AN INTEGRATED MODELING FRAMEWORK FOR CARBON MANAGEMENT TECHNOLOGIES

Volume 1 – Technical Documentation:
Amine-Based CO_2 Capture and Storage Systems
for Fossil Fuel Power Plant

Final Report of

by

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ABSTRACT

CO_2 capture and storage (CCS) is gaining widespread interest as a potential method to control greenhouse gas emissions from fossil fuel sources, especially electric power plants. Commercial applications of CO_2 separation and capture technologies are found in a number of industrial process operations worldwide. Many of these capture technologies also are applicable to fossil fuel power plants, although applications to large-scale power generation remain to be demonstrated. This report describes the development of a generalized modeling framework to assess alternative CO_2 capture and storage options in the context of multi-pollutant control requirements for fossil fuel power plants. The focus of the report is on post-combustion CO_2 capture using amine-based absorption systems at pulverized coal-fired plants, which are the most prevalent technology used for power generation today. The modeling framework builds on the previously developed Integrated Environmental Control Model (IECM). The expanded version with carbon sequestration is designated as IECM-cs. The expanded modeling capability also includes natural gas combined cycle (NGCC) power plants and integrated coal gasification combined cycle (IGCC) systems as well as pulverized coal (PC) plants.

This report presents details of the performance and cost models developed for an amine-based CO_2 capture system, representing the baseline of current commercial technology. The key uncertainties and variability in process design, performance and cost parameters which influence the overall cost of carbon mitigation also are characterized.

The new performance and cost models for CO_2 capture systems have been integrated into the IECM-cs, along with models to estimate CO_2 transport and storage costs. The CO_2 control system also interacts with other emission control technologies such as flue gas desulfurization (FGD) systems for SO_2 control. The integrated model is applied to study the feasibility and cost of carbon capture and sequestration at both new and existing PC plants as well as new NGCC plants. The cost of CO_2 avoidance using amine-based CO_2 capture technology is found to be sensitive to assumptions about the reference plant design and operation, as well as assumptions about the CO_2 capture system design. The

case studies also reveal multi-pollutant interactions and potential tradeoffs in the capture of CO_2, SO_2, NO_2 and NH_3.

The potential for targeted R&D to reduce the cost of CO_2 capture also is explored using the IECM-cs in conjunction with expert elicitations regarding potential improvements in key performance and cost parameters of amine-based systems. The results indicate that the performance of amine-based CO_2 capture systems can be improved significantly, and the cost of CO_2 capture reduced substantially over the next decade or two, via innovations such as new or improved sorbents with lower regeneration heat requirements, and improvements in power plant heat integration to reduce the (currently large) energy penalty of CO_2 capture. Future work will explore in more detail a broader set of advanced technology options to lower the costs of CO_2 capture and storage.

Volume 2 of this report presents a detailed User's Manual for the IECM-cs computer model as a companion to the technical documentation in Volume 1.

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1. INTRODUCTION

Today fossil fuels provide a major share of the total electricity generated in the United States The 300 GW of coal-fired power generation capacity in the U.S. provides 51% of all power generation and accounts for 79% of carbon emissions coming from electric utilities (EIA 2000). Even with the expected growth in natural gas for new generating capacity, coal's share of the electricity supply is still projected to be about 44% in 2020, and higher in absolute amount compared to today, according to recent DOE/EIA projections. Natural gas use is projected to account for 31% of power generation in 2020. At the same time, there is significant concern about the potential climate change impacts of increased CO_2 emissions to the atmosphere, to which fossil fuel power plants are the largest contributor. Thus, any new policies to significantly reduce CO_2 emissions during the next few decades must consider not only the technology options for new power plants, but also the retrofitting or repowering of existing coal and natural gas plants which will continue to operate for several decades to come. In all cases, improved modeling tools are needed to reliably estimate the performance and cost of alternative options, including the potential use of CO_2 capture and storage (CCS) as a CO_2 reduction strategy.

1.1 Technology Options for CO_2 Capture

A wide range of technologies currently exist for separation and capture of CO_2 from gas streams, although they have not been designed for power plant scale operations (Desideri and Corbelli 1998). Figure 1.1 and Table 1.1 briefly summarize the salient features of these technology options (Riemer, Audus et al. 1993; Hendriks 1994; Mimura, Satsumi et al. 1999; Audus 2000; Jeremy 2000; White, Strazisar et al. 2003). They are based on different physical and chemical processes including absorption, adsorption, membranes and cryogenics. Apart from these four types of processes that generate a concentrated stream of CO_2 product (which needs to be sequestered later), there is another set of biological systems (*e.g.* forestry, microbial/ algal processes etc.) that capture CO_2 and bio-chemically transform it into some other compounds (and hence also get sequestered simultaneously).

The choice of a suitable technology depends upon the characteristics of the gas stream from which CO_2 needs to be separated, which mainly depends on the power plant technology. At present, the majority of the plants in the U.S. are conventional combustion-based plants that generate a flue gas that is a dilute stream of CO_2 due to large amount of nitrogen in the combustion air. Future power plants may be designed so as to separate out CO_2 from coal before combustion (*e.g.* coal gasification systems), or they may employ pure oxygen combustion instead of air so as to obtain a concentrated CO_2 stream for treatment. **Error! Reference source not found.** shows the variety of power plant fuels and technologies that affect the choice of CO_2 capture systems.

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          ┌─────────────────────────────────┐
          │ Technology options to capture CO2│
          │   as a concentrated stream from  │
          │   fossil-fuel-based power plants │
          └─────────────────────────────────┘
                    │
        ┌───────────┴────────────┐
┌────────────────────┐  ┌─────────────────────────┐
│ Pre- or post-       │  │ Air separation          │
│ combustion CO2      │  │ followed by oxy-fuel    │
│ separation and      │  │ combustion to obtain    │
│ capture from        │  │ concentrated CO2 stream │
│ gaseous mixtures    │  └─────────────────────────┘
│ like syngas or      │
│ flue gas            │
└────────────────────┘
      │
 ┌────┬────┬────┬────┐
Absorption Adsorption Membranes Cryogenics
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Figure with boxes:

Technology options to capture CO_2 as a concentrated stream from fossil-fuel-based power plants

— Pre- or post-combustion CO_2 separation and capture from gaseous mixtures like syngas or flue gas
— Air separation followed by oxy-fuel combustion to obtain concentrated CO_2 stream

Absorption
— Chemical
— Physical

Adsorption
— Pressure swing
— Temperature swing

Membranes
— Gas separation
— Gas absorption

Cryogenics

Figure 1.1. Technology options for CO₂ separation and capture

1.2 Options for CO₂ Sequestration

Once the CO_2 is captured, it needs to be securely stored (sequestered). Again, there is a range of options potentially available, as shown in **Error! Reference source not found.**. Geologic formations including deep saline reservoirs, depleted oil and gas wells, and unmineable coal seams are some of the potentially attractive disposal sites (Adams, Ormerod et al. 1993; Byrer and Guthrie 1998; Lindeberg and Holloway 1998). Some of these options, such as enhanced oil recovery (EOR) and enhanced coal bed methane

(ECBM), can be revenue generating as well, as they lead to production of valuable products like oil and methane,.

Table 1.1. Comparison of technology options for CO_2 separation and capture

Technology Option	System Requirements	Advantages	Problems/ Drawbacks
Absorption (Chemical)	Absorber and stripper sections Chemical sorbent (*e.g.* MEA, HPC)	Suitable even for dilute CO_2 streams (typical flue gas from power plants) Operates at ordinary temperature & pressure Commercially available, proven technology	The heat of sorbent regeneration is very high Significant sorbent losses; pre-processing (*e.g.* S removal) may be required
Absorption (Physical)	Absorber and stripper sections Physical sorbent (*e.g.* Selexol)	Suitable only for gas streams with high partial pressure of CO_2 (typical syngas from gasification systems) Less energy required, assuming that the gas stream to be processed is already at high pressure Sorbents are less susceptible to the impurities in the gas stream	Requires high operating pressure Works better only with gas streams having high CO_2 content; so it is not suitable for flue gas processing
Adsorption	Adsorber bed(s)	Commercially available gas separation process	Low capacity and CO_2 selectivity of available adsorbents
Membranes	Membrane filter(s)	Upcoming, promising technology with diverse applications Space efficient	Requires high operating pressures Lower product purity; need for multiple stages/ recycle
Cryogenics	Refrigeration and distillation units	Direct production of liquid CO_2	Requires very large amount of energy for refrigeration (not suitable for dilute streams)
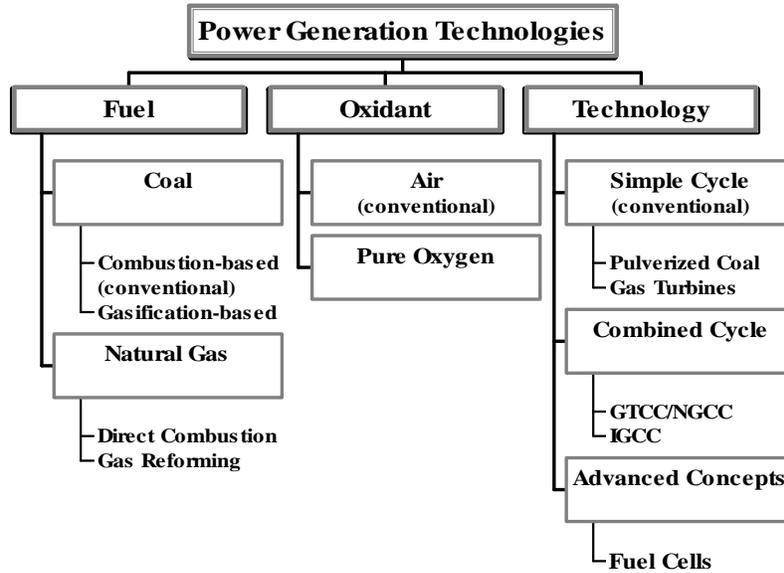
Figure 1.2. Technology options for fossil-fuel based power generation
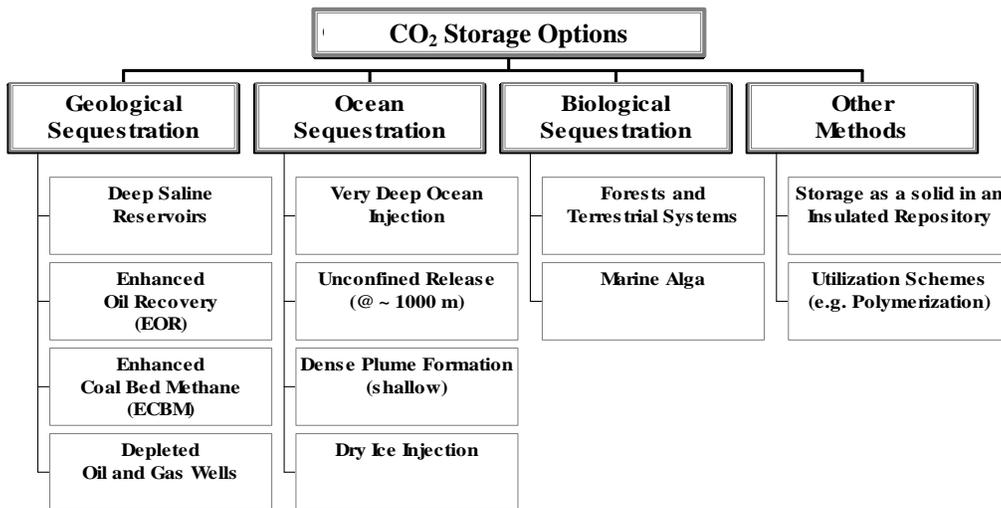
Figure 1.3. Options for CO₂ sequestration

respectively. Ocean storage is another option being studied (Herzog 1998; Spencer 1998)

The distance to a secure storage site, the availability and cost of transportation infrastructure, and the regulatory framework all affect the choice of disposal option. In general, studies indicate that geologic formations are the most plentiful and attractive

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option for U.S. power plants (Bergman and Winter 1995). While the economic costs of CO_2 storage appear to be low, its social and political acceptability are not yet clear, especially with regard to ocean sequestration (CO2-Experiment 2001; USDOE/NETL 2001). Issues related to geologic sequestration of CO_2, including. potential sites, storage capacities, environmental impacts, monitoring requirements and leakage are well summarized elsewhere (White, Strazisar et al. 2003).

1.3 Scope of this Project

Efforts by DOE/NETL and others to develop new or improved technologies for CO_2 capture and storage (CCS) can be assisted substantially by improved analytical tools to assess the overall effectiveness, costs, and carbon sequestration potential of alternative CCS approaches. Especially important for R&D planning and management is an ability to systematically compare advanced systems and concepts to current commercial technologies that also are evolving. To help provide this capability, this project described in this report builds upon the Integrated Environmental Control Model (IECM) developed under a prior DOE/NETL research contract (DE-AC22-92PC91346). The IECM was designed to provide preliminary performance, emissions and cost estimates for alternative environmental control options for coal-fired power plants, with a focus on SO_2, NO_x, particulates and (most recently) mercury emissions. The model simulates a single user-specified facility, and is built in a modular fashion that allows models of individual technologies to be incorporated into the overall framework. A user can then select different options to configure and evaluate a particular plant design involving current commercial technology or advanced emission control systems.

The current project extends the IECM capabilities to include technology options for carbon management. This version of the model is referred to as the IECM-cs. The CCS options include a set of "baseline" technologies representing currently available fossil-fuel power plants, plus a selected set of advanced process designs that promise improved performance and/or lower cost. In addition to coal combustion plants, the IECM-cs framework has been expanded to include natural gas combined cycle (NGCC) plants plus integrated coal gasification combined cycle (IGCC) systems. Pipeline transport and alternative CO_2 storage options also have been added. The IECM-cs thus provides a

comprehensive yet easy-to-use modeling tool for evaluating technological options in a total systems context.

The focus of this report is on the modeling of CCS systems for a pulverized coal-fired (PC) power plant, which is the most widely used technology for electric power generation. Models of NGCC and IGCC plants with CO_2 capture systems are undergoing further refinement as part of a continuing project which will issue reports on these technologies later in 2004. As detailed in the section below, the scope of the present report covers the development and illustrative applications of the baseline coal-fired power plant model employing an amine CO_2 capture system together with pipeline transport to a geological storage site. An advanced post-combustion CO_2 capture system design also is developed and evaluated to assess the potential benefits of R&D. Further details on the design and use of the IECM-cs software package can be found in Volume 2 of this report, which provides a complete User's Manual for the coal-fired plant model.

Some of the key research questions that are addressed in this report include:

- What technologies and design configurations are most suitable for capture and storage of CO_2 from coal or gas-fired power plants?
- What are the key parameters that affect the performance, cost and environmental acceptability of different options?
- How do alternative options compare in terms of these considerations?
- What are the technical and economic uncertainties associated with different options?
- What are the potential benefits of continued R&D on CCS technology?

1.4 Organization of this Report

The organization of this report is as follows. Chapter 2 gives a brief overview of amine-based CO_2 capture systems, including historical developments and process description. Chapter 3 documents the performance model developed in this research, including the model configurations, methodology, parameters and performance equations. Chapter 4 next describes the associated cost model, which is directly linked to the performance model. Then, Chapter 5 illustrates the applications of the new performance and cost models. Several case studies applying current amine-based CO_2 capture technology to coal and natural gas power plants (new and retrofits) are presented. Following this,

together with an expert elicitation study, Chapter 6, uses the model to explore potential future cost reductions and R&D management. Finally, Chapter 7 presents some concluding remarks and directions to future work. Additional details concerning amine systems, characterization of uncertainties and variability, probability distributions for model parameters, case study results, and expert elicitation study are included as Appendices A-E.

References (Chapter 1)

Adams, D., W. Ormerod, et al. (1993). Carbon Dioxide Disposal from Power Stations. UK, a report published by IEA Greenhouse Gas R&D Programme.

Audus, H. (2000). Leading options for the capture of CO2 at power stations. Presented at the Fifth International Conference on Greenhouse Gas Control Technologies, 13-16 August, Cairns, Australia.

Bergman, P. D. and E. M. Winter (1995). "Disposal of carbon dioxide in aquifers in the U.S." Energy Conversion and Management **36(6-9)**: 523-526.

Byrer, C. W. and H. D. Guthrie (1998). Coal deposits: potential geological sink for sequestering carbon dioxide emissions from power plants. in the proceedings of the AWMA's Second International Specialty Conference, Oct 13-15, 1998, Washington, DC.

CO2-Experiment (2001). Ocean sequestration of CO2: Field experiment (www.co2experiment.org).

Desideri, U. and R. Corbelli (1998). "CO2 capture in small size cogeneration plants: Technical and economical considerations." Energy Conversion and Management **39(9)**: 857-867.

EIA (2000). Annual Energy Outlook 2001 (with Projections to 2020), A report published by Energy Information Administration, U.S. Department of Energy, Washington, DC.

Hendriks, C. (1994). Carbon Dioxide Removal from Coal-fired Power Plants. The Netherlands, Kluwer Academic Publishers.

Herzog, H. J. (1998). Ocean sequestration of CO2 - an overview. in the proceedings of the AWMA's Second International Specialty Conference, Oct 13-15, 1998, Washington, DC.

Jeremy, D. (2000). Economic Evaluation of Leading Technology Options for Sequestration of Carbon Dioxide, M.S. Thesis, Massachusetts Institute of Technology, Cambridge, MA .

Lindeberg, E. and S. Holloway (1998). The next steps in geo-storage of carbon dioxide. Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September, Interlaken, Switzerland, Elsevier Science Ltd.

Mimura, T., S. Satsumi, et al. (1999). <u>Developments on energy saving technology for flue gas carbon dioxide recovery by the chemical absorption method and steam system in power plant</u>. Proceedings of the 4th International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September 1998, Interlaken, Switzerland, Elsevier Science Ltd.

Riemer, P., H. Audus, et al. (1993). Carbon Dioxide Capture from Power Stations. UK, a report published by IEA Greenhouse Gas R&D Programme.

Spencer, D. F. (1998). <u>Integration of an advanced CO2 separation process with methods for disposing of CO2 in oceans and terrestrial deep aquifers</u>. Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September, Interlaken, Switzerland, Elsevier Science Ltd.

USDOE/NETL (2001). Environmental Assessment: Ocean sequestration of CO_2 field experiment (report # DOE/EA-1336), U.S. Department of Energy, National Energy Technology Laboratory (NETL), Pittsburgh, PA 15236.

White, C. M., B. R. Strazisar, et al. (2003). "Separation and capture of CO2 from large stationary sources and sequestration in geological formations - coalbeds and deep saline aquifers." <u>Journal of the Air & Waste Management Association</u> **53**: 645-715.

2. AMINE-BASED CO2 CAPTURE SYSTEMS: AN OVERVIEW

In conventional coal plants, which use air for combustion, CO_2 can be separated from the post-combustion flue gas stream. Past studies have shown that amine-based CO_2 absorption systems are the most suitable for combustion-based power plants for the following reasons:

- These systems are effective for dilute CO_2 streams, such as coal combustion flue gas which typically contains about 10%-15 % CO_2 by volume.

- Amine-based CO_2 capture system is a proven technology that is commercially available today.

- Amine-based systems are similar to other post-combustion environmental control units used at power plants. These units are operated at ordinary temperature and pressure.

- A major effort is being made worldwide to improve this process in the light of its potential role in CO_2 abatement. Thus, one can anticipate future benefits from technology advances.

2.1 Historical Developments

Although the amine-based CO_2 capture technology has gained increasing attention in the recent times, this technology has been around for several decades now. Amine-based absorption/stripping has been a commercial technology for CO_2 removal from natural gas and hydrogen (Yeh 2003). The basic amine process was first patented by R. R. Bottom in 1930 as absorbents for acidic gas treating (Bottoms 1930). Alkanolamines (referred to as "amines" in this document) are a family of organic compounds that are derivatives of alkanols (compounds with functional group "OH" group, commonly called as alcoholic group), that also have an "amino" (NH_2) group attached to one of the carbon atoms. The amine-based gas treating basic process flow scheme has remained relatively unchanged over the years. Members of alkanolamine family were subsequently introduced into gas purification market and still remain the technology of choice for the removal of H_2S and CO_2 for gas purification (Kohl and Blohm 1950; Kohl and Nielsen 1997).

The idea of separating CO_2 from flue gas streams started back in 1970's, not with concern about the greenhouse effect, but as a possible economic source of CO_2, mainly for enhanced oil recovery (EOR) operations. Flue gas, a gaseous product stream resulting from the combustion of fossil fuels in air, mainly contains nitrogen, carbon dioxide, and water vapor along with small quantities of many other gases. Depending upon the carbon content of the fuel and the quantity of air used for combustion of the fuel, the flue gas stream may contain as high as 15% CO_2 and is a potential source of CO_2 which is available at no cost. Even today, about 80% of CO_2 production is used for EOR (Chapel et al., 1999), most of which is obtained from natural CO_2 domes. CO_2 is also produced for other industrial applications such as carbonation of brine and production of products like dry ice, urea and beverages.

