	1	
CO <sub>2</sub> Drying and Compression Unit	1	1	
Auxiliary Natural Gas Boiler	1	1	
Auxiliary Steam Turbine	1	1	

<sup>1</sup>All Units in retro \$/new \$

**Table 4.6: Set Parameters CO<sub>2</sub> Capture / CO<sub>2</sub> Capture System Process / Capital Cost Menu**

Parameter	IECM 5.2 Default	IECM 6.2 Default	References
Construction Time (years)	3	3	
General Facilities Capital (%PFC)	10	10	
Engineering & Home Office Fees (%PFC)	7	7	
Project Contingency Cost (%PFC)	15	15	
Process Contingency Cost (%PFC)	5	5	
Royalty Fees (%PFC)	0.5	0.5	
Months of Fixed O&M (months)	1	1	
Months of Variable O&M (months)	1	1	
Misc. Capital Cost (%TPI)	2	2	
Inventory Capital (%TPC)	0.5	0.5	
TCR Recovery Factor (%)	100	100	

**Table 4.7: Set Parameters / CO<sub>2</sub> Capture / CO<sub>2</sub> Capture System Process / O&M Cost Menu**

Parameter	IECM 5.2 Default	IECM 6.2 Default	References
Sorbent Cost (\$/ton)	1379	2142	DOE/NETL, 2007a
Inhibitor Cost (% of MEA)	20	0	DOE/NETL, 2007a
Activated Carbon Cost (\$/ton)	1411	2000	DOE/NETL, 2007a
Caustic (NaOH) Cost (\$/ton)	666.6	413	DOE/NETL, 2007a
Water Cost (\$/kgal)	0.8874	1.03	DOE/NETL, 2007a
Reclaimer Waste Disposal Cost (\$/ton)	201.2	211.6	
Electricity Price (Base Plant)	43.27	59.97	
Number of Operating Jobs (jobs/shift)	2	2	
Number of Operating Shifts (shifts/day)	4.75	4.75	
Operating Labor Rate (\$/hr)	24.82	33	DOE/NETL, 2007a
Total Maintenance Cost (%TPC)	2.5	2.5	
Maint. Cost Allocated to Labor (% total)	40	40	
Administrative & Support Cost (% total labor)	30	30	
CO <sub>2</sub> Transportation Cost (\$/ton)	2.4	2.38	IECM <sup>1</sup>
CO <sub>2</sub> Storage Cost (\$/ton)	5.75	6.047	IECM <sup>1</sup>

<sup>1</sup>The IECM calculates updated values based on other values supplied as inputs. These values may change with different plant configurations.

#### 4.4. Model Adjustments for the Advanced Amine-Based CO<sub>2</sub> Capture System

This section outlines a number of model adjustments that have been made to the amine-based CO<sub>2</sub> capture system. These include adjustments to heat integration, the impact of removing steam from the power plant steam cycle on the turbine power output (the heat-to-electricity conversion efficiency), the electrical use of the CO<sub>2</sub> capture system, the base plant costs equations, and the amine usage of the CO<sub>2</sub> capture system. For extended descriptions of the original models or of terminology, the reader can refer to the existing IECM documentation (IECM V6.2, IECM V7.0).

### The Modified Heat Integration Equation

In order to reflect the new regeneration heat requirement of the Advanced Amine CO<sub>2</sub> capture process, the original regression equation was multiplied by a scaling factor. In the original model, regeneration heat is calculated based on the following regression equation:

$$\begin{aligned} \text{Regen Heat } \left\{ \frac{\text{Btu}}{\text{lb CO}_2} \right\} &= \left[ \frac{\text{Sorbent Circulation } \left\{ \frac{\text{tons}}{\text{hr}} \right\} * 429.9046}{\text{Sorbent Molecular Weight } \left\{ \frac{\text{lb}}{\text{lb mole}} \right\} * \text{CO}_2 \text{ Captured } \left\{ \frac{\text{tons}}{\text{hr}} \right\}} \right] \\ &* \exp \left[ 2.5919 + 0.0259 * \text{Reagent Concentration } \{ \% \} - 6.3536 \right] \\ &* \text{Lean CO}_2 \text{ Loading } \left\{ \frac{\text{mol CO}_2}{\text{mol sorb}} \right\} - 0.0015 \\ &* \text{Actual CO}_2 \text{ Removal Efficiency } \{ \% \} - 0.0059 * \text{CO}_2 \text{ Flue Gas } \left\{ \frac{\text{lb moles}}{\text{hr}} \right\} \\ &* \left. \frac{100}{\text{Total Gas } \left\{ \frac{\text{lb moles}}{\text{hr}} \right\}} \right] \end{aligned}$$

This regression equation was multiplied by a scaling factor of 0.7639 to approximately match the regeneration energy of the advanced amine system of 1516 Btu/ lb CO<sub>2</sub> (DOE/NETL, 2007a), from 1984 Btu/ lb CO<sub>2</sub> for conventional MEA systems (Rao, 2002). There are multiple other ways that this equation could have been modified to give the same result. However, directly scaling the equation leaves the functional relationships between the variables in the equation intact which would be qualitatively representative of a CO<sub>2</sub> capture system with improved heat integration. For example, increases in solvent circulation or decreases in lean solvent CO<sub>2</sub>

loading still increases the regeneration energy requirement for the advanced amine system, though the impact is smaller than for the conventional MEA system. With this change, some of the effects are as follows: (1) the advanced CO<sub>2</sub> capture system uses less steam in the reboiler than the conventional MEA system. (2) For a given gross plant size, this in turn leads to less steam use from the steam cycle and therefore a smaller and less costly power plant.

### **Modified Heat-to-Electricity Conversion Efficiency Equation**

The effect on the turbine power output of removing steam from the power plant steam cycle and redirecting it to the CO<sub>2</sub> capture system is changed by the heat-to-electricity conversion efficiency of the steam cycle. To put this another way, the heat-to-electricity conversion efficiency is a variable that links the steam used by the CO<sub>2</sub> capture system to a reduction in steam turbine output. A higher heat-to-electricity conversion efficiency implies that the steam drawn off the turbine for the CO<sub>2</sub> capture system more negatively impacts the power produced by the turbine, and this would typically corresponds to a higher quality steam being drawn off the steam cycle and used in the CO<sub>2</sub> capture process.

In previous versions of the IECM, the Heat-to-Electricity Conversion Efficiency (or equivalence factor) was selected from a range of values in the literature as 14%. For the new advanced amine system, the heat-to-electricity conversion efficiency was estimated from data obtained using the NETL 2007 Baseline report (DOE/NETL, 2007a). In the NETL Baseline report, the following data is available for a subcritical plant without CO<sub>2</sub> Capture: For a subcritical plant without CO<sub>2</sub> capture and a gross plant size of 583 MWe, the coal flow rate is 437,699 lb coal/hr. Therefore, approximately 750 lb coal/hr is burned for each gross MWe produced for this plant. For a subcritical plant with CO<sub>2</sub> capture, the gross plant size is 680 MWe and the coal flow rate is 646,589 lb coal/hr. Based on the estimate of 750 lb coal/hr per gross MWe produced from the

plant without CO<sub>2</sub> capture and the 646,589 lb coal/hr flow rate of coal for the plant with CO<sub>2</sub> capture, the plant with CO<sub>2</sub> capture should produce approximately 862 MWe gross absent the CO<sub>2</sub> capture system. Therefore, 862 MWe - 680 MWe = approximately 182 MWe gross was lost due to steam drawn off the steam cycle for the CO<sub>2</sub> capture system.

For the subcritical plant with the CO<sub>2</sub> capture system, 1,995,300 lb steam/hr of quality 1397.7 Btu/lb is diverted from the low pressure turbine and this steam flows to the reboiler of the advanced amine-based CO<sub>2</sub> capture unit. Therefore the Heat-to-Electricity Conversion Efficiency can be calculated as:

Heat – to – electricity Conversion Efficiency

$$= \frac{182\{\text{MW}\}}{1995300 \left\{ \frac{\text{lb steam}}{\text{hr}} \right\} * 1397.7 \left\{ \frac{\text{Btu}}{\text{lb steam}} \right\} * 2.93\text{E}^{-7} \left\{ \frac{\text{MW}}{\text{Btu/hr}} \right\}}$$

$$= 0.222 \text{ or } 22\%,$$

Where  $2.93\text{E}^{-7} \{ \text{MW}/(\text{Btu/hr}) \}$  equals the MW equivalent of 1 Btu/hr.

Similar results are shown using the NETL Baseline Report for supercritical plants. For the advanced amine-based CO<sub>2</sub> capture system, the heat-to-electricity conversion efficiency as well as the steam quality parameters were updated for both the MEA and Advanced Amine CO<sub>2</sub> capture systems to reflect these results.

### **Modified Equivalent Electrical Loss Due to the Capture System**

In previous versions of the IECM, the electrical equivalent loss on the power plant from the conventional CO<sub>2</sub> capture system was based on multiplying the regeneration energy requirement by the heat-to-electricity conversion efficiency. However, with the advanced amine-based CO<sub>2</sub>

capture system this calculation was modified to reflect a more realistic process. The electrical equivalent loss (MW) due to the steam requirements in the reboiler are calculated in two steps, first by calculating the steam flow rate directly, and then by using this value and the heat-to-electricity conversion efficiency to calculate the electrical equivalent loss:

The steam flow rate is calculated as:

$$\dot{M}_{\text{Steam Flow}} = \frac{E_{\text{Regeneration}} * \text{CO}_2\text{Flow}}{\text{Enthalpy}_{\text{Steam Inlet}} - \text{Enthalpy}_{\text{Steam Condensate}}}$$

Where:

$$\dot{M}_{\text{Steam Flow}} = \text{Steam Flow Rate \{lb/hr\}}$$

$$E_{\text{Regeneration}} = \text{Regeneration Heat Energy \{Btu/ lb CO}_2\}$$

$$\text{CO}_2\text{Flow} = \text{Flow Rate of CO}_2 \text{ Captured \{lb CO}_2\text{/hr\}}$$

$$\text{Enthalpy}_{\text{Steam Inlet}} = \text{Enthalpy of Steam at Reboiler Inlet \{Btu/lb steam\}} = 1373$$

$$\text{Enthalpy}_{\text{Steam Outlet}} = \text{Enthalpy of Steam at Reboiler Outlet \{Btu/lb steam\}} = 319.5$$

The equivalent electrical loss (MWeq.) is calculated as:

$$\text{MW}_{\text{Eq.}} = (\text{Heat - to - Electricity Efficiency}) * \dot{M}_{\text{Steam Flow}} * \text{Enthalpy}_{\text{Steam Inlet}} * 2.97\text{E}^{-7}$$

Where:

$$\text{MW}_{\text{Eq.}} = \text{Equivalent Steam Usage \{MW\}}$$

$$\text{Heat - to - Electricity Efficiency} = \text{Energy Conversion Efficiency \{dimensionless\}}$$

$$\text{CO}_2\text{Flow} = \text{Flow Rate of CO}_2 \text{ Captured \{lb CO}_2\text{/hr\}}$$

$$\text{Enthalpy}_{\text{Steam Inlet}} = \text{Enthalpy of Steam at Reboiler Inlet \{Btu/lb steam\}}$$

$$2.97\text{E}^{-7} = \text{Conversion Factor for Btu to MW \{MW/[Btu/hr\}}$$

### **Modified Base Plant Cost Equations for the Case with CO<sub>2</sub> Capture**

In past versions of the IECM, the costs of the equipment in the base plant depended on the gross power produced (MWe) by the power plant. With the new advanced amine system, for cases with CO<sub>2</sub> capture, the majority of equipment in the base plant is instead sized on the gross power produced (MWe) plus the equivalent electrical loss (MWeq.), which is called the “gross power produced effective” internally in the IECM. This new calculation accounts for the steam produced by the base plant that is not used to generate electricity but is instead used to regenerate amine in the CO<sub>2</sub> capture system. The only exception is the turbine island, which is still sized based on the gross power produced. The turbine island does not need be sized based on the equivalent electrical loss because it is not part of the set of equipment that produces steam. This adjustment more realistically captures the cost equations of the base plant equipment for cases with CO<sub>2</sub> capture.

### **Modified Amine Usage for the Advanced Amine-Based CO<sub>2</sub> Capture System**

For the advanced amine-based CO<sub>2</sub> capture system, the nominal sorbent loss (lb/ton CO<sub>2</sub>) represents the amount of sorbent that has been lost due to unwanted polymerization and oxidation reactions. The total sorbent loss, which requires makeup sorbent (called Sorbent (lb/hr) on the Get Results / CO<sub>2</sub> Capture / Diagram screen), now represents the nominal sorbent loss (lb/ton CO<sub>2</sub>) minus the amount of CO<sub>2</sub> regenerated in the reclaimer.

#### **4.5. Layout of the IECM User Screens for the Amine System**

The IECM CO<sub>2</sub> Capture screens have been updated to include the changes listed in the previous sections, and the critical changes are highlighted below.

## Modified CO<sub>2</sub> Capture / Config Menu

The new advanced amine process has been included in the Set Parameters / CO<sub>2</sub> Capture / Config Menu, and the user can now select between traditional MEA and Advanced Amine (FG+) processes. An additional update to the Set Parameters / CO<sub>2</sub> Capture / Config Menu is the inclusion of a polishing unit intended to lower the SO<sub>2</sub> concentration in the feed gas. SO<sub>2</sub> can react with MEA to produce undesirable and irreversible byproducts, therefore causing a loss of amine in the system that requires makeup (DOE/NETL, 2007a). The polishing unit uses caustic to capture the SO<sub>2</sub>, reducing the cost of makeup chemicals. The user can choose whether or not to include the polishing unit as necessary. The modified Set Parameters / CO<sub>2</sub> Capture / Config Menu is shown in Figure 4.1.

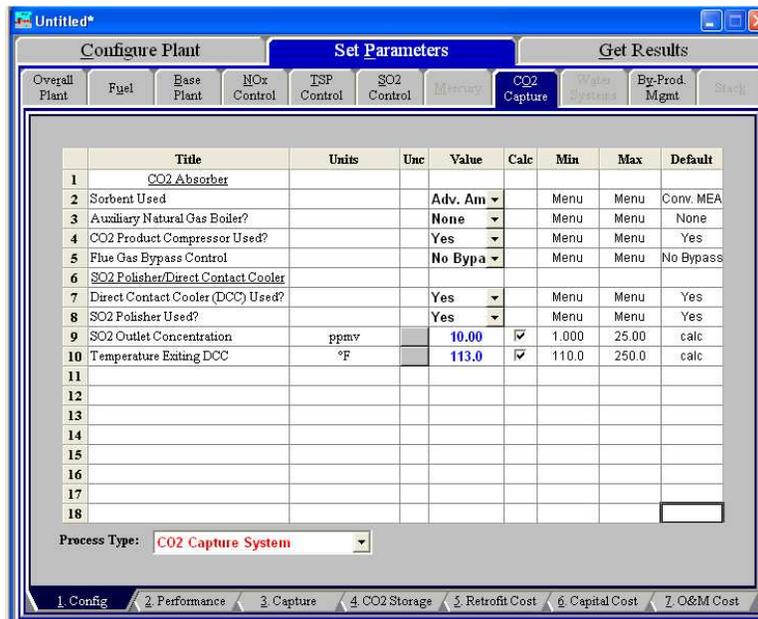


Figure 4.1: Modified CO<sub>2</sub> Capture / Config Menu

## Modified CO<sub>2</sub> Capture / Capture Menu

The CO<sub>2</sub> Capture menu for the advanced amine system (Figure 4.2) has been modified to show only the nominal sorbent loss. The sorbent oxidation loss line has been collapsed into nominal

sorbent loss to reflect the lack of detailed information on this parameter available for advanced amine capture systems.



Figure 4.2: Modified CO<sub>2</sub> Capture / Capture Menu

#### 4.6. Comparison of Plants with Conventional vs. Advanced Amine-Based Capture

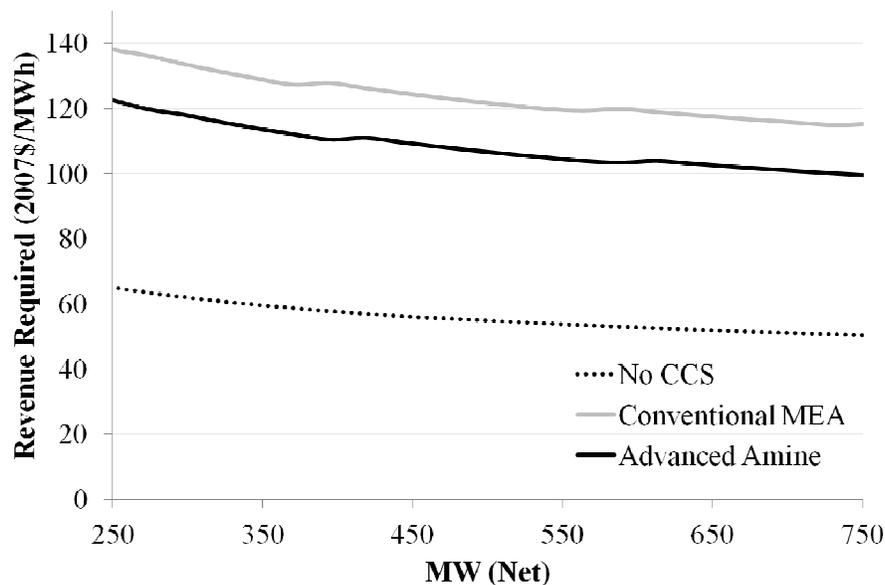
A comparison of supercritical coal-fired power plant with a conventional MEA CO<sub>2</sub> capture system and a similar power plant with the new advanced amine-based CO<sub>2</sub> capture system is shown below in Table 4.8. Both plants have the standard suite of environmental controls, similar to Case 2 as explained early in this document, and both plants follow the assumptions outlined in Table 2.1.

Table 4.8: Comparison of power plants with conventional and advanced amine-based CO<sub>2</sub> capture systems

	Net Plant	Net Plant	CO <sub>2</sub>	Capital Cost	Revenue
--	-----------	-----------	-----------------	--------------	---------

	Output (MWe)	Efficiency (%)	Emissions (million tonnes/year)	(\$/kW -net)	Required (\$/MWh)
Case 2: PC Supercritical	550.0	39.1%	2.9	1608	53.7
PC Supercritical+Conv.MEA	550.0	24.7%	0.45	3161	124.2
PC Supercritical+Advanced Ammonia	550.0	27.1%	0.42	2857	104.5

As shown in Table 4.8, the plant with the advanced amine-based CO<sub>2</sub> capture system produces the same amount of power as the plant with the conventional MEA-based CO<sub>2</sub> capture system, but also has a higher net plant efficiency, lower CO<sub>2</sub> emissions due to a reduced coal flow rate, and a lower revenue requirement. The analysis in Table 4.8 was also extended for various power plant sizes using the updated models in the IECM, as shown in Figure 4.3. As shown in the figure, the power plant with the advanced amine-based CO<sub>2</sub> capture system has a lower revenue required than the plant with the conventional MEA system. In producing Figure 4.3, the default train size for both CO<sub>2</sub> capture systems is kept constant at 208.7 tonnes of CO<sub>2</sub>/hr.



**Figure 4.3: Net plant efficiency by power plant and coal type for plants with conventional MEA and advanced amine CO<sub>2</sub> capture systems**

#### 4.7. Case Studies of PC and NGCC Plants with Advanced Amine-Based Capture

#### 4.7.1. Introduction to the Case Studies

This section presents three case studies for power plants with advanced amine-based post-combustion CO<sub>2</sub> capture systems: a subcritical PC plant (Case 7), a supercritical PC plant (Case 8), and an NGCC plant (Case 9). For these case studies, the models for the advanced amine-based post-combustion CO<sub>2</sub> capture system that were developed in this chapter were used along with the baseline plant assumptions in Table 2.1 and the IECM to develop performance and cost estimates for these plants.

#### 4.7.2. Summary Results from Case Studies 7-9

A summary of the performance results for the plants is given below in Table 4.9. The NGCC plant (Case 9) has the highest net efficiency, lowest CO<sub>2</sub> emissions, and lowest revenue required of the three plants with advanced amine-based CO<sub>2</sub> capture. Of the pulverized coal plants, the supercritical plant (Case 8) has a higher net plant efficiency, lower CO<sub>2</sub> emissions, and a lower revenue requirement than the subcritical plant (Case 7).

**Table 4.9: Summary of case study results for power plants with advanced amine-based CO<sub>2</sub> capture**

Cases	Gross Plant Output (MWe)	Net Plant Output (MWe)	Net Plant Efficiency (%)	CO <sub>2</sub> Emissions (million tonnes/year)	Capital Cost (\$/kW -net)	Revenue Required (\$/MWh)
Case 7: PC Subcritical	672.2	550.0	25.0%	0.45	2849	106.9
Case 8: PC Supercritical	668.2	550.0	27.0%	0.42	2857	104.5
Case 9: NGCC	489.2	453.6	42.7%	0.13	1080	84.5

#### 4.8. Chapter Conclusion

In this chapter the conventional MEA-based CO<sub>2</sub> capture system models in the IECM were modified to reflect significant improvements in amine-based CO<sub>2</sub> capture technology. Critical updates included: (1) parameters on the IECM Config, Performance, Capture, and O&M Cost menus. (2) Updated heat integration, heat-to-electricity efficiency, electrical loss, cost, and amine

usage. (3) And an updated layout of the IECM User Screens. The resulting advanced amine-based CO<sub>2</sub> capture system shows considerable performance and cost improvements over the conventional MEA-based system. In the next chapter, the new advanced amine-based system will be compared in detail against the chilled ammonia-based CO<sub>2</sub> capture system modeled in Chapter 2 and 3 of this thesis.

## **Chapter 5. Comparing Chilled Ammonia and Advanced Amine-Based CO<sub>2</sub> Capture**

### **5.1. Introduction**

This chapter uses the new IECM models developed in Chapter 2, 3 and 4 to compare ammonia and amine-based CO<sub>2</sub> capture technologies. First, performance and cost estimates detailed in the significant power plant process areas are presented for the 9 baseline case studies [(3 plant types) x (no capture, advanced amine, ammonia)]. Then, for the supercritical PC and NGCC cases, a sensitivity analysis is presented by fuel type, fuel cost, plant size, and CO<sub>2</sub> capture system train size. Finally, the probabilistic cost analysis presented for the ammonia-based CO<sub>2</sub> capture system in Chapter 2 is expanded into a probabilistic cost difference analysis between the two CO<sub>2</sub> capture systems on the supercritical PC plants. A probabilistic cost difference analysis can be used to compare two technologies with uncertainty, as it takes correlations between the uncertain variables common between the two technologies into account (Frey and Rubin, 1992).

### **5.2. Baseline Model Comparisons**

The baseline performance and cost estimates detailed to the significant power plant process areas for all 9 power plant case studies is shown in Table 5.1 and Table 5.2 respectively. All of the cases use the baseline plant assumptions from Table 2.1. It is important to recognize that changes in these assumptions can potentially have a large effect on the results that follow. For example, the natural gas cost of \$6.4/GJ used in this study was the 2010 U.S. national average cost delivered to utilities in 2007 dollars (DOE/NETL, 2007a). In the past these prices have fluctuated significantly. An assumed natural gas cost of \$3.2/GJ or \$12.8/GJ changes the revenue required of the Case 3 NGCC plant to \$34.0/MWh and \$103.1/MWh respectively. This illustrates how significantly the assumptions can effect the final results.

In Table 5.1 all the PC power plants (Cases 1-2, 4-5, 7-8) are configured to produce 550MW<sub>net</sub> electrical power output, while the NGCC plants (Cases 3, 6, 9) are configured with two General Electric 7FB gas turbines. Since the gas turbines are only available commercially in discrete sizes, the net electrical power output for each NGCC plant varies but the natural gas flow rate into the plant is fixed. As shown in these tables, the CO<sub>2</sub> capture systems for the plants with both ammonia-based CO<sub>2</sub> capture (Cases 4-6) and advanced amine-based CO<sub>2</sub> capture (Cases 7-9) are configured for 90% CO<sub>2</sub> removal, and all the systems have significant steam and electrical requirements. Therefore the PC plants with CO<sub>2</sub> capture (Cases 4-5, 7-8) are larger and the NGCC plants with CO<sub>2</sub> capture (Cases 6, 9) have a reduced power output compared to their counterparts without CO<sub>2</sub> capture (Cases 1-3).

**Table 5.1: Power plant performance estimates for all case studies.**

	No CCS			Ammonia CCS 90% Capture			Amine CCS 90% Capture		
	Case 1 PCSub	Case 2 PCSuper	Case 3 NGCC	Case 4 PCSub	Case 5 PCSuper	Case 6 NGCC	Case 7 PCSub	Case 8 PCSuper	Case 9 NGCC
Fuel Flow Rate									
Coal Flow Rate (tonnes/hr)	198.3	186.8		286.4	263.8		290.5	268.7	
Natural Gas Flow Rate (tonnes/hr)			72.2			72.2			72.2
Auxiliary Steam Load (MWe equivalent)				118.0	109.6	59.2	181.6	168.0	56.8
Steam Turbine Power (Gross Power) <sup>1</sup>	582.8	581.3	546.1	723.9	711.2	486.9	672.2	668.2	489.2
Auxiliary Electrical Load									
Base Plant (MWe)	20.3	18.8	10.9	30.7	26.7	9.8	28.5	25.0	9.8
Hot Side SCR (MWe)	0.1	0.1		0.1	0.1		0.1	0.1	
Fabric Filter (MWe)	0.1	0.1		0.1	0.1		0.1	0.1	
Wet FGD (MWe)	4.1	4.1		6.0	5.9		5.6	5.5	
Cooling Tower (MWe)	8.2	8.2	3.5	20.2	19.9	2.5	18.8	18.7	2.5
CO <sub>2</sub> Capture System (MWe)				98.4	91.6	62.4	11.2	15.4	7.1
CO <sub>2</sub> Compression (MWe)				18.3	16.9	5.2	57.9	53.3	16.2
Plant Net Power	550.0	550.0	531.6	550.0	550.0	407.1	550.0	550.0	453.6
Plant Efficiency (% HHV)	36.8%	39.1%	50.0%	25.5%	27.7%	38.3%	25.1%	27.1%	42.7%
Plant Derating (%)				30.7%	29.2%	23.4%	31.2%	30.6%	14.6%
CO <sub>2</sub> Emissions (million tonnes/year)	3.09	2.91	1.27	0.46	0.41	0.13	0.45	0.42	0.13

<sup>1</sup>This is the gross plant power output.

**Table 5.2: Power plant cost estimates for all case studies. All values are in 2007 \$Millions**

	No CCS			Ammonia CCS 90% Capture			Amine CCS 90% Capture		
	Case 1 PCSub	Case 2 PCSuper	Case 3 NGCC	Case 4 PCSub	Case 5 PCSuper	Case 6 NGCC	Case 7 PCSub	Case 8 PCSuper	Case 9 NGCC
CO <sub>2</sub> System (TCR)				526.2	497.6	397.6	427.3	395.8	181.9
Base Plant (TCR) <sup>1</sup>	627.8	671.9	334.6	844.3	890.7	277.1	835.6	886.5	277.8
Cooling Tower (TCR)	41.6	37.7	22.7	75.2	72.9	51.9	70.8	66.2	30.3
NO <sub>x</sub> Control (TCR)	26.0	25.0		35.4	33.4		35.7	33.9	
TSP Control (TCR)	39.0	37.5		52.7	49.3		53.3	50.0	
SO <sub>2</sub> Control (TCR)	115.3	112.3		143.3	138.0		144.2	139.2	
CO <sub>2</sub> System O&M/Year				21.1	19.5	16.3	25.1	23.3	10.3
Balance of Plant O&M/Year	98.2	94.5	169.9	136.6	128.9	172.4	139.0	129.7	170.9
Plant Total Capital Requirement	849.7	1186	357.3	1677	1682	726.6	1567	1572	490.0
Total O&M Costs/Year Capital Required (\$/kW-net)	98.2	94.5	169.9	157.7	148.4	188.7	164.1	153.0	181.2
Revenue Required (\$/MWh)	53.7	53.7	58.5	109.9	107.6	109.3	106.9	104.5	84.5

<sup>1</sup>The base plant cost for the ammonia-based CO<sub>2</sub> capture system design is higher than for the amine system because a larger steam turbine is required.

For the subcritical PC power plants with CO<sub>2</sub> capture (Cases 4, 7) and the supercritical PC power plants with CO<sub>2</sub> capture (Cases 5, 8) each of the CO<sub>2</sub> capture technologies show similar overall performance and cost impacts on the power plants despite the significant differences in the two CO<sub>2</sub> capture systems. In these cases, the ammonia-based CO<sub>2</sub> capture systems have a lower auxiliary steam load but a higher electrical load, while the advanced amine-based CO<sub>2</sub> capture systems have a lower electrical load but a higher steam load. Both CO<sub>2</sub> capture technologies also have substantial impacts on the NGCC plants (Cases 6, 9). However, for the NGCC case the ammonia system actually has a higher auxiliary steam use than the advanced amine system. This is primarily a consequence of a high ammonia slip above the absorber for Case 6 (~9000 ppm compared to ~3000 ppm for the PC plants) and subsequent steam loads for ammonia cleanup. The high ammonia slip is a result of the complex interactions between the flue gas CO<sub>2</sub> concentration, the ammonia concentration in the lean solvent, and the lean solvent flow rate. A more detailed discussion of the auxiliary stream use for the NGCC plant with ammonia-based

CO<sub>2</sub> capture can be found in Appendix B. The ammonia system on the NGCC plant also has a substantially higher electrical usage than the advanced amine system due to the high cooling requirements of ammonia-based CO<sub>2</sub> capture and the large gas volumes and relatively low CO<sub>2</sub> concentrations in NGCC plant flue gases.

Overall, the efficiency of the PC and NGCC plants dropped by approximately 11 and 12 percentage points with the ammonia-based CO<sub>2</sub> capture system and by approximately 12 and 7 percentage points for the advanced amine-based CO<sub>2</sub> capture system respectively. The ammonia-based CO<sub>2</sub> system benefits from lower steam loads in the regenerator and reduced compressor power requirements, but the chilling loads and associated costs largely offset these benefits.

For the PC plants, the revenue requirement for the two CO<sub>2</sub> capture technologies is similar given the assumptions used, with the amine technology having slightly lower revenue requirements. Adding either CO<sub>2</sub> capture technology onto the PC plants results in at least a 40% increase in the revenue requirement over the costs of the plants without CO<sub>2</sub> capture, and substantially less electricity is produced for the capital invested. For the NGCC plants with CO<sub>2</sub> capture, there is also a substantial increase in the revenue requirement compared to the plants without CO<sub>2</sub> capture. In the case of the NGCC plant with an ammonia-based CO<sub>2</sub> capture system, the capital costs of the CO<sub>2</sub> system exceed the capital costs of the base plant. This combined with the poor performance of the overall system makes NGCC and ammonia-based CO<sub>2</sub> capture unlikely to be competitive.

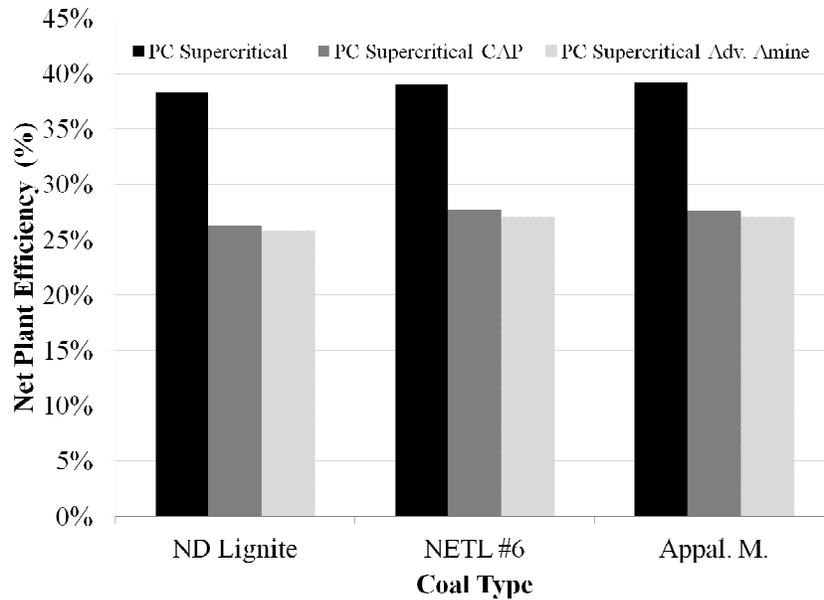
### **5.3. Sensitivity Analysis Across Critical Parameters**

Using the updated IECM, a sensitivity analysis was done over a number of critical variables in order to better understand the impact of changes in different assumptions or variables on the

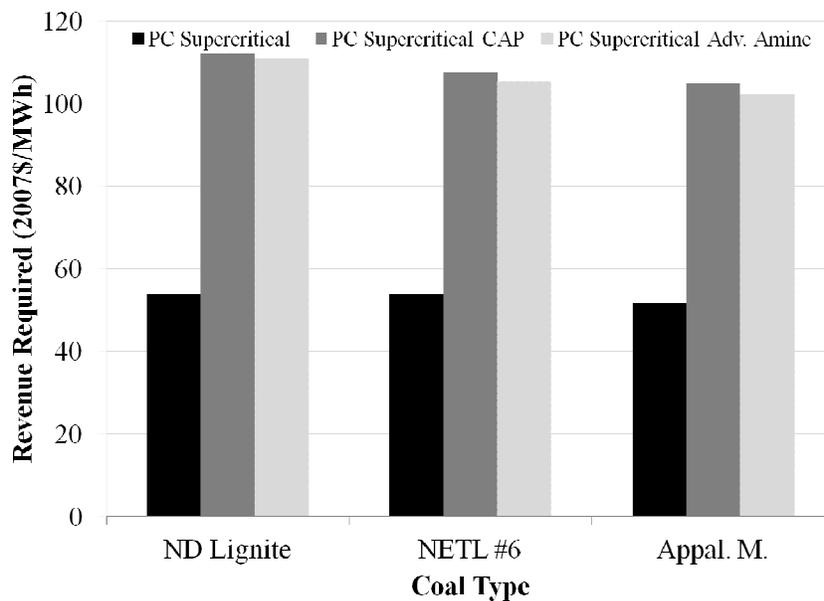
performance and costs of the power plants. These critical variables include fuel type, fuel cost, plant size, and train size. Each of these sets of analysis is shown in the sections below. For brevity some of the cases are not shown in the sensitivity analysis. For example, typically the supercritical PC plant is evaluated but not the subcritical PC plant with the expectation that the results will be qualitatively similar and that the subcritical plant will have slightly poorer performance and higher costs.

### **5.3.1. Sensitivity Analysis by Fuel Type and Cost**

In the sensitivity analysis by fuel type, Illinois No.6, North Dakota Lignite, and Appalachian Medium Sulfur coals are investigated with base prices of \$1.705/GJ, \$1.203/GJ, and \$1.617/GJ (IECM V7.0) respectively for the PC plants with and without CO<sub>2</sub> capture technology, and everything else was held constant in the analysis. The resulting net plant efficiency and revenue requirement are shown in Figure 5.1 and Figure 5.2 respectively. The trends established in Table 5.1 hold for the power plants across the fuel types, with changes in fuel type resulting in small changes in power plant efficiency and revenue required. Generally, the PC plants with ammonia-based CO<sub>2</sub> capture systems have a higher efficiency, but also higher revenue required compared to the plants with advanced amine-based CO<sub>2</sub> capture systems.



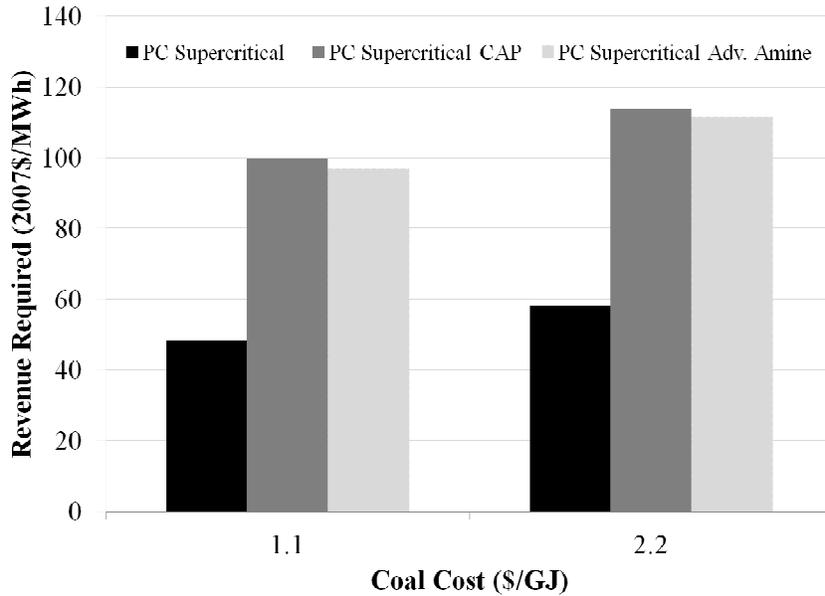
**Figure 5.1: Net plant efficiency by power plant and coal type**



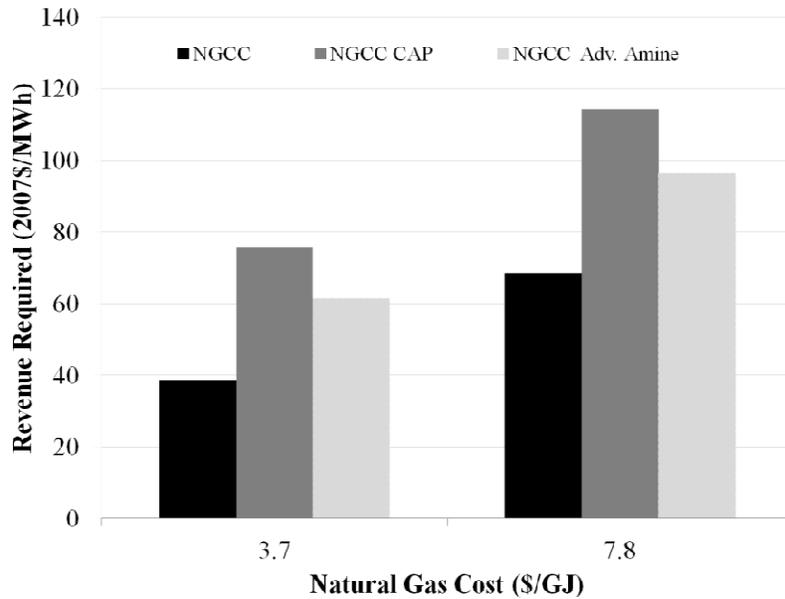
**Figure 5.2: Revenue requirement by power plant and coal type**

In the sensitivity analysis by fuel cost, Illinois No.6 coal was used for the PC plants with a range of prices from \$27-54 per short ton (approximately \$1.1-2.2/GJ for Illinois No.6), the range of prices in 2005 real dollars for bituminous coal from 2000 to 2010 in the U.S. (EIA, 2010a). For natural gas, the price range used was \$3.99-8.47 per thousand cubic feet (approximately \$3.7-

7.8/GJ), the range of prices in 2005 real dollars for the electric power sector from 2000 to 2010 in the U.S. (EIA, 2010a). The resulting revenue requirement for the plants is shown in Figure 5.3 and Figure 5.4 respectively.



**Figure 5.3: Revenue requirement by power plant and coal cost**

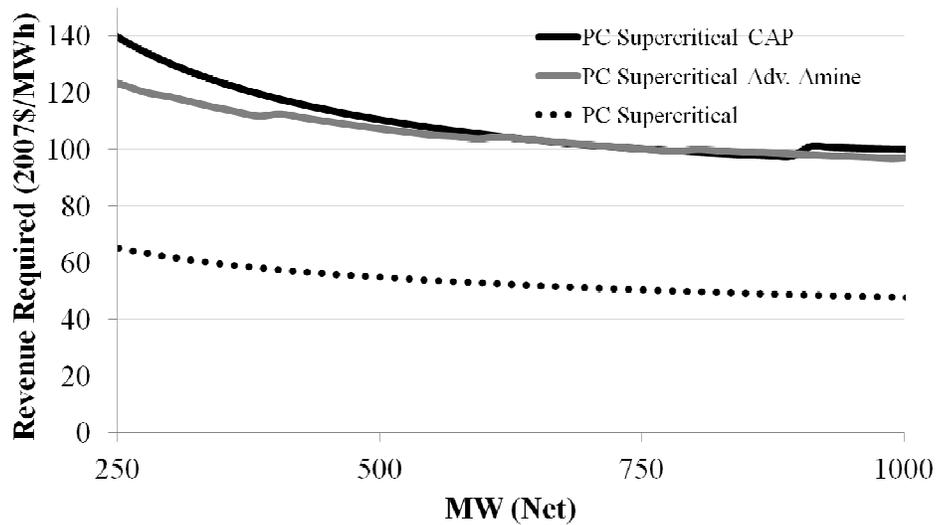


**Figure 5.4: Revenue requirement by power plant and natural gas cost**

As shown in the figures, the impact of fuel costs significantly impacts all the power plants, both with and without CCS, with natural gas price changes in particular significantly impacting the revenue requirement of the NGCC plants. At a natural gas price of \$3.7/GJ, the revenue requirement of the NGCC plant with advanced amine-based CO<sub>2</sub> capture is lower than the revenue requirement of the NGCC plant without CO<sub>2</sub> capture at a natural gas price of \$7.8/GJ, demonstrating how high natural gas prices negatively impact the cost of NGCC plants with and without CO<sub>2</sub> capture. However, the revenue requirement of the NGCC plant with advanced amine-based CO<sub>2</sub> capture at the high natural gas price is still roughly comparable to the revenue requirement of the PC plant with advanced amine-based CO<sub>2</sub> capture at the low coal price.

### **5.3.2. Sensitivity Analysis by Plant Size and Train Size**

In addition to fuel type and fuel cost, a sensitivity analysis was also done for the power plants across a range of net power plant sizes and CO<sub>2</sub> capture system train sizes. This analysis for the PC plants with and without CO<sub>2</sub> capture and for the default train sizes for each CO<sub>2</sub> capture system is shown in Figure 5.5, using the assumptions outlined in Table 2.1

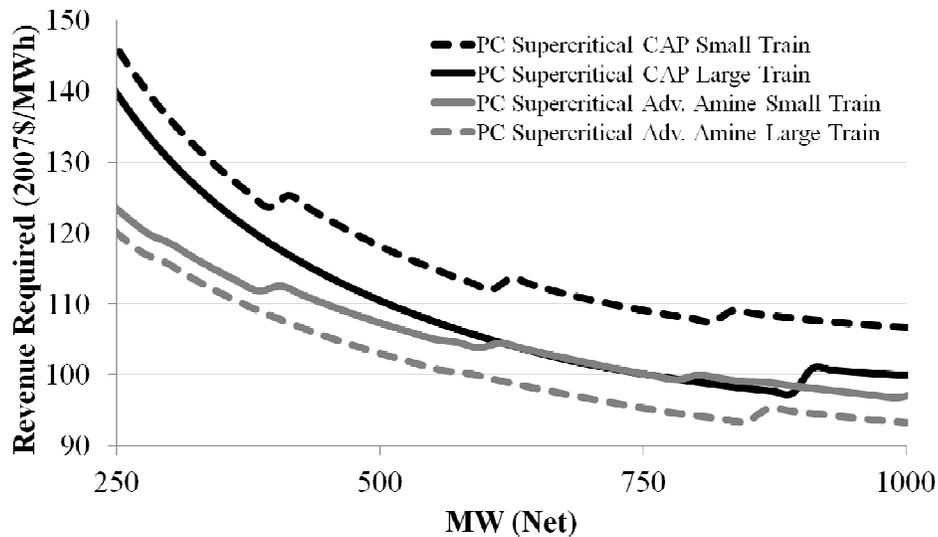


**Figure 5.5: Revenue requirement by PC power plants with and without CO<sub>2</sub> capture as a function of plant size. The CO<sub>2</sub> capture systems use the default train sizes.**

The slow decline of the revenue requirement of the all PC plants as the net power output increases is characteristic of the reduced costs due to economies of scale. For the PC plants with CO<sub>2</sub> capture, the decline in revenue requirement as plant size increases is occasionally interrupted by a small but sharp increase when an additional train must be added to the CO<sub>2</sub> capture system. For small scales below about 500MW<sub>net</sub>, the advanced amine-based CO<sub>2</sub> capture system has a lower revenue requirement than the ammonia-based CO<sub>2</sub> capture system, whereas the revenue requirements are similar for larger power plants sizes. For smaller power plants, a larger capital investment is required for the ammonia-based CO<sub>2</sub> capture system compared to the power plants with the advanced amine-systems, and this additional capital investment overtakes the efficiency benefits of the ammonia-based CO<sub>2</sub> capture system at the smaller plant sizes.

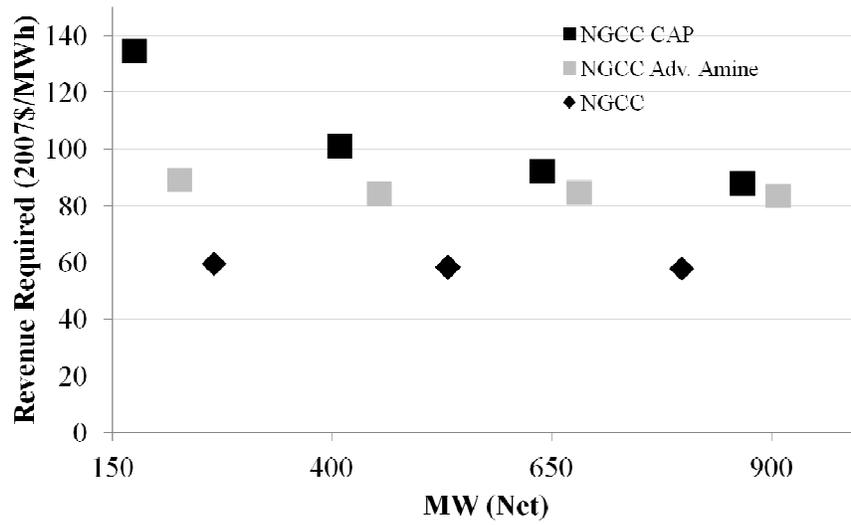
The results for the CO<sub>2</sub> capture systems in Figure 5.5 were extended for two different train sizes for each of the ammonia and advanced amine-based CO<sub>2</sub> capture systems, as shown in Figure

5.6. In the figure, each CO<sub>2</sub> capture system uses a small train size (208.7 tonnes/hr, the advanced amine system default) and a large train size (907.2 tonnes/hr, the ammonia-system default). The ammonia-system has a much larger default size because the absorber in the ammonia system is envisioned as a spray tower as opposed to a packed column as in the case of an advanced amine system. As shown in Figure 5.6, the train size can significantly impact the revenue required of these systems for all plant sizes.

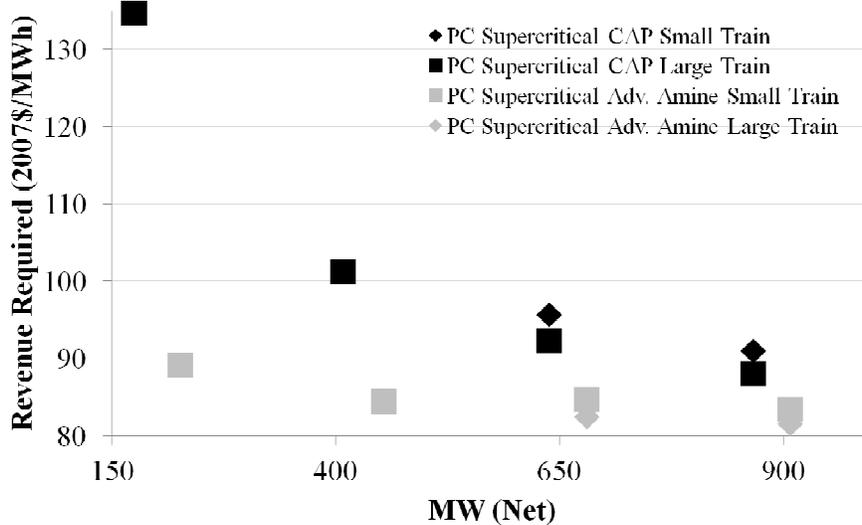


**Figure 5.6: Revenue requirement by PC power plants with CO<sub>2</sub> capture as a function of plant size. Each CO<sub>2</sub> capture system uses a small train size of 208.7 tonnes/hr and a large train size of 907.2 tonnes/hr.**

A similar sensitivity analysis was done for the NGCC plants with and without CO<sub>2</sub> capture across a range of net power plant sizes. The results are shown in Figure 5.7 and Figure 5.8. Since the NGCC plants are available only in discrete sizes and because the CO<sub>2</sub> captured at an NGCC plant is much less than for a PC plant, the effect of train size as the plant size increases is less apparent than for the PC plants.



**Figure 5.7: Revenue requirement by NGCC power plants with and without CO<sub>2</sub> capture system as a function of plant size. The CO<sub>2</sub> capture systems use the default train sizes.**



**Figure 5.8: Revenue requirement by NGCC power plants with and without CO<sub>2</sub> capture system as a function of plant size. Each CO<sub>2</sub> capture system uses a small train size of 208.7 tonnes/hr and a large train size of 907.2 tonnes/hr.**

As shown in Figure 5.7 and Figure 5.8, the impact of economies of scale is apparent for the NGCC plants with CO<sub>2</sub> capture as the revenue required decreases with increasing power plant

size. However, the effect of differences in CO<sub>2</sub> system train size is not as pronounced as for the PC plants because of the reduced CO<sub>2</sub> to be captured and therefore the reduced number of trains needed for each CO<sub>2</sub> capture system.

## **5.4. Probabilistic Cost Difference for Ammonia and Amine-Based Systems**

### **5.4.1. Introduction**

In Chapter 2, an uncertainty analysis was done to estimate the impact of uncertainty on the revenue required for ammonia-based CO<sub>2</sub> capture technologies. However, when two technologies are compared and uncertainty is included for both technologies, a more sophisticated measure for comparing these technologies is to look at the probabilistic revenue required differences between the two technologies. In this type of analysis, uncertainty distributions are applied to the key performance and cost variables of power plants with both technologies, and Monte Carlo simulations are undertaken to calculate cumulative probability distributions for both sets of systems. However, several of the variables in each of the Monte Carlo simulations will be perfectly correlated with one another. For example, these can include labor rates and fuel costs for plants that are located in the same region. In a probabilistic revenue required difference approach, these correlated variables are held at the same value through a Monte Carlo model run across both technologies, while the other variables which are not correlated are varied. In this section a probabilistic revenue required difference analysis was done between advanced amine and ammonia-based CO<sub>2</sub> capture technologies using these methods.

#### **5.4.2. Selection of Uncertainty Distributions**

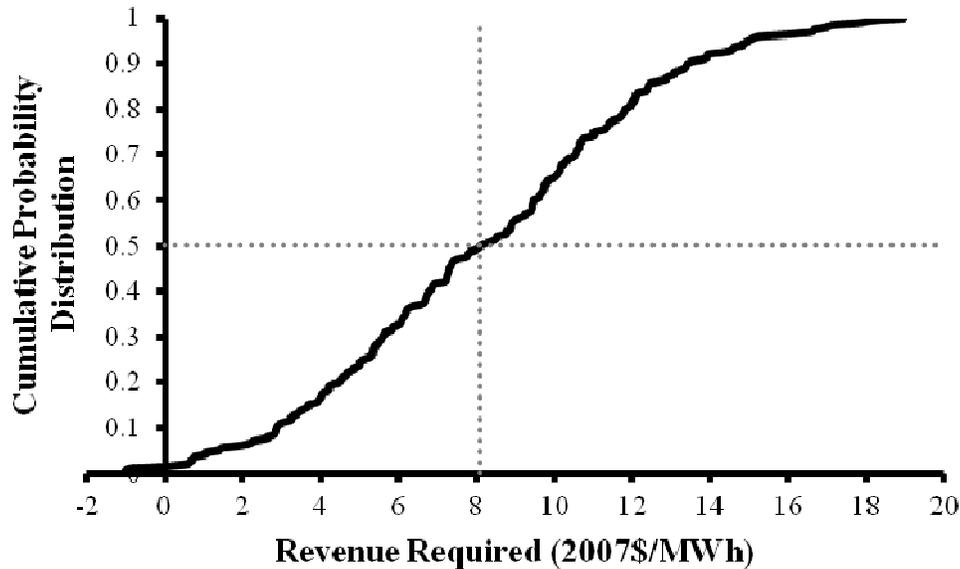
The values and uncertainty parameters used in this probabilistic cost difference analysis are shown in Table 5.3. As shown in the table, there are uncertainties for both the advanced amine and for the ammonia systems. Some of the uncertainties are technology specific and based on process design and chemistry, for example the regeneration heat energy for both CO<sub>2</sub> capture systems are calculated independently. Some of the uncertainties are likely to be highly correlated for both technologies, for example the pump and fan efficiencies in the CO<sub>2</sub> capture systems and the capacity factors for both power plants. In Table 5.3, both sets of uncertainties are listed. The ranges for the uncertainties in Table 5.3 are meant to be meaningful for a new standalone power plant at a landlocked greenfield site in the Midwestern US. Unique site-specific factors (e.g. inexpensive chilling from an external source) may lead to different overall results. However, these site-specific factors are not considered to be characteristic of a typical standalone power plant, and would require designed integration with these technologies. Therefore, these do not affect the uncertainty ranges in the table.

**Table 5.3: Nominal values and uncertainty parameters assessed in the CO<sub>2</sub> Capture systems**

Parameter	Units	Nominal (x)		Values (or $\sigma$ as % of x)	References
		Amine	Ammonia		
Capture System Performance					
Chilling Load Required	Tons Refrigeration	N/A	108,500	Normal(x, 10%)	Author Estimate
Chiller Electrical Use	kW/Ton Refrigeration	N/A	0.55	Uniform(0.47, 0.55)	Platts 2004
CO <sub>2</sub> Regeneration Heat Requirement	kJ/kg CO <sub>2</sub>	3631	2421	Normal(x, 10%)	Author Estimate
Pump Efficiency <sup>2</sup>	%	75	75	Uniform(70,75)	Rao 2002
$\Delta P$ Across CO <sub>2</sub> Capture System	kPa		20.7	Triangular(14,26,30)	Rao 2002
Blower Efficiency <sup>2</sup>	%	75	75	Uniform(70,75)	Author Estimate
CO <sub>2</sub> Compression, 1.0 bar to 152.7 bar	kWh/kg CO <sub>2</sub>	0.093		Normal(x, 10%)	Rao 2002
CO <sub>2</sub> Compression, 27.5 bar to 152.7 bar	kWh/kg CO <sub>2</sub>		0.030	Normal(x, 10%)	Author Estimate
Capture System Cost					
Reference Chilling Equipment Costs (PFC)	\$2007/Ton Refrigeration	N/A	441	Uniform(0.7x-1.3x)	Author Estimate
CO <sub>2</sub> Absorber Costs (PFC)	\$2007		105.1	Uniform(0.7x-2.5x)	Zhuang et al. 2011
General Facilities Capital <sup>2</sup>	% of PFC	95		Uniform(0.7x-1.3x)	Author Estimate
Eng. & Home Office Fees <sup>2</sup>	% of PFC	1.57 <sup>1</sup>	1.57 <sup>1</sup>	Normal(x, 10%)	Berkenpas et al. 1999
Project Contingency Cost <sup>2</sup>	% of PFC	9.37 <sup>1</sup>	9.37 <sup>1</sup>	Triangular(0.7x, 1x, 1.5x)	Berkenpas et al. 1999
Process Contingency Cost <sup>2</sup>	% of PFC	16.38 <sup>1</sup>	16.38 <sup>1</sup>	Normal(x, 20%)	Berkenpas et al. 1999
CO <sub>2</sub> System TS&M <sup>2</sup>	\$/tonne CO <sub>2</sub>	4.67 <sup>1</sup>	4.67 <sup>1</sup>	Normal(x, 30%)	Berkenpas et al. 1999
Plant Financing & Utilization					
Power Plant Fixed Charge Factor <sup>2</sup>	Fraction	3.75	3.75	Uniform(0.7x-1.3x)	Author Estimate
Power Plant Levelized Capacity Factor <sup>2</sup>	--	0.143	0.143	Uniform(0.130, 0.180)	Rubin and Zhai, 2011
		0.75	0.75	Uniform(0.65, 0.85)	Rubin and Zhai, 2011

<sup>1</sup>From DOE/NETL, 2007a. <sup>2</sup>Variables that are assumed to be perfectly correlated in the probabilistic cost difference analysis.