Several commercial CO_2 capture plants were constructed in the US in the late 1970's and early 1980's (Kaplan 1982; Pauley, Simiskey et al. 1984). Some of these plants are still in operation today. But all these plants are much smaller than a typical power plant in terms of tonnage of CO_2 handled. Figure 2.1 gives a rough idea about the various industrial applications of CO_2 capture technologies and their relative magnitude of operations.

Unless there is an application that can make use of the captured CO_2, it has to be securely stored (sequestered) to prevent it from entering the atmosphere. However, there is little experience with CO_2 sequestration technologies other than EOR. The first commercial CO_2 sequestration facility started in Norway in September 1996 in response to a Norwegian carbon tax. Since then, Statoil has been storing about 1 million tonnes of CO_2 per year from the Sleipner West gas field into a sandstone aquifer 1000 m beneath the North Sea (DOE 1999; Statoil 2001). The international research community is closely monitoring this facility to enhance the understanding of this geological experiment.

All these plants (Figure 2.1) capture CO_2 with processes based on chemical absorption using a monoethanolamine (MEA) based sorbent. MEA is a primary amine with high pH

	EOR	Fertilizer	Soda Ash	Food-grade	Sleipner Gasfield	500MW gross Coal Plant
status	shut down	operational	operational	operational	operational	prospective
year	'80s	'90s	1978	'90's	1996	2010?
tonne CO2/day	1200	150	800	320	3000	9600
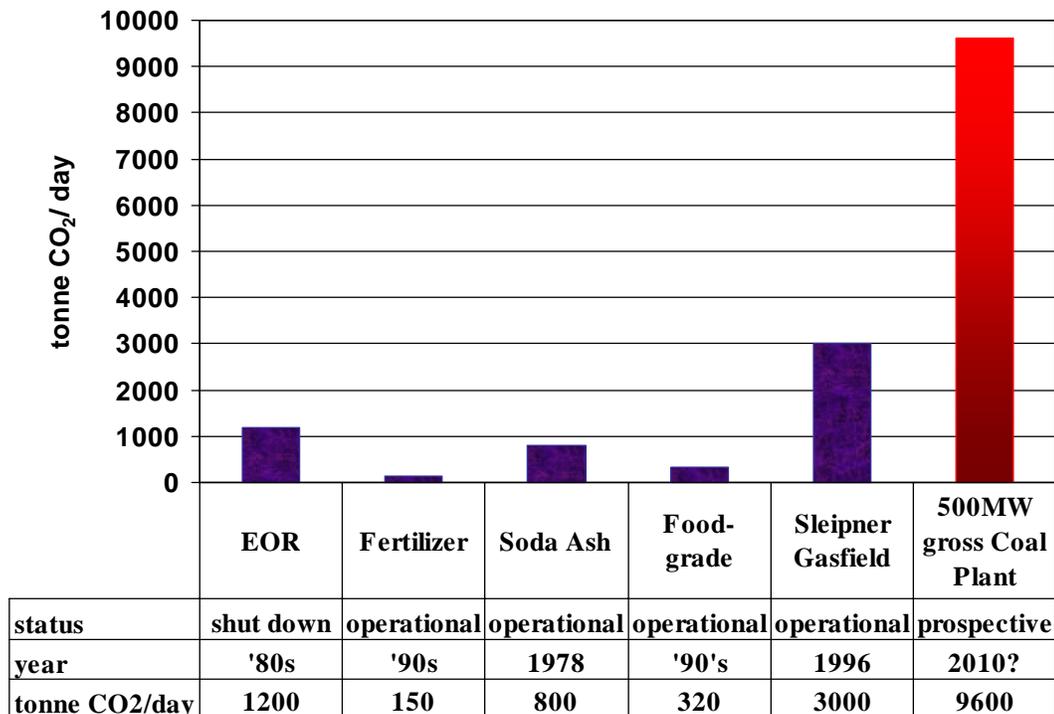
Figure 2.1. Major industrial applications of amine-based CO_2 capture systems

and has the lowest molecular weight of the available amines. It is completely soluble in water and is readily biodegradable (Dow 2000). It is a fairly reactive compound and is classified as a corrosive material. It was developed over 60 years ago as a general, non-selective sorbent to remove acidic gas impurities (e.g. H_2S, CO_2) from natural gas streams (Herzog 1998). The process was then adapted to treat flue gas streams for CO_2 capture. Dow Chemical Co. (and later Fluor Daniel Inc.), Kerr-McGee Chemical Corp. and ABB Lummus Crest Inc., are some of the key players in the development of MEA-based technology for CO_2 capture. About 75%-96% CO_2 may be captured using this technology to yield a fairly pure (>99%) CO_2 product stream.

Today there are two main MEA-based processes available for commercial CO_2 recovery plants: the Fluor Daniel Econamine FG process and the ABB Lummus Crest MEA process (Wong, Gunter et al. 2000). The brief commercial history of these processes is as follows.

2.1.1. Fluor Daniel's Econamine FG process

Dow Chemical and Union Carbide developed inhibited 30 wt.% MEA processes. These absorption processes use a sorbent containing MEA diluted in water to 30 % w/w (by weight) *i.e.,* each 100 kg of total sorbent contains 30 kg of MEA, for the recovery of CO_2 from flue gases in the 1970's and 1980's primarily for the EOR market. Dow Chemical's process was known as the GAS/SPEC FT-1^{TM} process till 1989. Several plants were built in West Texas to recover CO_2 from boiler flue gas for EOR purposes between 1982 and 1986. These plants were technical and commercial successes and were shut down only because of the crude oil price collapse in 1986. As the price of crude oil dropped dramatically in 1986, the EOR market disappeared and these licensors (Dow Chemical and Union Carbide) became less active. Dow Chemical sold its GAS/SPEC FT-1™ process to Fluor Daniel, Inc. in 1989. Fluor Daniel renamed the technology as Econamine FG℠. Sixteen commercial plants have been built, including nine large plants (60 tonne/d and larger) and seven skid-mounted plants (Chapel, Ernst et al. 1999). Seven of the large plants are still operating. Fluor Daniel licensed three of these commercial plants: a 320 tonne/d CO_2 plant in Bellingham, Massachusetts for Northeast Energy Associates, a 150 tonne/d plant for Sumitomo Chemical in Chiba, Japan, and a 90 tonne/d plant for Prosint Produtos Sintéticos in Rio de Janeiro, Brazil. Fluor Daniel also constructed the Bellingham plant. Prior to this, Dow had licensed six commercial plants ranging from 6 to 1000 tonne/d. The seven facilities employing skid-mounted Econamine FG process units have capacities ranging from 6 to 40 tonne/d and were manufactured by the Wittemann Company of Palm Coast, Florida.

It may be noted that all of the large commercial plants use flue gas resulting from the combustion of natural gas, except the Sumitomo plant that fires a variety of fuels including heavy fuel oil. The Econamine FG process has also been demonstrated with coal-derived flue gas in three pilot plants: a 4.5 tonne/d plant in Yokosuka, Japan for Tokyo Electric Power Co, a 2 tonne/d unit at the Sundance Generating Plant in Alberta, and a 4 tonne/d unit at the Boundary Dam Power Plant in Saskatchewan (IEA-GHGRDP 2002). Fluor Daniel has also licensed a natural gas-fired 2 tonne/d pilot plant in Osaka, Japan for KEPCO. Commercial operations at the Warrior Run Power Plant in Maryland

started in February 2000. The Econamine FG scrubber system is used to capture CO_2 from a portion of the flue gases from this 180 MWe AES cogeneration plant that burns only Western Maryland coal with a clean coal technology using a circulating fluidized bed boiler. The extracted CO_2 from this commercial-scale production facility is used for food processing and related processes. Recently, Fluor has announced an improved version of this process named as "Fluor Econamine FG Plus" that addresses some of the problems in the Econamine FG process (Reddy and Roberts 2003), although no commercial plants are yet in existence.

The performance and cost model developed in this thesis for current commercial systems is based primarily on the Econamine FG process, which uses 30% w/w MEA solution with an oxygen inhibitor. The inhibitor helps in two ways – reduced sorbent degradation and reduced equipment corrosion (Chapel, Ernst et al. 1999). It may be noted that this process is *not* applicable to reducing gas streams that contain large amounts of CO and H_2, or contain more than 1 ppm of H_2S, or contain less than 1% O_2 v/v.

2.1.2. The Kerr-McGee/ ABB Lummus amine process

Kerr-McGee started up their 800 tonne/d CO_2 recovery unit in 1978. The Trona unit has been fed flue gases from boilers fired with natural gas, coal and coke. During the first several years of operation, the installation was improved in terms of reliability and cost effectiveness. In late 1990, Kerr-McGee and Lummus Global concluded a joint licensing agreement whereby Lummus gained worldwide exclusive marketing rights to Kerr-McGee's CO_2 recovery technology and became responsible for marketing and basic engineering. Kerr-McGee maintains the continuing role in technology transfer, process improvement, quality control of new designs, operator training, and licensing(ALSTOM, AEP et al. 2001).

Two units have been licensed using this technology. They are:

1)   Applied Energy System, Poteau, Oklahoma. This 300 MW coal-fired co-generation plant incorporates a 200 TPD food-grade liquid CO_2 unit as the steam host. Startup was completed in January 1991. Lummus Crest was responsible for engineering, procurement and construction of the entire facility, including the

power plant.   Lummus Crest made several design improvements to the AES facility.

2)   Soda Ash Botswana, Pty. Ltd., Sue pan, Botswana. This soda ash facility, incorporating a 300 TPD CO_2 unit, started up in March 1991.

The Kerr-McGee/ ABB Lummus amine process uses a 15%-20% w/w MEA solution without any inhibitor (Marion, Nsakala et al. 2001). This technology can capture more than 96% of the CO_2 from flue gases, but the lower sorbent concentration leads to economic disadvantages in terms of greater capital requirements due to larger equipment size and higher energy requirements due to higher amount of dilution water per unit of sorbent. At the same time, it may be noted that this process has already been proven for coal-fired flue gas applications at commercial scale.

2.2 Process Description

Amine-based CO_2 capture process is based on the principle of chemical absorption. Chemical absorption systems tend to be more efficient than the other systems shown in Figure 1.1, as the process is accompanied by a chemical reaction that enhances the overall mass transfer from gas phase to liquid phase. Here, a continuous scrubbing system is used to separate CO_2 from a gaseous stream. The system consists of two main elements, an absorber, where CO_2 is absorbed into a sorbent and a regenerator (or stripper), where CO_2 is released (in concentrated form) and the original sorbent is recovered.

In a power plant application (Figure 2.2) cooled flue gases flow vertically upwards through the absorber countercurrent to the sorbent (MEA in a water solution, with some additives). The MEA reacts chemically with the CO_2 in the flue gases to form a weakly bonded compound (carbamate). The scrubbed gases are then washed and vented to the atmosphere. The CO_2-rich sorbent stream leaves the absorber and passes through a heat exchanger, then further heated in a reboiler using low-pressure steam. The weakly bonded compound formed during absorption is broken down by the application of heat, regenerating the sorbent, and producing a concentrated CO_2 stream. The hot CO_2-lean sorbent is then returned to the heat exchanger, where it is cooled, and then sent back to

the absorber. Some fresh MEA is added to make up for the losses incurred in the process (see section 2.5.2 for details).
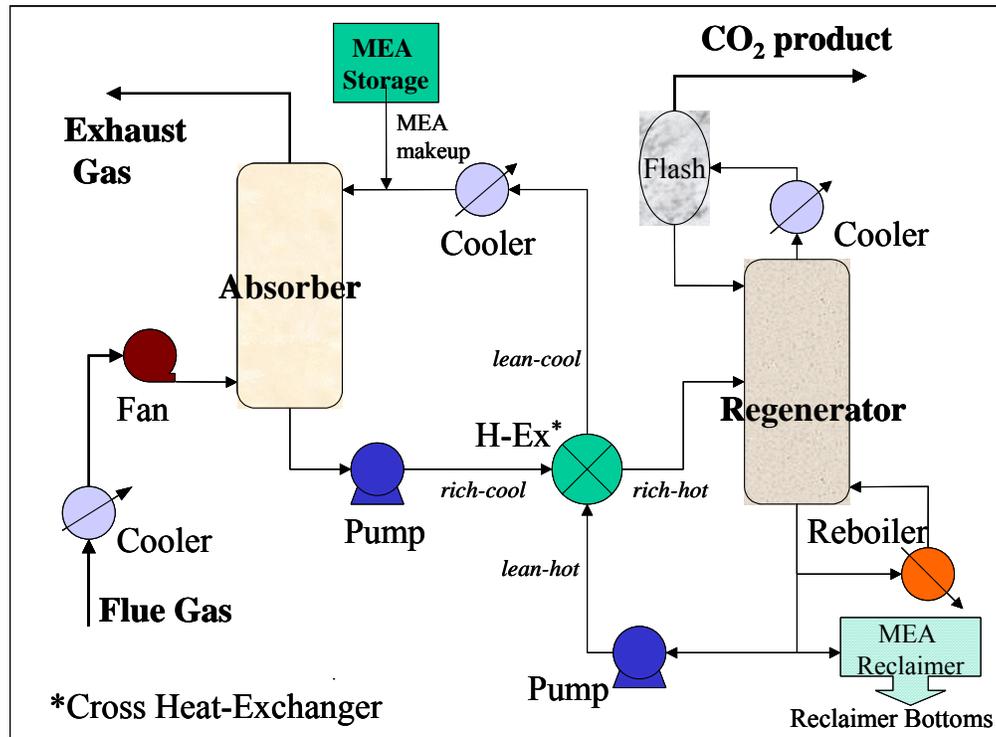
Figure 2.2. Flowsheet for CO₂ capture from flue gases using amine-based system

The CO_2 product is separated from the sorbent in a flash separator, and then taken to the drying and compression unit. It is compressed to very high pressures (about 2000 psig) so that it is liquefied and easily transported over long distances to the designated storage or disposal facility.

2.3 Process Chemistry

There are mainly three compounds, *viz.* carbon dioxide (CO_2), amine (MEA) and water (H_2O) that are active in this system. The following equilibrium reactions occur in the bulk of the liquid (Al-Baghil, Pruess et al. 2001):

Water hydrolysis: $2H_2O \leftrightarrow H_3O^+ + OH^-$

Bicarbonate formation: $2H_2O + CO_2 \leftrightarrow H_3O^+ + HCO_3^-$

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Carbonate formation: $H_2O + HCO_3^- \leftrightarrow H_3O^+ + CO_3^{-2}$

Amine protonation: $H_3O^+ + R\text{-}NH_2 \leftrightarrow H_2O + R\text{-}NH_3^+$

Carbamate formation: $R\text{-}NH_2 + HCO_3^- \leftrightarrow H_2O + R\text{-}NH\text{-}COO^-$
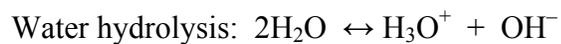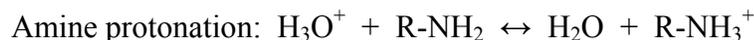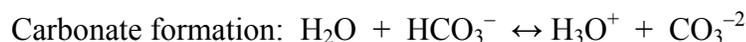
Here, MEA has been represented as $R\text{-}NH_2$, where "R" stands for $HO\text{-}CH_2CH_2$.

The process chemistry is complex, but the main reactions taking place are (Desideri and Paolucci 1999):

CO_2 Absorption: $2\,R\text{-}NH_2 + CO_2 \rightarrow R\text{-}NH_3^+ + R\text{-}NH\text{-}COO^-$

MEA Regeneration: $R\text{-}NH\text{-}COO^- + R\text{-}NH_3^+ + (\text{Heat}) \rightarrow CO_2 + 2\,R\text{-}NH_2$
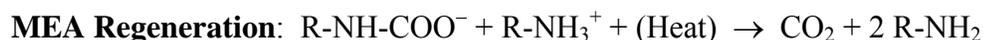
Pure MEA (with $R = HO\text{-}CH_2CH_2$) is an "unhindered" amine that forms a weakly bonded intermediate called "carbamate" that is fairly stable. Only half a mole of CO_2 is absorbed per mole of amine, as shown in the CO_2 absorption equation above. On application of heat, this carbamate dissociates to give back CO_2 and amine sorbent, as shown in the second equation above. Since the carbamate formed during absorption is quite stable, it takes a large amount of heat energy to break the bonds and to regenerate the sorbent. When MEA is used as the sorbent, the theoretical minimum heat requirement to reverse this reaction is about 1900 kJ/ $kgCO_2$.

MEA is a reactive compound. The reversible reactions allow us to make use of the same sorbent mass repeatedly, by capturing CO_2 in the absorber column and then regenerating the sorbent in the stripper or the regenerator column. In spite of dilution with water and use of inhibitors, a small quantity of MEA is lost through various unwanted reactions, mainly the polymerization reaction (to form long-chained compounds) and the oxidation reaction forming organic acids and liberating ammonia. Appropriate measures must be taken in order to avoid accumulation of these unwanted chemical species in the circulating sorbent.

Acid gas impurities that might be present in the flue gas are another potential source of sorbent loss. This is especially true for the flue gas from coal-fired combustion. Acid

gases like SO_2, HCl, and NO_2 react with MEA to form heat-stable salts that reduce the CO_2 absorption capacity of the sorbent. Thus, very low concentrations of these gases (on the order of 10 ppm) are desirable to avoid excessive loss of expensive sorbent. The problem is especially acute for SO_2 because its concentration in flue gases is typically 700 to 2500 ppm at coal-fired plants. NO_x is less of a problem because most of the NO_x is nitric oxide (NO), whereas only NO_2 (typically about 5% of total NO_x) is reactive with MEA. Hence, from a multi-pollutant perspective, there are important interactions between the CO_2 capture system and the control of other air pollutants, especially SO_2 and NO_x emissions.

For other "hindered" amines (e.g., where R is a bulky group), the carbamate formed is not stable, and an alternate reaction leads to formation of bicarbonate ions and hence a higher theoretical capacity of one mole of CO_2 per mole of amine, as shown in the CO_2 absorption equation below (Sartori, Ho et al. 1994; Herzog, Drake et al. 1997).

$$\textbf{CO}_2 \textbf{ Absorption: } R\text{-}NH_2 + CO_2 + H_2O \rightarrow R\text{-}NH_3^+ + HCO_3^-$$

$$\textbf{MEA Regeneration: } HCO_3^- + R\text{-}NH_3^+ + (\text{less Heat}) \rightarrow CO_2 + H_2O + R\text{-}NH_2$$
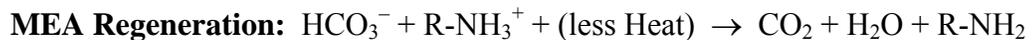
The regeneration of these amines requires lesser amount of heat energy as compared to the unhindered amines. But the CO_2 uptake rate of hindered amines is very low. Efforts are underway to formulate better sorbents by combining favorable properties of these two groups of amines.

2.4 Process Areas

The CO_2 capture and separation system consists of the following process equipment:

Direct contact cooler: The flue gases coming out of a power plant are quite hot. The temperature of flue gas may be ranging from as low as 60 deg. C (in case of coal-fired power plants with wet SO_2 scrubbers) to more than 550 deg. C (in case of natural gas fired simple cycle power plants). It is desirable to cool down the flue gases to about 45-50 deg. C, in order to improve absorption of CO_2 into the amine sorbent, to minimize sorbent losses, and to avoid excessive loss of moisture with the exhaust gases. Absorption being an exothermic process is favored by low temperatures, and higher

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temperature may also lead to sorbent losses due to evaporation and degradation. In case of gas-fired power plants or majority of coal-fired power plants without wet scrubbers for SO_2 removal, a direct contact cooler needs to be installed to reduce the temperature of the flue gas stream to acceptable levels. In case of coal-fired power plant applications that have a wet FGD (flue gas desulfurization) unit upstream of the amine system, the wet scrubber helps in substantial cooling of the flue gases, and additional cooler may not be required.

Flue gas blower: The flue gas enters at the bottom of the absorber column and flows upwards, countercurrent to the sorbent flow. Thus, it needs to overcome a substantial pressure drop as it passes through a very tall absorber column. Hence the cooled flue gas must be pressurized using a blower before it enters the absorber.

Absorber: This is the vessel where the flue gas is made to contact with the MEA-based sorbent, and where the CO_2 from the flue gas is dissolved into the sorbent. The column may be plate-type or packed. Most of the CO_2 absorbers are packed columns using some kind of polymer-based packing to provide large interfacial area.