The resulting probabilistic revenue required difference is shown in Figure 5.9. The probabilistic revenue required difference was completed by running Monte Carlo simulations on the power plants using the uncertainties shown in Table 5.3. For the Monte Carlo simulation, the values of the correlated variables were linked so that they varied together and were matched during individual runs. The resulting two cumulative probability distributions for revenue requirement for each power plant was subtracted, one from the other, to create Figure 5.9.



**Figure 5.9: Probabilistic revenue required difference between the supercritical PC plant with the ammonia-based CO<sub>2</sub> capture system and the supercritical PC plant with the advanced amine-based CO<sub>2</sub> capture system.**

Figure 5.9 indicates that there is only about a 2% chance that the ammonia system will be less expensive than the amine system under these conditions. The analysis indicates that there will be approximately a 50% chance that the revenue requirement for the plant with the advanced amine system will be at least \$8/MWh less than the revenue requirement for the plant with the ammonia system. Perhaps this result should not be too surprising given that the revenue requirement for the plants with ammonia-based CO<sub>2</sub> capture systems has been higher than for comparable plants with advanced amine-based CO<sub>2</sub> capture systems for all of the results in this chapter.

### 5.5. Chapter Conclusion

Through a variety of estimates, this chapter has shown that pulverized coal power plants with ammonia-based CO<sub>2</sub> capture systems could potentially operate with a higher net plant efficiency than pulverized coal power plants with advanced amine-based CO<sub>2</sub> capture systems. However, the revenue required was comparable or consistently higher for the plants with ammonia-based CO<sub>2</sub> capture in all cases. For the NGCC plants with CCS, the revenue required for the plant with

the ammonia-based CO<sub>2</sub> capture system was significantly higher than the revenue required for a similar plant with an advanced amine-based CO<sub>2</sub> capture system. Qualitatively, these results held across a sensitivity analysis by fuel type, fuel cost, plant size, train size, and for probabilistic revenue required difference between the two CO<sub>2</sub> capture technologies. The public policy implications of these estimates are explored in more detail in the next chapter.

## Chapter 6. Policy Implications

### 6.1. Introduction

This thesis is intended to be a starting point for estimating costs for post-combustion CO<sub>2</sub> capture processes, a starting point that will help policy makers be more informed about the costs of CO<sub>2</sub> emission mitigation options. Specific policy relevant issues that could be addressed in part with the results from this thesis may include:

- 1) An initial benchmarking of the performance and costs of advanced amine and ammonia-based CO<sub>2</sub> capture technologies, information that could be used:
  - a. To compare alternative technologies for CO<sub>2</sub> capture at power plants
  - b. To construct marginal abatement cost curves for different levels of mitigation (e.g. Creyts et al., 2007).
  - c. To assess the costs of reducing climate risk (Schaeffer et al., 2008).
  - d. To construct timelines for a rollout of post-combustion CO<sub>2</sub> capture systems.
  - e. In larger energy models to inform emissions targets and CO<sub>2</sub> prices.
- 2) An identification of the potential issues and opportunities associated with advanced amine and ammonia-based CO<sub>2</sub> capture, such as the conditions under which particularly high or low costs may be expected. An example of this is the impact of fuel cost on the overall revenue requirement of power plants with CCS.
- 3) The ability to extend the baseline results for specific application. The major results in this thesis are reproducible using the IECM and can be extended. This makes these

models a useful tool for policy makers in that it is flexible enough to provide results of power plant performance and cost under a range of conditions.

This chapter will briefly explore some of these policy relevant issues including the potential use of these estimates for informing EPA guidelines, for investigating the impact of CO<sub>2</sub> prices on the economics of power plants with and without CCS, for investigating unique site-specific factors when siting power plants, and for investigating potential retrofit costs.

## **6.2. The Potential Effect of Performance and Cost Estimates on Policy Decisions**

The direct performance and cost comparisons of power plants with ammonia and advanced amine-based CO<sub>2</sub> capture is useful for helping decide which technologies to support. The comparisons and models created in previous chapters can be used in a number of ways, some of which are explored as examples in the sections below.

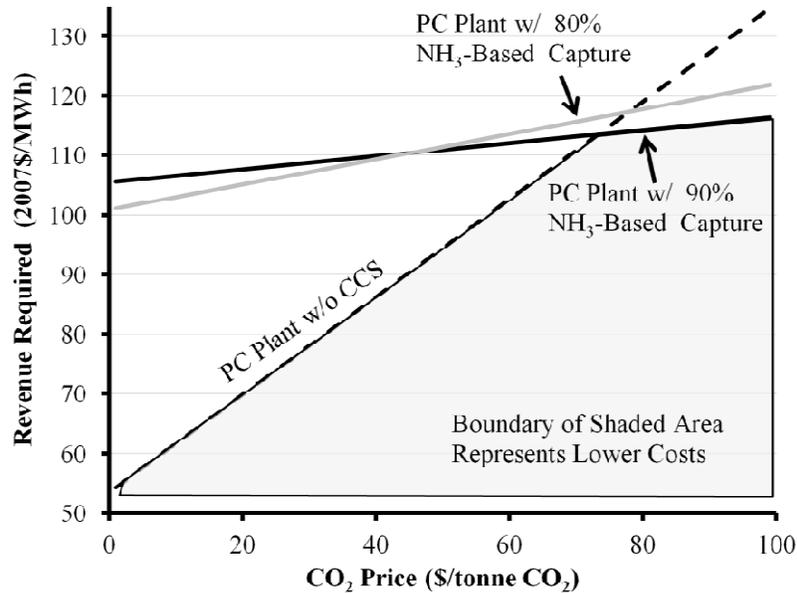
### **6.2.1. Information for EPA Guidelines**

In 2007, the Supreme Court issued an endangerment finding for CO<sub>2</sub>, ruling therefore that CO<sub>2</sub> must be regulated by the Environmental Protection Agency (EPA) under the Clean Air Act (CAA) (*Massachusetts v. EPA*, 549 U.S. 497, 2007) (EPA, 2011). Since this ruling, EPA has announced that it will enact an output based standard, limiting CO<sub>2</sub> emissions for new power plants over 25 MWe to 1000 lb CO<sub>2</sub>/MWh. In regulating for both new and in some cases existing sources, the EPA has the authority to consider both the technologies available and the costs of such technologies in order to implement guidelines (Richardson et al., 2010). The technology specific analysis in this thesis and other analysis like it should influence the setting of these guidelines.

It is plausible that the Environmental Protection Agency will investigate both advanced amine and ammonia-based CO<sub>2</sub> capture in assessing control technologies to meet New Source Performance Standards under the Clean Air Act. The emissions reductions under NSPS must be based on technologies that have been demonstrated (ibid), and therefore the standards that are applicable under this piece proposed regulation are tied to the maturity, commercial availability, performance, and cost of these technologies, all of which have been investigated in this thesis.

### **6.2.2. Effect of CO<sub>2</sub> Prices**

The economics and degree of CO<sub>2</sub> capture for plants with CCS depend critically on the restrictions or requirements imposed by policy and regulations limiting the emissions of CO<sub>2</sub>. Emissions are often expected to be tied to a “carbon price” (or tax) that an emission source must pay for each tonne of CO<sub>2</sub> emitted. For a power plant emitting CO<sub>2</sub> this additional cost increases the revenue required in proportion to the level of CO<sub>2</sub> emissions. Figure 6.1 shows the effect of CO<sub>2</sub> price on the revenue required for three cases: (1) a supercritical power plant with an ammonia-based CO<sub>2</sub> capture system capturing 90% CO<sub>2</sub> (Case 5) (2) the same plant capturing 80% CO<sub>2</sub>, and (3) the uncontrolled power plant with no CCS system (Case 2).



**Figure 6.1: Effect of a CO<sub>2</sub> price on the levelized cost of electricity for: (a) a supercritical power plant with an ammonia-based system that captures 90% CO<sub>2</sub> (Case 5); (b) the same plant with an ammonia-based system that captures 80% CO<sub>2</sub> (IECM); and (c) a supercritical power plant without a CCS System (Case 2).**

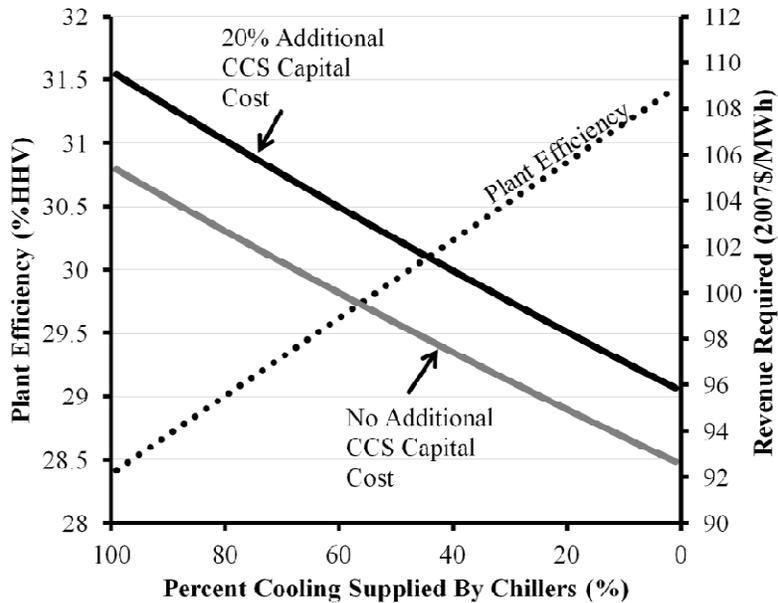
For a CO<sub>2</sub> price of up to \$73/tonne CO<sub>2</sub>, the supercritical power plant without CCS is the lowest cost option. At higher prices the supercritical power plant with an ammonia-based system capturing 90% CO<sub>2</sub> is the least costly of these options. In this analysis the plant with only 80% capture is more expensive than the other two options at all carbon prices. This is consistent with the results in Figure 2.12 showing a higher cost of CO<sub>2</sub> avoided as the capture efficiency decreases.

### 6.2.3. The Effect of Site Specific Factors

Every power plant project is in some ways unique, and there may be a variety of specific circumstances in which one technology may be chosen over another, though these may be in non-standard applications. These situations can be simulated and investigated using the IECM and the models developed in this thesis. Consider, for example, the situation in which the availability of inexpensive chilling favorably shifts the economics of an ammonia-based CO<sub>2</sub>

capture process. This situation could be a scenario in which a plant is located with direct access to a large cooling sink such as a deep water lake or ocean, or the re-gasification process of a liquefied natural gas facility, and the additional cooling at this location can be used to reduce the parasitic energy demands for the process chillers. While this scenario was not considered in the comparison between advanced amine and ammonia systems in Chapter 5 because it is a situation that is more site specific than technology specific, it could significantly improve the performance and cost outlook of ammonia-based CO<sub>2</sub> capture in these applications.

To illustrate this type of analysis, this situation was investigated using the IECM for the PC supercritical power plant with ammonia-based CO<sub>2</sub> capture (Case 5) where the revenue required as a function of percent cooling supplied by the chillers was calculated. The analysis is shown in Figure 6.2. In the figure, cooling not supplied by the chillers is assumed to come from an external source. The change in power plant efficiency and revenue required is shown with two cases, one without an addition cost to the CCS system for integration with the external chilling source, and one with a 20% additional cost to the CO<sub>2</sub> capture system for integration (\$74M applied to the process facilities cost).



**Figure 6.2: Effect of available cooling on the plant efficiency and the revenue required for the supercritical power plant with an ammonia-based system that captures 90% CO<sub>2</sub> (Case 5). Shown in the graph are two cases, one with no additional capital cost, and one with 20% additional cost applied to the process facilities cost of the CO<sub>2</sub> capture system.**

As shown in Figure 6.2 the original revenue required of the power plant (\$105/MWh) may be significantly reduced in this type of situation, even if a substantial additional capital cost is included. This is not meant to suggest that this will necessarily be a realistic situation for power plants in the future, but rather that these types of situations can be easily investigated using the new models in the IECM.

### 6.3. Potential Retrofit Costs

As described in Chapter 1, there is a large number of existing coal and natural gas-fired power plants, and many of these plants have several decades' worth of useful remaining operational life. As such, the costs of retrofitting CCS onto these existing facilities can and should be considered. Retrofitting an existing power plant with CCS may lead to reduced output and additional cost due to the likelihood of reduced opportunities for heat integration and added capital expenses

associated with installing CCS on a plant with a constrained footprint (Lucquiaud and Gibbins, 2009).

In a preliminary assessment of the costs of retrofitting a coal-fired power plant with both ammonia and advanced amine-based CO<sub>2</sub> capture technology, the supercritical PC power plants with CCS (Cases 5 and 8) were modified to simulate a retrofit scenario. In this assessment the fuel flow rate into the plants was held constant at 260 tonnes of Illinois No.6 coal per hour, the plant capital cost was assumed to be completely amortized except for the CO<sub>2</sub> capture system, the heat-to-electricity conversion efficiency was increased from 0.22 to 0.25, and a capital cost retrofit factor of 1.25 was used to scale the costs of the capture unit. A comparison of the results is shown in Table 6.1.

**Table 6.1: Summary of case study results for the retrofit analysis**

	No CCS (Prior to Retrofit)	Ammonia CCS 90% Capture	Amine CCS 90% Capture
	Case 2 PCSuper	Case 5 - Retrofit PCSuper	Case 8 - Retrofit PCSuper
<b>Performance</b>			
Coal Flow Rate (tonnes/hr)	260.0	260.0	260.0
Auxiliary Steam Load (MWe equivalent)		123.0	184.7
Steam Turbine Power (Gross Power)	808.9	686.0	624.3
Auxiliary Electrical Load			
Plant Net Power	765.4	528.0	513.9
Plant Efficiency (%HHV)	39.0%	26.9%	26.2%
CO <sub>2</sub> Emissions (million tonnes/year)	4.0	0.4	0.4
<b>Cost</b>			
CO <sub>2</sub> System (TCR)		616.1	484.5
Base Plant (TCR)	0.0	0.0	0.0
Cooling Tower (TCR)	0.0	0.0	0.0
NO <sub>x</sub> Control (TCR)	0.0	0.0	0.0
TSP Control (TCR)	0.0	0.0	0.0
SO <sub>2</sub> Control (TCR)	0.0	0.0	0.0
Capital Requirement (\$M)	0.0	616.1	484.5
Total O&M Costs/Year (\$M/Year)	106.4	128.2	131.1
Revenue Required (\$/MWh)	21.2	62.3	59.3

As shown in Table 6.1, retrofitting a facility with a CO<sub>2</sub> capture system adds significantly to the costs of generating electricity. This is a very simplified analysis and the potential difficulties with retrofitting can be significant. For example, due to system reliability concerns older plants that are slated for retirement may be put into reliability must run mode causing delays in scheduling a plant stoppage for a retrofit, especially in larger plants (Rose et al., 2011).

#### **6.4. Technological Maturity of Ammonia and Amine Technologies**

Many policies are justified through the economic efficiencies gained through learning by doing, such as with the California Solar Initiative (Gillingham et al., 2010). As shown in Table 1.2 and Table 1.3 of Chapter 1, both ammonia and advanced amine-based CO<sub>2</sub> capture technologies are or have recently been active at the demonstration stages, and commercial amine technologies are running though not yet at large power plant scales. Technological learning for CO<sub>2</sub> capture systems on coal and natural gas-fired power plants requires developing the technology at these different scales, into a variety of steam cycles, and on various flue gases.

It appears that technological learning will continue for amine systems, as these systems are now being developed by multiple companies for a wide variety of applications. The trend in recent years is for companies who are developing amine-based CO<sub>2</sub> capture systems to keep the details of their technologies as trade secrets (e.g. Fluor, Mitsubishi Heavy Industries). Keeping these technologies proprietary makes sense because of the significant development investment required in these technologies and the fact that in order for these companies to make a return on their investment they have to wait for a strong climate policy to be put in place.

The outlook is more uncertain for ammonia-based CO<sub>2</sub> capture. Only one large company, Alstom Power, appears to be developing ammonia systems, and since Alstom has recently halted plans to

move forward with developing a demonstration facility at the Mountaineer plant, it appears that learning rates of ammonia-based CO<sub>2</sub> capture systems are slowing. Historical evidence with flue gas desulfurization is instructive in the context of CO<sub>2</sub> capture technologies, and shows that many initially promising technologies do not get the opportunity to travel far along a learning curve. A large number of systems and schemes have been proposed but few ever develop to the commercial stage and fewer still gain a significant amount of market share. For example, an assessment of 189 different flue gas desulfurization processes in 1987 showed that only 11 were being used in power plants, with more than 65% of those being wet limestone or lime FGD systems (Kohl and Nielsen, 1997). Utilities often choose the most mature technologies, making it difficult for vendors of less mature technologies to win contracts (ibid). The patents filed in 2008 and 2009 for the chilled ammonia process by Alstom Power are aging and in addition, the plant derating estimates for ammonia-based CO<sub>2</sub> capture that can be calculated from Case 2 and Case 5 in Table 5.1 (approximately 29%) are substantially higher than estimates in the 9-20% range estimated by Alstom as shown in Table 2.3. These maturity issues should also be considered by policy makers when investigating these technologies.

## **6.5. Chapter Conclusion**

This chapter briefly looked at some of the potential policy implications of the work presented in this thesis. Examples of the potential impact of performance and cost estimates on policy decisions was investigated, potential retrofit costs were investigated, and comments were made on the maturity of ammonia and amine-based CO<sub>2</sub> capture technologies. The analysis in this chapter was done in the IECM, and all of the results should be reproducible to users of the model.

## **Chapter 7. Summary and Conclusions**

### **7.1. Thesis Summary**

This thesis is intended to be a starting point for estimating the costs for post-combustion CO<sub>2</sub> capture processes, a starting point that will help policy makers, researchers, technology developers, and others to be more informed about the costs of CO<sub>2</sub> emission mitigation options.

Chapter 1 briefly introduced the connection between CO<sub>2</sub> emissions and global climate change, and established that post-combustion based CCS could be a useful technology for significantly reducing CO<sub>2</sub> emissions at large point sources. Chapter 1 also limited the scope of this thesis to an investigation of the performance and economics of the two leading post-combustion CO<sub>2</sub> capture technologies: ammonia and amine-based CO<sub>2</sub> capture systems.

Chapter 2 provided an assessment of ammonia-based CO<sub>2</sub> capture. The performance and cost of two ammonia-based post-combustion CO<sub>2</sub> capture systems operating at a new supercritical coal-fired power plant were modeled. The results showed that for a fixed coal input, the plant derating of a CO<sub>2</sub> capture system operating with high ammonia concentrations (HighNH<sub>3</sub> Case) was found to be 28.5%. The plant derating of a CO<sub>2</sub> capture system operating with low ammonia concentrations (LowNH<sub>3</sub> Case) was substantially higher at 39.4%. Preliminary estimates of the revenue requirement of the plants with HighNH<sub>3</sub> and LowNH<sub>3</sub> systems were \$U.S. 133/MWh and \$U.S. 103/MWh, respectively. For the HighNH<sub>3</sub> ammonia system the absorber CO<sub>2</sub> capture efficiency, NH<sub>3</sub> slip, and solids precipitation were evaluated for changes in lean solution NH<sub>3</sub> concentration, NH<sub>3</sub>/CO<sub>2</sub> ratio, and absorber temperature. Reductions in NH<sub>3</sub> slip were also assessed for changes in absorber temperature and water wash flow rate. For 90% CO<sub>2</sub> capture the revenue required for the final plant design was estimated at \$US 105/MWh. The revenue

required for this system was found to depend strongly on the fraction of CO<sub>2</sub> captured as well as on key process design parameters such as lean solution NH<sub>3</sub> concentration. Uncertainties in system performance and cost also were estimated probabilistically. Assumptions about plant financing and utilization, as well as uncertainties in cooling costs and chemical reaction rates that affect absorber cost were found in particular to produce a wide range of cost estimates for ammonia-based CO<sub>2</sub> capture systems. With uncertainties included, costs ranged from \$80/MWh to \$160/MWh, with the 95% confidence interval ranging from \$95/MWh to \$143/MWh.

Chapter 3 then described the development of a general ammonia-based CO<sub>2</sub> capture model for use in the Integrated Environmental Control Model (IECM). The IECM is a freely available tool used by researchers, engineers, policy makers, and others for developing preliminary cost estimates for power plants burning coal and natural gas under a wide variety of plant configurations. The chapter outlined work done to integrate an ammonia-based post-combustion CO<sub>2</sub> capture into the IECM with newly-developed modules for this technology. A general response surface model of an ammonia-based CO<sub>2</sub> capture system was presented using a dataset created using the Aspen Plus® model developed in Chapter 2. The response surface model consisted of a large set of algebraic equations which were implemented as a module in the IECM.

In Chapter 4, a model of an existing amine system in the IECM was modified to reflect more modern designs of amine-based CO<sub>2</sub> systems. In this chapter the conventional MEA-based CO<sub>2</sub> capture system models in the IECM were modified to reflect significant improvements in amine-based CO<sub>2</sub> capture technology. Critical updates included: (1) parameters on the IECM Config, Performance, Capture, and O&M Cost menus. (2) Updated heat integration, heat-to-electricity efficiency, electrical loss, cost, and amine usage. (3) And an updated layout of the IECM User

Screens. The resulting advanced amine-based CO<sub>2</sub> capture system showed considerable performance and cost improvements over the conventional MEA-based system.

In Chapter 5, the ammonia and advanced amine-based systems are compared. Through a variety of estimates, this chapter showed that pulverized coal power plants with ammonia-based CO<sub>2</sub> capture systems could potentially operate with a higher net plant efficiency than pulverized coal power plants with advanced amine-based CO<sub>2</sub> capture systems. However, the revenue requirement was comparable or consistently higher for the plants with ammonia-based CO<sub>2</sub> capture in all cases. Qualitatively, these results held across a sensitivity analysis by fuel type, fuel cost, plant size, and train size. A probabilistic analysis for the difference in revenue requirement for the two CO<sub>2</sub> capture technologies also showed a very high likelihood that the ammonia-based plant would be more costly.

Chapter 6 then briefly described the policy implications of this comparison. Examples of the potential impact of performance and cost estimates on policy decisions was investigated, along with the potential retrofit costs. Comments also were made about the maturity of ammonia and amine-based CO<sub>2</sub> capture technologies, and its relevance for cost analysis. The primary conclusions are summarized below.

## **7.2. Main Results and Implications**

This thesis has modeled and compared ammonia and advanced amine-based CO<sub>2</sub> capture systems and has found that the advanced amine system will likely have a lower revenue requirement in most situations for a wide variety of power plant configurations. While the ammonia-based CO<sub>2</sub> capture system was found to require less steam from the steam cycle (except for NGCC plants) and less electricity for compression than the amine-based CO<sub>2</sub> capture

system, these energy benefits are offset by the large electrical demand for chilling the flue gases. The additional capital costs of the ammonia-based CO<sub>2</sub> capture system along with the high energy demands of flue gas cooling and ammonia cleanup lead to higher revenue requirements than for the advanced amine systems. The ammonia-based system modeled here is most competitive on pulverized coal-fired power plants, but still lead to a higher revenue requirement than the plant with an advanced amine-based system. The probabilistic revenue required difference analysis explored departures from the baseline performance and cost assumptions and supported the general conclusion that the revenue required was higher for plants with ammonia-based CO<sub>2</sub> capture systems.

The advanced amine technology modeled in this thesis looked more promising than the ammonia-based system, in particular on natural gas combined cycle plants. Despite larger steam use, advanced amine technologies benefit from lower electrical use and lower costs in all the CCS cases investigated in this thesis. In addition, advanced amine technologies are more mature than other post-combustion CO<sub>2</sub> capture technologies, and are commercially offered by large companies.

Finally, an important caveat is that the results of this thesis are highly dependent on the specific assumptions used, and may vary for different scenarios. For example, power plant performance and costs for the ammonia-based system would change for improved heat integration of the CO<sub>2</sub> capture system, use of alternate fuels, siting plants in other regions, distance and availability of alternative sequestration sites, and other factors. Similarly, further improvements in the process could alter its cost relative to amine-based systems. Models created in the IECM can be used to explore these and other options.

## Chapter 8. References

Aspentech, "Rate-Based Model of the CO<sub>2</sub> Capture Process by NH<sub>3</sub> Using Aspen Plus" Solution ID 129521, <http://www.aspentech.com>, Accessed September 13, 2010.

Bai H., Yeh A.C., "Removal of CO<sub>2</sub> Greenhouse Gas by Ammonia Scrubbing", *Industrial & Engineering Chemistry Research*, Vol. 36, pp 2490-2493, 1997

Berkenpas M.B., Frey H.C., Fry J.J., Kalagnanam J., and Rubin E.S. "Technical Documentation: Integrated Environmental Control Model", 1999, [http://www.cmu.edu/epp/iecm/IECM\\_Publications/1999ra%20Berkenpas%20et%20al,%20IECM%20Tech.pdf](http://www.cmu.edu/epp/iecm/IECM_Publications/1999ra%20Berkenpas%20et%20al,%20IECM%20Tech.pdf), Accessed May 16, 2011

Bollinger R., Muraskin D., Hammond M., Kozak F., Spitznogle G., Cage M., Sherrick B., Varner M., "CCS Project with Alstom's Chilled Ammonia Process at AEP's Mountaineer Plant", Alstom Power Systems, Technical Paper No.72, [http://secure.awma.org/presentations/Mega08/Papers/a167\\_1.pdf](http://secure.awma.org/presentations/Mega08/Papers/a167_1.pdf), Accessed November 15, 2010

Bottoms, R.R., "Separating Acid Gases", U.S. Patent 1,753,901, 1930

Buchanan T., DeLallo M., Schoff R., White J., "Evaluation of Innovative Fossil Fuel Power Plants with CO<sub>2</sub> Removal" EPRI, Report No. 1000316, 2000

Center for American Progress, "Issues, Energy and Environment, Carbon Capture and Sequestration 101", 2009, [http://www.americanprogress.org/issues/2009/03/ccs\\_101.html](http://www.americanprogress.org/issues/2009/03/ccs_101.html), Accessed March 1, 2012

Clean Air Task Force (CATF), "Coal without Carbon: An Investment Plan for Federal Action" Boston, MA, 2009, [http://www.catf.us/publications/reports/Coal\\_Without\\_Carbon.pdf](http://www.catf.us/publications/reports/Coal_Without_Carbon.pdf), 2009, Accessed April 30, 2010

Chang T., "Developing Chemical Additives for Aqueous Ammonia to Reduce CO<sub>2</sub> Capture Cost", Annual NETL CO<sub>2</sub> Capture Technology for Existing Plants R&D Meeting, Pittsburgh PA USA, March 24-26, 2009

Chen C., Rubin E.S., "CO<sub>2</sub> Control Technology Effects on IGCC Plant Performance and Cost", *Energy Policy*, Vol. 37, pp 915-924, 2009

Chung T., Patino-Echeverri D., "Expert Assessments of CO<sub>2</sub> Capture Technologies, Retrofitting Existing Coal-fired Power Plants", 8th Annual Conference on Carbon Capture & Sequestration, Pittsburgh PA USA, May 4-7, 2009

Ciferno J.P., DiPietro P., Tarka T., "An Economic Scoping Study for CO<sub>2</sub> Capture Using Aqueous Ammonia", National Energy Technology Laboratory, 2005, <http://www.transactionsmagazine.com/ArgonneLabCommonSense.pdf>, Accessed April 2, 2010

Creys J., Derkach A., Nyquist S., Ostrowski K., Stephenson J. “Reducing U.S. Greenhouse Gas Emissions: How Much at What Cost?”, McKinsey & Company, 2007

Darde V., van Well W. J.M., Stenby E.H., Thomsen K., “CO<sub>2</sub> Capture Using Aqueous Ammonia: Kinetic Study and Process Simulation”, Energy Procedia, Vol. 4, pp 1443-1450, 2011

Derks P.W.J., Versteeg G.F., “Kinetics of Absorption of Carbon Dioxide in Aqueous Ammonia Solutions”, Energy Procedia, Vol. 1, pp 1139-1146, 2009

Department of Energy National Energy Technology Laboratory (DOE/NETL), “How to Buy an Energy-Efficient Water-Cooled Electric Chiller”, Federal Energy Management Program, 2004, [http://www1.eere.energy.gov/femp/pdfs/wc\\_chillers.pdf](http://www1.eere.energy.gov/femp/pdfs/wc_chillers.pdf), Accessed July 28, 2011

Department of Energy National Energy Technology Laboratory (DOE/NETL, 2007a) Woods M.C., Capicotto P., Haslbeck J., Kuehn N., Matuszewski M., Pinkerton L. L., Rutkowski M.D., Schoff R. L., Vaysman V., “Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity Final Report”, National Energy Technology Laboratory, 2007

Department of Energy National Energy Technology Laboratory (DOE/NETL, 2007b) “Carbon Dioxide Capture from Existing Coal-Fired Power Plants” November 2007.

Department of Energy National Energy Technology Laboratory (DOE/NETL, 2010a), Ciferno J.P., Plasynski S.I., “DOE/NETL Advanced Carbon Dioxide Capture R&D Program: Technology Update”, September 2010

Department of Energy National Energy Technology Laboratory (DOE/NETL, 2010b). “NETL Carbon Capture and Storage Database,” 2010, [http://www.netl.doe.gov/technologies/carbon\\_seq/database/index.html](http://www.netl.doe.gov/technologies/carbon_seq/database/index.html), Accessed March 7, 2010.

Department of Energy National Energy Technology Laboratory (DOE/NETL, 2010c) Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2, US National Energy Technology Laboratory, 2010

Energy Information Administration (EIA, 2010a), “Electric Power Annual 2010”, November 2010, <http://205.254.135.7/electricity/annual/pdf/epa.pdf>, Accessed April 1, 2012

Energy Information Administration (EIA, 2010b), “1990 - 2010 Existing Nameplate and Net Summer Capacity by Energy Source, Producer Type and State (EIA-860)”, 2010

Environment Canada (Environment Canada, 2011a), “Facility and GHG Information”, [http://www.ec.gc.ca/pdb/ghg/onlinedata/facility\\_info\\_e.cfm?ghg\\_id=G10198&year=2009](http://www.ec.gc.ca/pdb/ghg/onlinedata/facility_info_e.cfm?ghg_id=G10198&year=2009), Accessed February 21, 2011, 2011

Environment Canada (Environment Canada, 2011b), “National Greenhouse Gas Emissions, 1990 to 2009”, <http://www.ec.gc.ca/indicateurs-indicators/default.asp?lang=en&n=FBF8455E-1>, Accessed February 21, 2011, 2011

Environmental Protection Agency (EPA), “Environmental Footprints and Costs of Coal-Based Integrated Gasification Combined Cycle and Pulverized Coal Technologies”, EPA Report 430-R-06-006, 2006

Environmental Protection Agency (EPA), “Inventory of U.S. Greenhouse Gas Emissions and Sinks: 1990-2008”, EPA Report 430-R-10-006, 2010

Environmental Protection Agency (EPA), “Endangerment and Cause or Contribute Findings for Greenhouse Gases under Section 202(a) of the Clean Air Act”, <http://www.epa.gov/climatechange/endangerment.html>, Accessed April 30, 2011

Electric Power Research Institute (EPRI), “Technical Assessment Guide”, Palo Alto, CA, 1993.

Electric Power Research Institute (EPRI), “Advanced Coal Power Systems with CO<sub>2</sub> Capture: EPRI’s CoalFleet for Tomorrow Vision” Report No.1016877, Palo Alto, CA, September 2008

Electric Power Research Institute (EPRI), “Post-Combustion CO<sub>2</sub> Capture Technology Development”, Report No. 10117644, Technical update, Palo Alto, CA, December 2009.

Feron P.H.M, "Progress in SP2 CO<sub>2</sub> Post-combustion Capture," Presentation at the ENCAP/CASTOR Seminar, March 2006.

Figueroa J.D., Fout T., Plasynski S., McIlvried H., Srivastava R.D., “Advances in CO<sub>2</sub> capture technology—The U.S. Department of Energy’s Carbon Sequestration Program,” International Journal of Greenhouse Gas Control, Vol. 2, pp. 9-20, 2008.

Frey H.C., Rubin E.S., “Modelling IGCC System Performance, Emissions, and Cost Using Probabilistic Engineering Models”, Proceedings of the Eighth Annual International Pittsburgh Coal Conference, Pittsburgh PA USA, October 14-18, 1991.

Frey H.C., Rubin, E.S., “Evaluation of Advanced Coal Gasification Combined-Cycle Systems under Uncertainty”, Industrial & Engineering Chemistry Research, Vol. 31, pp 1299-1307, 1992

Gal E., “Chilled-Ammonia Post Combustion CO<sub>2</sub> Capture System - Laboratory and Economic Evaluation Results”, EPRI, Report No. 1012797, 2006

Gal E., “Ultra Cleaning of Combustion Gas Including Removal of CO<sub>2</sub>” U.S. Patent Application No. US 2008/0072762 A1, Publication Date March 27, 2008.

Global Carbon Capture and Storage Institute (GCCSI), “Strategic Analysis of the Global Status of Carbon Capture and Storage”, Prepared by WorleyParsons, 2009, <http://www.globalccsinstitute.com/publications/strategic-analysis-global-status-carbon-capture-storage>, Accessed April 1, 2012

Gillingham K., van Benthem A., Sweeney J.J., “Learning by Doing and the California Solar Initiative”, In “Issues of the Day: 100 Commentaries on Climate, Energy, the Environment, Transportation, and Public Health Policy”, Edited by Perry I.W.H., and Day F., Resources for the Future, 2010, 229 pages

Grant P.M., "Plugged into the matrix: the rise and potential fall of the US electricity grid", *Nature*, Vol. 447, 2007, pp 145-146

Greenpeace USA, /<http://www.greenpeace.org/>, Accessed March 20, 2009

Haszeldine R.S., "Carbon Capture and Storage: How Green Can Black Be?", *Science*, Vol. 325, pp 1647-1652, 2009

Hilton R. G., "Mountaineer CCS Project Begins Operations" Alstom Webinar presentation, September 11, 2009

International Energy Agency Greenhouse Gas R&D Programme (IEAGHG), "CO<sub>2</sub> Capture and Storage," 2010, <http://www.co2captureandstorage.info/co2db.php>, Accessed April 3, 2010.

Integrated Environmental Control Model (IECM V6.2), Carnegie Mellon University, <http://www.iecm-online.com/> Accessed April 1, 2012. This is the public release of this model as of the access date.

Integrated Environmental Control Model (IECM V7.0), Carnegie Mellon University, <http://www.iecm-online.com/> Accessed April 1, 2012. This is the beta release of this model as of the access date.

Intergovernmental Panel on Climate Change (IPCC), "Climate Change 2007: Synthesis Report". In: Contribution of Working Groups I, II and III to the Fourth Assessment Report of the Intergovernmental Panel on Climate Change [Solomon, S., D. Qin, M. Manning, Z. Chen, M. Marquis, K.B. Averyt, M. Tignor and H.L. Miller (eds.)]. Cambridge University Press, Cambridge, United Kingdom and New York, NY, USA, 2007

Interagency Task Force on Carbon Capture and Sequestration (ITFCCS), "Report of the Interagency Task Force on Carbon Capture and Storage", August 2010, <http://www.epa.gov/climatechange/downloads/ES-CCS-Task-Force-Report-2010.pdf>, Accessed April 1, 2012

Jensen J.N., Knudsen J.N., "Experience with the CASTOR/CESAR Pilot Plant," Presentation at the Workshop on Operating Flexibility of Power Plants with CCS, Imperial College, London, November 2009.

Kishimoto S., Hirata T., Iijima M., Oshishi T., Higaki K., Mitchell R., "Current Status of MHI's CO<sub>2</sub> Recovery Technology and Optimization of CO<sub>2</sub> Recovery Plant with a PC Fired Power Plant", *Energy Procedia*, Vol. 1, pp 1091-1098, 2009

Kohl A.L., Nielsen R.B., "Gas Purification", Houston, TX, USA: Gulf Publishing Company, 1375 pages, 1997

Lee D.H., Choi W.J., Moon S.J., Ha S.H., Kim I.G., Oh K.J., "Characteristics of Absorption and Regeneration of Carbon Dioxide in Aqueous 2-Amino-2-Methyl-1-Propanol/Ammonia Solutions", *Korean Journal of Chemical Engineering*, Vol. 25, pp 279-284, 2008

Lucquiaud M., Gibbins J., "Retrofitting CO<sub>2</sub> capture ready fossil plants with post-combustion capture. Part 1: requirements for supercritical pulverized coal plants using solvent-based flue gas

scrubbing”, Proceedings of the IMechE, Part A: Journal of Power and Energy, Vol. 223, pp 213-226, 2009

“Marshall & Swift Equipment Cost Index.” Chemical Engineering, Vol. 116, pp 64, 2009

Massachusetts Institute of Technology (MIT), “The Future of Coal”, ISBN 978-0-615-14092-6, 192 pages, 2007

Massachusetts Institute of Technology (MIT), “Carbon Capture and Sequestration Technologies at MIT”, Massachusetts Institute of Technology Energy Initiative, <http://sequestration.mit.edu>, Accessed March, 2010

Massachusetts Institute of Technology (MIT), “The Future of the Electric Grid”, ISBN 978-0-9828008-6-7, 280 pages, 2011

Mathias P.M., Reddy S., Connell J. P., “Quantitative Evaluation of the Aqueous-Ammonia Process for CO<sub>2</sub> Capture Using Fundamental Data and Thermodynamic Analysis”, Energy Procedia, Vol. 1, No. 1, pp 1227-1234, 2009

McLarnon C.R., Jones M.D., “Testing of Ammonia Based CO<sub>2</sub> Capture with Multi-Pollutant Control Technology,” Proceedings of the 9th International Conference on Greenhouse Gas Control Technologies, 2008

The National Academies Press (NAP, 2011a), “Understanding Earth’s Deep Past: Lessons for Our Climate Future”, Committee on the Importance of Deep-Time Geologic Records for Understanding Climate Change Impacts; National Research Council of the National Academies, ISBN 978-0-309-20915-1, 212 pages, 2011

The National Academies Press (NAP, 2011b), “America's Climate Choices”, Committee on America's Climate Choices; National Research Council of the National Academies, ISBN: 0-309-14586-4, 144 pages, 2011

Peltier R., “Alstom’s chilled ammonia CO<sub>2</sub>-capture process advances toward commercialization”, Power, [http://www.powermag.com/environmental/Alstoms-chilled-ammonia-CO2-capture-process-advances-toward-commercialization\\_86.html](http://www.powermag.com/environmental/Alstoms-chilled-ammonia-CO2-capture-process-advances-toward-commercialization_86.html), February 15, 2008, Accessed December 1, 2009

Pew Center, “A Program to Accelerate the Deployment of CO<sub>2</sub> Capture and Storage (CCS): Rationale, Objectives and Costs”, October, <http://www.pewclimate.org/docUploads/CCS-Deployment.pdf>, 2007, Accessed April 30, 2010

Phase Change Products Ltd. (PCP) “PC0 Technical Data Sheet”, [http://www.pcpenergy.com.au/pc\\_0.html](http://www.pcpenergy.com.au/pc_0.html), Accessed July 27, 2011

Platts, “HVAC: Centrifugal Chillers”, 2004, [http://www.reliant.com/en\\_US/Platts/PDF/P\\_PA\\_14.pdf](http://www.reliant.com/en_US/Platts/PDF/P_PA_14.pdf), Accessed July 13, 2011

Powerspan Website (Powerspan, 2012), <http://powerspan.com/corporate-profile/>, Accessed April 1, 2012.

- Qin F., Wang S., Hartono A., Svendsen H.F., Chen C. “Kinetics of CO<sub>2</sub> absorption in aqueous ammonia solution”, *International Journal of Greenhouse Gas Control*, Vol. 4, pp 729-738, 2010
- Qin F., Wang S., Kim I., Svendsen H. F., Chen C., “Heat of Absorption of CO<sub>2</sub> in Aqueous Ammonia and Ammonium Carbonate/Carbamate Solutions”, *International Journal of Greenhouse Gas Control*, Vol. 5, pp 405-412, 2011
- Rao A., “Details of a Technical, Economic and Environmental Assessment of Amine-based CO<sub>2</sub> Capture Technology for Power Plant Greenhouse Gas Control”, Appendix to Annual Technical Progress Report for USDOE contract No. DE-FC26-00NT40935, 2002
- Resource Dynamics Corporation (RDC) “Cooling, Heating, and Power for Industry: A Market Assessment”, 2003
- Resnik K.P., Yeh J. T., Pennline H.W., “Aqua Ammonia Process for Simultaneous Removal of CO<sub>2</sub>, SO<sub>2</sub> and NO<sub>x</sub>”, *International Journal of Environmental Technology and Management*, Vol. 4, Nos. 1/2, pp 89-103, 2004
- Richardson N., Fraas A., Burtraw D. “Discussion Paper: Greenhouse Gas Regulation under the Clean Air Act”, *Resources for the Future*, DP 10-23, April 2010
- Roberts C. A., Gibbins J., Panesar R., Kelsall G. “Potential for Improvement in Power Generation with Post Combustion Capture of CO<sub>2</sub>” 7th international conference on greenhouse gas control technologies, Vancouver, 5 - 9 September, 2004
- Rochelle G., “Amine Scrubbing for CO<sub>2</sub> Capture,” *Science*, Vol. 325, No.25, pp 1652-1654, 2009
- Rose J., Collison K., Parmar H., “Retiring Coal Plants While Protecting System Reliability”, White Paper, ICF International, pp 1-5, 2011
- Rubin E.S., Marks A., Mantripragada H., Versteeg P., Kitchin J., “Prospects for Improved Carbon Capture Technology”, July 2010
- Rubin E.S., Zhai H., “The Cost of CCS for Natural Gas-Fired Power Plants”, 10<sup>th</sup> Annual Conference on Carbon Capture and Storage, Pittsburgh PA USA, May 2-5, 2011
- Schaeffer M., Kram T., Meinshausen M., van Vuuren D.P., Hare W.L., “Near-linear cost increase to reduce climate change risk,” *Proceedings of the National Academy of Sciences*, Vol. 105, No. 52, pp 20621–20626, 2008
- Soper D., “A-priori Sample Size Calculator for Multiple Regression”, DanielSoper.com, website: <http://www.danielsoper.com/statcalc/calc01.aspx>, Accessed October 23, 2010.
- Sourcewatch, "Southern Company abandons carbon capture and storage project," 2010, [http://www.sourcewatch.org/index.php?title=Southern\\_Company#cite\\_note-11](http://www.sourcewatch.org/index.php?title=Southern_Company#cite_note-11), Accessed April 1, 2012
- Statistics Canada, “Electric Power Generation, Transmission, and Distribution”, Manufacturing and Energy Division, Government of Canada, 57-202-X, 2009

Stephens J.C., Justo S. "Assessing innovation in emerging energy technologies: Socio-technical dynamics of carbon capture and storage (CCS) and enhanced geothermal systems (EGS) in the USA", *Energy Policy*, Vol. 38, pp 2020-2031, 2010

Stultz S.C., Kitto J.B., "Steam: Its Generation and Use", Babcock & Wilcox Company, 41<sup>st</sup> Edition, 1064 pages, 2005.

Valenti G., Bonalumi D., Macci E., "Energy and exergy analyses for the carbon capture with the Chilled Ammonia Process", *Energy Procedia*, Vol. 1, No. 1, pp 1059-1066, 2009

van Straelen J., Geuzebroek F., Goodchild N., Protopapas G., Mahony L., "CO<sub>2</sub> capture for refineries, a practical approach", *International Journal of Greenhouse Gas Control*, Vol. 4, pp 316-320, 2010

Versteeg P.L., Rubin E.S., "Technical and Economic Assessment of Ammonia-Based Post-Combustion CO<sub>2</sub> Capture", *Energy Procedia*, Vol. 4, pp 1957-1964, 2011

Versteeg, P.L., Rubin, E.S., "A technical and economic assessment of ammonia-based post-combustion CO<sub>2</sub> capture at coal-fired power plants", *International Journal of Greenhouse Gas Control*, Vol. 5, 1596-1605, 2011

Wise M.A., Dooley J.J., "The value of post-combustion carbon dioxide capture and storage technologies in a world with uncertain greenhouse gas emissions constraints," *International Journal of Greenhouse Gas Control*, Vol. 3, 39-48, 2009

European Technology Platform for Zero Emission Fossil Fuel Power Plants (ZEP), "EU Demonstration Programme for CO<sub>2</sub> Capture and Storage (CCS)," November 2008, <http://www.zeroemissionsplatform.eu/index>, Accessed March 5, 2010.

Zhuang Q., "Comparative Kinetics of Ammonia-Based CO<sub>2</sub> Capture and Amine CO<sub>2</sub> Capture Technologies: Absorber Sizing", 10<sup>th</sup> Annual Conference on Carbon Capture & Sequestration, Pittsburgh PA USA, May 2-5, 2011

## Appendix A. Case Study Generation

### Case Study No.1 Subcritical PC plants without CO<sub>2</sub> capture

For the Subcritical plant without a CO<sub>2</sub> Capture System, a number of default parameters in the IECM were changed to generate Case No.1. These changes are shown in Table A.1.

**Table A.1: IECM parameters changed for Case 1: subcritical plants without CO<sub>2</sub> capture**

	Parameter	IECM Default	IECM Case No.1, Non Calculated Default	
Configure Plant Overall Plant	NO <sub>x</sub> Control	None	Hot-Side SCR	
	Particulates	None	Fabric Filter	
	SO <sub>2</sub> Control	None	Wet FGD	
	Cooling System	Once-Through	Wet Cooling Tower	
Fuel Properties	Fuel Name	Appalachian Medium Sulfur	Illinois #6	
	Fuel Cost (\$/GJ)	1.617	1.705	
Overall Plant Performance	Capacity Factor (%)	75.00	75.00	
	Ambient Air Temperature (Avg.) (°C)	25.0	21.1	
Overall Plant Financing	Year Costs Reported	2010	2007	
	Discount Rate (Before Taxes) (fraction)	0.0709	1.0E-4	
	Fixed Charge Factor (FCF) (fraction)	0.1128	0.1130	
	Gross Electrical Output (MWg)	650	583.2	
Base Plant Performance	Unit Type:	Supercritical	Subcritical	
	Boiler Efficiency (%)	89.64	89.00	
	Leakage Air at Preheater (% stoich.)	10.0	5.5	
	Coal Pulverizer (% MWg)	0.4840	0.5800	
	General Facilities Capital (%PFC)	10.00	2.515	
	Engineering & Home Office Fees (%PFC)	6.50	9.3	
Base Plant Capital Cost	Project Contingency Cost (%PFC)	11.67	13.8	
	Process Contingency Cost (%PFC)	0.3	0.0	
	Royalty Fees (%PFC)	7.00E-2	0.0	
	Fixed Operating Cost (months)	1.00	0.0	
	Variable Operating Cost (months)	1.00	0.0	
	Misc. Capital Cost (%TPI)	2.0	0.0	
	Inventory Capital (%TPC)	6.00E-2	0.0	
	Base Plant O&M Cost	Waste Disposal Cost (\$/tonne)	10.32	17.03
		Total Maintenance Cost (%TPC)	1.896	1.60
	NO <sub>x</sub> Control Performance	Actual NO <sub>x</sub> Removal Efficiency (%)	76.66	86.00
Hot-Side SCR Power Requirement (% MWg)		0.5088	1.00e-2	
NO <sub>x</sub> Control Capital Cost	General Facilities Capital (%PFC)	10.00	2.515	
	Engineering & Home Office Fees (%PFC)	10.00	9.30	

	Project Contingency Cost (%PFC)	10.00	13.80
	Process Contingency Cost (%PFC)	6.495	0.0
	Months of Fixed O&M (months)	1	0
	Months of Variable O&M (months)	1	0
	Misc. Capital Cost (%TPI)	2	0
	Inventory Capital (%TPC)	0.5	0
NO <sub>x</sub> Control O&M Cost	Total Maintenance Cost (%TPC)	2.00	1.60
TSP Control Performance	Particulate Removal Efficiency (%)	99.32	99.90
	Fabric Filter Power Requirement (% MWg)	0.331	1.70E-2
TSP Control Capital Cost	General Facilities Capital (%PFC)	1.00	2.515
	Engineering & Home Office Fees (%PFC)	5	9.30
	Project Contingency Cost (%PFC)	20	13.80
	Process Contingency Cost (%PFC)	0	0.0
	Royalty Fees (%PFC)	0	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.0	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.5	0.0
TSP Control O&M Cost	Total Maintenance Cost (%TPC)	1.124	1.600
SO <sub>2</sub> Control Performance	Scrubber SO <sub>2</sub> Removal Efficiency (%)	80.86	98.00
	Wet FGD Power Requirement (% MWg)	1.594	0.71
SO <sub>2</sub> Control Capital Cost	General Facilities Capital (%PFC)	10	2.515
	Engineering & Home Office Fees (%PFC)	10	9.30
	Project Contingency Cost (%PFC)	15	13.80
	Process Contingency Cost (%PFC)	2	0.0
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.2964	0.0
SO <sub>2</sub> Control O&M Cost	Total Maintenance Cost (%TPC)	4.325	1.60
Water Systems Performance	Ambient Air Temp (Dry Bulb Avg.) (°C)	25.0	21.1
	Power Requirement (% MWg)	1.250	1.410
Water Systems Capital Cost	General Facilities Capital (%PFC)	10	2.515
	Engineering & Home Office Fees (%PFC)	10	9.30
	Project Contingency Cost (%PFC)	15	13.80
	Process Contingency Cost (%PFC)	0	0.0
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.0	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.5	0.0
Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

## Case Study #2: PC Supercritical without CO<sub>2</sub> Capture

For the Subcritical plant without a CO<sub>2</sub> Capture System, a number of default parameters in the IECM were changed to generate Case No.3. These changes are shown in Table A.2.

**Table A.2: IECM parameters changed for Case 2: supercritical plants without CO<sub>2</sub> capture**

	Parameter	IECM Default	IECM Case No.2, Non Calculated Default
Configure Plant Overall Plant	NO <sub>x</sub> Control	None	Hot-Side SCR
	Particulates	None	Fabric Filter
	SO <sub>2</sub> Control	None	Wet FGD
	Cooling System	Once-Through	Wet Cooling Tower
Fuel Properties	Fuel Name	Appalachian Medium Sulfur	Illinois #6
	Fuel Cost (\$/GJ)	1.617	1.705
Overall Plant Performance	Capacity Factor (%)	75.00	75.00
	Ambient Air Temperature (Avg.) (°C)	25.0	21.1
Overall Plant Financing	Year Costs Reported	2010	2007
	Discount Rate (Before Taxes) (fraction)	0.0709	1.0E-4
	Fixed Charge Factor (FCF) (fraction)	0.1128	0.1130
Base Plant Performance	Gross Electrical Output (MWg)	650	581.3
	Unit Type:	Supercritical	Supercritical
	Boiler Efficiency (%)	89.64	89.00
	Leakage Air at Preheater (% stoich.)	10.0	5.5
	Coal Pulverizer (% MWg)	0.4840	0.5500
Base Plant Capital Cost	General Facilities Capital (%PFC)	10.00	2.515
	Engineering & Home Office Fees (%PFC)	6.50	9.3
	Project Contingency Cost (%PFC)	11.67	13.8
	Process Contingency Cost (%PFC)	0.3	0.0
	Royalty Fees (%PFC)	7.00E-2	0.0
	Fixed Operating Cost (months)	1.00	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	6.00E-2	0.0
	Base Plant O&M Cost	Waste Disposal Cost (\$/tonne)	10.32
Total Maintenance Cost (%TPC)		1.896	1.60
NO <sub>x</sub> Control Performance	Actual NO <sub>x</sub> Removal Efficiency (%)	76.66	86.00
	Hot-Side SCR Power Requirement (% MWg)	0.5088	1.00e-2
NO <sub>x</sub> Control Capital Cost	General Facilities Capital (%PFC)	10.00	2.515
	Engineering & Home Office Fees (%PFC)	10.00	9.30
	Project Contingency Cost (%PFC)	10.00	13.80
	Process Contingency Cost (%PFC)	6.495	0.0

	Months of Fixed O&M (months)	1	0
	Months of Variable O&M (months)	1	0
	Misc. Capital Cost (%TPI)	2	0
	Inventory Capital (%TPC)	0.5	0
NO <sub>x</sub> Control O&M Cost	Total Maintenance Cost (%TPC)	2.00	1.60
TSP Control Performance	Particulate Removal Efficiency (%)	99.32	99.90
	Fabric Filter Power Requirement (% MWg)	0.331	1.70E-2
TSP Control Capital Cost	General Facilities Capital (%PFC)	1.00	2.515
	Engineering & Home Office Fees (%PFC)	5	9.30
	Project Contingency Cost (%PFC)	20	13.80
	Process Contingency Cost (%PFC)	0	0.0
	Royalty Fees (%PFC)	0	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.0	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.5	0.0
TSP Control O&M Cost	Total Maintenance Cost (%TPC)	1.124	1.600
SO <sub>2</sub> Control Performance	Scrubber SO <sub>2</sub> Removal Efficiency (%)	80.86	98.00
	Wet FGD Power Requirement (% MWg)	1.594	0.71
SO <sub>2</sub> Control Capital Cost	General Facilities Capital (%PFC)	10	2.515
	Engineering & Home Office Fees (%PFC)	10	9.30
	Project Contingency Cost (%PFC)	15	13.80
	Process Contingency Cost (%PFC)	2	0.0
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.2964	0.0
SO <sub>2</sub> Control O&M Cost	Total Maintenance Cost (%TPC)	4.325	1.60
Water Systems Performance	Ambient Air Temp (Dry Bulb Avg.) (°C)	25.0	21.1
	Power Requirement (% MWg)	1.250	1.410
Water Systems Capital Cost	General Facilities Capital (%PFC)	10	2.515
	Engineering & Home Office Fees (%PFC)	10	9.30
	Project Contingency Cost (%PFC)	15	13.80
	Process Contingency Cost (%PFC)	0	0.0
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.0	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.5	0.0
Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

### Case Study #3: NGCC Plant without CO<sub>2</sub> Capture

For the NGCC plant without a CO<sub>2</sub> Capture System, a number of default parameters in the IECM were changed to generate Case No.3. These changes are shown in Table A.3.