Rich/lean cross heat exchanger: The CO_2-loaded sorbent must be heated in order to strip off CO_2 and regenerate the sorbent. On the other hand, the regenerated (lean) sorbent coming out of the regenerator must be cooled down before it can be circulated back to the absorber column. Hence, these two sorbent streams are passed through a cross heat exchanger, where the rich (CO_2-loaded) sorbent gets heated and the lean (regenerated) sorbent gets cooled.

Regenerator: This is the column where the weak intermediate compound (carbamate) formed between the MEA-based sorbent and dissolved CO_2 is broken down by the application of heat. The CO_2 is separated from the sorbent to leave reusable sorbent behind. In case of unhindered amines like MEA, the carbamate formed is stable and requires large amounts of energy to dissociate. It also consists of a flash separator, where CO_2 is separated from most of the moisture and evaporated sorbent to produce a fairly concentrated CO_2 stream.

Reboiler: The regenerator is connected with a reboiler that is basically a heat exchanger where low-pressure steam extracted from the power plant is used to heat the loaded sorbent.

Steam extractor: In case of coal-fired power plants that generate electricity in a steam turbine, a part of the LP (low pressure)/IP (intermediate pressure) steam has to be diverted to the reboiler for sorbent regeneration. Steam extractors are installed to take out steam from the steam turbines.

MEA reclaimer: The presence of acid gas impurities (SO_2, SO_3, NO_2 and HCl) in the flue gas leads to formation of heat stable salts in the sorbent stream, which can not be dissociated even on application of heat. In order to avoid accumulation of these salts in the sorbent stream and to recover some of this lost MEA sorbent, a part of the sorbent stream is periodically distilled in this vessel. Addition of a strong alkali such as caustic helps in release of some of the MEA. The recovered MEA is recycled to the sorbent stream while the bottom sludge (reclaimer waste) is sent for proper disposal.

Sorbent processing area: The regenerated sorbent needs to be further cooled even after passing through the rich/lean cross heat exchanger using a cooler. This is so that the sorbent temperature can be brought back to an acceptable level (about 40 deg C). In order to make up for the sorbent losses, a small quantity of fresh MEA sorbent must be added to the sorbent stream. So, the sorbent processing area primarily consists of a sorbent cooler, MEA storage tank, and a mixer. It also consists of an activated carbon bed filter that adsorbs impurities (degradation products of MEA) from the sorbent stream.

CO_2 drying and compression unit: The CO_2 product may have to be carried over very long distances via pipelines. Hence it is desirable that it does not contain any moisture in order to avoid corrosion in the pipelines. Of course, the CO_2 product specifications may vary depending upon the end use (or storage/ disposal method) and the material of construction of the pipeline. Also, it has to be compressed to very high pressures so that it gets liquefied and can overcome the pressure losses during the pipeline transport. The multi-stage compression unit with inter-stage cooling and drying yields a final CO_2

product at the specified pressure (about 2000 psig) that contains acceptable levels of moisture and other impurities (e.g. N_2).

CO_2 transport facility: The CO_2 captured at the power plant site has to be carried to the appropriate storage/ disposal site. Considering the scale of the operation (thousands of tonnes of CO_2 per day), pipelines are the suitable mode of transportation. There is fair amount of industrial experience and expertise in the field of the construction (and operation) of pipelines for CO_2 transport. Recently, a 325-km pipeline carrying CO_2 from the Great Plains Synfuels Plant in Beulah, North Dakota (owned by Dakota Gasification Company of Bismarck, North Dakota) to the Weyburn oil fields in Saskatchewan, Canada went operational (EnCana 2003). In case of retrofit applications, where construction of new pipelines might be prohibitively expensive (and questionable in terms of public acceptance, especially in densely populated regions), transport via tankers may have to be considered.

CO_2 disposal facility: Once the CO_2 is captured, it needs to be securely stored (sequestered). Again, there are a wide range of options potentially available (see Figure 1.3). Geologic formations such as underground deep saline reservoirs, depleted oil and gas wells, and abandoned coal seams are some of the potentially attractive disposal sites (Adams, Ormerod et al. 1993; Byrer and Guthrie 1998; Lindeberg and Holloway 1998). Ocean disposal and terrestrial sinks are additional options being studied (Herzog 1998; Spencer 1998). The distance to a secure storage site and the availability and cost of transportation infrastructure also affect the choice of disposal option. In general, studies indicate that geologic formations are the most plentiful and attractive option for U.S. power plants (Bergman and Winter 1995). Transport of CO_2 to a storage site is typically assumed to be via pipeline. While the economic costs of CO_2 storage appear to be low, the social and political acceptability of different options are not yet clear.

2.5 Limitations of the MEA Process

Although MEA-based absorption process is the most suitable technology available today for post-combustion capture of CO_2 from power plant flue gases, it has several important drawbacks. The main problems may be listed as follows:

2.5.1. Energy Penalty

The stable carbamate ion requires substantial energy to break the bonds. Because MEA is diluted with lot of water, a large amount of heat is required to regenerate the sorbent. Substantial energy also is needed to compress the captured CO_2 for pipeline transport to a storage site. This heat and electricity requirement reduces the net efficiency of the power plant if it is extracted internally by de-rating the power plant. Alternatively, a much bigger power plant must be built in order to achieve the same "net" power generation capacity as it would have been without CO_2 capture. In case of retrofit applications, an auxiliary gas-fired boiler and steam turbine might be added to maintain the power output.

2.5.2. Loss of Sorbent

Sorbent is lost during the process because of a variety of reasons including degradation, entrainment, vaporization and mechanical losses (Stewart and Lanning 1994; Stewart and Lanning 1994). All the sorbent entering the stripper does not get regenerated. Flue gas impurities, especially oxygen, sulfur oxides and nitrogen dioxide react with MEA to form heat-stable salts, thus reducing the CO_2-absorption capacity of the sorbent. Proprietary inhibitors are available that make the sorbent tolerant to oxygen. Flue gas NO_x is not a major problem since nitric oxide (NO) is the predominant form ($\sim$ 90-95%) of total NO_x in the flue gas, and does not react with inhibited amines (Suda, Fujii et al. 1992; Leci 1996). But, SO_2 does degenerate MEA sorbent, so very low inlet concentrations (10 ppm) are desirable to avoid excessive loss of sorbent. However, untreated flue gases of coal-fired power plants contain about 700 to 2500 ppm SO_2 (plus roughly 10-40 ppm NO_2). The interaction of SO_2 with CO_2 control system is thus particularly important. The heat-stable salts that are formed may be treated in a side stream MEA-reclaimer, which can regenerate some of the MEA. Technologies such as electro-dialysis are also being proposed for this purpose (Yagi, Shibuya et al. 1992).

Most of the previous studies on reactions of MEA had focused on natural gas purification processes and were conducted with pure gases under laboratory-controlled conditions. However, the flue gas generated by a fossil-fuel-fired boiler is a complex mixture of various chemical species. It has been shown that there are chemical degradation reactions that occur under plant conditions which do not occur in laboratory experiments

with pure gases (Strazisar, Anderson et al. 2003). Better understanding of the mechanisms and chemical pathways associated with MEA degradation in actual power plant applications may help reduce or eliminate its negative impacts.

2.5.3. Corrosion

MEA is a reactive compound and the MEA solution in water in the presence of oxygen and carbon dioxide is a highly corrosive system. Irreversible side reactions with CO_2 and other flue gas components lead to the formation of various degradation products. These byproducts are associated with increased corrosion in this system. Hence, corrosion control is very important in amine systems processing oxygen-containing flue gases. In order to reduce corrosion rates, corrosion inhibitors, lower concentrations of MEA, appropriate materials of construction, and mild operating conditions are recommended (Barchas and Davis 1992).

2.5.4. Environmental Emissions

In case of MEA-based CO_2 capture systems, environmental problems may arise from the spent sorbent slurry discharged from the MEA reclaimer (also called "reclaimer bottoms") and the emissions of MEA and ammonia (NH_3) carried by the treated flue gas. Details about the potential environmental impacts of these emissions are discussed in *Appendix A*. The spent sorbent is treated as a hazardous waste. Entrainment of MEA with the treated flue gas is at most a few ppm and is likely to go down with the improvements in the absorber design (wash section at the top). The emission rate of NH_3 from a CO_2 capture plant is substantially higher than that from a power plant without amine-based CO_2 capture system. However, these NH_3 emissions are much smaller than those coming from other sources (animals, farms). At this time, it is not very clear if these emissions are likely to have significant environmental impacts. Hence, these aspects must be carefully looked into before a widespread application of this technology.

References (Chapter 2)

Adams, D., W. Ormerod, et al. (1993). Carbon Dioxide Disposal from Power Stations. UK, a report published by IEA Greenhouse Gas R&D Programme.

Al-Baghil, N. A., S. A. Pruess, et al. (2001). "A rate-based model for the design of gas absorbers for the removal of CO2 and H2S using aqueous solutions of MEA and DEA." Fluid Phase Equilibria **185**(2001): 31-43.

ALSTOM, AEP, et al. (2001). Engineering feasibility and economics of CO2 capture on an existing coal-fired power plant, Final report prepared by ALSTOM Power Inc., ABB Lummus Global Inc., ALSTOM Power Environmental Systems and American Electric Power (AEP). report no. (PPL-01-CT-09) submitted to Ohio Department of Development, Columbus, OH and US Department of Energy/ NETL, Pittsburgh, PA**:** 163.

Barchas, R. and R. Davis (1992). "The Kerr-McGee/ ABB Lummus Crest technology for the recovery of CO2 from stack gases." Energy Conversion and Management **33**(5-8): 333-340.

Bergman, P. D. and E. M. Winter (1995). "Disposal of carbon dioxide in aquifers in the U.S." Energy Conversion and Management **36(6-9)**: 523-526.

Bottoms, R. R. (1930). Process for Separating Acid Gases.

Byrer, C. W. and H. D. Guthrie (1998). Coal deposits: potential geological sink for sequestering carbon dioxide emissions from power plants. in the proceedings of the AWMA's Second International Specialty Conference, Oct 13-15, 1998, Washington, DC.

Chapel, D., J. Ernst, et al. (1999). Recovery of CO2 from flue gases: commercial trends (paper no. 340). presented at the Canadian Society of Chemical Engineers Annual Meeting, 4-6 October, Saskatoon, Saskatchewan, Canada.

Desideri, U. and A. Paolucci (1999). "Performance modelling of a carbon dioxide removal system for power plants." Energy Conversion and Management **40**: 1899-1915.

DOE (1999). Carbon sequestration: research and development, A U.S. Department of Energy Report, Office of Science, Office of Fossil Energy, U.S. Department of Energy, Washington, DC.

Dow (2000). "Specialty Alkanolamines": Product information available at the website of The Dow Chemical Company (www.dow.com/alkanolamines/).

EnCana (2003). Weyburn carbon dioxide miscible flood project, report available online at: http://www.encana.com/operations_and_projects/weyburn.shtml.

Herzog, H. J. (1998). The economics of CO2 capture. Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September, Interlaken, Switzerland, Elsevier Science Ltd.

Herzog, H. J. (1998). Ocean sequestration of CO2 - an overview. in the proceedings of the AWMA's Second International Specialty Conference, Oct 13-15, 1998, Washington, DC.

Herzog, H. J., E. Drake, et al. (1997). CO2 capture, reuse and storage technologies for mitigating global climate change. A White Paper (Final Report), January 1997.

IEA-GHGRDP (2002). International test network for CO2 capture: report no. PH4/11 on 3rd workshop (May 16-17 2002, Apeldoorn, The Netherlands), IEA Greenhouse Gas R&D Programme, UK.

Kaplan, L. J. (1982). "Cost-saving process recovers CO2 from power-plant flue gas." Chemical Engineering **89**(24): 30-31.

Kohl, A. L. and C. L. Blohm (1950). "Technical aspects of glycol-amine gas treating." Petroleum Engineer **22(June)**: C-37.

Kohl, A. L. and R. B. Nielsen (1997). Gas Purification. Houston, TX, Gulf Publishing Company.

Leci, C. L. (1996). "Financial implications on power generation costs resulting from the parasitic effect of CO2 capture using liquid scrubbing technology from power station flue gases." Energy Conversion and Management **37(6-8)**: 915-921.

Lindeberg, E. and S. Holloway (1998). The next steps in geo-storage of carbon dioxide. Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September, Interlaken, Switzerland, Elsevier Science Ltd.

Marion, J., N. Nsakala, et al. (2001). Engineering feasibility of CO2 capture on an existing US coal-fired power plant. in the proceedings of the Twenty-sixth International Conference on Coal Utilization and Fuel Systems, 5-8 March, Clearwater, FL, USA.

Pauley, C. R., P. L. Simiskey, et al. (1984). "N-Ren recovers CO2 from flue gas economically." Oil and Gas Journal **82**(20): 87-92.

Reddy, S. and C. A. Roberts (2003). ECONOMINE FGSM Plus: An Enhanced Amine Based CO$_2$ Capture Process. 2nd Annual Conference on Carbon Sequestration, Alexandria, VA, USA.
Sartori, G., W. A. Ho, et al. (1994). Sterically hindered amines for acid gas absorption. Carbon Dioxide Chemistry: Environmental Issues. J. Paul and C. Pradier. Cambridge, UK, The Royal Society of Chemistry.

Spencer, D. F. (1998). Integration of an advanced CO2 separation process with methods for disposing of CO2 in oceans and terrestrial deep aquifers. Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September, Interlaken, Switzerland, Elsevier Science Ltd.

Statoil (2001). CO2 storage 1000 m down, report available online at: http://www.statoil.com/STATOILCOM/SVG00990.nsf/web/sleipneren?opendocument.

Stewart, E. J. and R. A. Lanning (1994). "Reduce amine plant solvent losses (Part 1)." Hydrocarbon Processing **May**: 67-81.

Stewart, E. J. and R. A. Lanning (1994). "Reduce amine plant solvent losses (Part 2)." Hydrocarbon Processing **June**: 51-54.

Strazisar, B. R., R. R. Anderson, et al. (2003). "Degradation pathways for monoethanolamine in a CO_2 capture facility." Energy & Fuels **17**(4): 1034-1039.

Suda, T., M. Fujii, et al. (1992). "Development of flue gas carbon dioxide recovery technology." Energy Conversion and Management **33**(5-8): 317-324.

Wong, S., W. D. Gunter, et al. (2000). Economics of CO2 sequestration in coal bed methane reservoirs. presented at the 2000 SPE/CERI Gas Technology Symposium, 3-5 April, Calgary, Alberta, Canada.

Yagi, T., H. Shibuya, et al. (1992). "Application of chemical absorption process to CO2 recovery from flue gas generated in power plants." Energy Conversion and Management **33**(5-8): 349-355.

Yeh, S. (2003). The historical developments in amine-based CO2 capture technology - personal communication.

3. PERFORMANCE MODEL DEVELOPMENT

A performance model of a chemical technology essentially consists of the simulation of mass and energy balances around the system. In case of an environmental control technology such as amine-based CO_2 capture, it keeps track of various material flows (e.g. flow of flue gas and its individual components, flow of the sorbent and its degradation products, flow of other chemical reagents) and energy flows (e.g. thermal energy requirement for sorbent regeneration, electrical energy required for mechanical devices such as pumps, fans, compressors).

The rate of removal of CO_2 from flue gas using an amine scrubber depends on the gas-liquid mass transfer process. The chemical reactions that permit diffusion of CO_2 in the liquid film at the gas-liquid interface enhance the overall rate of mass transfer. So, the CO_2 removal efficiency in the absorber is a function of various parameters that affect the gas-liquid equilibrium (*e.g.*, flow rates, temperature, pressure, flue gas composition, MEA concentration, equipment design, *etc*.). Absorption of CO_2 in an alkaline medium may be considered as a first order reaction. Higher CO_2 concentration thus improves the efficiency of the absorption system. Even at low concentration of CO_2, MEA has great affinity for CO_2. The solubility of CO_2 in MEA is much higher as compared to many other conventional sorbents. Similarly, the design of and conditions in the regenerator affect the energy requirement and the overall performance of the system.

A number of previous studies have reported some cost and/or performance data for specific amine-based systems, including hypothetical applications to coal-fired power plants (Smelser, Stock et al. 1991; Hendriks 1994; Kohl and Nielsen 1997; Chakma and Tontiwachwuthikul 1998; Herzog 1998; Simbeck 1998; Chapel, Ernst et al. 1999; Marion, Nsakala et al. 2001). However, there are no generally available process models that can be used or modified for detailed studies of CO_2 removal options. Cost data also are relatively limited and often incomplete.

The performance model for amine-based CO_2 capture system was developed for use in the IECM modeling framework, which simulates the performance and cost of a coal-fired power plant equipped with multiple emission control systems (Rubin, Kalagnanam et al.

1997). The performance model of the CO_2 capture system is a response-surface model with about two dozen parameters; developed from numerous runs of a detailed process simulation model. The performance model estimates all of the key flow rates and energy requirements. The output from the performance model is then used by the cost model to estimate various costs of the system.

In this chapter, the process simulation packages used are described in section 3.1, the overall model development methodology is laid out in section 3.2 and the various performance parameters are discussed in section 3.3. The actual performance equations employed in the model are listed in section 3.4 and the model outputs are explained in section 3.5. Finally, section 3.6 discusses the characterization of uncertainty in the performance parameters.

3.1   Process Simulation Model

Two process simulators, viz. *ASPEN-Plus* and *ProTreat* have been used to derive the performance equations. The CO_2 capture and separation model is based on the *ProTreat* simulations while the CO_2 compression model is based on the *ASPEN-Plus* simulations.

- ***ProTreat*** is a software package for simulating processes for the removal of H_2S, CO_2, and mercaptans from a variety of high and low pressure gas streams by absorption into thermally regenerable aqueous solutions containing one or more amines (ProTreat^{TM} 2002).   The ***ProTreat*** package makes exclusive use of a column model that treats the separation as a mass transfer rate process.

- *ASPEN-Plus* is a widely used process engineering tool for the design and steady-state simulation and optimization of process plants (Aspen-Plus$^{(R)}$ 2002).

3.2   Methodology

A large number of process simulation runs have been conducted to cover a reasonable range of values for the key parameters. The details are presented in the following sections.

3.2.1. ProTreat Simulation Runs for CO_2 capture and separation from flue gas

The CO_2 capture and separation system consists of a flue gas cooler, compressor, absorber, heat exchangers, regenerator, MEA reclaimer, sorbent circulation pumps and other accessories. Based on the literature review, parameters relevant to the overall performance of the CO_2-capture process were identified and were varied within reasonable ranges in the *ProTreat* model runs. Table 3.1 summarizes the parameters that were varied and the ranges for each parameter.

Table 3.1. ProTreat parameter ranges (total number of simulation runs = 1983)

No.	Parameter	Type	Units	Range
1	CO_2 content in flue gas (y_{CO2})	Input	mole %	3.5-13.5
2	Flue gas flow rate (G)	Input	kmole/hr	9000-24000
3	Inlet flue gas temperature (T_{fg})	Input	deg C	40-65
4	MEA concentration (C)	Input	wt %	15-40
5	Sorbent flow rate (L)	Input	kmole/hr	16000-70000
6	L/G (calc.)	Input	-	0.7-5.6
7	Reboiler heat duty (Q)	Input	GJ/hr	95-664
8	Q/L (calc.)	Input	MJ/kmole	2.4-22.5
9	CO_2 capture efficiency	Output	%	40-99
10	CO_2 product flow rate	Output	kmole/hr	333-2840
11	Lean sorbent CO_2 loading	Output	moleCO_2/mole MEA	0.05-0.34
12	Rich sorbent CO_2 loading	Output	moleCO_2/mole MEA	0.27-0.55
13	Absorber diameter	Output	ft	26-42
14	Regenerator diameter	Output	ft	12-42
15	Exhaust flue gas temperature	Output	deg C	40-72

The following set of parameters related to the design/configuration of the CO_2 capture system were held constant. These values were set for those of a typical commercial application based on experts' suggestions.

- Absorber height: 40 ft

- Absorber packing: Rasching rings, metallic, 1-inch packing size

- Inlet flue gas pressure: 3 psi

- Sorbent pumping pressure: 30 psi

- Number of trays in regenerator: 24 (tray spacing = 2 ft, weir height = 3 inches)

3.2.2. ASPEN-Plus Simulation Runs for CO₂ Compression

The concentrated CO_2 product stream obtained from sorbent regeneration is compressed and dried using a multi-stage compressor with inter-stage cooling. The *ASPEN-Plus* module used for this simulation consists of 4 stages of compression with inter-stage cooling that deliver the compressed product at $35^{\circ}C$. The compressor efficiency, CO_2 product pressure and purity were used as the main control variables. These parameters were varied over the following ranges

- Compressor efficiency: 60-100 %

- CO_2 product pressure: 500-2500 psi

- CO_2 stream purity: 99-100 %

3.2.3. Regressions to derive performance equations

A set of algebraic equations were derived from the process simulation runs to provide a performance model for use in the IECM (Rubin, Kalagnanam et al. 1997; IECM 2001). The key performance output variables were regressed against all the input variables to obtain a set of relationships among them. The data collected from the process simulation runs was used to carry out these multivariate regressions using a statistical package called SAS (SAS 1999-2001). Only those variables with significance value greater than 0.9995 were retained in the performance equations. Details are presented later in section 3.5.