**Table A.3: IECM parameters changed for Case 3: NGCC plants without CO<sub>2</sub> capture**

Parameter		IECM Default	IECM Case No.3, Non Calculated Default
Configure Plant Overall	CO <sub>2</sub> Capture	None	None
Plant	Cooling System	Once-Through	Wet Cooling Tower
Overall Plant	Capacity Factor (%)	75.00	75.00
Performance	Ambient Air Temperature (Avg.) (°C)	25.0	21.1
Overall Plant Financing	Year Costs Reported	2010	2007
	Discount Rate (Before Taxes) (fraction)	0.0709	1.0E-4
	Fixed Charge Factor (FCF) (fraction)	0.1128	0.1130
Fuel Properties	Fuel Name	Natural Gas (93.1% CH <sub>4</sub> )	Natural Gas (93.9% CH <sub>4</sub> )
	Fuel Cost (\$/GJ)	6.5	6.4
Power Block Capital Cost	General Facilities Capital (%PFC)	15.00	2.515
	Engineering & Home Office Fees (%PFC)	10.00	9.300
	Project Contingency Cost (%PFC)	15.00	13.80
	Process Contingency Cost (%PFC)	7.501	0.0
	Royalty Fees (%PFC)	0.5000	0.0
	Months of Fixed O&M (Months)	1.000	0.0
	Months of Variable O&M (Months)	1.000	0.0
	Misc. Capital Cost (%TPI)	2.000	0.0
	Inventory Capital (%TPC)	0.5000	0.0
Water Systems	Ambient Air Temp (Dry Bulb Avg.) (°C)	25.0	21.1
Performance	Power Requirement (% MWg)	1.250	1.410
Water Systems Capital	General Facilities Capital (%PFC)	10	2.515
Cost	Engineering & Home Office Fees (%PFC)	10	9.30
	Project Contingency Cost (%PFC)	15	13.80
	Process Contingency Cost (%PFC)	0	0.0
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.0	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.5	0.0
Water Systems O&M	Total Maintenance Cost (%TPC)	2.0	1.600
Cost			

## Case Study #4: PC Subcritical with Ammonia-Based CO<sub>2</sub> Capture

For the subcritical plant with an ammonia-based CO<sub>2</sub> capture system, a number of default parameters in the IECM were changed so that these parameters matched the NETL 2007 Baseline Report. These changes are shown in Table A.4.

**Table A.4: IECM parameters changed for Case 4: subcritical plants with ammonia-based CO<sub>2</sub> capture**

Parameter		IECM Default	IECM Case 4
Configure Plant	NO <sub>x</sub> Control	None	Hot-Side SCR
Overall Plant	Particulates	None	Fabric Filter
	SO <sub>2</sub> Control	None	Wet FGD
	CO <sub>2</sub> Capture	None	Ammonia System
	Cooling System	Once-Through	Wet Cooling Tower
	Fuel	Fuel Name	Appalachian Medium
Properties		Sulfur	
	Fuel Cost (\$/GJ)	1.617	1.705
Overall Plant Performance	Capacity Factor (%)	75.00	75.00
	Ambient Air Temperature (Avg.) (°C)	25.0	21.1
Overall Plant Financing	Year Costs Reported	2010	2007
	Discount Rate (Before Taxes) (fraction)	7.09e-2	1.0E-4
	Fixed Charge Factor (FCF) (fraction)	0.1128	0.1430
Base Plant Performance	Gross Electrical Output (MWg)	650	582.8
	Unit Type:	Supercritical	Subcritical
	Boiler Efficiency (%)	89.64	89.00
	Leakage Air at Preheater (% stoich.)	10	5.5
	Coal Pulverizer (% MWg)	0.4840	0.5800
Base Plant Capital Cost	General Facilities Capital (%PFC)	10.00	1.570
	Engineering & Home Office Fees (%PFC)	6.50	9.37
	Project Contingency Cost (%PFC)	11.67	16.38
	Process Contingency Cost (%PFC)	0.3	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Fixed Operating Cost (months)	1.00	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	6.00E-2	0.0
Base Plant O&M Cost	Waste Disposal Cost (\$/tonne)	10.32	17.03
	Total Maintenance Cost (%TPC)	1.896	1.60
NO <sub>x</sub> Control Performance	Actual NO <sub>x</sub> Removal Efficiency (%)	76.66	86.00
	Hot-Side SCR Power Requirement (% MWg)	0.5088	1.100e-2
NO <sub>x</sub> Control Capital Cost	General Facilities Capital (%PFC)	10.00	1.57
	Engineering & Home Office Fees (%PFC)	10.00	9.37
	Project Contingency Cost (%PFC)	10.00	16.38
	Process Contingency Cost (%PFC)	6.397	4.670

	Royalty Fees (%PFC)	7.00E-2	0.0
	Months of Fixed O&M (months)	1	0
	Months of Variable O&M (months)	1	0
	Misc. Capital Cost (%TPI)	2	0
	Inventory Capital (%TPC)	0.5	0
NO <sub>x</sub> Control O&M Cost	Total Maintenance Cost (%TPC)	2.00	1.60
TSP Control Performance	Particulate Removal Efficiency (%)	99.32	99.90
	Fabric Filter Power Requirement (% MWg)	0.331	1.70E-2
TSP Control Capital Cost	General Facilities Capital (%PFC)	1.00	1.57
	Engineering & Home Office Fees (%PFC)	5	9.37
	Project Contingency Cost (%PFC)	20	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
TSP Control O&M Cost	Total Maintenance Cost (%TPC)	1.124	1.600
SO <sub>2</sub> Control Performance	Scrubber SO <sub>2</sub> Removal Efficiency (%)	80.86	98.00
	Wet FGD Power Requirement (% MWg)	1.607	0.83
SO <sub>2</sub> Control Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	2	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.3545	0
SO <sub>2</sub> Control O&M Cost	Total Maintenance Cost (%TPC)	4.398	1.60
CO <sub>2</sub> Capture Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	7	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	5	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
CO <sub>2</sub> Capture O&M Cost	Reclaimer Waste Disposal Cost (\$/tonne)	183.4	0.0
	Total Maintenance Cost (%TPC)	2.5	1.600
	CO <sub>2</sub> Transportation Cost (\$/tonne)	1.684	0.0
	CO <sub>2</sub> Storage Cost (\$/tonne)	3.145	3.75
Water Systems Performance	Ambient Air Temp (Dry Bulb Avg.) (°C)	25.0	21.1
	Power Requirement (% MWg)	2.8	2.800
Water Systems	General Facilities Capital (%PFC)	10	1.57

Capital Cost	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

### Case Study #5: PC Supercritical with Ammonia-Based CO<sub>2</sub> Capture

For the supercritical plant with an ammonia-based CO<sub>2</sub> capture system, a number of default parameters in the IECM were changed so that these parameters matched the NETL 2007 Baseline Report. These changes are shown in Table A.5.

**Table A.5: IECM parameters changed for Case 5: supercritical plants with ammonia-based CO<sub>2</sub> capture**

Parameter		IECM Default	IECM Case 5
Configure Plant	NO <sub>x</sub> Control	None	Hot-Side SCR
Overall Plant	Particulates	None	Fabric Filter
	SO <sub>2</sub> Control	None	Wet FGD
	CO <sub>2</sub> Capture	None	Ammonia System
	Cooling System	Once-Through	Wet Cooling Tower
	Fuel	Fuel Name	Appalachian Medium
Fuel Properties		Sulfur	
	Fuel Cost (\$/GJ)	1.617	1.705
Overall Plant	Capacity Factor (%)	75.00	75.00
Performance	Ambient Air Temperature (Avg.) (°C)	25.0	21.1
Overall Plant	Year Costs Reported	2010	2007
Financing	Discount Rate (Before Taxes) (fraction)	7.09e-2	1.0E-4
	Fixed Charge Factor (FCF) (fraction)	0.1128	0.1430
Base Plant Performance	Gross Electrical Output (MWg)	650	582.8
	Unit Type:	Supercritical	Supercritical
	Boiler Efficiency (%)	89.64	89.00
	Leakage Air at Preheater (% stoich.)	10	5.5
	Coal Pulverizer (% MWg)	0.4840	0.5800
Base Plant Capital Cost	General Facilities Capital (%PFC)	10.00	1.570
	Engineering & Home Office Fees (%PFC)	6.50	9.37
	Project Contingency Cost (%PFC)	11.67	16.38
	Process Contingency Cost (%PFC)	0.3	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Fixed Operating Cost (months)	1.00	0.0

	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	6.00E-2	0.0
Base Plant O&M Cost	Waste Disposal Cost (\$/tonne)	10.32	17.03
	Total Maintenance Cost (%TPC)	1.896	1.60
NO <sub>x</sub> Control Performance	Actual NO <sub>x</sub> Removal Efficiency (%)	76.66	86.00
	Hot-Side SCR Power Requirement (% MWg)	0.5088	1.100e-2
NO <sub>x</sub> Control Capital Cost	General Facilities Capital (%PFC)	10.00	1.57
	Engineering & Home Office Fees (%PFC)	10.00	9.37
	Project Contingency Cost (%PFC)	10.00	16.38
	Process Contingency Cost (%PFC)	6.397	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Months of Fixed O&M (months)	1	0
	Months of Variable O&M (months)	1	0
	Misc. Capital Cost (%TPI)	2	0
	Inventory Capital (%TPC)	0.5	0
NO <sub>x</sub> Control O&M Cost	Total Maintenance Cost (%TPC)	2.00	1.60
TSP Control Performance	Particulate Removal Efficiency (%)	99.32	99.90
	Fabric Filter Power Requirement (% MWg)	0.331	1.70E-2
TSP Control Capital Cost	General Facilities Capital (%PFC)	1.00	1.57
	Engineering & Home Office Fees (%PFC)	5	9.37
	Project Contingency Cost (%PFC)	20	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
TSP Control O&M Cost	Total Maintenance Cost (%TPC)	1.124	1.600
SO <sub>2</sub> Control Performance	Scrubber SO <sub>2</sub> Removal Efficiency (%)	80.86	98.00
	Wet FGD Power Requirement (% MWg)	1.607	0.83
SO <sub>2</sub> Control Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	2	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.3545	0
SO <sub>2</sub> Control O&M Cost	Total Maintenance Cost (%TPC)	4.398	1.60
CO <sub>2</sub> Capture Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	7	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	5	4.670
	Royalty Fees (%PFC)	0.5	0.0

	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
CO <sub>2</sub> Capture	Reclaimer Waste Disposal Cost (\$/tonne)	183.4	0.0
O&M Cost	Total Maintenance Cost (%TPC)	2.5	1.600
	CO <sub>2</sub> Transportation Cost (\$/tonne)	1.684	0.0
	CO <sub>2</sub> Storage Cost (\$/tonne)	3.145	3.75
Water Systems	Ambient Air Temp (Dry Bulb Avg.) (°C)	25.0	21.1
Performance	Power Requirement (% MWg)	2.8	2.800
Water Systems	General Facilities Capital (%PFC)	10	1.57
Capital Cost	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
Water Systems	Total Maintenance Cost (%TPC)	2.0	1.600
O&M Cost			

### Case Study #6: NGCC Supercritical with Ammonia-Based CO<sub>2</sub> Capture

For the NGCC plant with an ammonia-based CO<sub>2</sub> Capture System, a number of default parameters in the IECM were changed to generate Case No.3. These changes are shown in Table A.6.

**Table A.6: IECM parameters changed for Case 6: NGCC plants with ammonia-based CO<sub>2</sub> capture**

Parameter		IECM Default	IECM Case No.6, Non Calculated Default
Configure Plant Overall	CO <sub>2</sub> Capture	None	None
Plant	Cooling System	Once-Through	Wet Cooling Tower
Overall Plant	Capacity Factor (%)	75.00	75.00
Performance	Ambient Air Temperature (Avg.) (°C)	25.0	21.1
Overall Plant Financing	Year Costs Reported	2010	2007
	Discount Rate (Before Taxes) (fraction)	7.09e-2	1.0E-4
	Fixed Charge Factor (FCF) (fraction)	0.1128	0.1430
Fuel Properties	Fuel Name	Natural Gas (93.1% CH <sub>4</sub> )	Natural Gas (93.9% CH <sub>4</sub> )
	Fuel Cost (\$/GJ)	6.5	6.4
Power Block Capital Cost	General Facilities Capital (%PFC)	15.00	1.57
	Engineering & Home Office Fees (%PFC)	10.00	9.37
	Project Contingency Cost (%PFC)	15.00	16.38

	Process Contingency Cost (%PFC)	7.501	4.670
	Royalty Fees (%PFC)	0.5000	0.0
	Months of Fixed O&M (Months)	1.000	0.0
	Months of Variable O&M (Months)	1.000	0.0
	Misc. Capital Cost (%TPI)	2.000	0.0
	Inventory Capital (%TPC)	0.5000	0.0
CO <sub>2</sub> Capture System	Ammonia Concentration (wt%)	14.4	15.6 <sup>1</sup>
CO <sub>2</sub> Capture Capital Cost	General Facilities Capital (%PFC)	0	1.57
	Engineering & Home Office Fees (%PFC)	9.37	9.37
	Project Contingency Cost (%PFC)	16.38	16.38
	Process Contingency Cost (%PFC)	4.67	4.670
	Royalty Fees (%PFC)	0.0	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.0	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
CO <sub>2</sub> Capture O&M Cost	Reclaimer Waste Disposal Cost (\$/tonne)	183.4	0.0
	Total Maintenance Cost (%TPC)	2.5	1.600
	CO <sub>2</sub> Transportation Cost (\$/tonne)	1.684	0.0
	CO <sub>2</sub> Storage Cost (\$/tonne)	3.145	3.75
Water Systems Performance	Ambient Air Temp (Dry Bulb Avg.) (°C)	25.0	21.1
	Power Requirement (% MWg)	1.250	1.410
Water Systems Capital Cost	General Facilities Capital (%PFC)	10	2.515
	Engineering & Home Office Fees (%PFC)	10	9.30
	Project Contingency Cost (%PFC)	15	13.80
	Process Contingency Cost (%PFC)	0	0.0
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.0	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.5	0.0
Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

<sup>1</sup>See Figure Figure B.1

### Case Study #7: PC Subcritical with Amine-Based CO<sub>2</sub> Capture

For the Subcritical plant with a CO<sub>2</sub> Capture System, a number of default parameters in the IECM were changed so that these parameters matched the NETL 2007 Baseline Report. These changes are shown in Table A.8.

**Table A.7: IECM parameters changed for Case 7: subcritical plants with amine-based CO<sub>2</sub> capture**

	Parameter	IECM Default	IECM Case 7
Configure Plant	NO <sub>x</sub> Control	None	Hot-Side SCR
Overall Plant	Particulates	None	Fabric Filter
	SO <sub>2</sub> Control	None	Wet FGD
	CO <sub>2</sub> Capture	None	Amine System
	Cooling System	Once-Through	Wet Cooling Tower
Fuel	Fuel Name	Appalachian Medium	Illinois #6
Properties		Sulfur	
Overall Plant	Capacity Factor (%)	75.00	75.00
Performance	Ambient Air Temperature (Avg.) (°C)	25.0	21.1
Overall Plant	Discount Rate (Before Taxes) (fraction)	7.09e-2	1.0E-4
Financing	Fixed Charge Factor (FCF) (fraction)	0.1128	0.1430
	Plant or Project Book Life (years)	30.00	20.00
Overall Plant	Natural Gas Cost (\$/mscf)	5.99	7.58
O&M Cost			
Base Plant	Gross Electrical Output (MWg)	500	663.3
Performance	Unit Type:	Sub-Critical	Supercritical
	Boiler Efficiency (%)	88.89	89.00
	Leakage Air at Preheater (% stoich.)	10	5.5
	Coal Pulverizer (% MWg)	0.5897	0.6700
Base Plant Capital	General Facilities Capital (%PFC)	10.00	1.570
Cost	Engineering & Home Office Fees (%PFC)	6.50	9.37
	Project Contingency Cost (%PFC)	11.67	16.38
	Process Contingency Cost (%PFC)	0.3	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Fixed Operating Cost (months)	1.00	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	6.00E-2	0.0
Base Plant O&M	Waste Disposal Cost (\$/ton)	9.36	15.45
Cost	Total Maintenance Cost (%TPC)	1.896	1.60
NO <sub>x</sub> Control	Actual NO <sub>x</sub> Removal Efficiency (%)	76.66	86.00
Performance	Hot-Side SCR Power Requirement (% MWg)	0.6294	1.100e-2
NO <sub>x</sub> Control	General Facilities Capital (%PFC)	10.00	1.57
Capital Cost	Engineering & Home Office Fees (%PFC)	10.00	9.37
	Project Contingency Cost (%PFC)	10.00	16.38
	Process Contingency Cost (%PFC)	6.397	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Months of Fixed O&M (months)	1	0
	Months of Variable O&M (months)	1	0
	Misc. Capital Cost (%TPI)	2	0
	Inventory Capital (%TPC)	0.5	0
NO <sub>x</sub> Control O&M	Total Maintenance Cost (%TPC)	2.00	1.60
Cost			
TSP Control	Particulate Removal Efficiency (%)	99.10	99.90
Performance	Fabric Filter Power Requirement (% MWg)	0.1958	1.70E-2

TSP Control	General Facilities Capital (%PFC)	1.00	1.57
Capital Cost	Engineering & Home Office Fees (%PFC)	5	9.37
	Project Contingency Cost (%PFC)	20	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
TSP Control O&M	Total Maintenance Cost (%TPC)	1.723	1.600
Cost			
SO <sub>2</sub> Control	Scrubber SO <sub>2</sub> Removal Efficiency (%)	85.49	98.00
Performance	Wet FGD Power Requirement (% MWg)	1.607	0.83
SO <sub>2</sub> Control	General Facilities Capital (%PFC)	10	1.57
Capital Cost	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	2	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.3545	0
SO <sub>2</sub> Control O&M	Total Maintenance Cost (%TPC)	4.398	1.60
Cost			
CO <sub>2</sub> Capture	Amine Scrubber Power Requirement (%	13.98	10.28
Performance	MWg)		
CO <sub>2</sub> Capture	General Facilities Capital (%PFC)	10	1.57
Capital Cost	Engineering & Home Office Fees (%PFC)	7	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	5	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
CO <sub>2</sub> Capture	Reclaimer Waste Disposal Cost (\$/tonne)	244.6	0.0
O&M Cost	Total Maintenance Cost (%TPC)	2.5	1.600
	CO <sub>2</sub> Transportation Cost (\$/tonne)	1.684	0.0
	CO <sub>2</sub> Storage Cost (\$/tonne)	3.145	3.75
Water Systems	Ambient Air Temp (Dry Bulb Avg.) (°F)	77.00	70.00
Performance	Power Requirement (% MWg)	2.8	2.800
Water Systems	General Facilities Capital (%PFC)	10	1.57
Capital Cost	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0

	Inventory Capital (%TPC)	0.5	0
Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

### Case Study #8: PC Supercritical with Amine-Based CO<sub>2</sub> Capture

For the Supercritical plant with a CO<sub>2</sub> Capture System, a number of default parameters in the IECM were changed so that these parameters matched the NETL 2007 Baseline Report. These changes are shown in Table A.8.

**Table A.8: IECM parameters changed for Case 8: supercritical plants with amine-based CO<sub>2</sub> capture**

	Parameter	IECM Default	IECM Case 8
Configure Plant	NO <sub>x</sub> Control	None	Hot-Side SCR
Overall Plant	Particulates	None	Fabric Filter
	SO <sub>2</sub> Control	None	Wet FGD
	CO <sub>2</sub> Capture	None	Amine System
	Cooling System	Once-Through	Wet Cooling Tower
Fuel	Fuel Name	Appalachian Medium	Illinois #6
Properties		Sulfur	
Overall Plant	Capacity Factor (%)	75.00	75.00
Performance	Ambient Air Temperature (Avg.) (°C)	25.0	21.1
Overall Plant	Discount Rate (Before Taxes) (fraction)	7.09e-2	1.0E-4
Financing	Fixed Charge Factor (FCF) (fraction)	0.1128	0.1430
	Plant or Project Book Life (years)	30.00	20.00
Overall Plant	Natural Gas Cost (\$/mscf)	5.99	7.58
O&M Cost			
Base Plant	Gross Electrical Output (MWg)	500	663.3
Performance	Unit Type:	Sub-Critical	Supercritical
	Boiler Efficiency (%)	88.89	89.00
	Leakage Air at Preheater (% stoich.)	10	5.5
	Coal Pulverizer (% MWg)	0.5897	0.6700
Base Plant Capital	General Facilities Capital (%PFC)	10.00	1.570
Cost	Engineering & Home Office Fees (%PFC)	6.50	9.37
	Project Contingency Cost (%PFC)	11.67	16.38
	Process Contingency Cost (%PFC)	0.3	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Fixed Operating Cost (months)	1.00	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	6.00E-2	0.0
Base Plant O&M	Waste Disposal Cost (\$/ton)	9.36	15.45
Cost	Total Maintenance Cost (%TPC)	1.896	1.60
NO <sub>x</sub> Control	Actual NO <sub>x</sub> Removal Efficiency (%)	76.66	86.00

Performance	Hot-Side SCR Power Requirement (% MWg)	0.6294	1.100e-2
NO <sub>x</sub> Control	General Facilities Capital (%PFC)	10.00	1.57
Capital Cost	Engineering & Home Office Fees (%PFC)	10.00	9.37
	Project Contingency Cost (%PFC)	10.00	16.38
	Process Contingency Cost (%PFC)	6.397	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Months of Fixed O&M (months)	1	0
	Months of Variable O&M (months)	1	0
	Misc. Capital Cost (%TPI)	2	0
	Inventory Capital (%TPC)	0.5	0
NO <sub>x</sub> Control O&M	Total Maintenance Cost (%TPC)	2.00	1.60
Cost			
TSP Control	Particulate Removal Efficiency (%)	99.10	99.90
Performance	Fabric Filter Power Requirement (% MWg)	0.1958	1.70E-2
TSP Control	General Facilities Capital (%PFC)	1.00	1.57
Capital Cost	Engineering & Home Office Fees (%PFC)	5	9.37
	Project Contingency Cost (%PFC)	20	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
TSP Control O&M	Total Maintenance Cost (%TPC)	1.723	1.600
Cost			
SO <sub>2</sub> Control	Scrubber SO <sub>2</sub> Removal Efficiency (%)	85.49	98.00
Performance	Wet FGD Power Requirement (% MWg)	1.607	0.83
SO <sub>2</sub> Control	General Facilities Capital (%PFC)	10	1.57
Capital Cost	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	2	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.3545	0
SO <sub>2</sub> Control O&M	Total Maintenance Cost (%TPC)	4.398	1.60
Cost			
CO <sub>2</sub> Capture	Amine Scrubber Power Requirement (%	13.98	10.28
Performance	MWg)		
CO <sub>2</sub> Capture	General Facilities Capital (%PFC)	10	1.57
Capital Cost	Engineering & Home Office Fees (%PFC)	7	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	5	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
CO <sub>2</sub> Capture	Reclaimer Waste Disposal Cost (\$/tonne)	244.6	0.0

O&M Cost	Total Maintenance Cost (%TPC)	2.5	1.600
	CO <sub>2</sub> Transportation Cost (\$/tonne)	1.684	0.0
	CO <sub>2</sub> Storage Cost (\$/tonne)	3.145	3.75
Water Systems Performance	Ambient Air Temp (Dry Bulb Avg.) (°C)	25.0	21.1
	Power Requirement (% MWg)	2.8	2.800
Water Systems Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

### Case Study #9: NGCC with Amine-Based CO<sub>2</sub> Capture

For the Subcritical plant with a CO<sub>2</sub> Capture System, a number of default parameters in the IECM were changed so that these parameters matched the NETL 2007 Baseline Report. These changes are shown in Table A.9.

**Table A.9: IECM parameters changed for Case 9: NGCC plants with amine-based CO<sub>2</sub> capture**

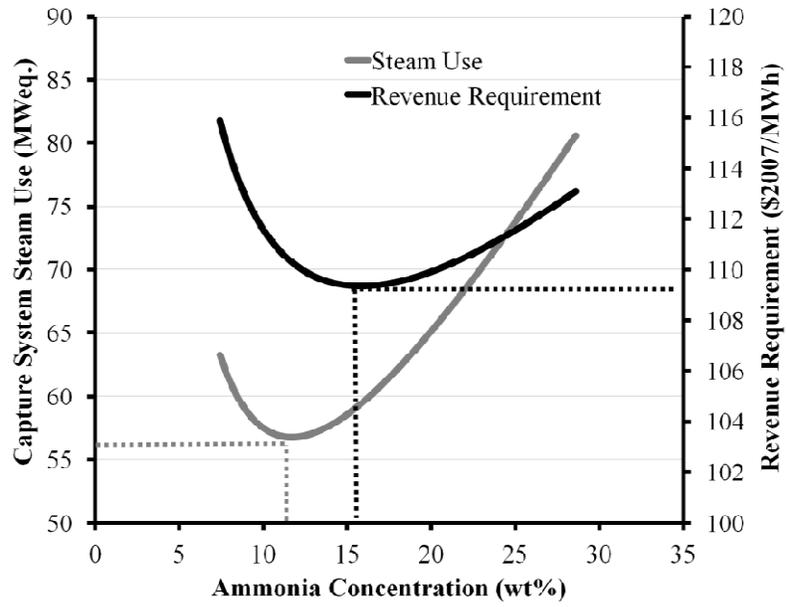
	Parameter	IECM Default	IECM Case 9
Configure Plant Overall Plant	NO <sub>x</sub> Control	None	Hot-Side SCR
	Particulates	None	Fabric Filter
	SO <sub>2</sub> Control	None	Wet FGD
	CO <sub>2</sub> Capture	None	Amine System
	Cooling System	Once-Through	Wet Cooling Tower
Overall Plant Performance	Capacity Factor (%)	75.00	75.00
	Ambient Air Temperature (Avg.) (°C)	25.0	21.1
Overall Plant Financing	Discount Rate (Before Taxes) (fraction)	7.09e-2	1.0E-4
	Fixed Charge Factor (FCF) (fraction)	0.1128	0.1430
Fuel Properties	Fuel Name	Natural Gas (93.1% CH <sub>4</sub> )	Natural Gas (93.9% CH <sub>4</sub> )
	Fuel Cost (\$/GJ)	6.5	6.4
	Inventory Capital (%TPC)	0.3545	0
Power Block Capital Cost	General Facilities Capital (%PFC)	15.00	1.57
	Engineering & Home Office Fees (%PFC)	10.00	9.37
	Project Contingency Cost (%PFC)	15.00	16.38
	Process Contingency Cost (%PFC)	7.501	4.670

	Royalty Fees (%PFC)	0.5000	0.0
	Months of Fixed O&M (Months)	1.000	0.0
	Months of Variable O&M (Months)	1.000	0.0
	Misc. Capital Cost (%TPI)	2.000	0.0
	Inventory Capital (%TPC)	0.5000	0.0
CO <sub>2</sub> Capture Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	7	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	5	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
CO <sub>2</sub> Capture O&M Cost	Reclaimer Waste Disposal Cost (\$/tonne)	244.6	0.0
	Total Maintenance Cost (%TPC)	2.5	1.600
	CO <sub>2</sub> Transportation Cost (\$/tonne)	1.684	0.0
	CO <sub>2</sub> Storage Cost (\$/tonne)	3.145	3.75
Water Systems Performance	Ambient Air Temp (Dry Bulb Avg.) (°C)	25.0	21.1
Water Systems Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

---

## **Appendix B. Ammonia Based-CO<sub>2</sub> Capture at an NGCC Plant**

The optimal ammonia concentration (14.4 wt%) for the ammonia-based CO<sub>2</sub> capture system for the PC power plants was estimated using Aspen Plus® in Chapter 2. In contrast, the updated IECM models were used to find an appropriate ammonia concentration when the CO<sub>2</sub> capture system was added to the NGCC power plant. In the IECM, a sensitivity analysis was done to compare the ammonia concentration vs. the CO<sub>2</sub> capture system steam use and the overall revenue requirement for the power plant. The results are shown in Figure B.1. As shown in the figure, the lowest steam use (56MWeq. at 11.7 wt%) does not correspond to the lowest revenue requirement (\$109.3/MWh at 15.6 wt%) for the power plant. At an ammonia concentration of 11.7 wt% vs. 15.6 wt%, the plant efficiency is higher (38.6% vs 38.3% HHV), but CO<sub>2</sub> capture system solvent flow rate is also higher (2385 vs. 1655 tonnes/hr). This leads to increases in heat exchanger cost, circulation pump cost, and sorbent regenerator cost, substantially increasing the total cost of the CO<sub>2</sub> capture system (\$M421.4 vs. \$M397.6) and correspondingly leading to a higher revenue requirement.



**Figure B.1: Ammonia concentration vs. the CO<sub>2</sub> capture system steam use and plant revenue requirement.**

## Appendix C. Aspen Plus Code

The following is a generic printout of the Aspen Plus® flowsheet input for the chilled ammonia process presented in Chapter 2 and used in Chapter 3 to develop the response surface model in the IECM. Code lines that start with “F ” denote Fortran code lines programmed into the Aspen Plus® model to aid in calculations. A user should be able to use this as a starting point for reproducing the Aspen Plus® results of the thesis.

```
;  
;Input Summary created by Aspen Plus Rel. 24.0 at 06:45:42 Tue Mar 27, 2012  
;Directory  Filename C:\Documents and Settings\Peter\Desktop\Recent Research\Research\Improving Power  
Plants\PhD Dissertation\Aspen Text Files\chilled ammonia - september 10 2011 - paper response.inp  
;
```

DYNAMICS

DYNAMICS RESULTS=ON

IN-UNITS SI MOLE-FLOW='mol/sec' ENTHALPY-FLO='Btu/hr' &  
PRESSURE=atm TEMPERATURE=F DELTA-T=F PDROP=atm

DEF-STREAMS CONVEN ALL

DIAGNOSTICS

HISTORY INSERT=NO SORTED=NO

SIM-OPTIONS

IN-UNITS ENG

SIM-OPTIONS FLASH-MAXIT=400 GAMUS-BASIS=AQUEOUS &  
OLD-DATABANK=NO

RUN-CONTROL MAX-ERRORS=1000000 MAX-FORT-ERR=1000000

DATABANKS 'APV72 PURE20' / 'APV72 AQUEOUS' / 'APV72 SOLIDS' / &  
'APV72 INORGANIC' / NOASPENPCD

PROP-SOURCES 'APV72 PURE20' / 'APV72 AQUEOUS' / 'APV72 SOLIDS' &  
'APV72 INORGANIC'

COMPONENTS

AR AR /

CO2 CO2 /

H2 H2 /

H2O H2O /

N2 N2 /

NH3 H3N /

NH4HCO3S NH4HCO3 /

NH4HCO3 NH4HCO3 /

O2 O2 /  
SO2 O2S /  
H3O+ H3O+ /  
OH- OH- /  
NH4+ NH4+ /  
NH2COO- NH2COO- /  
HCO3- HCO3- /  
CO3-- CO3-2

HENRY-COMPS NH3 NH3 CO2 N2 O2 AR

CHEMISTRY NH3

IN-UNITS SI  
DISS NH4HCO3 NH4+ 1. / HCO3- 1.  
STOIC 1 H2O -2. / H3O+ 1. / OH- 1.  
STOIC 2 CO2 -1. / H2O -2. / HCO3- 1. / H3O+ 1.  
STOIC 3 HCO3- -1. / H2O -1. / CO3-- 1. / H3O+ 1.  
STOIC 4 NH3 -1. / H2O -1. / NH4+ 1. / OH- 1.  
STOIC 5 NH3 -1. / HCO3- -1. / NH2COO- 1. / H2O 1.  
K-STOIC 1 A=132.899 B=-13445.9 C=-22.4773  
K-STOIC 2 A=231.465 B=-12092.1 C=-36.7816  
K-STOIC 3 A=216.049 B=-12431.7 C=-35.4819  
K-STOIC 4 A=-1.256563 B=-3335.7 C=1.4971 D=-0.0370566  
K-STOIC 5 A=-4.5834 B=2900. C=0.  
SALT NH4HCO3S NH4+ 1. / HCO3- 1.  
K-SALT NH4HCO3S A=-914.00821 B=38648.2117 C=136.174996

FLOWSHEET

BLOCK HPPUMP IN=10 OUT=11  
BLOCK HEATER1 IN=12 OUT=13  
BLOCK FLASH2 IN=14 OUT=15 16  
BLOCK HTX3 IN=17 11 OUT=18 12  
BLOCK STRIPPER IN=13 16 OUT=14 17  
BLOCK MIXER IN=27 24 18 32 OUT=28  
BLOCK HTX4 IN=25 20 OUT=26 21  
BLOCK NH3STRIP IN=21 23 OUT=22 25  
BLOCK WATWASH IN=19 9 OUT=33 20  
BLOCK BLOWER IN=3 OUT=4  
BLOCK COOLER1 IN=28 30 OUT=29 31  
BLOCK B1 IN=22 OUT=24 23  
BLOCK HTX1 IN=37 38 OUT=40 39  
BLOCK HTX2 IN=4 5 OUT=7 6  
BLOCK DCC1 IN=1 35 OUT=3 2  
BLOCK DCC2 IN=40 33 OUT=36 35  
BLOCK ABSORBER IN=7 8 OUT=9 10

PROPERTIES ELECNRTL HENRY-COMPS=NH3 CHEMISTRY=NH3 TRUE-COMPS=YES

PROP-REPLACE ELECNRTL ELECNRTL

MODEL VAQCLK 1 1  
MODEL MUL2JONS 1 1 1 2  
MODEL KL2RDL  
MODEL DL1NST 1 1  
MODEL SIG2ONSG 1 -9 1  
MODEL DL0NST 1 1

PROP-DATA CPSP01-1  
 IN-UNITS SI MOLE-FLOW='mol/sec' ENTHALPY-FLO='Btu/hr' &  
 PRESSURE=atm TEMPERATURE=F DELTA-T=F PDROP=atm  
 PROP-LIST CPSP01  
 PVAL AR 23279.2

PROP-DATA SIGDIP-1  
 IN-UNITS SI  
 PROP-LIST SIGDIP  
 PVAL NH3 .0948191945 1.152342130 0.0 0.0 0.0 195.4100000 &  
 405.6500000

PROP-DATA VLQKIJ-1  
 IN-UNITS SI  
 PROP-LIST VLQKIJ  
 BPVAL H2O NH3 -.3003897000  
 BPVAL NH3 H2O -.3003897000

PROP-DATA HENRY-1  
 IN-UNITS SI  
 PROP-LIST HENRY  
 BPVAL NH3 H2O -133.4630000 -157.5520000 28.10010000 &  
 -.0492270000 273.0000000 498.0000000 0.0  
 BPVAL CO2 H2O 170.7126000 -8477.711000 -21.95743000 &  
 5.78074800E-3 273.0000000 500.0000000 0.0  
 BPVAL N2 NH3 19.64450100 0.0 0.0 0.0 293.0000000 &  
 293.0000000 0.0  
 BPVAL N2 H2O 176.5070000 -8432.770000 -21.55800000 &  
 -8.4362400E-3 273.0000000 346.0000000 0.0  
 BPVAL O2 NH3 20.04800000 0.0 0.0 0.0 293.0000000 &  
 293.0000000 0.0  
 BPVAL O2 H2O 155.9210000 -7775.060000 -18.39740000 &  
 -9.4435400E-3 274.0000000 348.0000000 0.0  
 BPVAL NH3 H2 1.000000000000000e+035 1.000000000000000e+035 &  
 1.000000000000000e+035 1.000000000000000e+035 293.0000000 &  
 293.0000000 1.000000000000000e+035  
 BPVAL AR H2O 180.9910000 -8137.130000 -23.25470000 &  
 3.06357000E-3 274.0000000 347.0000000 0.0

PROP-DATA MUKIJ-1  
 IN-UNITS SI  
 PROP-LIST MUKIJ  
 BPVAL H2O NH3 3.318036770 0.0 0.0 0.0 0.0 298.1500  
 BPVAL NH3 H2O 3.318036770 0.0 0.0 0.0 0.0 298.1500

PROP-DATA MULIJ-1  
 IN-UNITS SI  
 PROP-LIST MULIJ  
 BPVAL H2O NH3 -.9537000450 0.0 0.0 0.0 0.0 298.1500  
 BPVAL NH3 H2O .9537000450 0.0 0.0 0.0 0.0 298.1500

PROP-DATA NRTL-1  
 IN-UNITS SI  
 PROP-LIST NRTL  
 BPVAL H2O NH3 -.5440720000 1678.469000 .2000000000 0.0 0.0 &  
 0.0 273.1500000 473.1500000

BPVAL NH3 H2O -.1642422000 -1027.525000 .2000000000 0.0 &  
0.0 0.0 273.1500000 473.1500000  
BPVAL H2O CO2 10.06400000 -3268.135000 .2000000000 0.0 0.0 &  
0.0 273.1500000 473.1500000  
BPVAL CO2 H2O 10.06400000 -3268.135000 .2000000000 0.0 0.0 &  
0.0 273.1500000 473.1500000

PROP-DATA GMELCC-1

IN-UNITS SI  
PROP-LIST GMELCC  
PPVAL H2O ( H3O+ OH- ) 8.045000000  
PPVAL H2O ( H3O+ HCO3- ) 8.045000000  
PPVAL H2O ( H3O+ CO3-- ) 8.045000000  
PPVAL H2O ( H3O+ NH2COO- ) 8.045000000  
PPVAL H2O ( NH4+ OH- ) 8.045000000  
PPVAL H2O ( NH4+ HCO3- ) 8.012939770  
PPVAL H2O ( NH4+ CO3-- ) 8.045000000  
PPVAL H2O ( NH4+ NH2COO- ) 7.832378950  
PPVAL ( H3O+ OH- ) H2O -4.072000000  
PPVAL ( H3O+ HCO3- ) H2O -4.072000000  
PPVAL ( H3O+ CO3-- ) H2O -4.072000000  
PPVAL ( H3O+ NH2COO- ) H2O -4.072000000  
PPVAL ( NH4+ OH- ) H2O -4.072000000  
PPVAL ( NH4+ HCO3- ) H2O -3.653583560  
PPVAL ( NH4+ CO3-- ) H2O -4.072000000  
PPVAL ( NH4+ NH2COO- ) H2O -3.856105710  
PPVAL ( NH4+ NH2COO- ) NH3 -2.000000000  
PPVAL CO2 ( H3O+ OH- ) 15.000000000  
PPVAL CO2 ( H3O+ HCO3- ) 15.000000000  
PPVAL CO2 ( H3O+ CO3-- ) 15.000000000  
PPVAL CO2 ( NH4+ HCO3- ) 35.00355480  
PPVAL CO2 ( NH4+ NH2COO- ) 20.20611250  
PPVAL ( H3O+ OH- ) CO2 -8.000000000  
PPVAL ( H3O+ HCO3- ) CO2 -8.000000000  
PPVAL ( H3O+ CO3-- ) CO2 -8.000000000  
PPVAL ( NH4+ HCO3- ) CO2 -.6581496790  
PPVAL ( NH4+ NH2COO- ) CO2 5.055714020  
PPVAL NH3 ( NH4+ NH2COO- ) 10.000000000  
PPVAL NH3 ( NH4+ CO3-- ) 10  
PPVAL ( NH4+ CO3-- ) NH3 -2  
PPVAL NH3 ( NH4+ HCO3- ) 10.000000000  
PPVAL ( NH4+ HCO3- ) NH3 -2.000000000

PROP-DATA GMELCD-1

IN-UNITS SI  
PROP-LIST GMELCD  
PPVAL H2O ( NH4+ HCO3- ) 0.0  
PPVAL ( NH4+ HCO3- ) H2O 0.0  
PPVAL CO2 ( H3O+ OH- ) 0.0  
PPVAL ( H3O+ OH- ) CO2 0.0  
PPVAL CO2 ( H3O+ HCO3- ) 0.0  
PPVAL ( H3O+ HCO3- ) CO2 0.0  
PPVAL CO2 ( H3O+ CO3-- ) 0.0  
PPVAL ( H3O+ CO3-- ) CO2 0.0

PROP-DATA GMELCE-1

```

IN-UNITS SI
PROP-LIST GMELCE
PPVAL CO2 ( H3O+ OH- ) 0.0
PPVAL ( H3O+ OH- ) CO2 0.0
PPVAL CO2 ( H3O+ HCO3- ) 0.0
PPVAL ( H3O+ HCO3- ) CO2 0.0
PPVAL CO2 ( H3O+ CO3-- ) 0.0
PPVAL ( H3O+ CO3-- ) CO2 0.0
PPVAL CO2 ( NH4+ HCO3- ) 0.0
PPVAL ( NH4+ HCO3- ) CO2 0.0
PPVAL CO2 ( NH4+ NH2COO- ) 0.0
PPVAL ( NH4+ NH2COO- ) CO2 0.0

PROP-DATA GMELCN-1
IN-UNITS SI
PROP-LIST GMELCN
PPVAL H2O ( H3O+ OH- ) 0.2
PPVAL H2O ( H3O+ HCO3- ) 0.2
PPVAL H2O ( H3O+ CO3-- ) 0.2
PPVAL H2O ( H3O+ NH2COO- ) 0.2
PPVAL H2O ( NH4+ OH- ) 0.2
PPVAL H2O ( NH4+ HCO3- ) 0.2
PPVAL H2O ( NH4+ CO3-- ) 0.2
PPVAL H2O ( NH4+ NH2COO- ) 0.2
PPVAL CO2 ( H3O+ OH- ) .1000000000
PPVAL CO2 ( H3O+ HCO3- ) .1000000000
PPVAL CO2 ( H3O+ CO3-- ) .1000000000
PPVAL CO2 ( H3O+ NH2COO- ) 0.1
PPVAL CO2 ( NH4+ OH- ) 0.1
PPVAL CO2 ( NH4+ HCO3- ) 0.1
PPVAL CO2 ( NH4+ CO3-- ) 0.1
PPVAL CO2 ( NH4+ NH2COO- ) 0.1
PPVAL NH3 ( H3O+ OH- ) 0.1
PPVAL NH3 ( H3O+ HCO3- ) 0.1
PPVAL NH3 ( H3O+ CO3-- ) 0.1
PPVAL NH3 ( H3O+ NH2COO- ) 0.1
PPVAL NH3 ( NH4+ OH- ) 0.1
PPVAL NH3 ( NH4+ HCO3- ) 0.1
PPVAL NH3 ( NH4+ CO3-- ) 0.1

PROP-SET CUMFLOW VOLFLMX UNITS='cum/sec' SUBSTREAM=MIXED

PROP-SET MOLEFLOW
IN-UNITS ENG
PROPNAME-LIS MOLEFLOW SUBSTREAM=MIXED PHASE=T

PROP-SET PARTPRES
IN-UNITS SI
PROPNAME-LIS PPMX SUBSTREAM=MIXED PHASE=V

PROP-SET PH
IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &
DELTA-T=F PDROP=atm
PROPNAME-LIS PH SUBSTREAM=MIXED PHASE=L

PROP-SET SOLINDEX

```

IN-UNITS SI  
 PROPNAME-LIS SOLINDEX SUBSTREAM=MIXED COMPS=NH4HCO3S PHASE=L

PROP-SET TRUEMASS  
 IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
 DELTA-T=F PDROP=atm  
 PROPNAME-LIS WTRUE SUBSTREAM=MIXED PHASE=L S

STREAM 1  
 IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
 DELTA-T=F PDROP=atm  
 SUBSTREAM MIXED TEMP=135. PRES=1.  
 MOLE-FLOW AR 240.35658 / CO2 3983.91031 / H2 0. / H2O &  
 5011.43469 / N2 20102.8234 / NH3 0. / NH4HCO3S 0. / &  
 NH4HCO3 0. / O2 706.047453 / SO2 0. / H3O+ 0. / &  
 OH- 0. / NH4+ 0. / NH2COO- 0. / HCO3- 0. / CO3-- &  
 0.

STREAM 2  
 IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
 DELTA-T=F PDROP=atm  
 SUBSTREAM MIXED TEMP=41. PRES=1.  
 MOLE-FLOW AR 0.5 / CO2 125. / H2O 10000. / N2 9. / &  
 O2 0.8

STREAM 3  
 IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
 DELTA-T=F PDROP=atm  
 SUBSTREAM MIXED TEMP=80.1024297 PRES=1.  
 MOLE-FLOW AR 240.371887 / CO2 3974.41256 / H2 0. / H2O &  
 909.737959 / N2 20103.3675 / NH3 7.5722E-006 / &  
 NH4HCO3S 0. / NH4HCO3 0. / O2 706.087847 / SO2 0. / &  
 H3O+ 0. / OH- 0. / NH4+ 0. / NH2COO- 0. / HCO3- &  
 0. / CO3-- 0.

STREAM 5  
 SUBSTREAM MIXED TEMP=37. PRES=1.  
 MASS-FLOW H2O 1000.

STREAM 8  
 IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
 DELTA-T=F PDROP=atm  
 SUBSTREAM MIXED TEMP=47. PRES=1. FREE-WATER=NO NPHASE=2 &  
 PHASE=V  
 MOLE-FLOW CO2 2209.21 / H2O 16394.78 / NH3 6312.02

STREAM 9  
 SUBSTREAM MIXED TEMP=50. PRES=1.

STREAM 10  
 IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
 DELTA-T=F PDROP=atm  
 SUBSTREAM MIXED TEMP=50. PRES=1.

STREAM 11  
 IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &

DELTA-T=F PDROP=atm  
SUBSTREAM MIXED TEMP=51.9348532 PRES=31.9816031 TOL=0.0001

STREAM 13

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
SUBSTREAM MIXED TEMP=163. PRES=30

STREAM 14

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
SUBSTREAM MIXED TEMP=188.912135 PRES=30  
MOLE-FLOW CO2 4000. / H2O 30.

STREAM 17

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
SUBSTREAM MIXED TEMP=293.376376 PRES=30

STREAM 19

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
SUBSTREAM MIXED TEMP=80. PRES=1.  
MOLE-FLOW H2O 2000. / NH3 0.06

STREAM 20

SUBSTREAM MIXED TEMP=52.5461336 PRES=1.  
MOLE-FLOW AR 0.00055376 / CO2 0.01417088 / H2 0. / H2O &  
1875.72463 / N2 0.02097169 / NH3 2.622138 / NH4HCO3S &  
0. / NH4HCO3 0. / O2 0.00148406 / SO2 0. / H3O+ &  
6.5567E-008 / OH- 0.00019394 / NH4+ 40.4333844 / &  
NH2COO- 11.1047049 / HCO3- 22.4998175 / CO3-- &  
3.41433403

STREAM 21

SUBSTREAM MIXED TEMP=120. PRES=1.  
MOLE-FLOW AR 0.00055376 / CO2 0.01417088 / H2 0. / H2O &  
1875.72463 / N2 0.02097169 / NH3 2.622138 / NH4HCO3S &  
0. / NH4HCO3 0. / O2 0.00148406 / SO2 0. / H3O+ &  
6.5567E-008 / OH- 0.00019394 / NH4+ 40.4333844 / &  
NH2COO- 11.1047049 / HCO3- 22.4998175 / CO3-- &  
0.93824404

STREAM 22

SUBSTREAM MIXED TEMP=200. PRES=1.  
MOLE-FLOW CO2 100. / H2O 50. / NH3 100.

STREAM 27

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
SUBSTREAM MIXED TEMP=80. PRES=1.  
MOLE-FLOW CO2 1. / H2O 0. / NH3 0.

STREAM 30

SUBSTREAM MIXED TEMP=37. PRES=1.  
MOLE-FLOW H2O 5000.

STREAM 32

SUBSTREAM MIXED TEMP=80. PRES=1.  
MOLE-FLOW H2O 1.

STREAM 37

SUBSTREAM MIXED TEMP=133. PRES=1.  
MOLE-FLOW CO2 50. / H2O 80508.348

STREAM 38

SUBSTREAM MIXED TEMP=80. PRES=1.  
MASS-FLOW H2O 5000.

BLOCK MIXER MIXER

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
PARAM PRES=1.

BLOCK HEATER1 HEATER

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
PARAM TEMP=165.5 PRES=30.

BLOCK ABSORBER FLASH2

PARAM TEMP=50. PRES=1.  
BLOCK-OPTION SIM-LEVEL=5 PROP-LEVEL=5 STREAM-LEVEL=5 &  
TERM-LEVEL=5 ENERGY-BAL=YES  
REPORT REPORT  
UTILITY UTILITY-ID=U-1

BLOCK B1 FLASH2

PARAM TEMP=150. PRES=1.

BLOCK FLASH2 FLASH2

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
PARAM TEMP=85. PRES=400. <psi>

BLOCK COOLER1 HEATX

PARAM DELT-HOT=10. U-OPTION=PHASE F-OPTION=CONSTANT &  
CALC-METHOD=SHORTCUT  
FEEDS HOT=28 COLD=30  
PRODUCTS HOT=29 COLD=31  
HOT-SIDE DP-OPTION=CONSTANT  
COLD-SIDE DP-OPTION=CONSTANT

BLOCK HTX1 HEATX

PARAM DELT-HOT=10. U-OPTION=PHASE F-OPTION=CONSTANT &  
CALC-METHOD=SHORTCUT  
FEEDS HOT=37 COLD=38  
PRODUCTS HOT=40 COLD=39  
HOT-SIDE DP-OPTION=CONSTANT  
COLD-SIDE DP-OPTION=CONSTANT

BLOCK HTX2 HEATX

PARAM DELT-HOT=5. U-OPTION=PHASE F-OPTION=CONSTANT &

CALC-METHOD=SHORTCUT  
FEEDS HOT=4 COLD=5  
PRODUCTS HOT=7 COLD=6  
FLASH-SPECS 7 NPHASE=1 PHASE=V FREE-WATER=NO  
FLASH-SPECS 6 NPHASE=1 PHASE=L FREE-WATER=NO  
HOT-SIDE DP-OPTION=CONSTANT  
COLD-SIDE DP-OPTION=CONSTANT

BLOCK HTX3 HEATX

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
PARAM DELT-HOT=10. MIN-TAPP=1.8 U-OPTION=PHASE &  
F-OPTION=CONSTANT CALC-METHOD=SHORTCUT  
FEEDS HOT=17 COLD=11  
PRODUCTS HOT=18 COLD=12  
FLASH-SPECS 12 NPHASE=1 PHASE=L FREE-WATER=NO  
HOT-SIDE DP-OPTION=CONSTANT  
COLD-SIDE DP-OPTION=CONSTANT

BLOCK HTX4 HEATX

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
PARAM DELT-HOT=10. MIN-TAPP=1.8 U-OPTION=PHASE &  
F-OPTION=CONSTANT CALC-METHOD=SHORTCUT  
FEEDS HOT=25 COLD=20  
PRODUCTS HOT=26 COLD=21  
FLASH-SPECS 26 NPHASE=1 PHASE=L FREE-WATER=NO  
FLASH-SPECS 21 NPHASE=1 PHASE=L FREE-WATER=NO  
HOT-SIDE DP-OPTION=CONSTANT  
COLD-SIDE DP-OPTION=CONSTANT

BLOCK DCC1 RADFRAC

PARAM NSTAGE=8 ALGORITHM=STANDARD INIT-OPTION=STANDARD &  
ABSORBER=YES  
COL-CONFIG CONDENSER=NONE REBOILER=NONE  
FEEDS 1 9 / 35 1  
PRODUCTS 2 8 L / 3 1 V  
P-SPEC 1 1.  
COL-SPECS

BLOCK DCC2 RADFRAC

PARAM NSTAGE=8 ALGORITHM=STANDARD INIT-OPTION=STANDARD &  
ABSORBER=YES  
COL-CONFIG CONDENSER=NONE REBOILER=NONE  
FEEDS 40 1 / 33 9  
PRODUCTS 36 1 V / 35 8 L  
P-SPEC 1 1.  
COL-SPECS

BLOCK NH3STRIP RADFRAC

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
PARAM NSTAGE=10 ALGORITHM=STANDARD INIT-OPTION=STANDARD  
COL-CONFIG CONDENSER=NONE  
FEEDS 21 5 / 23 1 ON-STAGE  
PRODUCTS 25 10 L / 22 1 V

P-SPEC 1 1.  
COL-SPECS MOLE-D=250.

BLOCK STRIPPER RADFRAC

PARAM NSTAGE=8 ALGORITHM=STANDARD INIT-OPTION=STANDARD &  
MAXOL=200 FLASH-MAXIT=50 DAMPING=MILD  
COL-CONFIG CONDENSER=NONE  
FEEDS 13 1 ON-STAGE / 16 1 ON-STAGE  
PRODUCTS 17 8 L / 14 1 V  
P-SPEC 1 400. <psi>  
COL-SPECS MOLE-D=4550.