3.3. Model Configuration Options

For post-combustion CO_2 capture from flue gas, the amine-based CO_2 capture system requires substantial thermal energy in the form of heat or steam for sorbent regeneration. Depending upon how this energy is supplied, there are three configuration options available in the model. These are shown graphically in Figure 3.1 and described below.

a)   **Reference base plant**

b)   **Capture plant with internal derating**

c)   **Capture plant with an auxiliary boiler and a secondary steam turbine**

d)   **Post-capture CO$_2$ transport and disposal options**
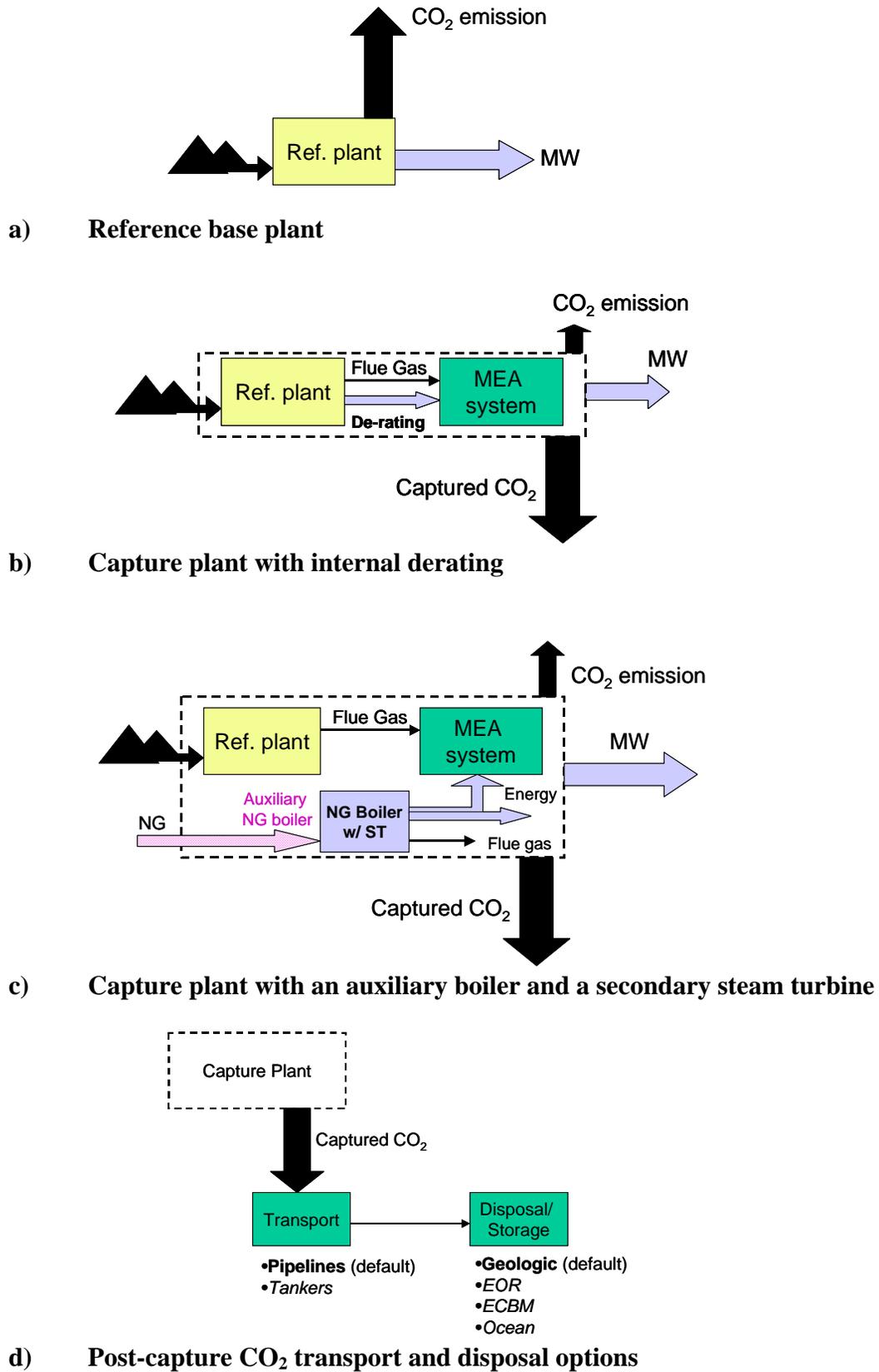
Figure 3.1. CO$_2$ capture plant configuration options

Base plant de-rating: Heat extracted from the base plant can provide energy for the amine system. Here, the low-pressure (LP) steam is extracted from the steam cycle of the power plant and supplied to the reboiler for sorbent regeneration. Extraction of steam leads to loss of power generation capacity, and the net plant output decreases substantially. In case of a new plant to be designed with a CO_2 capture system, it is possible to optimally design the steam cycle to take care of the steam requirement of the amine system, and proper heat integration may help in reducing the energy penalty. In case of an existing coal plant to be retrofitted with an amine system, optimal heat integration may not be achievable, and is likely to lead to a much higher energy penalty due to steam extraction.

Auxiliary Natural Gas Boiler (w/ Steam Turbine):   Another potential option to provide the energy for the amine system is an auxiliary natural gas (NG)-fired boiler. Often it would be combined with a steam turbine to generate some additional power (*e.g.* to supply electrical energy demand of CO_2 capture unit), where the LP steam would be then used for sorbent regeneration. Thus, the original steam cycle of the power plant remains unperturbed and the net power generation capacity of the power plant does not get adversely affected. Again, it comes at an additional cost of capital requirement for the boiler (and turbine) and the cost of supplemental natural gas fuel. Also, the combustion of natural gas leads to additional CO_2 emissions (and NO_x emissions). So, there are at least two possible sub-options available:

1. The flue gas from the auxiliary boiler is cooled down to acceptable exhaust temperature and then directly vented to the atmosphere. Here, the <u>net</u> CO_2 capture efficiency of the system is substantially lowered because of the additional CO_2 emissions from natural gas boiler. Also, the total NO_x emissions may exceed the allowable levels of emission.

2. The CO_2 capture system may be designed so as to capture CO_2 from the auxiliary boiler as well. In this case, the secondary flue gas stream (after cooling and NO_x removal, if required) may be merged with the main flue gas stream before it enters the CO_2 capture system. With higher amount of flue gas to be treated (and

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more CO_2 to be captured), the amine-system would require more steam and thus a still bigger auxiliary NG boiler would be required (which means more secondary flue gas!). The auxiliary NG boiler size may be determined by an iterative calculation procedure, so that it matches the sorbent regeneration steam requirement of the CO_2 capture system treating the total flue gas. Thus, the CO_2 capture level is maintained at the originally desired level, but it requires a large auxiliary NG boiler facility. Our preliminary analysis as well as other studies suggest that this option is quite costly (Simbeck and McDonald 2000), hence in the present version of IECM, this configuration option is not included.

In terms of the configuration of the CO_2 capture system shown in Figure 3.1, the user can make the following choices as well.

Direct contact cooler: The default design of an amine system includes a direct contact cooler (DCC) to cool the flue gas before it enters the amine system. The temperature of the flue gas affects the absorption reaction because absorption of CO_2 in MEA sorbent is an exothermic process favored by lower temperatures. Also, the volumetric flow rate of the flue gas stream, which is a key determinant of the sizes of various equipments (*e.g.*, direct contact cooler, flue gas blower, absorber), is directly related to the flue gas temperature. Hence lower flue gas temperature is desired. The typically acceptable range of flue gas temperature is less than 50-60 ^{o}C. If the flue gas is coming from wet sulfur scrubber, additional DCC may not be required. But in case of flue gas from NG-fired boiler, which often does not pass through a sulfur scrubber, DCC is essential. Exclusion of the DCC is an option available in the model.

Choice of sorbent: At this time, MEA is the only sorbent used in the model. The nominal values of various parameters are based on a process simulation model that uses only MEA. However, the users can overwrite the nominal values of these parameters if they wish to use a different sorbent and have the relevant data. The model could include other sorbents in the future by providing the appropriate values for the key parameters.

CO_2 transport: As shown in Figure 3.1(d), the default mode of CO_2 transport is via pipelines. This is the suggested mode of transporting large quantities of CO_2 that would

be captured in a typical power plant application. The user can specify the distance over which CO_2 needs to be carried to, and the unit cost of CO_2 transportation. This module may be expanded in the future to include more detailed parameters regarding pipeline transport and other transport options.

CO_2 storage/ disposal: The default option for CO_2 disposal is underground geological storage (see Figure 3.1(d)). Studies indicate that geologic formations are the most plentiful and attractive option for U.S. power plants (Bergman and Winter 1995). A nominal cost of $5/ tonne CO_2 has been suggested, which can be changed by the user to match the specific details about the location. If CO_2 is being used as a byproduct for EOR or ECBM activity, it may generate some revenue. This module, which is represented by a single cost parameter, may be expanded in future to include details about the various storage/ disposal options.

3.4. Performance Parameters

There are three types of input parameters to the CO_2 performance model:

- **Parameters from the "reference plant"**: These include the flow rate, temperature, pressure and composition of the flue gas inlet to the CO_2 absorber, and the gross power generation capacity of the power plant.

- **Parameters to configure the CO_2 system**: The CO_2 module provides a menu of options from which the user may select a CO_2 capture technology, CO_2 product pressure, mode and distance of CO_2 product transport, and CO_2 storage/ disposal method. At this stage, a model of the MEA-based absorption system with pipeline transport and geologic sequestration has been developed; other options are still under construction.

- **Parameters controlling the performance of the CO_2 system**: The main parameters include the CO_2 capture efficiency, MEA concentration, lean sorbent CO_2 loading, regeneration heat requirement, pressure drop across the system, MEA make-up requirement, pump efficiency, compressor efficiency and several others.

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These parameters are used to calculate the sorbent flow rate, MEA requirement, and energy penalty of the CO_2 system.

Functional relationships and default values for all model parameters were developed based on engineering fundamentals, a detailed review of the literature, and contacts with experts in the field. All of these performance parameters directly affect the cost of the system. Here is a brief description of the various input parameters to the CO_2 system.

3.4.1. Parameters obtained from the "reference base plant"

The amine-based CO_2 capture system gets the following inputs from the (reference) base plant:

- Gross plant size = MW_g

- Net plant size (prior to environmental controls) = MW_{noctl}

- Flue gas composition and flow rate (as entering into the amine system): This is an array of molar flow rates of different gas components that include N_2, O_2, H_2O, CO_2, CO, HCl, SO_2, SO_3, NO, NO_2 and mass flow rate of particulates.

- The total molar flow rate of the flue gas = G,

- The molar fraction of CO_2 in the flue gas = y_{CO2}.

- Temperature of flue gas = T_{fg}

- Plant capacity factor = PCF (%)

- Annual hours of operation = HPY = (PCF/100)*365.25*24 hrs/yr

3.4.2. Parameters to configure the CO_2 system

These are the choices the user can make in order to configure the CO_2 capture system. These configuration choices have been described in section 3.3, and could be summarized as follows:

- Flue gas cooler: Whether to include DCC (default) or exclude it

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- Sorbent regeneration steam source: Steam extraction from the base plant (default, internal derating) or steam generated from an auxiliary natural gas boiler with (or without) a secondary steam turbine

- Mode of CO_2 product transportation: Via pipelines (default) or any other means.

- Mode of CO_2 storage/ disposal: Underground geologic reservoir (default), EOR, ECBM, Depleted oil/gas wells, or Ocean

3.4.3. *Parameters controlling the performance of the CO₂ system*

Input parameters are given to allow the model user to control the performance of the CO2 capture system. These parameters then determine the intermediate and final output results. It may be noted that the user can override any of these values, but may want to change values of all the relevant parameters to avoid inconsistencies. These are the key input parameters below:

i) CO₂ capture efficiency (η_{CO_2})
The overall CO_2 capture efficiency of the system is the fraction of CO_2 present in the incoming flue gas stream captured in this system.

$$\eta_{CO_2} = 100 \times (\text{Moles } CO_2 \text{ in } - \text{ Moles } CO_2 \text{ out}) / (\text{Moles } CO_2 \text{ in})$$

Most of the studies report the CO_2 capture efficiency of the amine-based systems to be 90%, with few others reporting as high as 96% capture efficiency. Here, it has been assumed to be 90% as the nominal value, but the user can specify the desired level of CO_2 capture efficiency.

ii) MEA concentration (C_{MEA})
The sorbent used for CO_2 absorption is a mixture of mono-ethanol-amine (MEA) with water. The presence of water enhances the chemical absorption reaction of MEA with CO_2. MEA is a highly corrosive liquid, especially in the presence of oxygen and carbon dioxide, and hence needs to be diluted. Today the commercially available MEA-based technology supplied by Fluor Daniel uses 30% w/w MEA sorbent with the help of some corrosion inhibitors. Other suppliers, who do not use this inhibitor, prefer to use lower

MEA concentrations in the range of 15%-20% w/w. Here we use 30% as the nominal value for the sorbent concentration with an acceptable range between 15-40%.

iii) Lean sorbent CO₂ loading (ϕ_{min})

iii) Lean sorbent CO$_2$ loading (ϕ_{min})
Ideally, the sorbent will be completely regenerated on application of heat in the regenerator section. Actually, even on applying heat, not all the MEA molecules are freed from CO_2. So, the regenerated (or lean) sorbent contains some "left-over" CO_2. The level of lean sorbent CO_2 loading mainly depends upon the initial CO_2 loading in the sorbent and the amount of regeneration heat supplied, or alternatively, the regeneration heat requirement depends on the allowable level of lean sorbent loading. Here we use a nominal value of 0.2 based on the values reported in the literature, and the user may specify any desired value in the range (0.1-0.25).

iv) Liquid to gas ratio (L/G)
The liquid to gas ratio is the ratio of total molar flow rate of the liquid (MEA sorbent plus water) to the total molar flow rate of flue gas being treated in the absorber. This is a calculated parameter and is derived by the process simulation model.

v) Liquid flow rate (L)
The liquid flow rate is the total molar flow rate of sorbent plus dilution water being circulated in the CO_2 capture system. It is obtained by multiplying (L/G), which is derived from the process simulation model, by the total flue gas flow rate (G) entering the CO_2 capture system.

$$L = (L/G) \times (G)$$

vi) Acid gases removal efficiency ($\eta_{acid\ gas}$) and stoichiometric MEA loss ($n_{MEA,acidgas}$)
As discussed before, MEA is an alkaline sorbent that has strong affinity for various acid gases. In fact, gases such as hydrogen chloride and oxides of sulfur are much more reactive towards MEA than carbon dioxide itself. These gases form heat stable salts (HSS) with MEA that can not be broken down even after application of heat. So, they cause a permanent loss of MEA sorbent that may be estimated according the stoichiometry of their reactions with MEA. The typical removal efficiencies of these gases in the absorber using MEA sorbent designed for 90% removal of CO_2 are given in Table 3.2.

Table 3.2. Removal efficiency of acid gases due to MEA sorbent (90% CO_2 removal)

Acid gas	Removal efficiency (%)	MEA loss (mole MEA/mole acid gas)
SO_2	$\eta_{SO_2} = 99.5\%$	$n_{MEA, SO_2} = 2$
SO_3	$\eta_{SO_3} = 99.5\%$	$n_{MEA, SO_3} = 2$
NO_2	$\eta_{NO_2} = 25\%$	$n_{MEA, NO_2} = 2$
NO	$\eta_{NO} = 0$	$n_{MEA, NO} = 0$
HCl	$\eta_{HCl} = 95\%$	$n_{MEA, HCl} = 1$

v) Temperature of the flue gas entering the CO_2 capture system ($T_{fg,in}$)

The desirable temperature of the flue gas entering the CO_2 capture system is about 45-50 deg C. If a direct contact cooler is installed upstream of CO_2 capture system, then this temperature level may be achieved.

The temperature of the flue gas affects the absorption reaction as absorption of CO_2 in MEA sorbent is an exothermic process favored by lower temperatures. Also, the flue gas temperature directly affects the volumetric flow rate of the flue gas stream, which is a key determinant of the sizes of various equipments (*e.g.*, direct contact cooler, flue gas blower, absorber).

vi) Nominal MEA loss ($\dot{m}_{MEA, nom}$)

MEA is a reactive sorbent. In spite of dilution with water and use of inhibitors, a small quantity of MEA is lost through various unwanted reactions, mainly the polymerization reaction to form long-chained compounds and the oxidation reaction forming organic acids and liberating ammonia. In general, this nominal loss of MEA is estimated as about 1.5 kg MEA/ tonne CO_2.

It is assumed that 50 % of this MEA loss is due to polymerization:

$$\dot{m}_{MEA, polym} = = 50\% \text{ of } (\dot{m}_{MEA, nom})$$

The remaining 50% of the MEA loss is due to oxidation to acids:

$$\dot{m}_{MEA, oxid} = 50\% \text{ of } (\dot{m}_{MEA, nom})$$

vii) NH₃ Generation (n_{NH₃})

The oxidation of MEA to organic acids (oxalic, formic, etc.) also leads to formation of NH_3. Each mole of MEA lost in oxidation liberates a mole of ammonia (NH_3).

$$\text{Rate of ammonia generation, } n_{NH_3} = \frac{1\,mole\,NH_3}{mole\,MEA\,oxidized}$$

viii) Heat-Stable Salts (HSS)

The organic acids (product of MEA oxidation) combine with MEA to form some other heat stable salts (HSS). The exact nature of these salts is not known. The most conservative estimate, assuming that the organic acids are mono-basic, is that each mole of organic acid takes up one mole of fresh MEA. So, each mole of MEA lost in oxidation takes up additional mole of MEA in HSS formation.

$$n_{MEA, organics} = \frac{1\,mole\,MEA}{mole\,organic\,acid}$$

ix) Caustic Consumption in Reclaimer ($\dot{m}_{NaOH}$)

Caustic in the form of NaOH is added in the reclaimer so that some of the MEA could be regenerated from HSS. $\dot{m}_{NaOH}$ is the quantity (mass) of caustic (as NaOH) consumed in MEA reclaimer per tonne of CO_2 captured. A typical value is 0.13 kg NaOH/ tonne CO_2.

x) Reclaimed MEA

Caustic regenerates stoichiometric amount of MEA from the HSS in the reclaimer. Each mole of NaOH regenerates 1 mole of MEA, and adds the corresponding sodium salt of organic acid to the reclaimer bottoms.

$$\dot{n}_{MEA, reclaimed} = \text{no. of moles of MEA reclaimed using caustic}$$

$$= \text{no. of moles of caustic added}$$

$$= \dot{n}_{NaOH}$$

$$= \dot{m}_{NaOH} \, / \, (\text{Molecular Weight of NaOH})$$

$$= \dot{m}_{NaOH} \, / \, 40$$

xi) Removal efficiency for particulates (η_{partic})

Amine-based absorption system for CO_2 removal is a wet scrubbing operation. So, it also leads to removal of particulate matter from the flue gas to certain extent. Based on the experience of other scrubbing systems, the removal efficiency for particulates has been assumed to be 50%.

xii) Density of sorbent ($\rho_{sorbent}$)

MEA has a density of 1.022 g/cc that is similar to that of water. So, the overall density of the MEA based sorbent (with almost 70% water) is assumed to be same as that of water ~1 tonne/ m^3.

xiii) Activated Carbon Requirement ($\dot{m}_{act\text{-}C}$)

An activated carbon bed in the sorbent circuit removes the long chained/ cyclic polymeric compounds formed from the degenerated MEA. Over a period of time (~3-6 months) the carbon-bed needs to be replaced and the used bed is sent back to the suppliers. $\dot{m}_{act\text{-}C}$ is the average amount of activated carbon consumed per tonne of CO_2 captured. This consumption is estimated to be about 0.075 kg C/ tonne CO_2.

xiv) Total moles of CO_2 captured (n_{CO2})

This is the molar flow rate of CO_2 captured from the flue gas. It is obtained by multiplying the total CO_2 content in the inlet flue gas (kmole CO_2/ hr) by the CO_2 capture efficiency of the system.

$$n_{CO2} = (\eta_{CO_2} \, / \, 100)*(\text{Moles } CO_2 \text{ in}) = (\eta_{CO_2} \, / \, 100)*(G*y_{CO2})$$

Since the molecular weight of CO_2 is 44, the total amount of CO_2 captured (m_{CO2}, tonne/ hr) is

$$m_{CO2} = n_{CO2} * (44/1000)$$

xv) CO_2 product purity (α)

The final CO_2 product has to meet certain specifications depending upon the mode of transport and final destination. Impurities such as nitrogen are undesirable as they may pose problems during compression and liquefaction of CO_2. In order to avoid corrosion in the pipelines during transport, the moisture levels must be controlled. The acceptable level of purity of CO_2 product for most of the applications is about 99.8%.

xvi) Reboiler duty per mole of liquid (Q/L)

This is the total amount of heat energy input required for the regeneration of the sorbent per unit of liquid circulated. This is mainly dependent on lean sorbent loading, CO_2 capture efficiency, MEA concentration and CO_2 content of the flue gas and is a calculated value derived from the process simulation model.

xvii) Total heat requirement for sorbent regeneration (Q)

This is the total amount of heat energy required in the reboiler for sorbent regeneration. It is obtained by multiplying (Q/L) above, by the total sorbent circulation molar flow rate (MEA sorbent plus dilution water) in the CO_2 capture system.

$$Q = (Q/L) \times (L)$$

xviii) Unit heat of sorbent regeneration (q_{regen})

This is the amount of heat required for the regeneration of the MEA sorbent (loaded with CO_2) in the stripper/ regenerator section. It is expressed as the amount of heat (in kJ or Btu) per unit mass (kg or lb) of CO_2 captured. Theoretically, the heat of reaction that needs to be supplied in order to reverse the absorption reaction between CO_2 and MEA is about 1900 kJ/ kg CO_2 (*i.e.* about 825 Btu/ lb CO_2). The actual amount of heat required for regeneration of the sorbent is much higher, about 2-3 times higher than this theoretical minimum. This is because of the large amount of latent heat taken up by the dilution water in the sorbent. A wide range of numbers have been reported for the regeneration heat requirement of MEA system. The majority of the sources report a heat requirement of about 4000 kJ/kg CO_2. Here it is obtained by dividing the total heat requirement for sorbent regeneration (Q) by the total amount of CO_2 captured (m_{CO2}).

$$q_{regen} = Q / m_{CO2}$$

xix) Enthalpy of regenerating steam (h_{steam})

The regeneration heat is provided in the form of LP (low pressure) steam extracted from the steam turbine in case of coal-fired power plants and combined-cycle gas plants, through the reboiler (a heat exchanger). In case of simple cycle natural gas fired power plants, a heat recovery unit maybe required. (h_{steam}) is the enthalpy or heat content of the steam used for sorbent regeneration. Typically, the LP steam is around 300°C and 60-80 psi. From the steam-tables, the enthalpy (heat content) of such steam is found to be about 2000 kJ/ kg steam.

xx) Heat-to-electricity equivalence factor (F_{HE})

The extraction of LP (low pressure) steam for sorbent regeneration results in some loss of power generation capacity of the plant. The factor (F_{HE}) gives the equivalent loss of power generation capacity due to the heat requirement for sorbent regeneration.