BLOCK WATWASH RADFRAC

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
PARAM NSTAGE=8 ALGORITHM=STANDARD INIT-OPTION=STANDARD &  
ABSORBER=YES  
COL-CONFIG CONDENSER=NONE REBOILER=NONE  
RATESEP-ENAB CALC-MODE=EQUILIBRIUM  
FEEDS 19 1 / 9 9  
PRODUCTS 20 8 L / 33 1 V  
P-SPEC 1 1.  
COL-SPECS

BLOCK HPPUMP PUMP

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
PARAM PRES=30. PUMP-TYPE=PUMP

BLOCK BLOWER COMPR

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
PARAM TYPE=ISENTROPIC PRES=1.2 NPHASE=2  
BLOCK-OPTION FREE-WATER=NO

UTILITY U-1 GENERAL

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
COST PRICE=0.01  
PARAM UTILITY-TYPE=WATER PRES=1. PRES-OUT=1. TIN=37. &  
TOUT=50. CALOPT=FLASH CONST-CHECK=INFORMATION

DESIGN-SPEC COOLER

IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
DELTA-T=F PDROP=atm  
DEFINE WTRFLO MASS-FLOW STREAM=30 SUBSTREAM=MIXED &  
COMPONENT=H2O  
DEFINE OUTEMP STREAM-VAR STREAM=29 SUBSTREAM=MIXED &  
VARIABLE=TEMP  
SPEC "OUTEMP" TO "47"  
TOL-SPEC "0.001"  
VARY STREAM-VAR STREAM=30 SUBSTREAM=MIXED VARIABLE=MASS-FLOW  
LIMITS "1" "8000"

EO-CONV-OPTI

```

CALCULATOR MIXBAL
  DEFINE MIXCO2 MOLE-FLOW STREAM=27 SUBSTREAM=MIXED &
    COMPONENT=CO2
  DEFINE MIXNH3 MOLE-FLOW STREAM=27 SUBSTREAM=MIXED &
    COMPONENT=NH3
  DEFINE MIXH2O MOLE-FLOW STREAM=27 SUBSTREAM=MIXED &
    COMPONENT=H2O
  DEFINE ASCO2 MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &
    COMPONENT=CO2
  DEFINE ASH2O MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &
    COMPONENT=H2O
  DEFINE ASNH3 MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &
    COMPONENT=NH3
  DEFINE RECO2 MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
    COMPONENT=CO2
  DEFINE REH2O MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
    COMPONENT=H2O
  DEFINE RENH3 MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
    COMPONENT=NH3
  DEFINE ABCO2I MOLE-FLOW STREAM=1 SUBSTREAM=MIXED &
    COMPONENT=CO2
  DEFINE ABH2OI MOLE-FLOW STREAM=1 SUBSTREAM=MIXED &
    COMPONENT=H2O
  DEFINE ABNH3I MOLE-FLOW STREAM=1 SUBSTREAM=MIXED &
    COMPONENT=NH3
  DEFINE ABCO2O MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
    COMPONENT=CO2
  DEFINE ABH2OO MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
    COMPONENT=H2O
  DEFINE ABNH3O MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
    COMPONENT=NH3
  F IF (ABNH3O+RENH3-ASNH3 .LT. 0) THEN
  F MIXNH3=0
  F ELSE
  F MIXNH3=ABNH3O+RENH3-ASNH3
  F ENDIF
  F
  F IF (ABCO2O+RECO2-ABCO2I-ASCO2 .LT. 0) THEN
  F MIXCO2=0
  F ELSE
  F MIXCO2=ABCO2O+RECO2-ABCO2I-ASCO2
  F ENDIF
  F
  F
  F IF (ABH2OO+REH2O-ABH2OI-ASH2O .LT. 0) THEN
  F MIXH2O=0
  F ELSE
  F MIXH2O=ABH2OO+REH2O-ABH2OI-ASH2O
  F ENDIF
  READ-VARS ASNH3 ASCO2 ASH2O RENH3 RECO2 REH2O ABNH3I &
    ABCO2I ABH2OI ABCO2O ABH2OO ABNH3O
  WRITE-VARS MIXCO2 MIXNH3 MIXH2O

```

```

CALCULATOR PURGEBAL
  DEFINE PURCO2 MOLE-FLOW STREAM=32 SUBSTREAM=MIXED &
    COMPONENT=CO2

```

```

DEFINE PURNH3 MOLE-FLOW STREAM=32 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE PURH2O MOLE-FLOW STREAM=32 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE ASCO2 MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE ASH2O MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE ASNH3 MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE RECO2 MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE REH2O MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE RENH3 MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE ABCO2I MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE ABH2OI MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE ABNH3I MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE ABCO2O MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE ABH2OO MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE ABNH3O MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
  COMPONENT=NH3
F IF ((ABNH3O+RENH3-ASNH3)*(-1) .LT. 0) THEN
F PURNH3=0
F ELSE
F PURNH3=(ABNH3O+RENH3-ASNH3)*(-1)
F ENDIF
F
F IF ((ABCO2O+RECO2-ABCO2I-ASCO2)*(-1) .LT. 0) THEN
F PURCO2=0
F ELSE
F PURCO2=(ABCO2O+RECO2-ABCO2I-ASCO2)*(-1)
F ENDIF
F
F IF ((ABH2OO+REH2O-ABH2OI-ASH2O)*(-1) .LT. 0) THEN
F PURH2O=0
F ELSE
F PURH2O=(ABH2OO+REH2O-ABH2OI-ASH2O)*(-1)
F ENDIF
READ-VARS ASNH3 ASCO2 ASH2O RENH3 RECO2 REH2O ABNH3I &
  ABCO2I ABH2OI ABCO2O ABH2OO ABNH3O
WRITE-VARS PURCO2 PURNH3 PURH2O

SENSITIVITY ABSADDIT
PARAM WIDE=NO BASE-CASE=NO CASES=YES
DEFINE COLWTF STREAM-VAR STREAM=38 SUBSTREAM=MIXED &
  VARIABLE=MASS-FLOW
DEFINE CHLWTF STREAM-VAR STREAM=5 SUBSTREAM=MIXED &
  VARIABLE=MASS-FLOW

```

```

DEFINE CHLBTU BLOCK-VAR BLOCK=HTX2 VARIABLE=CALC-DUTY &
  SENTENCE=RESULTS
DEFINE ABSCO2 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE ABSNH3 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE ABSH2O MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE ABSAR MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &
  COMPONENT=AR
DEFINE ABSN2 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &
  COMPONENT=N2
DEFINE ABSO2 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &
  COMPONENT=O2
DEFINE ABSKG STREAM-VAR STREAM=9 SUBSTREAM=MIXED &
  VARIABLE=MASS-FLOW
DEFINE TOTCO2 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE TOTNH3 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE TOTH2O MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE TOTAR MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &
  COMPONENT=AR
DEFINE TOTN2 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &
  COMPONENT=N2
DEFINE TOTO2 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &
  COMPONENT=O2
DEFINE RICHSD MOLE-FLOW STREAM=10 SUBSTREAM=MIXED &
  COMPONENT=NH4HCO3S
DEFINE ABSBTU BLOCK-VAR BLOCK=ABSORBER VARIABLE=NET-DUTY &
  SENTENCE=PARAM
DEFINE RICHM STREAM-VAR STREAM=10 SUBSTREAM=MIXED &
  VARIABLE=MASS-FLOW
DEFINE WWWFLO MASS-FLOW STREAM=19 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE DCC1V STREAM-PROP STREAM=1 PROPERTY=CUMFLOW
DEFINE DCC2V STREAM-PROP STREAM=33 PROPERTY=CUMFLOW
DEFINE ABSV STREAM-PROP STREAM=7 PROPERTY=CUMFLOW
DEFINE ABSWFL UTILITY-VAR UTILITY=U-1 &
  VARIABLE=REQUIREMENT SENTENCE=RESULTS
DEFINE ABCO2I MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE ABNH3I MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE ABH2OI MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE ABARIN MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=AR
DEFINE ABN2IN MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=N2
DEFINE ABO2IN MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=O2
DEFINE WWCO2 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE WWNH3 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &

```

```

COMPONENT=NH3
DEFINE WWH2O MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
COMPONENT=H2O
DEFINE WWAR MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
COMPONENT=AR
DEFINE WWN2 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
COMPONENT=N2
DEFINE WWO2 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
COMPONENT=O2
DEFINE HTX1 BLOCK-VAR BLOCK=HTX1 VARIABLE=CALC-AREA &
SENTENCE=RESULTS
DEFINE HTX2 BLOCK-VAR BLOCK=HTX2 VARIABLE=CALC-AREA &
SENTENCE=RESULTS
TABULATE 1 "COLWTF"
TABULATE 2 "CHLWTF"
TABULATE 3 "CHLBTU"
TABULATE 4 "ABSCO2"
TABULATE 5 "ABSNH3"
TABULATE 6 "ABSH2O"
TABULATE 7 "ABSAR"
TABULATE 8 "ABSN2"
TABULATE 9 "ABSO2"
TABULATE 10 "ABSKG"
TABULATE 11 "TOTCO2"
TABULATE 12 "TOTNH3"
TABULATE 13 "TOTH2O"
TABULATE 14 "TOTAR"
TABULATE 15 "TOTN2"
TABULATE 16 "TOTO2"
TABULATE 17 "RICHSD"
TABULATE 18 "ABSBTU"
TABULATE 19 "LNBTU"
TABULATE 20 "RICHM"
TABULATE 21 "WWWFLO"
TABULATE 22 "MAKNH3"
TABULATE 23 "MAKH2O"
TABULATE 24 "MAKCO2"
TABULATE 25 "STRCO2"
TABULATE 26 "STRNH3"
TABULATE 27 "STRH2O"
TABULATE 28 "STRAR"
TABULATE 29 "STRN2"
TABULATE 30 "STRO2"
TABULATE 31 "HTRBTU"
TABULATE 32 "STRBTU"
TABULATE 33 "FSHBTU"
TABULATE 34 "NH3BTU"
TABULATE 35 "DCC1V"
TABULATE 36 "DCC2V"
TABULATE 37 "ABSV"
TABULATE 38 "ABSWFL"
TABULATE 39 "HTX3A"
TABULATE 40 "STRFLO"
TABULATE 41 "CMPFLO"
TABULATE 42 "MIXNH3"
TABULATE 43 "MIXCO2"

```

TABULATE 44 "MIXH2O"  
 TABULATE 45 "PURNH3"  
 TABULATE 46 "PURCO2"  
 TABULATE 47 "PURH2O"  
 TABULATE 48 "CLRARA"  
 TABULATE 49 "CLRWAT"  
 TABULATE 50 "HTX4A"  
 TABULATE 51 "STPUTP"  
 TABULATE 52 "STPBTP"  
 TABULATE 53 "ABCO2I"  
 TABULATE 54 "ABNH3I"  
 TABULATE 55 "ABH2OI"  
 TABULATE 56 "ABARIN"  
 TABULATE 57 "ABN2IN"  
 TABULATE 58 "ABO2IN"  
 TABULATE 59 "WWCO2"  
 TABULATE 60 "WWNH3"  
 TABULATE 61 "WWH2O"  
 TABULATE 62 "WWAR"  
 TABULATE 63 "WWN2"  
 TABULATE 64 "WWO2"  
 TABULATE 65 "HPPMW"  
 TABULATE 66 "HTX1"  
 TABULATE 67 "HTX2"  
 VARY STREAM-VAR STREAM=1 SUBSTREAM=MIXED VARIABLE=TEMP  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=CO2  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=N2  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=H2O  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=O2  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=AR  
 VARY STREAM-VAR STREAM=38 SUBSTREAM=MIXED VARIABLE=TEMP  
 VARY MOLE-FLOW STREAM=37 SUBSTREAM=MIXED COMPONENT=H2O  
 VARY STREAM-VAR STREAM=5 SUBSTREAM=MIXED VARIABLE=TEMP  
 VARY MOLE-FLOW STREAM=8 SUBSTREAM=MIXED COMPONENT=NH3  
 VARY MOLE-FLOW STREAM=8 SUBSTREAM=MIXED COMPONENT=CO2  
 VARY MOLE-FLOW STREAM=8 SUBSTREAM=MIXED COMPONENT=H2O  
 VARY BLOCK-VAR BLOCK=ABSORBER VARIABLE=TEMP SENTENCE=PARAM  
 VARY BLOCK-VAR BLOCK=ABSORBER VARIABLE=PRES SENTENCE=PARAM  
 VARY BLOCK-VAR BLOCK=ABSORBER VARIABLE=PRES SENTENCE=PARAM  
 VARY BLOCK-VAR BLOCK=ABSORBER VARIABLE=PRES SENTENCE=PARAM  
 VARY MOLE-FLOW STREAM=19 SUBSTREAM=MIXED COMPONENT=H2O  
 VARY BLOCK-VAR BLOCK=ABSORBER VARIABLE=PRES SENTENCE=PARAM  
 VARY BLOCK-VAR BLOCK=ABSORBER VARIABLE=PRES SENTENCE=PARAM  
 REINIT BLOCKS=ALL STREAMS=ALL  
 CASES 1 VALUES= 131. 4842. 20565. 2911. 706.047453 &  
     240.35658 71. 99151. 37. 8820. 3528. 19545. 50. 1. &  
     1. 1. 19569. 1. 1.  
 CASES 2 VALUES= 153. 5434. 22352. 2360. 706.047453 &  
     240.35658 86. 63905. 37. 4483. 1793.2 15580. 50. 1. &  
     1. 1. 6301. 1. 1.  
 CASES 3 VALUES= 145. 2737. 25093. 3135. 706.047453 &  
     240.35658 83. 51182. 37. 5178. 2071.2 16880. 50. 1. &  
     1. 1. 5597. 1. 1.  
 CASES 4 VALUES= 126. 2447. 15400. 4615. 706.047453 &  
     240.35658 89. 43131. 37. 3337. 1334.8 17575. 50. 1. &  
     1. 1. 13547. 1. 1.

CASES 5 VALUES= 155. 5835. 23309. 4951. 706.047453 &  
240.35658 89. 91228. 37. 4037. 1614.8 19916. 50. 1. &  
1. 1. 19093. 1. 1.

CASES 6 VALUES= 124. 5181. 18896. 2403. 706.047453 &  
240.35658 87. 44506. 37. 7849. 3139.6 6142. 50. 1. &  
1. 1. 19117. 1. 1.

CASES 7 VALUES= 122. 3373. 27415. 3068. 706.047453 &  
240.35658 85. 66317. 37. 7050. 2820. 16386. 50. 1. &  
1. 1. 19001. 1. 1.

CASES 8 VALUES= 124. 4493. 18905. 5426. 706.047453 &  
240.35658 89. 87215. 37. 4376. 1750.4 10990. 50. 1. &  
1. 1. 2869. 1. 1.

CASES 9 VALUES= 124. 2039. 17903. 1926. 706.047453 &  
240.35658 77. 60371. 37. 5692. 2276.8 18178. 50. 1. &  
1. 1. 2504. 1. 1.

CASES 10 VALUES= 129. 3746. 17268. 1365. 706.047453 &  
240.35658 73. 84107. 37. 8932. 3572.8 7936. 50. 1. &  
1. 1. 16321. 1. 1.

CASES 11 VALUES= 132. 5921. 21822. 2315. 706.047453 &  
240.35658 75. 91768. 37. 1068. 427.2 16829. 50. 1. &  
1. 1. 9028. 1. 1.

CASES 12 VALUES= 159. 5296. 23077. 1839. 706.047453 &  
240.35658 84. 46763. 37. 5748. 2299.2 20557. 50. 1. &  
1. 1. 12575. 1. 1.

CASES 13 VALUES= 137. 2525. 23352. 3354. 706.047453 &  
240.35658 88. 38940. 37. 9953. 3981.2 10566. 50. 1. &  
1. 1. 11090. 1. 1.

CASES 14 VALUES= 150. 2561. 21603. 1288. 706.047453 &  
240.35658 78. 63896. 37. 8928. 3571.2 13416. 50. 1. &  
1. 1. 19083. 1. 1.

CASES 15 VALUES= 152. 4621. 18054. 5801. 706.047453 &  
240.35658 83. 85736. 37. 7195. 2878. 14821. 50. 1. &  
1. 1. 15925. 1. 1.

CASES 16 VALUES= 150. 2550. 20232. 3180. 706.047453 &  
240.35658 74. 55874. 37. 6460. 2584. 16439. 50. 1. &  
1. 1. 18535. 1. 1.

CASES 17 VALUES= 149. 5232. 18885. 3127. 706.047453 &  
240.35658 73. 87160. 37. 6754. 2701.6 17349. 50. 1. &  
1. 1. 17771. 1. 1.

CASES 18 VALUES= 120. 5219. 21625. 1433. 706.047453 &  
240.35658 86. 61471. 37. 1737. 694.8 10139. 50. 1. &  
1. 1. 14053. 1. 1.

CASES 19 VALUES= 126. 2882. 22754. 1871. 706.047453 &  
240.35658 87. 40452. 37. 6427. 2570.8 19424. 50. 1. &  
1. 1. 12770. 1. 1.

CASES 20 VALUES= 137. 3605. 29942. 1327. 706.047453 &  
240.35658 71. 89646. 37. 4027. 1610.8 22586. 50. 1. &  
1. 1. 6497. 1. 1.

CASES 21 VALUES= 124. 2767. 29056. 2266. 706.047453 &  
240.35658 83. 39884. 37. 9714. 3885.6 22136. 50. 1. &  
1. 1. 2510. 1. 1.

CASES 22 VALUES= 144. 3654. 21548. 1094. 706.047453 &  
240.35658 78. 75574. 37. 7785. 3114. 6960. 50. 1. 1. &  
1. 19942. 1. 1.

CASES 23 VALUES= 140. 2245. 19115. 5812. 706.047453 &  
240.35658 70. 50465. 37. 4396. 1758.4 15529. 50. 1. &

1. 1. 6832. 1. 1.  
CASES 24 VALUES= 144. 4796. 17582. 1300. 706.047453 &  
240.35658 72. 70662. 37. 8764. 3505.6 11285. 50. 1. &  
1. 1. 7051. 1. 1.  
CASES 25 VALUES= 139. 3013. 21553. 5366. 706.047453 &  
240.35658 80. 72238. 37. 5242. 2096.8 11886. 50. 1. &  
1. 1. 10072. 1. 1.  
CASES 26 VALUES= 143. 3722. 22273. 4745. 706.047453 &  
240.35658 81. 85422. 37. 2905. 1162. 12946. 50. 1. &  
1. 1. 19123. 1. 1.  
CASES 27 VALUES= 132. 4134. 29628. 4417. 706.047453 &  
240.35658 71. 89503. 37. 7357. 2942.8 23384. 50. 1. &  
1. 1. 16654. 1. 1.  
CASES 28 VALUES= 146. 4553. 22161. 1423. 706.047453 &  
240.35658 80. 38885. 37. 8767. 3506.8 15539. 50. 1. &  
1. 1. 17955. 1. 1.  
CASES 29 VALUES= 159. 5964. 21745. 1707. 706.047453 &  
240.35658 71. 50728. 37. 5978. 2391.2 7371. 50. 1. &  
1. 1. 4175. 1. 1.  
CASES 30 VALUES= 126. 4410. 26258. 5749. 706.047453 &  
240.35658 78. 75571. 37. 2832. 1132.8 12456. 50. 1. &  
1. 1. 13967. 1. 1.  
CASES 31 VALUES= 149. 3242. 17740. 2021. 706.047453 &  
240.35658 77. 51167. 37. 6881. 2752.4 21971. 50. 1. &  
1. 1. 19264. 1. 1.  
CASES 32 VALUES= 149. 2463. 29919. 3423. 706.047453 &  
240.35658 79. 69713. 37. 8766. 3506.4 22062. 50. 1. &  
1. 1. 6420. 1. 1.  
CASES 33 VALUES= 145. 3900. 29004. 3838. 706.047453 &  
240.35658 79. 78950. 37. 7845. 3138. 13143. 50. 1. &  
1. 1. 19237. 1. 1.  
CASES 34 VALUES= 153. 5809. 18222. 1426. 706.047453 &  
240.35658 72. 36584. 37. 7304. 2921.6 9548. 50. 1. &  
1. 1. 5814. 1. 1.  
CASES 35 VALUES= 126. 5911. 22656. 2654. 706.047453 &  
240.35658 77. 46580. 37. 1696. 678.4 23698. 50. 1. &  
1. 1. 4226. 1. 1.  
CASES 36 VALUES= 149. 4614. 23273. 2042. 706.047453 &  
240.35658 81. 63660. 37. 4834. 1933.6 14417. 50. 1. &  
1. 1. 18841. 1. 1.  
CASES 37 VALUES= 150. 3357. 20419. 4430. 706.047453 &  
240.35658 72. 95643. 37. 1107. 442.8 9242. 50. 1. 1. &  
1. 18017. 1. 1.  
CASES 38 VALUES= 133. 3458. 21413. 3748. 706.047453 &  
240.35658 81. 54665. 37. 1209. 483.6 20104. 50. 1. &  
1. 1. 4885. 1. 1.  
CASES 39 VALUES= 125. 4362. 28560. 3206. 706.047453 &  
240.35658 75. 60900. 37. 2504. 1001.6 21283. 50. 1. &  
1. 1. 15163. 1. 1.  
CASES 40 VALUES= 128. 3832. 25916. 4102. 706.047453 &  
240.35658 73. 79291. 37. 8460. 3384. 5646. 50. 1. 1. &  
1. 11335. 1. 1.  
CASES 41 VALUES= 147. 3077. 20410. 1897. 706.047453 &  
240.35658 79. 44479. 37. 6224. 2489.6 8998. 50. 1. &  
1. 1. 1556. 1. 1.  
CASES 42 VALUES= 151. 4693. 16974. 1358. 706.047453 &

240.35658 89. 79349. 37. 3877. 1550.8 12268. 50. 1. &  
1. 1. 9585. 1. 1.  
CASES 43 VALUES= 160. 5436. 23447. 2927. 706.047453 &  
240.35658 84. 66226. 37. 4176. 1670.4 8703. 50. 1. &  
1. 1. 16324. 1. 1.  
CASES 44 VALUES= 158. 4454. 24553. 1800. 706.047453 &  
240.35658 90. 62633. 37. 6104. 2441.6 20948. 50. 1. &  
1. 1. 4885. 1. 1.  
CASES 45 VALUES= 139. 2737. 17132. 3547. 706.047453 &  
240.35658 75. 40975. 37. 7644. 3057.6 6566. 50. 1. &  
1. 1. 10590. 1. 1.  
CASES 46 VALUES= 146. 2959. 19088. 2882. 706.047453 &  
240.35658 80. 65552. 37. 4753. 1901.2 6847. 50. 1. &  
1. 1. 12025. 1. 1.  
CASES 47 VALUES= 144. 3525. 18318. 5970. 706.047453 &  
240.35658 80. 57123. 37. 6543. 2617.2 20980. 50. 1. &  
1. 1. 16854. 1. 1.  
CASES 48 VALUES= 142. 2680. 20953. 2809. 706.047453 &  
240.35658 84. 79894. 37. 9350. 3740. 18730. 50. 1. &  
1. 1. 12786. 1. 1.  
CASES 49 VALUES= 147. 5488. 28179. 5587. 706.047453 &  
240.35658 78. 76496. 37. 4841. 1936.4 11028. 50. 1. &  
1. 1. 13127. 1. 1.  
CASES 50 VALUES= 138. 4945. 29220. 4518. 706.047453 &  
240.35658 90. 60050. 37. 3369. 1347.6 6471. 50. 1. &  
1. 1. 8454. 1. 1.  
CASES 51 VALUES= 124. 4479. 22514. 5055. 706.047453 &  
240.35658 88. 32285. 37. 5789. 2315.6 15351. 50. 1. &  
1. 1. 16164. 1. 1.  
CASES 52 VALUES= 146. 3394. 22726. 5503. 706.047453 &  
240.35658 83. 64462. 37. 8091. 3236.4 19286. 50. 1. &  
1. 1. 15872. 1. 1.  
CASES 53 VALUES= 152. 4240. 23220. 1503. 706.047453 &  
240.35658 76. 49858. 37. 6657. 2662.8 17000. 50. 1. &  
1. 1. 12932. 1. 1.  
CASES 54 VALUES= 160. 5066. 27519. 4531. 706.047453 &  
240.35658 79. 47484. 37. 1543. 617.2 10201. 50. 1. &  
1. 1. 9648. 1. 1.  
CASES 55 VALUES= 132. 5391. 29300. 2703. 706.047453 &  
240.35658 75. 75955. 37. 6678. 2671.2 15057. 50. 1. &  
1. 1. 4883. 1. 1.  
CASES 56 VALUES= 160. 5693. 26838. 5935. 706.047453 &  
240.35658 73. 87707. 37. 4879. 1951.6 23357. 50. 1. &  
1. 1. 17659. 1. 1.  
CASES 57 VALUES= 159. 3243. 24490. 5943. 706.047453 &  
240.35658 72. 32429. 37. 5187. 2074.8 16104. 50. 1. &  
1. 1. 4409. 1. 1.  
CASES 58 VALUES= 133. 5207. 26859. 2136. 706.047453 &  
240.35658 85. 76671. 37. 3287. 1314.8 13126. 50. 1. &  
1. 1. 6607. 1. 1.  
CASES 59 VALUES= 149. 3107. 26854. 5613. 706.047453 &  
240.35658 77. 78153. 37. 1429. 571.6 7338. 50. 1. 1. &  
1. 6130. 1. 1.  
CASES 60 VALUES= 156. 2231. 22610. 2348. 706.047453 &  
240.35658 89. 31782. 37. 6526. 2610.4 8776. 50. 1. &  
1. 1. 2298. 1. 1.

CASES 61 VALUES= 140. 5984. 19042. 1695. 706.047453 &  
240.35658 86. 56112. 37. 6778. 2711.2 21230. 50. 1. &  
1. 1. 8982. 1. 1.

CASES 62 VALUES= 137. 2177. 26125. 5862. 706.047453 &  
240.35658 73. 90956. 37. 2247. 898.8 9791. 50. 1. 1. &  
1. 3369. 1. 1.

CASES 63 VALUES= 153. 2794. 24649. 1569. 706.047453 &  
240.35658 89. 99030. 37. 3136. 1254.4 21055. 50. 1. &  
1. 1. 2780. 1. 1.

CASES 64 VALUES= 138. 5583. 19270. 1061. 706.047453 &  
240.35658 74. 58325. 37. 2264. 905.6 6354. 50. 1. 1. &  
1. 7825. 1. 1.

CASES 65 VALUES= 124. 5500. 24336. 5491. 706.047453 &  
240.35658 76. 48571. 37. 3165. 1266. 21193. 50. 1. &  
1. 1. 10289. 1. 1.

CASES 66 VALUES= 159. 5391. 24390. 1431. 706.047453 &  
240.35658 72. 82315. 37. 3567. 1426.8 6573. 50. 1. &  
1. 1. 10310. 1. 1.

CASES 67 VALUES= 134. 4445. 27497. 3774. 706.047453 &  
240.35658 83. 75501. 37. 7935. 3174. 16488. 50. 1. &  
1. 1. 12788. 1. 1.

CASES 68 VALUES= 149. 3314. 29490. 3409. 706.047453 &  
240.35658 76. 65502. 37. 8484. 3393.6 6641. 50. 1. &  
1. 1. 14377. 1. 1.

CASES 69 VALUES= 141. 5994. 23209. 5964. 706.047453 &  
240.35658 88. 51913. 37. 7238. 2895.2 9854. 50. 1. &  
1. 1. 6923. 1. 1.

CASES 70 VALUES= 132. 3169. 20185. 1762. 706.047453 &  
240.35658 76. 89376. 37. 6399. 2559.6 22219. 50. 1. &  
1. 1. 14615. 1. 1.

CASES 71 VALUES= 160. 5582. 25158. 3236. 706.047453 &  
240.35658 78. 49793. 37. 1661. 664.4 7503. 50. 1. 1. &  
1. 15488. 1. 1.

CASES 72 VALUES= 121. 2805. 15658. 2707. 706.047453 &  
240.35658 86. 65076. 37. 7954. 3181.6 19622. 50. 1. &  
1. 1. 7957. 1. 1.

CASES 73 VALUES= 155. 3736. 23558. 3564. 706.047453 &  
240.35658 82. 38099. 37. 5441. 2176.4 8891. 50. 1. &  
1. 1. 15276. 1. 1.

CASES 74 VALUES= 127. 4189. 25880. 1419. 706.047453 &  
240.35658 80. 71283. 37. 9882. 3952.8 12224. 50. 1. &  
1. 1. 17354. 1. 1.

CASES 75 VALUES= 123. 3500. 20545. 3162. 706.047453 &  
240.35658 81. 62875. 37. 4981. 1992.4 15762. 50. 1. &  
1. 1. 7140. 1. 1.

CASES 76 VALUES= 129. 5758. 24354. 4018. 706.047453 &  
240.35658 79. 39407. 37. 6980. 2792. 16518. 50. 1. &  
1. 1. 3906. 1. 1.

CASES 77 VALUES= 129. 4957. 29161. 1144. 706.047453 &  
240.35658 70. 96428. 37. 6200. 2480. 22874. 50. 1. &  
1. 1. 17840. 1. 1.

CASES 78 VALUES= 133. 3374. 22168. 4369. 706.047453 &  
240.35658 88. 77419. 37. 4989. 1995.6 10266. 50. 1. &  
1. 1. 11576. 1. 1.

CASES 79 VALUES= 135. 3945. 23090. 2087. 706.047453 &  
240.35658 89. 36480. 37. 8517. 3406.8 8644. 50. 1. &

1. 1. 18328. 1. 1.  
CASES 80 VALUES= 126. 3352. 24834. 4837. 706.047453 &  
240.35658 73. 84477. 37. 6095. 2438. 17173. 50. 1. &  
1. 1. 8041. 1. 1.  
CASES 81 VALUES= 136. 2368. 15478. 4004. 706.047453 &  
240.35658 71. 74630. 37. 3916. 1566.4 15781. 50. 1. &  
1. 1. 9039. 1. 1.  
CASES 82 VALUES= 146. 4068. 22070. 5788. 706.047453 &  
240.35658 87. 66648. 37. 9809. 3923.6 14415. 50. 1. &  
1. 1. 14593. 1. 1.  
CASES 83 VALUES= 155. 2075. 21257. 5189. 706.047453 &  
240.35658 80. 67401. 37. 8000. 3200. 5094. 50. 1. 1. &  
1. 3287. 1. 1.  
CASES 84 VALUES= 130. 2544. 19312. 4831. 706.047453 &  
240.35658 86. 72653. 37. 5007. 2002.8 14821. 50. 1. &  
1. 1. 8101. 1. 1.  
CASES 85 VALUES= 137. 3281. 29670. 1603. 706.047453 &  
240.35658 80. 67418. 37. 9021. 3608.4 18938. 50. 1. &  
1. 1. 8359. 1. 1.  
CASES 86 VALUES= 143. 3040. 24926. 2264. 706.047453 &  
240.35658 80. 81999. 37. 2551. 1020.4 12604. 50. 1. &  
1. 1. 2162. 1. 1.  
CASES 87 VALUES= 155. 3029. 23042. 2489. 706.047453 &  
240.35658 72. 57198. 37. 1103. 441.2 11183. 50. 1. &  
1. 1. 18660. 1. 1.  
CASES 88 VALUES= 145. 3407. 28506. 3333. 706.047453 &  
240.35658 86. 62328. 37. 8459. 3383.6 18675. 50. 1. &  
1. 1. 12242. 1. 1.  
CASES 89 VALUES= 151. 3686. 21140. 2749. 706.047453 &  
240.35658 88. 35197. 37. 1035. 414. 6077. 50. 1. 1. &  
1. 6644. 1. 1.  
CASES 90 VALUES= 133. 3674. 21372. 5567. 706.047453 &  
240.35658 77. 43687. 37. 8294. 3317.6 5001. 50. 1. &  
1. 1. 16742. 1. 1.  
CASES 91 VALUES= 123. 2233. 24616. 5991. 706.047453 &  
240.35658 79. 46437. 37. 7749. 3099.6 8266. 50. 1. &  
1. 1. 18520. 1. 1.  
CASES 92 VALUES= 133. 5553. 16363. 3678. 706.047453 &  
240.35658 80. 30420. 37. 9911. 3964.4 23612. 50. 1. &  
1. 1. 18196. 1. 1.  
CASES 93 VALUES= 151. 2843. 16641. 2749. 706.047453 &  
240.35658 75. 88716. 37. 7223. 2889.2 19276. 50. 1. &  
1. 1. 11043. 1. 1.  
CASES 94 VALUES= 121. 2665. 26582. 5342. 706.047453 &  
240.35658 73. 80301. 37. 3060. 1224. 14396. 50. 1. &  
1. 1. 2586. 1. 1.  
CASES 95 VALUES= 154. 5537. 22835. 4251. 706.047453 &  
240.35658 70. 31133. 37. 9139. 3655.6 21676. 50. 1. &  
1. 1. 9132. 1. 1.  
CASES 96 VALUES= 140. 4366. 21308. 4804. 706.047453 &  
240.35658 78. 67311. 37. 2197. 878.8 9796. 50. 1. 1. &  
1. 4749. 1. 1.  
CASES 97 VALUES= 137. 4798. 21584. 4448. 706.047453 &  
240.35658 86. 34287. 37. 9084. 3633.6 9125. 50. 1. &  
1. 1. 13534. 1. 1.  
CASES 98 VALUES= 138. 4878. 21107. 4935. 706.047453 &

240.35658 74. 41569. 37. 4002. 1600.8 8534. 50. 1. &  
 1. 1. 10893. 1. 1.  
 CASES 99 VALUES= 128. 5514. 27105. 1667. 706.047453 &  
 240.35658 87. 97433. 37. 3829. 1531.6 7030. 50. 1. &  
 1. 1. 1967. 1. 1.  
 CASES 100 VALUES= 134. 2970. 21825. 2690. 706.047453 &  
 240.35658 77. 87450. 37. 7819. 3127.6 23942. 50. 1. &  
 1. 1. 2754. 1. 1.

SENSITIVITY AMMONIA

PARAM WIDE=NO BASE-CASE=NO  
 DEFINE COLWTF STREAM-VAR STREAM=38 SUBSTREAM=MIXED &  
 VARIABLE=MASS-FLOW  
 DEFINE CHLWTF STREAM-VAR STREAM=5 SUBSTREAM=MIXED &  
 VARIABLE=MASS-FLOW  
 DEFINE CHLBTU BLOCK-VAR BLOCK=HTX2 VARIABLE=CALC-DUTY &  
 SENTENCE=RESULTS  
 DEFINE ABSCO2 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=CO2  
 DEFINE ABSNH3 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=NH3  
 DEFINE ABSH2O MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=H2O  
 DEFINE ABSAR MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=AR  
 DEFINE ABSN2 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=N2  
 DEFINE ABSO2 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=O2  
 DEFINE ABSKG STREAM-VAR STREAM=9 SUBSTREAM=MIXED &  
 VARIABLE=MASS-FLOW  
 DEFINE TOTCO2 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &  
 COMPONENT=CO2  
 DEFINE TOTNH3 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &  
 COMPONENT=NH3  
 DEFINE TOTH2O MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &  
 COMPONENT=H2O  
 DEFINE TOTAR MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &  
 COMPONENT=AR  
 DEFINE TOTN2 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &  
 COMPONENT=N2  
 DEFINE TOTO2 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &  
 COMPONENT=O2  
 DEFINE RICHSD MOLE-FLOW STREAM=10 SUBSTREAM=MIXED &  
 COMPONENT=NH4HCO3S  
 DEFINE ABSBTU BLOCK-VAR BLOCK=ABSORBER VARIABLE=NET-DUTY &  
 SENTENCE=PARAM  
 DEFINE RICHM STREAM-VAR STREAM=10 SUBSTREAM=MIXED &  
 VARIABLE=MASS-FLOW  
 DEFINE WWWFLO MASS-FLOW STREAM=19 SUBSTREAM=MIXED &  
 COMPONENT=H2O  
 DEFINE MAKNH3 MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &  
 COMPONENT=NH3  
 DEFINE MAKH2O MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &  
 COMPONENT=H2O  
 DEFINE MAKCO2 MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &

```

COMPONENT=CO2
DEFINE NH3BTU BLOCK-VAR BLOCK=NH3STRIP VARIABLE=REB-DUTY &
  SENTENCE=RESULTS
DEFINE DCC1V STREAM-PROP STREAM=1 PROPERTY=CUMFLOW
DEFINE DCC2V STREAM-PROP STREAM=33 PROPERTY=CUMFLOW
DEFINE ABSV STREAM-PROP STREAM=7 PROPERTY=CUMFLOW
DEFINE ABSWFL UTILITY-VAR UTILITY= U-1 &
  VARIABLE=REQUIREMENT SENTENCE=RESULTS
DEFINE HTX4A BLOCK-VAR BLOCK=HTX4 VARIABLE=CALC-AREA &
  SENTENCE=RESULTS
DEFINE ABCO2I MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE ABNH3I MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE ABH2OI MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE ABARIN MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=AR
DEFINE ABN2IN MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=N2
DEFINE ABO2IN MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
  COMPONENT=O2
DEFINE WWCO2 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE WWNH3 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE WWH2O MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE WWAR MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
  COMPONENT=AR
DEFINE WWN2 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
  COMPONENT=N2
DEFINE WWO2 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
  COMPONENT=O2
DEFINE HTX1 BLOCK-VAR BLOCK=HTX1 VARIABLE=CALC-AREA &
  SENTENCE=RESULTS
DEFINE HTX2 BLOCK-VAR BLOCK=HTX2 VARIABLE=CALC-AREA &
  SENTENCE=RESULTS
TABULATE 1 "COLWTF"
TABULATE 2 "CHLWTF"
TABULATE 3 "CHLBTU"
TABULATE 4 "ABSCO2"
TABULATE 5 "ABSNH3"
TABULATE 6 "ABSH2O"
TABULATE 7 "ABSAR"
TABULATE 8 "ABSN2"
TABULATE 9 "ABSO2"
TABULATE 10 "ABSKG"
TABULATE 11 "TOTCO2"
TABULATE 12 "TOTNH3"
TABULATE 13 "TOH2O"
TABULATE 14 "TOTAR"
TABULATE 15 "TOTN2"
TABULATE 16 "TOTO2"
TABULATE 17 "RICHSD"
TABULATE 18 "ABSBTU"

```

TABULATE 19 "LNBTU"  
TABULATE 20 "RICHM"  
TABULATE 21 "WWWFLO"  
TABULATE 22 "MAKNH3"  
TABULATE 23 "MAKH2O"  
TABULATE 24 "MAKCO2"  
TABULATE 25 "STRCO2"  
TABULATE 26 "STRNH3"  
TABULATE 27 "STRH2O"  
TABULATE 28 "STRAR"  
TABULATE 29 "STRN2"  
TABULATE 30 "STRO2"  
TABULATE 31 "HTRBTU"  
TABULATE 32 "STRBTU"  
TABULATE 33 "FSHBTU"  
TABULATE 34 "NH3BTU"  
TABULATE 35 "DCC1V"  
TABULATE 36 "DCC2V"  
TABULATE 37 "ABSV"  
TABULATE 38 "ABSWFL"  
TABULATE 39 "HTX3A"  
TABULATE 40 "STRFLO"  
TABULATE 41 "CMPFLO"  
TABULATE 42 "MIXNH3"  
TABULATE 43 "MIXCO2"  
TABULATE 44 "MIXH2O"  
TABULATE 45 "PURNH3"  
TABULATE 46 "PURCO2"  
TABULATE 47 "PURH2O"  
TABULATE 48 "CLRARA"  
TABULATE 49 "CLRWAT"  
TABULATE 50 "HTX4A"  
TABULATE 51 "STPUTP"  
TABULATE 52 "STPBTP"  
TABULATE 53 "ABCO2I"  
TABULATE 54 "ABNH3I"  
TABULATE 55 "ABH2OI"  
TABULATE 56 "ABARIN"  
TABULATE 57 "ABN2IN"  
TABULATE 58 "ABO2IN"  
TABULATE 59 "WWCO2"  
TABULATE 60 "WWNH3"  
TABULATE 61 "WWH2O"  
TABULATE 62 "WWAR"  
TABULATE 63 "WWN2"  
TABULATE 64 "WWO2"  
TABULATE 65 "HPPMW"  
TABULATE 66 "HTX1"  
TABULATE 67 "HTX2"  
VARY STREAM-VAR STREAM=1 SUBSTREAM=MIXED VARIABLE=TEMP  
RANGE LIST=135.  
VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=CO2  
RANGE LIST=3983.91031  
VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=N2  
RANGE LIST=20102.8234  
VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=H2O

RANGE LIST=5011.43469  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=O2  
 RANGE LIST=706.047453  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=AR  
 RANGE LIST=240.35658  
 VARY STREAM-VAR STREAM=38 SUBSTREAM=MIXED VARIABLE=TEMP  
 RANGE LIST=80.  
 VARY MOLE-FLOW STREAM=37 SUBSTREAM=MIXED COMPONENT=H2O  
 RANGE LIST=80508.348  
 VARY STREAM-VAR STREAM=5 SUBSTREAM=MIXED VARIABLE=TEMP  
 RANGE LIST=37.  
 VARY MOLE-FLOW STREAM=8 SUBSTREAM=MIXED COMPONENT=NH3  
 RANGE LIST=9981.8  
 VARY MOLE-FLOW STREAM=8 SUBSTREAM=MIXED COMPONENT=CO2  
 RANGE LIST=3992.72  
 VARY MOLE-FLOW STREAM=8 SUBSTREAM=MIXED COMPONENT=H2O  
 RANGE LIST=36328.73  
 VARY BLOCK-VAR BLOCK=ABSORBER VARIABLE=TEMP SENTENCE=PARAM  
 RANGE LIST=50.  
 VARY BLOCK-VAR BLOCK=NH3STRIP VARIABLE=MOLE-D &  
 SENTENCE=COL-SPECS  
 RANGE LOWER="250" UPPER="1000" NPOINT="16"  
 VARY BLOCK-VAR BLOCK=B1 VARIABLE=TEMP SENTENCE=PARAM  
 RANGE LIST=150.  
 VARY MOLE-FLOW STREAM=19 SUBSTREAM=MIXED COMPONENT=H2O  
 RANGE LIST=4000.  
 REINIT BLOCKS=ALL STREAMS=ALL

#### SENSITIVITY FULL

PARAM WIDE=NO BASE-CASE=NO  
 DEFINE COLWTF STREAM-VAR STREAM=38 SUBSTREAM=MIXED &  
 VARIABLE=MASS-FLOW  
 DEFINE CHLWTF STREAM-VAR STREAM=5 SUBSTREAM=MIXED &  
 VARIABLE=MASS-FLOW  
 DEFINE CHLBTU BLOCK-VAR BLOCK=HTX2 VARIABLE=CALC-DUTY &  
 SENTENCE=RESULTS  
 DEFINE ABSO2 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=CO2  
 DEFINE ABSNH3 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=NH3  
 DEFINE ABSH2O MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=H2O  
 DEFINE ABSAR MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=AR  
 DEFINE ABSN2 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=N2  
 DEFINE ABSO2 MOLE-FLOW STREAM=9 SUBSTREAM=MIXED &  
 COMPONENT=O2  
 DEFINE ABSKG STREAM-VAR STREAM=9 SUBSTREAM=MIXED &  
 VARIABLE=MASS-FLOW  
 DEFINE TOTCO2 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &  
 COMPONENT=CO2  
 DEFINE TOTNH3 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &  
 COMPONENT=NH3  
 DEFINE TOTH2O MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &  
 COMPONENT=H2O

```

DEFINE TOTAR MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &
  COMPONENT=AR
DEFINE TOTN2 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &
  COMPONENT=N2
DEFINE TOTO2 MOLE-FLOW STREAM=36 SUBSTREAM=MIXED &
  COMPONENT=O2
DEFINE RICHSO MOLE-FLOW STREAM=10 SUBSTREAM=MIXED &
  COMPONENT=NH4HCO3S
DEFINE ABSBTU BLOCK-VAR BLOCK=ABSORBER VARIABLE=NET-DUTY &
  SENTENCE=PARAM
DEFINE LNBTU BLOCK-VAR BLOCK=COOLER1 VARIABLE=CALC-DUTY &
  SENTENCE=RESULTS
DEFINE RICHM STREAM-VAR STREAM=10 SUBSTREAM=MIXED &
  VARIABLE=MASS-FLOW
DEFINE WWWFLO MASS-FLOW STREAM=19 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE MAKNH3 MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE MAKH2O MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE MAKCO2 MOLE-FLOW STREAM=24 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE STRCO2 MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
  COMPONENT=CO2
DEFINE STRNH3 MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE STRH2O MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
  COMPONENT=H2O
DEFINE STRAR MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
  COMPONENT=AR
DEFINE STRN2 MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
  COMPONENT=N2
DEFINE STRO2 MOLE-FLOW STREAM=15 SUBSTREAM=MIXED &
  COMPONENT=O2
DEFINE HTRBTU BLOCK-VAR BLOCK=HEATER1 VARIABLE=NET-DUTY &
  SENTENCE=RESULTS
DEFINE STRBTU BLOCK-VAR BLOCK=STRIPPER VARIABLE=REB-DUTY &
  SENTENCE=RESULTS
DEFINE FSHBTU BLOCK-VAR BLOCK=FLASH2 VARIABLE=NET-DUTY &
  SENTENCE=PARAM
DEFINE NH3BTU BLOCK-VAR BLOCK=NH3STRIP VARIABLE=REB-DUTY &
  SENTENCE=RESULTS
DEFINE DCC1V STREAM-PROP STREAM=1 PROPERTY=CUMFLOW
DEFINE DCC2V STREAM-PROP STREAM=33 PROPERTY=CUMFLOW
DEFINE ABSV STREAM-PROP STREAM=7 PROPERTY=CUMFLOW
DEFINE ABSWFL UTILITY-VAR UTILITY= U-1 &
  VARIABLE=REQUIREMENT SENTENCE=RESULTS
DEFINE HTX3A BLOCK-VAR BLOCK=HTX3 VARIABLE=CALC-AREA &
  SENTENCE=RESULTS
DEFINE STRFLO STREAM-VAR STREAM=13 SUBSTREAM=MIXED &
  VARIABLE=MASS-FLOW
DEFINE CMPFLO STREAM-VAR STREAM=15 SUBSTREAM=MIXED &
  VARIABLE=MASS-FLOW
DEFINE MIXNH3 MOLE-FLOW STREAM=27 SUBSTREAM=MIXED &
  COMPONENT=NH3
DEFINE MIXCO2 MOLE-FLOW STREAM=27 SUBSTREAM=MIXED &

```

```

COMPONENT=CO2
DEFINE MIXH2O MOLE-FLOW STREAM=27 SUBSTREAM=MIXED &
COMPONENT=H2O
DEFINE PURNH3 MOLE-FLOW STREAM=32 SUBSTREAM=MIXED &
COMPONENT=NH3
DEFINE PURCO2 MOLE-FLOW STREAM=32 SUBSTREAM=MIXED &
COMPONENT=CO2
DEFINE PURH2O MOLE-FLOW STREAM=32 SUBSTREAM=MIXED &
COMPONENT=H2O
DEFINE CLRARA BLOCK-VAR BLOCK=COOLER1 VARIABLE=CALC-AREA &
SENTENCE=RESULTS
DEFINE CLRWAT MOLE-FLOW STREAM=30 SUBSTREAM=MIXED &
COMPONENT=H2O
DEFINE HTX4A BLOCK-VAR BLOCK=HTX4 VARIABLE=CALC-AREA &
SENTENCE=RESULTS
DEFINE STPUTP BLOCK-VAR BLOCK=STRIPPER VARIABLE=TEMP &
SENTENCE=PROFILE ID1=1
DEFINE STPBTP BLOCK-VAR BLOCK=STRIPPER VARIABLE=TEMP &
SENTENCE=PROFILE ID1=8
DEFINE ABCO2I MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
COMPONENT=CO2
DEFINE ABNH3I MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
COMPONENT=NH3
DEFINE ABH2OI MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
COMPONENT=H2O
DEFINE ABARIN MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
COMPONENT=AR
DEFINE ABN2IN MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
COMPONENT=N2
DEFINE ABO2IN MOLE-FLOW STREAM=7 SUBSTREAM=MIXED &
COMPONENT=O2
DEFINE WWCO2 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
COMPONENT=CO2
DEFINE WWNH3 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
COMPONENT=NH3
DEFINE WWH2O MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
COMPONENT=H2O
DEFINE WWAR MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
COMPONENT=AR
DEFINE WWN2 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
COMPONENT=N2
DEFINE WWO2 MOLE-FLOW STREAM=33 SUBSTREAM=MIXED &
COMPONENT=O2
DEFINE HPPMW BLOCK-VAR BLOCK=HPPUMP VARIABLE=NET-WORK &
SENTENCE=RESULTS
DEFINE HTX1 BLOCK-VAR BLOCK=HTX1 VARIABLE=CALC-AREA &
SENTENCE=RESULTS
DEFINE HTX2 BLOCK-VAR BLOCK=HTX2 VARIABLE=CALC-AREA &
SENTENCE=RESULTS
TABULATE 1 "COLWTF"
TABULATE 2 "CHLWTF"
TABULATE 3 "CHLBTU"
TABULATE 4 "ABSCO2"
TABULATE 5 "ABSNH3"
TABULATE 6 "ABSH2O"
TABULATE 7 "ABSAR"

```

TABULATE 8 "ABSN2"  
TABULATE 9 "ABSO2"  
TABULATE 10 "ABSKG"  
TABULATE 11 "TOTCO2"  
TABULATE 12 "TOTNH3"  
TABULATE 13 "TOTH2O"  
TABULATE 14 "TOTAR"  
TABULATE 15 "TOTN2"  
TABULATE 16 "TOTO2"  
TABULATE 17 "RICHSD"  
TABULATE 18 "ABSBTU"  
TABULATE 19 "LNBTU"  
TABULATE 20 "RICHM"  
TABULATE 21 "WWWFLO"  
TABULATE 22 "MAKNH3"  
TABULATE 23 "MAKH2O"  
TABULATE 24 "MAKCO2"  
TABULATE 25 "STRCO2"  
TABULATE 26 "STRNH3"  
TABULATE 27 "STRH2O"  
TABULATE 28 "STRAR"  
TABULATE 29 "STRN2"  
TABULATE 30 "STRO2"  
TABULATE 31 "HTRBTU"  
TABULATE 32 "STRBTU"  
TABULATE 33 "FSHBTU"  
TABULATE 34 "NH3BTU"  
TABULATE 35 "DCC1V"  
TABULATE 36 "DCC2V"  
TABULATE 37 "ABSV"  
TABULATE 38 "ABSWFL"  
TABULATE 39 "HTX3A"  
TABULATE 40 "STRFLO"  
TABULATE 41 "CMPFLO"  
TABULATE 42 "MIXNH3"  
TABULATE 43 "MIXCO2"  
TABULATE 44 "MIXH2O"  
TABULATE 45 "PURNH3"  
TABULATE 46 "PURCO2"  
TABULATE 47 "PURH2O"  
TABULATE 48 "CLRARA"  
TABULATE 49 "CLRWAT"  
TABULATE 50 "HTX4A"  
TABULATE 51 "STPUTP"  
TABULATE 52 "STPBTP"  
TABULATE 53 "ABCO2I"  
TABULATE 54 "ABNH3I"  
TABULATE 55 "ABH2OI"  
TABULATE 56 "ABARIN"  
TABULATE 57 "ABN2IN"  
TABULATE 58 "ABO2IN"  
TABULATE 59 "WWCO2"  
TABULATE 60 "WWNH3"  
TABULATE 61 "WWH2O"  
TABULATE 62 "WWAR"  
TABULATE 63 "WWN2"

TABULATE 64 "WWO2"  
 TABULATE 65 "HPPMW"  
 TABULATE 66 "HTX1"  
 TABULATE 67 "HTX2"  
 VARY STREAM-VAR STREAM=1 SUBSTREAM=MIXED VARIABLE=TEMP  
 RANGE LIST=135.  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=CO2  
 RANGE LIST=3983.91031  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=N2  
 RANGE LIST=20102.8234  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=H2O  
 RANGE LIST=5011.43469  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=O2  
 RANGE LIST=706.047453  
 VARY MOLE-FLOW STREAM=1 SUBSTREAM=MIXED COMPONENT=AR  
 RANGE LIST=240.35658  
 VARY STREAM-VAR STREAM=38 SUBSTREAM=MIXED VARIABLE=TEMP  
 RANGE LIST=80.  
 VARY MOLE-FLOW STREAM=37 SUBSTREAM=MIXED COMPONENT=H2O  
 RANGE LIST=80508.348  
 VARY STREAM-VAR STREAM=5 SUBSTREAM=MIXED VARIABLE=TEMP  
 RANGE LIST=37.  
 VARY MOLE-FLOW STREAM=8 SUBSTREAM=MIXED COMPONENT=NH3  
 RANGE LIST=9981.8  
 VARY MOLE-FLOW STREAM=8 SUBSTREAM=MIXED COMPONENT=CO2  
 RANGE LIST=3992.72  
 VARY MOLE-FLOW STREAM=8 SUBSTREAM=MIXED COMPONENT=H2O  
 RANGE LIST=36328.73  
 VARY BLOCK-VAR BLOCK=ABSORBER VARIABLE=TEMP SENTENCE=PARAM  
 RANGE LIST=50.  
 VARY BLOCK-VAR BLOCK=STRIPPER VARIABLE=MOLE-D &  
 SENTENCE=COL-SPECS  
 RANGE LOWER="4000" UPPER="4100" NPOINT="21"  
 VARY BLOCK-VAR BLOCK=NH3STRIP VARIABLE=MOLE-D &  
 SENTENCE=COL-SPECS  
 RANGE LIST=720.  
 VARY BLOCK-VAR BLOCK=B1 VARIABLE=TEMP SENTENCE=PARAM  
 RANGE LIST=150.  
 VARY MOLE-FLOW STREAM=19 SUBSTREAM=MIXED COMPONENT=H2O  
 RANGE LIST=4000.  
 VARY BLOCK-VAR BLOCK=HEATER1 VARIABLE=TEMP SENTENCE=PARAM  
 RANGE LIST=225. 230.  
 VARY BLOCK-VAR BLOCK=FLASH2 VARIABLE=TEMP SENTENCE=PARAM  
 RANGE LIST=85.  
 REINIT BLOCKS=ALL STREAMS=ALL

CONV-OPTIONS  
 WEGSTEIN MAXIT=400

STREAM-REPOR MOLEFLOW MOLEFRAC PROPERTIES=TRUEMASS SOLINDEX PH

PROPERTY-REP PCES

REACTIONS NH3-REA REAC-DIST  
 IN-UNITS SI  
 REAC-DATA 1

REAC-DATA 2  
 REAC-DATA 3  
 REAC-DATA 4 KINETIC  
 REAC-DATA 5 KINETIC  
 REAC-DATA 6 KINETIC  
 REAC-DATA 7 KINETIC  
 K-STOIC 1 A=-1.2566 B=-3335.7 C=1.4971 D=-0.0370566  
 K-STOIC 2 A=132.899 B=-13445.9 C=-22.4773  
 K-STOIC 3 A=216.049 B=-12431.7 C=-35.4819  
 RATE-CON 4 PRE-EXP=4.32E+013 ACT-ENERGY=13249. <cal/mol>  
 RATE-CON 5 PRE-EXP=2.38E+017 ACT-ENERGY=29451. <cal/mol>  
 RATE-CON 6 PRE-EXP=135000000000. ACT-ENERGY=11585. <cal/mol>  
 RATE-CON 7 PRE-EXP=4.75E+020 ACT-ENERGY=16529. <cal/mol>  
 STOIC 1 NH3 -1. / H2O -1. / NH4+ 1. / OH- 1.  
 STOIC 2 H2O -2. / H3O+ 1. / OH- 1.  
 STOIC 3 HCO3- -1. / H2O -1. / CO3-- 1. / H3O+ 1.  
 STOIC 4 CO2 -1. / OH- -1. / HCO3- 1.  
 STOIC 5 HCO3- -1. / CO2 1. / OH- 1.  
 STOIC 6 NH3 -1. / CO2 -1. / H2O -1. / NH2COO- 1. / &  
 H3O+ 1.  
 STOIC 7 NH2COO- -1. / H3O+ -1. / NH3 1. / CO2 1. / &  
 H2O 1.  
 POWLAW-EXP 4 CO2 1. / OH- 1.  
 POWLAW-EXP 5 HCO3- 1.  
 POWLAW-EXP 6 NH3 1. / CO2 1. / H2O 0.  
 POWLAW-EXP 7 NH2COO- 1. / H3O+ 1.  
 SALT NH4HCO3 NH4+ 1. / NH4HCO3 HCO3- 1.  
 K-SALT NH4HCO3 CONST-A=-914.00821 CONST-B=38648.2117 &  
 CONST-C=136.174996  
 SALT-DATA NH4HCO3

PROP-TABLE LIQNH3 FLASHCURVE  
 IN-UNITS SI MOLE-FLOW='mol/sec' PRESSURE=atm TEMPERATURE=F &  
 DELTA-T=F PDROP=atm  
 MOLE-FLOW CO2 0. / H2O 55.08 / NH3 14.  
 STATE TEMP=100. PRES=1.  
 VARY MOLE-FLOW COMP=CO2  
 RANGE LIST=1. 2. 3. 4. 5. 6. 7. 8. 9. 10. 11. 12. &  
 13. 14.  
 TABULATE PROPERTIES=PH MOLEFLOW PARTPRES SOLINDEX

DISABLE  
 BLOCK B1 COOLER1 FLASH2 HEATER1 HTX3 HTX4 MIXER NH3STRIP &  
 STRIPPER  
 STREAM 12 13 14 15 16 17 18 21 22 23 24 25 26 27 &  
 28 29 30 31 32