Let us try to understand what this factor stands for and how it is derived. According to the Second Law of Thermodynamics, thermal energy cannot be fully converted to work. Electrical energy is a form of work. The efficiency of converting heat into electricity depends upon the temperature and pressure conditions of the working fluid, *i.e.* steam in case of a conventional coal plant. Even the advanced steam plants (using super-critical or ultra-super-critical boilers) can achieve net plant efficiencies of about 42-45 percent by using high quality steam (steam at high pressure and temperature, *e.g.* 31 MPa and 590 °C) (Rubin 2001). In general, steam at lower pressure and temperature will have still lower efficiency to convert heat into electric energy. The steam used for regeneration of sorbent is low quality steam (low pressure, ~0.5 MPa and low temperature, ~120 °C), and hence the equivalent loss of power generation is much lower than the power plant efficiency. Again, the electric energy penalty due to steam extraction also depends on the overall heat integration within the plant. Factor (F_{HE}) may be looked upon as the power generation efficiency of the LP steam extracted from the steam cycle of the power plant for sorbent regeneration. So, the equivalent loss in power generation capacity of the base plant may be estimated by multiplying the total sorbent regeneration heat requirement by this factor.

From the data obtained from the available studies (Smelser, Stock et al. 1991; Hendriks 1994; Mimura, Simoyoshi et al. 1997; Bolland and Undrum 1998; Marion, Nsakala et al. 2001), this factor has been found to lie in the range (9, 19) for a new plant and (20, 25) for retrofit cases. So, the nominal value (for this new plant application) has been taken as 14%.

For example, if 10,000 kJ/hr is the regeneration heat requirement for the CO_2 capture operation, then the corresponding loss in power generation capacity of the power plant is estimated as 14% of 10,000 kJ/hr i.e. 1400 kJ/hr, or (1400/3600 =) 0.39 kW. It may be noted that, in case of retrofit applications, the energy penalty might be significantly higher due to poor heat integration, and F_{HE} may be approximately 22%.

xxi) Blower pressure head (ΔP_{fg})

Prior to entering into the absorber column, the flue gas has to be compressed in a flue gas blower so that it can overcome the pressure drop in the absorber tower. (ΔP_{fg}) is the pressure head that needs to be provided to the flue gas in the blower, and is about 14 kPa (~2 psi).

xxii) Blower (fan) efficiency (η_{blower})

This is the efficiency of the fan/blower to convert electrical energy input into mechanical work output. Typically, the value of blower efficiency (η_{blower}) is about 75%.

xxiii) Sorbent pumping head ($\Delta P_{sorbent}$)

The sorbent has to flow through the absorber column through packed media, countercurrent to the flue gas flowing upwards. So, some pressure loss is encountered in the absorber column and sufficient sorbent head has to be provided to overcome these pressure losses. ($\Delta P_{sorbent}$) is the pressure head to be provided to the sorbent using sorbent circulation pumps. A typical value is about 200 kPa (~ 30 psi).

xxiv) Pump efficiency (η_{pump})

This is the efficiency of the sorbent circulation pumps to convert electrical energy input into mechanical energy output. Typically, the value of (η_{pump}) is assumed to be 75%.

xxv) CO_2 product pressure (P_{CO_2})

The CO_2 product may have to be carried over long distances. Hence it is necessary to compress and liquefy it to very high pressures (P_{CO_2}), so that it may be delivered to the required destination in liquid form and with minimal recompression facilities en route. The critical pressure is the minimum pressure required to liquefy a gas at its critical temperature, which is the temperature above which it is not possible to liquefy a gas. For CO_2 this is about 1070 psig. The typically reported value of final pressure to which the product CO_2 stream is pressurized using compressors before it is transported is about 2000 psig (about 137 atm).

xxvi) Energy required for CO_2 compression (e_{comp})

This is the electrical energy required (kWh per tonne CO_2) to compress a unit mass of CO_2 product stream to the designated pressure (P_{CO_2}) expressed in psig. Compression of CO_2 to high pressures requires substantial energy, and is a principle contributor to the overall energy penalty of a CO_2 capture unit in a power plant.

xxvii) CO_2 compression efficiency (η_{comp})

This is the effective efficiency of the compressors used to compress CO_2 to the desirable pressure. Typically, the value of compressor efficiency (η_{comp}) is about 80%. It may be noted that the energy requirement calculated from the performance equation (e_{comp}) has to be corrected by this efficiency factor in order to get the total energy required for CO_2 compression.

The following set of parameters are relevant only if the CO_2 capture system has been configured to include an auxiliary NG boiler to supply sorbent regeneration heat.

xxviii) Heating value of natural gas (NGHV)

This is the high heating value (HHV, MJ/ kmole NG) of the natural gas used as fuel for the auxiliary boiler.

xxix) Density of natural gas (ρ_{NG})

This is the density (lb/ft^3) of the natural gas used as fuel for the auxiliary boiler.

xxx) Average molecular weight of natural gas (mw_NG)

This is the average molecular weight (kg/ kmole NG) of the natural gas used as fuel for the auxiliary boiler. This is a function of the molar composition of the natural gas.

xxxi) Flow rate of natural gas (m_NG)

This is the total molar flow rate (kmole NG/ hr) of the natural gas used as fuel for the auxiliary boiler. It is a function of the total heat requirement for sorbent regeneration in the amine system.

xxxii) Auxiliary NG boiler efficiency (η_{NGB})

This is the efficiency of the auxiliary boiler that uses natural gas as fuel input. It is defined as the ratio of total thermal energy (in the from of steam) delivered by the boiler divided by the total heat energy input (in the form of heating value of the natural gas input).

xxxiii) Secondary steam turbine power generation efficiency (η_{ST2})

This is the efficiency of the secondary steam turbine added with the auxiliary NG boiler to generate electrical power. It may be defined as the ratio of electrical energy generated (MW_{ST2}) by the steam turbine divided by the total thermal energy (in the form of steam) input from the auxiliary NG boiler. It is assumed that the rest of the thermal energy is contained in the LP exhaust steam from the turbine, which is sent to the reboiler for sorbent regeneration.

3.5. Performance Equations

The performance equations define the functional relationships among various key performance parameters. They have been derived as multivariate linear or log-linear regression equations from the data obtained from the process simulation model runs.

$$(L/G) = \exp (-1.4352 + 0.1239*y_{CO2} + 3.4863*\phi_{lean} + 0.0174*\eta_{CO2} - 0.0397*C + 0.0027*T_{fg,in}) \qquad [\text{adj. } R^2 = 0.92] \qquad (3\text{-}1)$$

$$(Q/L) = \exp (-2.4452 - 0.0037*y_{CO2} - 6.2743*\phi_{lean} + 0.0254*C) $$
$$[\text{adj. } R^2 = 0.96] \qquad (3\text{-}2)$$

$$(T_{fg,out}) = 41.15 + 0.062*T_{fg,in} + 1.307*y_{CO2} - 18.872*\phi_{lean} + 0.270*C$$
$$[\text{adj. } R^2 = 0.92] \qquad (3\text{-}3)$$

$$(mw_{lean}) = 16.907 + 2.333*\phi_{lean} + 0.204*C$$

[adj. $R^2 = 0.95$]   *(3-4)*

$$(e_{comp}) = -51.632 + 19.207*\ln(P_{CO2} + 14.7)$$

[adj. $R^2 > 0.99$]   *(3-5)*

where,

    L = total sorbent flow rate (kmole/ hr)

    G = total inlet flue gas flow rate (kmole/ hr)

    (L/G) = total liquid (sorbent) applied per unit flue gas flow rate in absorber (ratio of molar flow rates)

    Q = total sorbent regeneration heat requirement (GJ/ hr)

    (Q/L) = total regeneration heat supplied per unit of sorbent flow (GJ/ kmole)

    y_{CO2} = CO_2 concentration in the inlet flue gas (mole %)

    ϕ_{lean} = lean sorbent CO_2 loading (mole CO_2/ mole MEA)

    η_{CO2} = CO_2 capture efficiency (%)

    C = MEA concentration in the sorbent (wt %)

    $T_{fg,in}$ = Temperature of the flue gas entering the CO_2 absorber (deg C)

    $T_{fg,out}$ = Temperature of the flue gas leaving the CO_2 absorber (deg C)

    mw_{lean} = Average molecular weight of the lean sorbent (kg/ kmole sorbent)

    e_{comp} = Unit energy requirement for CO_2 compression (kWh/ tonne CO_2)

    P_{CO2} = Desired CO_2 product pressure (psig)

3.6. Model Outputs

The key outputs of the amine system performance model include:

- **MEA makeup requirement**. This depends mainly on the mass flow rate of CO_2 in the flue gas, the desired CO_2 capture efficiency, MEA concentration, and CO_2 loadings in the sorbent. Depending on the level of impurities in the flue gas, there is some loss of sorbent. If the power plant does not have emission controls for SO_x and NO_x, the cost imposed due to amine loss may be significant.

- **Energy requirement**. Heat for sorbent regeneration is derived from low-pressure steam available in the power plant, which decreases power generation efficiency. Additional electrical energy is required for CO_2 product compression, sorbent circulation, and other system requirements. The energy requirement is one of the most important results, as it dictates the net size of the power plant, and hence the net cost of power generation and CO_2 avoidance.

The following material and energy flows are estimated using the above stated inputs

1.1.1. *Total quantity of CO_2 captured:*

$$m_{CO2} \text{ (tonne/hr)} = n_{CO2} \times (MolWt)_{CO2}$$

$$= \eta_{CO2} \times n_{CO2,inlet} \times (MolWt)_{CO2}$$

where,

n_{CO2} = Total moles of CO_2 captured (kmole CO_2/ hr)

$n_{CO2,inlet}$ = Molar flow rate of CO_2 in the inlet flue gas (kmole CO_2/hr)

$(MolWt)_{CO2}$ = Molecular weight of CO_2 = 0.044 tonne/ kmole CO_2

3.6.2 *MEA makeup requirement:*

3.2.1.
$$\dot{m}_{MEA,makeup}$$

= Net loss of MEA

$$= \begin{pmatrix} \text{loss due to acid gas} \\ \textit{impurities} \end{pmatrix} + \begin{pmatrix} \text{loss due to} \\ \text{polymerization} \end{pmatrix}$$

$$+ \begin{pmatrix} \text{loss due to} & \text{Gain in} \\ \text{HSS formation} & - & \text{Reclaimer} \end{pmatrix} + \begin{pmatrix} \text{loss with} \\ \text{fluegas exhaust} \end{pmatrix}$$

3.6.3 Total sorbent circulation flow rate:

From the performance equations, we find

$L/G = f(y_{CO2}, \phi_{lean}, \eta_{CO2}, C, T_{fg,in})$, and $L = G*(L/G)$

Including the MEA makeup requirement gives the total sorbent flow rate (m^3/hr)

$$L_{tot,v} = \{G*(L/G)*mw_{lean} + \dot{m}_{MEA,makeup}*(100/C)\}*\rho_{sorbent}$$

3.6.4 Waste generated from reclaimer:

$$m_{waste} = \begin{pmatrix} MEA\,lost\,due\,to \\ acid\,gases \end{pmatrix} + \begin{pmatrix} Total\,qty\,of \\ acid\,gases\,removed \end{pmatrix} + \begin{pmatrix} MEA\,loss\,due \\ to\,oxidation \end{pmatrix}$$
$$+ \begin{pmatrix} MEA\,loss\,due\,to \\ HSS\,format^n \\ - Gain\,in\,reclaimer \end{pmatrix} + \begin{pmatrix} Caustic\,added \\ to \\ Reclaimer \end{pmatrix}$$

Considering ($f_{w,waste}$) as the water content (% w/w) in the waste, the actual mass flow rate of waste is obtained as:

$M_{waste,total} = m_{waste}/(1 - f_{w,waste})$ kg/hr

Typically, the reclaimer waste contains about 40% water.

3.6.5 Activated carbon consumption:

$m_{act-C} = \dot{m}_{act-C} \times m_{CO_2}$ kg act-C/hr

3.6.6 Caustic consumption in reclaimer:

$m_{Caustic} = \dot{m}_{NaOH} \times m_{CO_2}$ kg NaOH/ hr

3.6.7 Process water requirement:

Unit process water makeup = $\dot{m}_{pw}$ (tonne/ hr)/MW(net)

Typically, the value of $\dot{m}_{pw}$ is about 0.114 tonne/hr per MW(net) (Smelser, Stock et al. 1991). Therefore, the process water requirement is:

$$(M_{pw}) = \dot{m}_{pw} \times MW_{net} \quad \text{tonne/hr}$$

3.6.8 Cooling water requirement:

If there is a direct contact cooler installed, the required flow rate of cooling water is estimated based on the following assumptions

Specific heat of water, $SH_w = 4.2$ kJ/kg ^{o}C

Specific heat of flue gas $= SH_{fg}$ (Generally, this is around 1.2 kJ/kg ^{o}C)

Temperature rise in the cooling water (once through system) $= \Delta T_w$

Drop in flue gas temperature $= \Delta T_{fg} = (T_{fg,i} - T_{fg})$ ^{o}F

where,

$T_{fg,i} =$ Temperature of flue gas entering the direct contact cooler

$T_{fg} =$ Temperature of flue gas exiting the direct contact cooler

Mass flow rate of flue gas $= m_{fg}$ tonne/ hr

So, the required cooling water flow rate,

$$M_{cw} = m_{fg}*(\Delta T_{fg} / \Delta T_w)*(SH_{fg} / SH_w) \quad \text{tonne/hr}$$

3.6.9 Steam requirement:

LP steam is extracted from the power plant steam turbine or secondary steam turbine in order to provide the sorbent regeneration heat in the reboiler. Based on the regeneration heat requirement and enthalpy of regeneration steam, the flow rate of steam may be estimated as follows:

From the performance equations,

$$(Q/L) = f(y_{CO2}, \phi_{lean}, C)$$

August 2011 Update: In the most recent version of the IECM, thesteam use at the reboiler is calculated using a new set of formulas. Please see "IECM Technical Documentation Updates, Vol. I to V (Final Report to USDOE/NETL from Carnegie Mellon Univesrity)", for details.

Total regeneration heat requirement,

Q (MJ/ hr) $= (Q/L)*(L)$

Mass flow rate of steam,

m_{steam} (tonne/hr) $= Q / q_{steam}$

The equivalent energy penalty due to regeneration steam requirement is (E_{regen}). Depending upon the CO_2 capture system configuration (source of regeneration steam supply), E_{regen} has to be estimated in two different ways.

1. In case of steam extraction from the base plant steam cycle (derating)

$$E_{regen} = Q*F_{HE}$$

2. In case of steam supplied from an auxiliary NG boiler,

$$E_{regen} = - E_{ST2} = - (m_{NG}*NG_{HV}*\eta_{NGB}*\eta_{ST})$$

It maybe noted that in the case of auxiliary NG boiler, the energy penalty term is negative, implying that there is an increase in the net power generation of the plant.

3.6.10 Electrical energy requirement:

Total electricity requirement of CO_2 capture system is:

$$E_{CO_2, tot} = E_{regen} + E_{pumping} + E_{compr}$$

where,

E_{regen} = as explained above

$E_{pumping} = E_{blower} + E_{pump}$

$$E_{blower} \text{ (hp)} = \frac{144 \, Q_{fg} \Delta P_{fg}}{33000 \cdot \eta_{blower}}$$

where Q_{fg} and ΔP_{fg} are expressed in ft^3/min and psi respectively,

$$E_{pump} \text{ (hp)} = \frac{2Q_{solvent} \, \Delta P_{solvent}}{1714 \cdot \eta_{pump}}$$

where $Q_{sorbent}$ and $\Delta P_{sorbent}$ are expressed in gal/min and psi respectively, and

$$E_{compr} = e_{comp} * m_{CO2}/\eta_{comp}$$

where e_{comp} is given by equation *(3-5)*.

3.7. Characterization of Uncertainty and Variability

Any techno-economic analysis, and especially that of new energy and environmental control technologies that are still in the research phase, involves uncertainties regarding the performance and costs. These uncertainties come from incomplete information available and numerous assumptions and approximations built into simulations. Some parameters, especially the cost parameters, are influenced by a larger set of factors outside the scope of the particular study and fluctuations in these quantities may be seen as "inherent randomness" when viewed within this limited focus area. In addition, there may be significant variability in plant or process design assumptions across different studies or organizations.

One of the distinguishing features of this modeling effort is a probabilistic capability that allows model inputs to be represented by probability distributions rather than single deterministic values. Probability distributions for these parameters reflect the ranges of values reported in the literature, the evolving nature of the technology, and practical considerations in running such plants. Also, it is possible to use probability distributions for more than one parameter (or all the parameters together) simultaneously.

Table 3.3 lists the uncertainty distributions developed for performance model parameters based on the current literature on amine-based (MEA) systems. These distributions reflect both uncertainty and variability in system designs. Details are presented in *Appendix B* and *Appendix C*. The data sources from which the parameter values were

Table 3.3. Amine system performance model parameters and uncertainties

Performance Parameter	Units	Data (Range)	Nominal Value	Unc. Representation (Distribution Function)
CO_2 removal efficiency	%	Mostly 90	90	-
SO_2 removal efficiency	%	Almost 100	99.5	Uniform(99,100)
NO_2 removal efficiency	%	20-30	25	Uniform(20,30)
HCl removal efficiency	%	90-95	95	Uniform(90,95)
Particulate removal eff.	%	50	50	Uniform(40,60)
MEA concentration	wt%	15-50	30	Triangular (20,30,40)
Lean sorbent CO_2 loading	mol CO_2/mol MEA	0.15-0.30	0.2	Triangular(0.1,0.2,0.25)
Nominal MEA make-up	kg MEA/tonne CO_2	0.5-3.1	1.5	Triangular(0.5,1.5,3.1)
MEA loss (SO_2)	mol MEA/mol SO_2	2	2	-
MEA loss (NO_2)	mol MEA/mol NO_2	2	2	-
MEA loss (HCl)	mol MEA/mol HCl	1	1	-
NH_3 generation	mol NH_3/mol MEA oxidized	1	1	-
Caustic consumption in MEA reclaimer	kg NaOH/tonneCO_2	0.13	0.13	-
Activated carbon use	kg C/tonne CO_2	0.075	0.075	-
Gas-phase pressure drop	kPa	4-37	14	Triangular(4,14,37)
Fan efficiency	%	70-80	75	Uniform (70,80)
Sorbent pumping head	kPa	200	200	Triangular(150,200,250)
Pump efficiency	%	70-80	75	Uniform (70,80)
Equiv. elec. requirement	% regeneration heat	9-19	14[a]	Uniform (9,19)
CO_2 product pressure	MPa	7.58-15.16	13.79	Triangular(7.58,13.79,15.16)
Compressor efficiency	%	75-88	80	Uniform (75,88)

[a] For retrofit applications, nominal value is 22.

obtained also included peer-reviewed journal articles, conference papers, books, technical reports, and technical judgments given by experts.

Almost all the sources reported a CO_2 capture efficiency of 90%. Hence this value has been used as the default nominal value without any default probability distribution. Nonetheless, the user is allowed to use a probability distribution for this parameter as well.

References (Chapter 3)

Aspen-Plus[R] (2002). Aspen Technology, Inc. Cambridge, MA.

Bergman, P. D. and E. M. Winter (1995). "Disposal of carbon dioxide in aquifers in the U.S." Energy Conversion and Management **36(6-9)**: 523-526.

Bolland, O. and H. Undrum (1998). Removal of CO2 from gas turbine power plants: Evaluation of pre- and post-combustion methods. Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September, Interlaken, Switzerland, Elsevier Science Ltd.

Chakma, A. and P. T. Tontiwachwuthikul (1998). Designer solvents for energy efficient CO2 separation from flue gas streams. Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September, Interlaken, Switzerland, Elsevier Science Ltd.

Chapel, D., J. Ernst, et al. (1999). Recovery of CO2 from flue gases: commercial trends (paper no. 340). presented at the Canadian Society of Chemical Engineers Annual Meeting, 4-6 October, Saskatoon, Saskatchewan, Canada.

Hendriks, C. (1994). Carbon Dioxide Removal from Coal-fired Power Plants. The Netherlands, Kluwer Academic Publishers.

Herzog, H. J. (1998). The economics of CO2 capture. Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September, Interlaken, Switzerland, Elsevier Science Ltd.

IECM (2001). Integrated Environmental Control Model and User Documentation (available at www.iecm-online.com), Center for Energy and Environmental Studies, Carnegie Mellon University, Pittsburgh, PA.

Kohl, A. L. and R. B. Nielsen (1997). Gas Purification. Houston, TX, Gulf Publishing Company.

Marion, J., N. Nsakala, et al. (2001). Engineering feasibility of CO2 capture on an existing US coal-fired power plant. in the proceedings of the Twenty-sixth International Conference on Coal Utilization and Fuel Systems, 5-8 March, Clearwater, FL, USA.

Mimura, T., H. Simoyoshi, et al. (1997). "Development of energy saving technology for flue gas carbon dioxide recovery in power plants by chemical absorption method and steam system." Energy Conversion and Management **38(Suppl)**: S57-S62.

ProTreat^{TM} (2002). Optimized Gas Treating, Inc.,. Houston, TX.

Rubin, E. S. (2001). Introduction to engineering and the environment. New York, NY, The McGraw-Hill Companies, Inc.

Rubin, E. S., J. R. Kalagnanam, et al. (1997). "Integrated environmental control modeling of coal-fired power systems." Journal of Air and& Water Management Association **47**: 1180-1188.

SAS (1999-2001). SAS Institute Inc. Cary, NC, USA.

Simbeck, D. (1998). A portfolio selection approach for power plant CO2 capture, separation and R & D options. Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September, Interlaken, Switzerland, Elsevier Science Ltd.

Simbeck, D. and M. McDonald (2000). Existing coal power plant retrofit CO_2 control options analysis. presented at the Fifth International Conference on Greenhouse Gas Control Technologies, 13-16 August, Cairns, Australia.

Smelser, S. C., R. M. Stock, et al. (1991). Engineering and economic evaluation of CO2 removal from fossil-fuel-fired power plants, EPRI IE-7365, Volume 1, Project 2999-10, a research project final report prepared by Fluor Daniel Inc., for EPRI and IEA.

4. COST MODEL DEVELOPMENT

The CO_2 capture and sequestration system cost model is directly linked to the performance model. The cost model follows the framework used in the IECM (Rubin, Kalagnanam et al. 1997) to ensure consistency in economic calculations. There are four types of cost calculated by this model based on the available data (Smelser, Stock et al. 1991; Hendriks 1994; Leci 1996; Mariz 1998; Simbeck 1998; Chapel, Ernst et al. 1999; Desideri and Paolucci 1999; Jeremy and Herzog 2000). They are capital cost, operation and maintenance (O&M) cost, incremental cost of electricity (COE) and cost of CO_2 avoidance.

4.1 Capital Cost

The total capital requirement (TCR) of a system is calculated as the sum of direct equipment costs (which depend on one or more performance variables that determine the size or capacity of the component), plus various indirect costs that are estimated as fractions of the total direct cost following the EPRI cost estimating guidelines (TAG 1993; TAG 1999).

The capital cost model is primarily based on detailed information obtained from Fluor Daniel Inc. (Fluor 1998). To develop a generalized cost model applicable to different plant sizes and material flow rates, the Fluor Daniel data are used as a reference point for the application of engineering "scaling laws". For example, where multiple trains are required to perform the CO_2 capture operation, the maximum train size is taken to be 5000 tonnes per day of CO_2, based on current design. Based on the actual CO_2 capture rate (m_{CO2}) the number of trains required to be installed (N_{min}) is determined.

Different components have different maximum capacity limits. $E_{n,i}$ defines the number of components required per train. Each train consists of the following pieces of equipment ($E_{n,i}$):

- Direct contact cooler (DCC), flue gas blower, absorber, heat exchanger, regenerator, steam extractor, MEA reclaimer - 1 per each train

- Pumps - 2 per train

- Reboilers - 4 per train

Special cases:

1. Only one installation is required for the sorbent processing area, auxiliary natural gas boiler, secondary steam turbine, CO_2 transport facility and CO_2 disposal facility.

2. In case of CO_2 compressors, which have higher capacity (~ 7200 tonnes per day of CO_2), the number of compressors required is calculated accordingly.

The unit cost of each component of this system (absorber, regenerator, flue gas blower *etc*.) is scaled based on the flow rate of the material being handled by that particular device, using the 0.6 power law commonly used in chemical engineering costing (Peters and Timmerhaus 1991). For example, the cost of an absorber is scaled on the basis of the flue gas flow rate entering the CO_2 system, with the data obtained from Fluor Daniel serving as the reference basis for this scaling. Thus, in general, the capital cost of a component of arbitrary size (less than the maximum size) may be estimated as:

$$C_i = C_{i,ref} \cdot \left(\frac{X_i}{X_{i,ref}} \right)^{0.6}$$

where,

C_i = Component cost

$C_{i,ref}$ = Cost of reference size

X = actual flow rate (for scaling)

$X_{i,ref}$ = Flow rate of reference plant

Once the cost of a particular equipment (C_i) is calculated, it needs to be multiplied by the total number of equipment installed (Z_i) in order to get the total cost of installation for that process area (i).

The overall amine system is divided into a number of process areas for which some physical quantity is identified as the basis for scaling of the capital cost. For example, flue gas flow rate, sorbent flow rate, CO_2 product flow rate, CO_2 compression energy requirement, steam flow rate, and makeup MEA flow rate are used for scaling he capital cost of various process areas.

The direct capital cost (process facilities cost) of CO_2 capture and separation system consists of the following cost areas:

Direct contact cooler: In case of coal-fired power plant applications that have a wet FGD (flue gas desulfurization) unit upstream of the amine system, the wet scrubber helps in substantial cooling of the flue gases, and additional cooler may not be required. In case of gas-fired power plants or majority of coal-fired power plants that do not have wet scrubbers for SO_2 removal, a direct contact cooler has to be installed to bring down the temperature of the flue gas stream to acceptable levels. A direct contact cooler is a large vessel where the incoming hot flue gas is made to contact with the cooling water. The size of this unit is a function of the volumetric flow rate of the flue gas, which in turn depends upon the temperature and pressure conditions of the flue gas stream. The capital cost of the unit is estimated as:

$$C_{dcc} = C_{dcc,\ ref} \cdot \left(\frac{V_{fg}}{V_{fg,ref}} \cdot \frac{T_{fg}}{T_{fg,ref}} \right)^{0.6}$$

Flue gas blower: The cooled flue gas is pressurized using a blower before it enters the absorber. The size (and the cost) of the blower is again a function of the volumetric flow rate of the flue gas as it enters the blower. So, the cost may be estimated as:

$$C_{blower} = C_{blower,\ ref} \cdot \left(\frac{V_{fg,1}}{V_{fg,1,ref}} \cdot \frac{T_{fg,1}}{T_{fg,1,ref}} \right)^{0.6}$$

Absorber: This is the vessel where the flue gas is made to contact with the MEA-based sorbent, and some of the CO_2 from the flue gas gets dissolved in the sorbent. Again, the size of this unit is mainly a function of the volumetric flow rate of the flue gas, which in turn depends upon the temperature and pressure conditions of the flue gas stream, as it

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enters this vessel. The cost of the unit is again estimated on the basis of the cost information available for a particular reference case:

$$C_{absorber} = C_{absorber,\ ref} \cdot \left(\frac{V_{fg,in}}{V_{fg,in,ref}} \cdot \frac{T_{fg,in}}{T_{fg,in,ref}} \right)^{0.6}$$

Rich/lean cross heat exchanger: The rich (CO$_2$-loaded) and lean (regenerated) sorbent streams are passed through this cross heat exchanger, where the rich sorbent gets heated and the lean sorbent gets cooled. So, the size (and cost) of this unit is a function of the volumetric sorbent flow rate in the absorber. It is assumed that this volumetric flow rate is constant in the range of temperature and pressure conditions found in this system. The capital cost of the unit is estimated as:

$$C_{crossHEx} = C_{crossHEx,\ ref} \cdot \left(\frac{V_{solvent}}{V_{solvent,ref}} \right)^{0.6}$$

Regenerator: This is the column where the CO$_2$-loaded sorbent is regenerated with the application of heat. Sorbent flow rate is the main physical quantity that decides the size (and cost) of this unit, for a given residence time, which is a function of many parameters including the sorbent concentration, desired CO$_2$ capture efficiency, *etc*. So, the cost may be estimated as above:

$$C_{regenerator} = C_{regenerator,\ ref} \cdot \left(\frac{V_{solvent}}{V_{solvent,ref}} \right)^{0.6}$$

Reboiler: The regenerator is connected with a reboiler, which is a heat exchanger where low-pressure steam extracted from the power plant is used to heat the loaded sorbent. So, the size (and cost) of this unit is a function of mainly the flow rate of the sorbent as well as the flow rate of steam. The cost of the unit is estimated as:

$$C_{reboiler} = C_{reboiler,\ ref} \cdot \left(\frac{V_{solvent}}{V_{solvent,ref}} \cdot \frac{M_{steam}}{M_{steam,ref}} \right)^{0.6}$$

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It may be noted that the ratio of mass flow rates of LP steam (M_{steam}/ $M_{steam,ref}$) has been used in place of the ratio of volumetric flow rates of LP steam, assuming that the temperature and pressure conditions of the LP steam in both cases (actual and reference) are approximately the same.

Steam extractor: Steam extractors are installed to take LP steam from the steam turbines in the power plant. The size (and the cost) of the steam extractor is assumed to be a function of the steam flow rate.

$$C_{steam\_extractor} = C_{steam\_extractor,\ ref} \cdot \left(\frac{M_{steam}}{M_{steam,ref}} \right)^{0.6}$$

This cost item is included if the CO_2 capture system is configured to make use of steam extracted from the steam cycle of the base plant. Alternatively, an auxiliary NG boiler and a secondary steam turbine may be used, and the next two cost items (C_{NG_boiler} and C_{ST2}) are included in its place.

Auxiliary boiler <u>with steam turbine</u>:   The cost of the NG boiler is estimated on the basis of the steam flow rate generated by the boiler using a cost estimation formula reported by Simbeck (Simbeck and McDonald 2000):

$$C_{NG\_boiler} = \$15 \cdot (\text{steam flow rate expressed in lb/hr})$$

Since the steam flow rate (m_{steam}) in the model is given as tonnes/hr, the following expression is obtained after accounting for the unit conversions

$$C_{NG\_boiler} = \$33000 \cdot (m_{steam})$$

The cost of the secondary steam turbine is estimated on the basis of the electrical power generated from this new turbine again using a cost estimation formula reported by Simbeck (Simbeck and McDonald 2000):

$$C_{ST2} = \$300 \cdot (E_{ST2})$$

Where,

E_{ST2} = Power generation from secondary steam turbine expressed in MWe

Auxiliary boiler <u>without steam turbine</u>: If a NG boiler is installed to provide only LP steam (and not to generate any electrical power using a secondary steam turbine), then the boiler cost is lower than that of high pressure boiler discussed earlier. According to Simbeck (Simbeck 2002)its cost is estimated as:

$$C_{NG\_boiler\_only} = \$22000 \cdot (m_{steam})$$

MEA reclaimer: In order to avoid accumulation of the heat stable salts in the sorbent stream and to recover some of the lost MEA sorbent, a part of the sorbent stream is periodically distilled in this vessel. Addition of caustic helps in freeing of some of the MEA. The amount of MEA makeup requirement may be taken as an indicative of the amount of heat stable salts formed and the quantity of sorbent to be distilled in the reclaimer. So, the mass flow rate of makeup MEA requirement is used as a scaling parameter to estimate the cost of this unit:

$$C_{MEA\_reclaimer} = C_{MEA\_reclaimer,\ ref} \cdot \left(\frac{M_{MEA\_makeup}}{M_{MEA\_makeup,ref}} \right)^{0.6}$$

Sorbent processing area: The sorbent processing area primarily consists of sorbent cooler, MEA storage tank, and a mixer. It also consists of an activated carbon bed filter that adsorbs impurities (degradation products of MEA) from the sorbent stream. So, the size (and cost) of this unit (together) will be a function of the total sorbent flow rate, and may be estimated as follows:

$$C_{solvent\_proc} = C_{solvent\_proc,\ ref} \cdot \left(\frac{V_{solvent}}{V_{solvent,ref}} \right)^{0.6}$$

CO_2 drying and compression unit: The multi-stage compression unit with inter-stage cooling and drying yields the final CO_2 product at the specified pressure (about 2000 psig) that contains only acceptable levels of moisture and other impurities (e.g. N_2). The size (and cost) of this unit will be a function of the CO_2 product flow rate, and may be estimated as follows:

$$C_{CO2\_compr} = C_{CO2\_compr,\ ref} \cdot \left(\frac{M_{CO2}}{M_{CO2,ref}} \right)^{0.6}$$

The sum of all these individual process area equipment costs is termed as process facilities capital (PFC). The various indirect costs are then estimated as fractions of the PFC following the EPRI cost estimating guidelines (TAG 1993; TAG 1999). Table 4.1 lists the elements of total capital cost. Because of data limitations some of the indirect cost factors for the amine system are estimated based on other similar technologies.

Table 4.1. MEA capital cost model parameters and nominal values

	Capital Cost Elements	Value
A	Process Area Equipment Costs	$A_1, A_2, A_3, \ldots, A_{10}$
B	Total Process Facilities Capital (PFC)	ΣA_i
C	Engineering and Home Office	7% PFC
D	General Facilities	10% PFC
E	Project Contingency	15% PFC
F	Process Contingency	5% PFC
G	Total Plant Cost (TPC) = sum of above	B+C+D+E+F
H	AFUDC (interest during construction)	Calculated
I	Royalty Fees	0.5% PFC
J	Pre-production	1 month's fixed O&M cost
K	Pre-production	1 month's variable O&M cost
L	Inventory (startup) Cost	0.5% TPC
M	Total Capital Requirement (TCR)[a]	G+H+I+J+K+L

The total plant cost (TPC) is the sum of the process facilities capital (PFC), general facilities capital (GFC, which is the total construction cost of the general facilities, including roads, office buildings, shops, laboratories etc.), engineering and home office (EHO) overhead, and contingencies – project and process. The project contingency is a capital cost contingency factor covering the cost of additional equipment or other costs that would result from a more detailed design at an actual site. On the other hand, the process contingency is a capital cost contingency factor (added cost) applied to a technology to reflect its level of maturity. TPC is developed on the basis of instantaneous

("overnight") construction – occurring at a single point in time, and is generally expressed in mid-year dollars of a reference year.

The total capital requirement (TCR) includes all the capital necessary to complete the entire project, including interest during construction (AFUDC, allowance for funds during construction) and owner costs, which include royalties, startup costs, inventory capital *etc*.

4.2 O&M Cost

The major operating and maintenance (O&M) cost consists of fixed costs and variable cost elements as listed in Table 4.2

Table 4.2 MEA O&M cost model parameters and nominal values

O&M Cost Elements	Typical Value
Fixed O&M Costs	
Total Maintenance Cost	2.5% TPC
Maintenance Cost Allocated to Labor ($f_{maintlab}$)	40% of total maint. cost
Admin. & Support Labor Cost (f_{admin})	30% of total labor cost
Operating Labor (N_{labor})	2 jobs/shift
Variable O&M Costs	
Reagent (MEA) Cost	$1200/ tonne
Water Cost	$0.8/ 1000 gallon
Solid Waste Disposal Cost	$175/ tonne waste
CO_2 Transport Cost	$0.02/ tonne CO_2 per km
CO_2 Storage/Disposal Cost	$5/ tonne CO_2

4.2.1. Fixed O&M Costs

The *fixed O&M* (FOM) costs in the model include the costs of maintenance (materials and labor) and labor (operating labor, administrative and support labor). They are estimated on annual basis ($M/yr) as follows:

$$FOM = FOM_{labor} + FOM_{maint} + FOM_{admin}$$

$$FOM_{labor} = labor \times N_{labor} \times 40(hrs/week) \times 52(weeks/yr)$$

$$\text{FOM}_{maint} = \Sigma_i \, (f_{maint})_i \times \text{TPC}_i \quad \text{where } i = \text{process area}$$

$$\text{FOM}_{admin} = f_{admin} \times (\text{FOM}_{labor} + f_{maintlab} \times \text{FOM}_{maint})$$

where,

$labor$ = the hourly wages to the labor (\$/hr)

N_{labor} = number of operating labor required

$(f_{maint})_i$ = total annual maintenance cost expressed as the fraction of the total plant cost (TPC)

f_{admin} = the administrative labor cost expressed as the fraction of the total labor cost

4.2.2. *Variable O&M Costs*

The *variable O&M* (VOM) costs depend on the capacity factor (or load factor) of the plant. They include costs of chemicals consumed (MEA, inhibitor, other reagents such as caustic and activated carbon), utilities (water, steam, power), fuel (natural gas, in case of auxiliary boiler) and services used (waste disposal, CO_2 transport and storage). These quantities are determined in the performance model. The unit cost of each item (e.g., dollars per ton of reagent, or dollars per ton of CO_2 stored) is a parameter specified as a cost input to the model. The total annual cost of each item is then calculated by multiplying the unit cost by the total annual quantity used or consumed. Total annual quantities depend strongly on the plant capacity factor, which is defined as the ratio of total annual generation (kWh per year) to maximum possible generation (which is the product of plant capacity times total hours per year). The plant capacity factor is one of the crucial assumptions that influences the overall economics of the plant, especially in the case of a capital-intensive technology.

The individual components of variable O&M costs are as follows:

Cost of MEA reagent (VOM_{MEA}): The makeup MEA requirement estimated in the performance model is transformed into dollar amount by using the unit cost of MEA, which is a user-controlled cost input variable.

$$VOM_{MEA} = M_{MEA,makeup} \times UC_{MEA} \times HPY$$

where, UC_{MEA} is the unit cost of MEA, and HPY is the equivalent annual hours per year of plant operation at full capacity (*e.g.*, a capacity factor of 75% is equivalent to about 6575 hours per year, assuming an average of 365.25 days per year).

Cost of inhibitor ($VOM_{inhibitor}$): Addition of inhibitor makes it possible to use higher concentrations of MEA sorbent in the system with minimal corrosion problems. Inhibitors are special compounds that come at a cost premium. The cost of inhibitor is estimated as 20% of the cost of MEA.

$$VOM_{inhibitor} = 0.2 \times VOM_{MEA}$$

Cost of other reagents ($VOM_{reagents}$): The cost of other reagents, such as, caustic and activated carbon are also calculated from their physical quantities estimated in the performance model and the unit costs of these reagents.

$$VOMreagents = VOM_{Caustic} + VOM_{act-C}$$

$$= \{(m_{Caustic} \times UC_{Caustic}) + (m_{act-C} \times UC_{act-C})\} \times HPY$$

where $UC_{Caustic}$ and UC_{act-C} are the unit costs of the reagents caustic and activated carbon, respectively.

Cost of waste disposal (VOM_{waste}): Another important variable operating cost item is the cost incurred in disposal of the spent sorbent *i.e.*, the reclaimer waste. The quantity estimated in the performance model is:

$$VOM_{waste} = M_{waste,total} \times UC_{waste} \times HPY$$

where, UC_{waste} is the unit cost of waste disposal for the reclaimer waste.

Cost of CO₂ transport ($VOM_{transport}$): Transportation of CO_2 product is assumed to take place via pipelines. The cost of CO_2 transport is estimated on the basis of two user-specified parameters, *viz.*, transportation distance (TD, in km) and unit cost of transport ($UC_{transport}$, \$/km per tonne CO_2), plus the CO_2 product flow rate (calculated result from performance model).

$$VOM_{transport} = M_{CO2} \times UC_{transport} \times TD \times HPY$$

Cost of CO₂ storage ($VOM_{disposal}$): Depending upon the method of CO_2 disposal or storage, either there may be some revenue generated (as in enhanced oil recovery, or enhanced coal bed methane), or an additional cost (all other disposal methods). The total cost or revenue of CO_2 disposal/ storage is estimated from the unit cost and CO_2 product flow rate (UC_{disp}).

$$VOM_{disposal} = M_{CO2} \times UC_{disp} \times HPY$$

Cost of energy (VOM_{energy}): By default, all energy costs are handled internally in the model by de-rating the overall power plant based on the calculated power requirement. The CO_2 capture unit is charged for the total electricity production foregone because of CO_2 capture and compression ($E_{CO2,\ tot}$).

For power plants with multi-pollutant controls the desire to quantify costs for a single pollutant requires an arbitrary choice of how to charge or allocate certain costs. This is especially relevant for energy-intensive processes like CO_2 capture systems.

The unit cost of electricity (COE_{noctl}) is estimated by the base plant module, or may be overridden by a user-specified value if this energy is assumed to be supplied from an external source. Since energy cost is one of the biggest O&M cost items for the CO_2 unit, the way in which it is accounted for is important when calculating the mitigation cost.

$$VOM_{energy} = E_{CO2,tot} \times HPY \times COE_{noctl}$$

Alternatively, when regeneration steam and additional electricity is provided by an auxiliary NG boiler, the cost of energy is estimated from the total annualized cost of the

new boiler and secondary steam turbine, which takes into account their capital cost requirements and cost of natural gas fuel.

Cost of water (VOM_{water}):   Water is mainly required for process cooling and also as process makeup. Generally this is a minor cost item in the overall plant operation, but it is included over here for the sake of completeness, based on the amount of water needed (M_w) and the unit cost of water (UC_{water}):

$$VOM_{water} = M_w \times UC_{water} \times HPY$$

The total variable O&M (VOM, $/yr) cost is obtained by adding all these costs:

$$VOM = VOM_{MEA} + VOM_{reagents} + VOM_{waste} + VOM_{transport} + VOM_{disposal} +$$
$$VOM_{energy} + VOM_{water}$$

Finally, the total annual O&M cost (TOM, $/yr) may be obtained as:

$$TOM = FOM + VOM$$

4.3 Incremental Cost of Electricity

Once the total capital requirement and the total O&M costs are known, the total annualized cost of the power plant may be estimated as follows:

$$Total\ annual\ revenue\ requirement,\ TRR\ (\$/yr) = (TCR \times CRF) + TOM$$

where, TCR = Total capital requirement of the power plant ($), and

$$CRF = Capital\ recovery\ factor\ (fraction)$$

The capital recovery factor, or fixed charge factor (FCF), is the factor that annualizes the total capital requirement of the plant. It depends on the applicable interest rate (or discount rate) and useful lifetime of the plant.

It can be seen that a higher value of this factor (*e.g.* from assumptions of shorter plant life and/or higher interest rate) leads to a higher overall annualized cost. Hence the assumption about this factor (a user-defined parameter) is crucial in the overall economics of the plant.

The IECM framework calculates the cost of electricity (COE) for the overall power plant by dividing the total annualized plant cost ($/yr) by the net electricity generated (kWh/yr). Results are expressed in units of $/MWh (equivalent to mills/kWh).

Cost of electricity, COE ($/MWh) $= TRR / (MW_{net}*HPY)$

where, TRR = Total annual revenue requirement ($/yr)

MW_{net} = Net power generation capacity (MW)

HPY = Annual hours of operation (hrs/yr)

Note that the COE includes the cost of all environmental control systems, not just the CO_2 control system. Thus, by running two scenarios of the power plant model, one without CO_2 capture (reference plant) and one with CO_2 capture (CO_2 capture plant), we obtain the incremental capital costs, O&M costs, and total annualized costs attributed to CO_2 capture. The addition of a CO_2 capture and sequestration system increases the COE for the plant; this incremental cost of electricity is attributed to CO_2 control.

4.4 Cost of CO_2 Avoidance

Analysts often express the cost of an environmental control system in terms of the cost per unit mass of pollutant removed. However, for energy-intensive CO_2 controls there is a big difference between the cost per tonne CO_2 "removed" and the cost per tonne "avoided" based on *net* plant capacity. Since the purpose of adding a CO_2 unit is to reduce the CO_2 emissions per net kWh delivered, the "cost of CO_2 avoidance" is the economic indicator that is widely used in this field. It can be calculated as:

$$Cost\ of\ CO_2\ Avoided\ (\$/tonne) = \frac{(\$/kWh)_{after} - (\$/kWh)_{before}}{(tonne\ CO_2/kWh)_{before} - (tonne\ CO_2/kWh)_{after}} \quad (4\text{-}1)$$

In contrast, the cost per unit of CO_2 removed or captured is simply the additional expenses incurred in the capture of CO_2, divided by the total quantity of CO_2 captured.

This can be calculated as the difference between the total annualized cost of the plant (TRR, M$/yr) with and without CO_2 control, divided by the total quantity of CO_2 captured (tonne CO_2/ yr), with the net power generated by the two plants remaining the same. Hence, the CO_2 avoidance cost, as calculated in equation *4-1*, is quite different from the cost per unit of CO_2 captured. In case of CO_2 control using an energy-intensive technology like amine-scrubbing, the *cost of CO_2 avoidance* may be substantially higher than *cost of CO_2 capture*.

The cost of CO_2 avoidance has another interpretation in terms of a carbon-tax scenario. Consider a scenario where a power plant must pay a fixed tax (C-tax) that is proportional to its CO_2 emissions. Now consider a reference plant (that does not control its CO_2 emissions) and the same plant with CO_2 capture (*e.g.*, 90% of its CO_2 emissions). The COE for the reference plant is initially lower, but increases much faster as compared to the COE for the capture plant, in response to increasing levels of the C-tax. Eventually, a C-tax level is reached where the COE for both the plants are the same (see Figure 4.1).
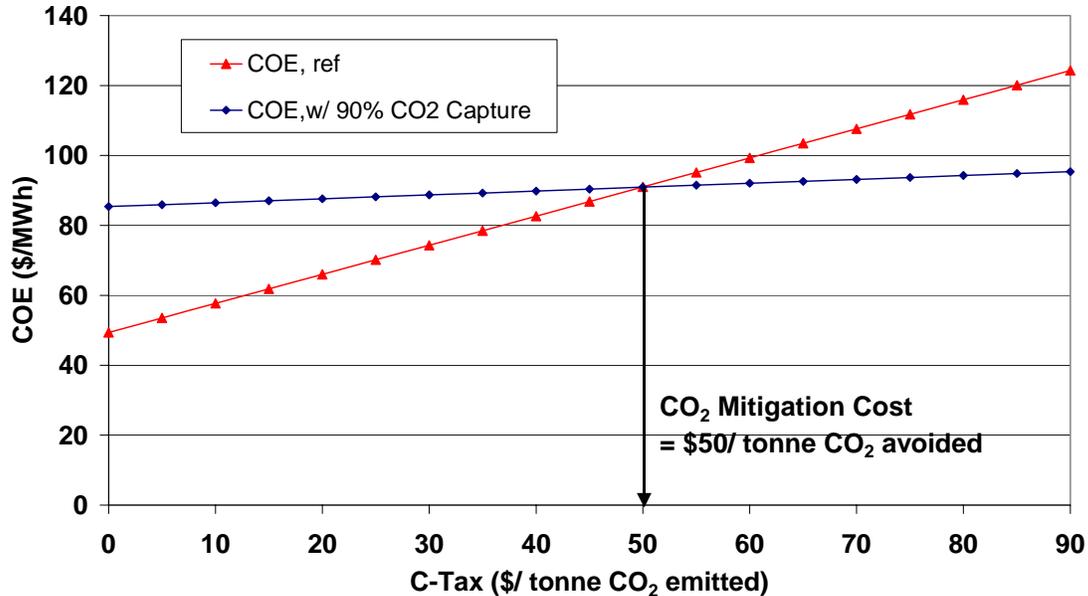
Figure 4.1. Cost of electricity (COE) as a function of carbon-tax

At this C-tax level, the power plant might be indifferent between paying the C-tax or incurring the cost of the CO_2 capture unit. The cost of CO_2 avoidance is this C-tax level,

where the COE for the reference plant and capture plant become equal. Note that this comparison assumes that both the reference plant and capture plant operate at the same annual capacity factor. In practice, load factors may vary for plants with and without emission controls in the context of a specific regional situation (Johnson 2002).

4.5 Cost of Electricity (COE) as a Function of Carbon Tax

As mentioned earlier, the IECM modeling framework has probabilistic capabilities that allow model inputs to be represented by probability distributions. In addition to uncertainties or variability in performance model parameters, there is also uncertainty and/ or variability in cost model parameters.

Table 4.3 lists the uncertainty distributions developed for the cost model parameters based on the current literature on amine-based (MEA) systems, the evolving nature of the technology, and views expressed by professionals working in this field. These distributions reflect both uncertainty and variability in the cost of these systems. Additional details are presented in *Appendix B* and *Appendix C*.

The last three parameters listed in Table 4.3 reflect the cost of CO_2 transport and storage. The storage cost or revenue expected from CO_2 product stream is the area of largest uncertainty. Various factors including the possibility of government regulation of CO_2 emission, the evolution of national and/or international markets for CO_2, and site-specific conditions will significantly influence the cost (or credits) for CO_2 disposal. The distributions in Table 4.3 reflect the variety of options and costs commonly cited in the current literature.

Other parameters that directly contribute to the overall distribution of cost estimates for this technology include fuel prices, plant capacity factor, and fixed charge factor or capital recovery factor. These are the parameters that essentially belong to the reference plant (base plant) itself, and have nothing to do with amine-based CO_2 capture. So, in order to assess the uncertainty associated with the amine-based CO_2 capture technology, distributions on these parameters as well for the CO_2 transport and disposal cost parameters should not be used.

Table 4.3. Amine system cost model parameters and uncertainties

Cost Parameter	Units	Nominal Value	Unc. Representation (Distribution Function)
Process facilities	M$	calc	*Normal(1.0,0.1)
Engineering and home office	% PFC	7	Triangular (5,7,10)
General facilities	% PFC	10	Triangular (5,10,15)
Project contingency	% PFC	15	Triangular (10,15,20)
Process contingency	% PFC	5	Triangular (2,5,10)
Royalty fees	% PFC	0.5	Triangular (0,0.5,0.5)
Startup cost	Month of TOM	1	Triangular(0.5,1,1)
Inventory cost	% TPC	0.5	Triangular(0.4,0.5,0.6)
Total maintenance cost	% TPC	2.5	Triangular(1,2.5,5)
Operating labor	Jobs/shift	2	Triangular(1,2,3)
MEA cost	$/ tonne MEA	1200	Uniform (1100,1300)
Waste disposal cost	$/ tonne waste	175	*Normal(1.0,0.1)
CO_2 transport cost	$/ tonne CO_2/ km	0.02	Triangular (0.004,0.02,0.08)
CO_2 storage/disposal cost	$/ tonne CO_2	5	-
Revenue from CO_2 usage in EOR or ECBM[b]	$/ tonne CO_2	-18[a]	-

*This function is used as a multiplier to the nominal value.

[a] The negative sign indicates that it is a revenue rather than cost to the system.

[b] Individual scenarios may be run with the assumptions about revenue generation from CO_2 product usage in EOR/ECBM applications, instead of additional cost for CO_2 storage/ disposal. Alternatively, a chance function may be used that will assign either cost or revenue for CO_2 product stream during each individual run.

References (Chapter 4)

Chapel, D., J. Ernst, et al. (1999). Recovery of CO2 from flue gases: commercial trends (paper no. 340). presented at the Canadian Society of Chemical Engineers Annual Meeting, 4-6 October, Saskatoon, Saskatchewan, Canada.

Desideri, U. and A. Paolucci (1999). "Performance modelling of a carbon dioxide removal system for power plants." Energy Conversion and Management **40**: 1899-1915.

Fluor (1998). Evaluation of technologies for the removal of CO2 from coal fired power plant flue gas, a report prepared by Fluor Daniel Canada Inc. for TransAlta, November 1998.

Hendriks, C. (1994). Carbon Dioxide Removal from Coal-fired Power Plants. The Netherlands, Kluwer Academic Publishers.

Jeremy, D. and H. J. Herzog (2000). The cost of carbon capture. Fifth Greenhouse Gas Control Technologies, Cairns, Australia.

Johnson, T. L. (2002). Electricity without carbon dioxide: assessing the role of carbon capture and sequestration in US electric markets. Department of Engineering and Public Policy. Pittsburgh, PA 15213, Carnegie Mellon University: 247.

Leci, C. L. (1996). "Financial implications on power generation costs resulting from the parasitic effect of CO2 capture using liquid scrubbing technology form power station flue gases." Energy Conversion and Management **37(6-8)**: 915-921.

Mariz, C. L. (1998). "Carbon dioxide recovery: large scale design trends." The Journal of Canadian Petroleum Technology **37**(7).

Peters, M. S. and K. D. Timmerhaus (1991). Plant design and economics. New York, NY, McGraw-Hill.

Rubin, E. S., J. R. Kalagnnam, et al. (1997). "Integrated environmental control modeling of coal-fired power systems." Journal of Air and& Water Management Association **47**: 1180-1188.

Simbeck, D. (1998). A portfolio selection approach for power plant CO2 capture, separation and R & D options. Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September, Interlaken, Switzerland, Elsevier Science Ltd.

Simbeck, D. (2002). "Personal communication."

Simbeck, D. and M. McDonald (2000). Existing coal power plant retrofit CO_2 control options analysis. Sixth Greenhouse Gas Control Technologies, Cairns, Australia.

Smelser, S. C., R. M. Stock, et al. (1991). Engineering and economic evaluation of CO2 removal from fossil-fuel-fired power plants, EPRI IE-7365, Volume 1, Project 2999-10, a research project final report prepared by Fluor Daniel Inc., for EPRI and IEA.

TAG (1999). Technical Assessment Guide. Palo Alto, CA., EPRI TR 102276, EPRI.

5. MODEL APPLICATIONS

One of the major objectives behind building the IECM-CS model is to provide an analytical tool to compare various technological options for controlling CO_2 emissions from fossil-fuel based power plants on a consistent basis. Such assessments of the technical, environmental, and economic impacts of including CO_2 control systems in power plant applications are important for greenhouse gas control policy analysis. Policy aspects of controlling CO_2 emissions from power plants might include questions such as:

1) Within the spectrum of fossil-fuel based power plants, are there any particular categories that are better candidates for applying this technology?

2) What levels of emission reduction are possible? Are there any secondary environmental benefits associated with this process? Are there any potential environmental concerns? Are there any local pollution control problems or benefits arising from this process?

3) How does CO_2 capture affect the overall performance (and cost) of the power plant?

4) How much would it cost in terms of capital requirement, incremental cost of electricity and cost per unit of CO_2 avoided? What are the uncertainties associated with these estimates?

5) In case of conventional combustion-based power plants, is it cheaper to control CO_2 emissions from a coal plant than from a gas plant?

6) Is it feasible to retrofit the existing power plants with this technology? If yes, are there any crucial issues of system integration in this case? How would the costs for a retrofit application compare to that for a greenfield case?

7) What are the key factors that affect these costs? Is there any scope for improvement in this technology through targeted R&D efforts? What are the possible cost reductions that could be realized through such improvements?

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The IECM-CS model may be used to generate various scenarios that can provide useful insights into these questions. It may be noted that all these questions address issues related to the techno-economic and environmental feasibility of CO_2 capture and separation alone. There is another whole set of questions related to the issues concerning the large-scale transport and storage/disposal of concentrated CO_2 stream generated in this process, *viz.* technical feasibility, long-term environmental impacts, public perceptions, potential risks, uncertainties in the cost estimates *etc*. However, these are beyond the scope of the present study.

This chapter will address such questions by means of specific case studies of greenfield and retrofit applications of this technology presented in subsequent sections. The last question will be addressed later in the next chapter.

5.1. Case Study of a New Coal-fired Power Plant

Let us consider the case of a new conventional coal-fired power plant, and impact of adding a post-combustion amine-based CO_2 capture unit to it. The basic assumptions and input parameters are listed in Table 5.1.

Table 5.1 Design parameters for case study of new pulverized coal plant

Parameter	Value	Parameter	Value
Gross plant size (MW)	500, 670	Emission standards	2000 NSPS[d]
Base plant steam cycle type	SC[a]	NO$_x$ Controls	LNB[e] +SCR[f]
Gross plant heat rate (kJ/kWh)	8359[a]	Particulate Control	ESP[g]
Plant capacity factor (%)	75	SO$_2$ Control	FGD[h]
Coal characteristics		CO$_2$ Control	MEA[i]
Rank	Sub-bit.	CO$_2$ capture efficiency (%)	90
HHV (kJ/kg)	19,346	CO$_2$ product pressure (kPa)	13,790[j]
% S	0.48	Distance to storage (km)	165
% C	47.85		
Mine-mouth cost ($/tonne)	13.73	Cost year basis (constant dollars)	2000
Delivered cost ($/tonne)	23.19[c]	Fixed charge factor	0.15[k]

[a]Nominal case is a super-critical unit. Unc. = Uniform(8167, 8560) ; [b]Unc = Triangular(65,75,85); [c]Unc = Triangular(15.94,23.19,26.81); [d]NO$_x$ = 65 ng/J, PM = 13 ng/J, SO$_2$ = 70% removal (upgraded to 99% with MEA systems); [e]LNB = Low- NO$_x$ Burner; [f]SCR = Selective Catalytic Reduction; [g]ESP = Electrostatic Precipitator; [h]FGD = Flue Gas Desulfurization; [i]MEA = Monoethanolamine system;
[j]Unc. = Triangular(7580,13790,15160); [k]Corresponds to a 30-year plant lifetime with a 14.8% real interest rate (or, a 20-year life with 13.9% interest); Unc = Uniform(0.10,0.20)

The reference plant (without CO_2 control unit) is New Source Performance Standard (NSPS) compliant coal-fired power plant and the complete plant with multi-pollutant environmental controls is simulated using IECM. The FGD system is assumed to remove 95% of the SO_2 emissions, reflecting the best available technology, yielding lower emissions than what is required under the present NSPS. This assumption is crucial as the "CO_2 avoidance cost" depends on the reference plant design as well. Wyoming Powder River Basin coal has been assumed to be used. The model outputs are presented later in Table 5.2 in comparison with the estimates for the CO_2 capture plant.

In case of the CO_2 capture plant, the following changes have been assumed as compared to the reference plant:

1) Air leakage has been reduced to 10% (reference plant uses the default value which is about 19%).

2) The FGD system is assumed to be further upgraded to 99% SO_2 removal efficiency.

3) MEA-based CO_2 capture system including CO_2 product compression, transport and storage has been added.

The values for other parameters are listed in Tables 3.3, 4.1 and 4.2 in the previous chapters.

Since we know that the CO_2 capture system is going to consume a lot of energy, we will be comparing two alternative cases of CO_2 capture plant:

A) Using the same coal input (same MW_{gross} size), so that the CO_2 capture plant will end up with lower net power generation, called as *Case-A*.

B) Starting with a bigger base plant (higher MW_{gross} size, larger coal input), so that the CO_2 capture plant produces the same net power as generated by the reference plant. The exact size of this plant has to be found by trial-and-error method, and this case will be referred as *Case-B*. From a practical point of view, this is a more useful case to consider, since the

ultimate objective of a power plant is to generate (a certain amount of) electricity. Hence, this case will be used as a base case in further analysis.

5.1.1. Deterministic Results

We examine the model first for the reference power plant (without CO_2 capture) and then for the same plant including CO_2 capture (case A) and for a similar but bigger plant including CO_2 capture (case B). Table 5.2 quantifies the impact of the CO_2 unit on plant performance, plant discharges and plant costs. More detailed results (including those for Case-A) have been provided in *Appendix D*. Here we refer to the results from the case-B capture plant in comparison to the reference plant.

The CO_2 capture system is energy intensive, and puts a large parasitic load on the base plant. Hence, the net heat rate of the capture plant (11,550 Btu/kWh) is higher than that of the reference plant (8,657 Btu/kWh). Energy requirements consume about 22% of gross plant capacity, mostly for sorbent regeneration (54%) and CO_2 product compression (36%). Sorbent circulation and fan power account for the remaining share (10%) of the total energy consumption of CO_2 capture unit.

It may be observed that the new CO_2 capture plant is significantly bigger (+34%) than the reference plant. This implies that the capture plant has greater environmental impacts in terms of higher emissions of solid wastes (bottom ash, fly ash, FGD waste etc.) and ammonia in absolute quantities. This plant uses much more coal, hence it is also responsible for the externalities associated with extraction and transport of this extra coal, although these environmental impacts are beyond the scope of this study. It must be noted that the capture plant has lower SO_x, NO_x, and particulate emissions, in addition to lower CO_2 emissions. However, it has higher emission rate of NO_x (mass of NO_x emitted per unit of power generated, 0.78 g/kWh) as compared to the reference plant (0.59 g/kWh). This happens because of the large energy penalty of the CO_2 capture system. Finally, the capture plant generates two new discharge streams, *viz.* spent sorbent (which is treated as a hazardous waste) and a large concentrated CO_2 stream. It is assumed that the CO_2 product is transported via pipelines to the final storage or disposal site.

Table 5.2. Case study results for new pulverized coal plants

Parameter	Reference Plant	Capture Plant (B)	% change
Gross capacity (MW)	500	670	+34%
FGD efficiency (%)	95	99	
CO_2 capture efficiency (%)	-	90	
Net power generation (MW)	458	458	NO CHANGE
Net cycle heat rate (Btu/kWh)	8,657	11,550	+33%
Coal consumption (tonne/hr)	215.5	287.5	+33%
Limestone consumption (tonne/hr)	2.3	3.2	+39%
CO_2 emission (tonne/ hr)	381.3	50.9	-87%
NO_x emission (kg/ hr)	269.5	355.2	+32%
SO_x emission (kg/ hr)	76.4	0.14	-99.8%
NH_3 emission (kg/ hr)	5.35	40.0	+648%
CO_2 emission (gCO_2/ kWh)	833.3	111.2	-87%
NO_x emission (gNO_x/ kWh)	0.59	0.78	+32%
SO_x emission (gNO_x/ kWh)	0.17	0.0003	-99.8%
FGD waste (tonne/hr)	4.2	5.8	+39%
Spent MEA sorbent (tonne/hr)	-	1.4	
CO_2 product (tonne/ hr)	-	458	
TCR (M$)	616	963	+56%
TCR ($/ kW)	1345	2104	+56%
COE ($/ MWh)	49.4	85.7	+73%
$/ tonne CO_2 avoided	-	50.2	
$/ tonne CO_2 delivered	-	36.2	

Because the CO_2 capture plant is a bigger plant (670 MW gross), the total plant capital cost increases from \$616M (for the reference plant) to \$963M. It maybe noted that this plant benefits from economies of scale. This is evident from the fact that the total capital requirement for case A (500 MW_{gross} with CO_2 capture) is \$760M (\$2228/kW). In terms of operational costs, the steam and power cost and CO_2 transport and storage/ disposal cost are two most important cost items, and contribute to about 34% and 23% to the total

annualized cost of CO_2 capture. MEA makeup requirements contribute to about 8% of this cost. Figure 5.1 shows the relative contributions from various cost areas in the overall annualized cost of CO_2 capture unit.

The presence of SO_2 impurities can substantially increase the reagent costs, so the cost of CO_2 avoidance is sensitive to the SO_2 removal efficiency of the FGD unit located upstream of the MEA system. Tradeoff studies showed that overall plant costs were minimized by reducing the SO_2 concentration to about 10 ppm inlet to the MEA absorber. Therefore, in this case study the SO_2 removal system was upgraded to 99% removal with CO_2 controls, as against 95% SO_2 removal in the reference case. The costs of this upgrade were charged to the CO_2 unit.

Figure 5.2 shows graphically the relationship of electricity cost to CO_2 emissions for the two cases; the slope of the line connecting the two points represents the cost of CO_2 avoided, which is \$50/tonne for this case. Most of this cost (77%) is associated with the CO_2 capture process (including product compression); CO_2 transport (9%) and storage (14%) account for the remainder.
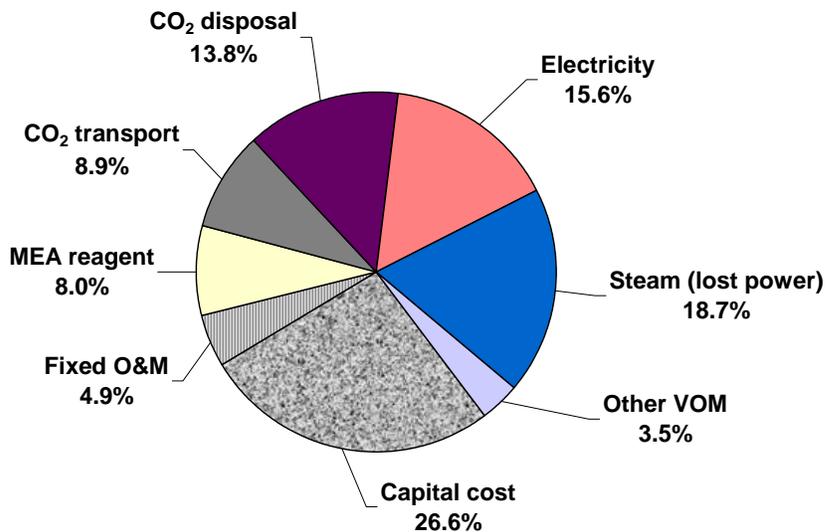
Figure 5.1. Breakup of annualized cost of CO_2 capture unit
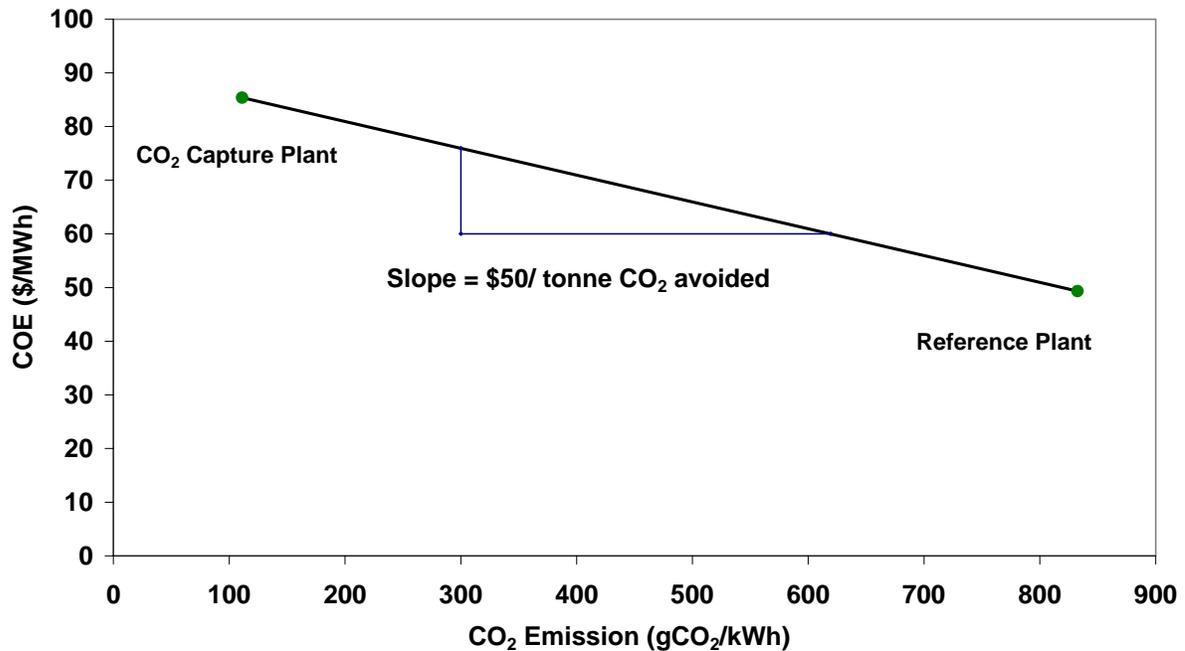
Figure 5.2. Estimation of CO$_2$ avoidance cost for a new low-S coal plant

It may be noted that the CO$_2$ avoidance cost (\$50/ tonne CO$_2$ avoided) is much higher than the CO$_2$ capture cost (\$36/ tonne CO$_2$ delivered).

5.1.2. Probabilistic Results

How much is it likely to cost to avoid CO$_2$ emissions using MEA-based CO$_2$ capture systems from a new pulverized coal (PC) plant in the US today? The question sounds quite simple; however one needs to make a lot of assumptions in order to answer it. Different assumptions about power plant design, coal properties, plant operation and CO$_2$ capture system design can have a significant effect on CO$_2$ mitigation costs, and are responsible for many of the cost differences found in the literature. To systematically characterize the effects of different assumptions, parameters that vary across design studies are represented here by probability distributions, with values selected based on the current literature, experts' opinions and (in some cases) my own judgment. Note that the term "uncertainty" is used loosely here to include parameter variability. More discussion about this issue may be found in *Appendix B* and *Appendix C.*

We can set the characteristics of the PC plant (*e.g.*, configuration, heat rate, capacity factor) and that of the fuel used (*e.g.*, coal properties, coal price) in order to define the reference case, *i.e.* power plant without CO$_2$ capture. Now we want to estimate the cost of CO$_2$ avoidance for this plant using MEA-based CO$_2$ capture system. Again, there are many parameters that need to be defined for this system. If we use a

single value for each of them, we are looking at a unique system configuration with well defined design variables and operating conditions. Thus we get a deterministic cost estimate. However, if we are considering the whole population of potential configurations of MEA-based CO_2 capture systems, most of those parameters would rather be represented as ranges of values (probability distributions) in place of the precise values. Various combinations of these parameter values then represent the set of possible configurations of a new commercial CO_2 capture system if built today.

Figure 5.3 first shows the effect of considering uncertainties and design variability only in the performance parameters of the amine system (from Table 3.3). The resulting distribution for cost of CO_2 avoidance has a 95-percentile range of \$34-54/ tonne CO_2 avoided. The main contributors to this range are the lean sorbent CO_2 loading, sorbent concentration (both of which determine the reboiler duty), and equivalent electrical penalty of the regeneration steam requirement. It may be noted that most of the distributions for the input parameters are triangular, which is also reflected in this result to a certain extent. The tapering tails with low probability generally imply some combinations of design conditions and performance parameter values that are either too idealistic (*e.g.* high sorbent concentration, high lean loading) or too conservative (*e.g.* lower sorbent concentration, lower lean loading). For example, as explained earlier, amine systems with high sorbent concentration and high lean loading would have to deal with more severe corrosion problems, making them less practical at this time. Such interactions are not well quantified and, hence, not reflected in the current model. However, R&D efforts could be targeted to make these potentially "better performing, low cost systems" practical in future. The median value (and its vicinity) represents a more realistic and near-optimal set of design conditions for current commercial systems.

Figure 5.4 shows the effect of uncertainty in CO_2 cost model parameters, excluding the costs of the CO_2 pipeline transport and geologic storage (see Table 4.3). Also included is the distribution obtained by considering uncertainties only in the cost parameters of the capture system. It shows that the uncertainties in the cost parameters of amine system have a very small effect on the overall distribution of the cost of CO_2 avoidance, which is dominated by the variability and uncertainty in design and performance variables.

Next, we consider the effect of assumptions about the cost of transport and storage/ disposal of CO_2 product on the distribution of the cost of CO_2 avoidance developed in Figure 5.4. Figure 5.5 shows the effect of considering costs of different options to store or dispose the CO_2 product stream. Curve A(0) represents a case that assumes zero storage cost. Next, curve B(5) is the case where storage of CO_2 stream (*e.g.*, in a geologic formation such as underground saline reservoir) incurs a cost of $5/ tonne CO_2. Finally, curve C(-18) shows a case that assumes a revenue of $18/ tonne CO_2 generated through EOR activity, represented as negative cost. The assumption about the cost of disposal of CO_2 product has a significant impact on the distribution of CO_2 avoidance cost. The 95-percentile ranges for these 3 cases are $36-60 (case A(0)), $43-68 (case B(5)) and $12-31 (case C(-18)) per tonne CO_2 avoided. These may be compared to the 95-percentile range of $34-54/ tonne CO_2 avoided for the case presented earlier in Figure 5.4. So the option B(5), that assumes an additional cost of $5/tonne CO_2 for storage, not only increases the overall cost of CO_2 avoidance (shifting the curve to the right), but also increases the 95-percentile range. On the other hand, option C(-18) leads to lower overall cost of CO_2 avoidance (shifting the curve to the left) as well as slightly narrower range.

Figure 5.6 shows the effect of additional uncertainties (or variability) for the assumed power plant efficiency (heat rate), plant capacity factor, fixed charge factor, and coal price (see Table 3.1). These parameter values are identical for the reference plant and capture plant. Here we are now looking at the probability distribution of CO_2 mitigation cost for a general population of pulverized coal plants using amine systems. The COE of each plant is strongly affected by assumptions about the base plant parameters. Also, the range in the CO_2 avoidance cost increases for all the three options A(0), B(5) and C(-18) to $33-63/tonne, $40-71 and $8-36/tonne, respectively.
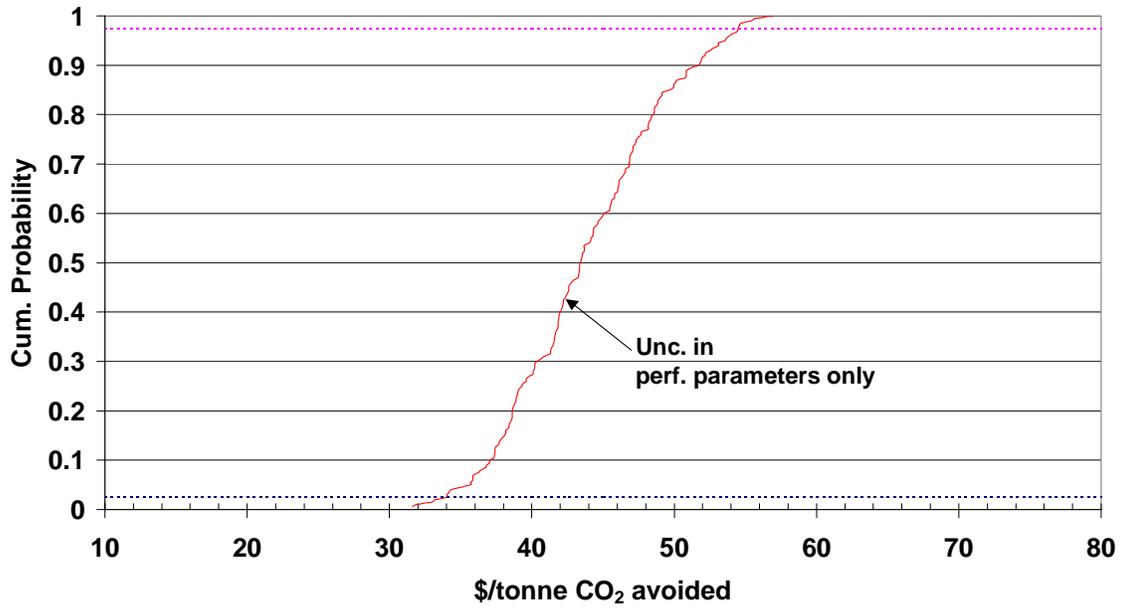
Figure 5.3. Probability distribution for the CO_2 avoidance cost, with only the uncertainties in performance parameters of amine system considered.
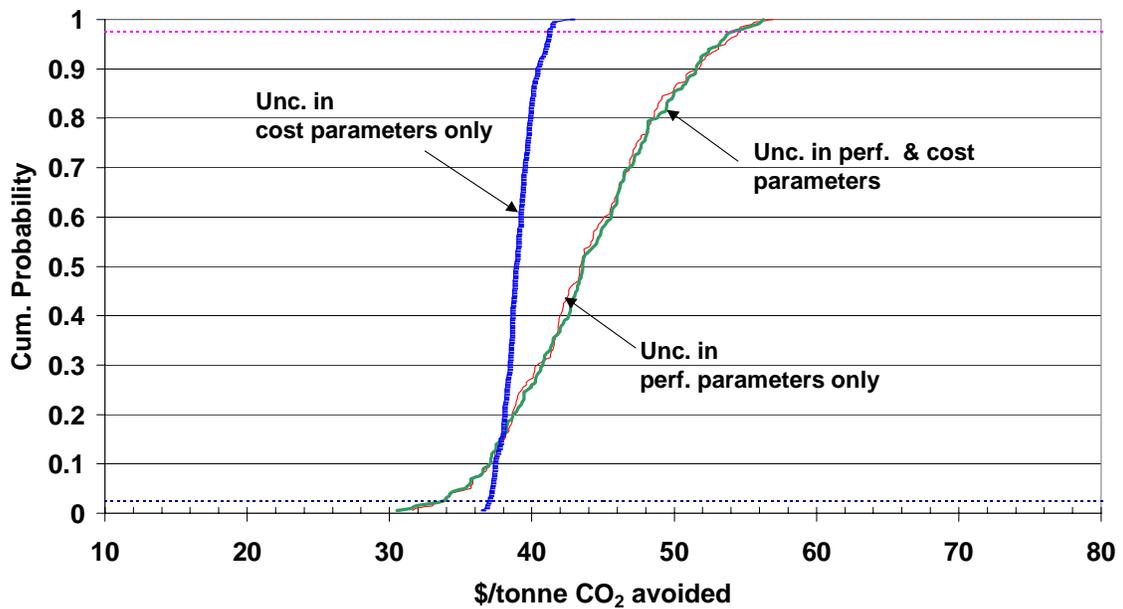
Figure 5.4. Probability distribution for the CO_2 avoidance cost, with the uncertainties in performance and cost parameters of amine system considered.
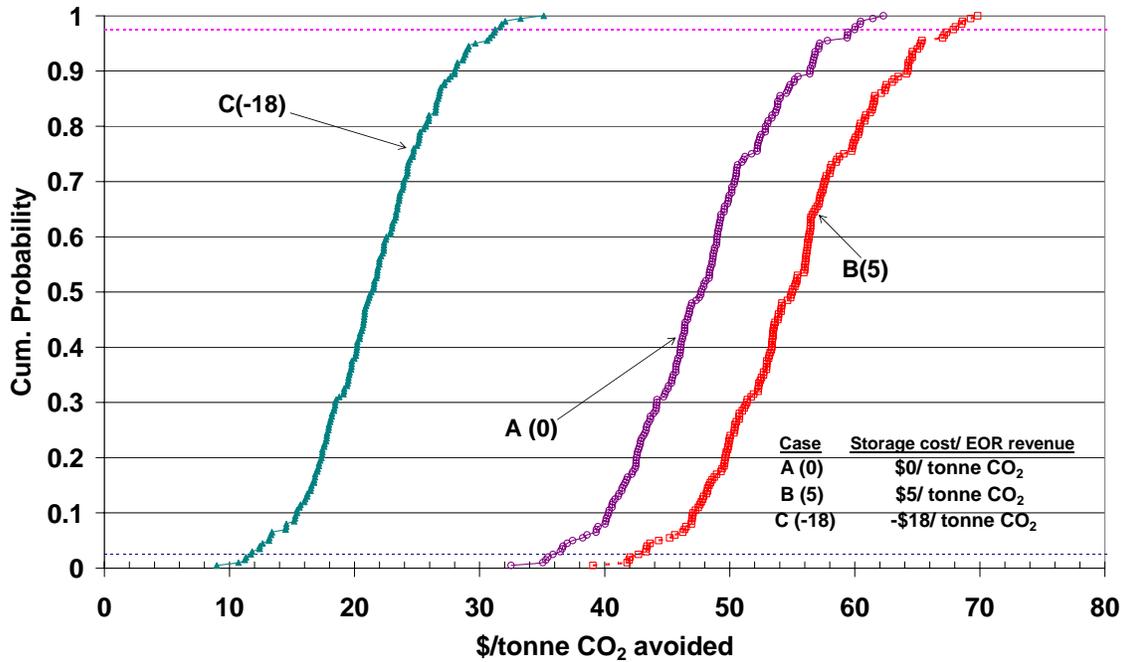
Figure 5.5. Effect of considering three different options for CO$_2$ storage/disposal, assuming either no cost (A(0)), storage cost of $5/tonne CO$_2$ (B(5)), or revenue of $18/ tonne CO$_2$ from EOR (C(-18)), on the distribution of the CO$_2$ avoidance cost (with uncertainties in amine system parameters).
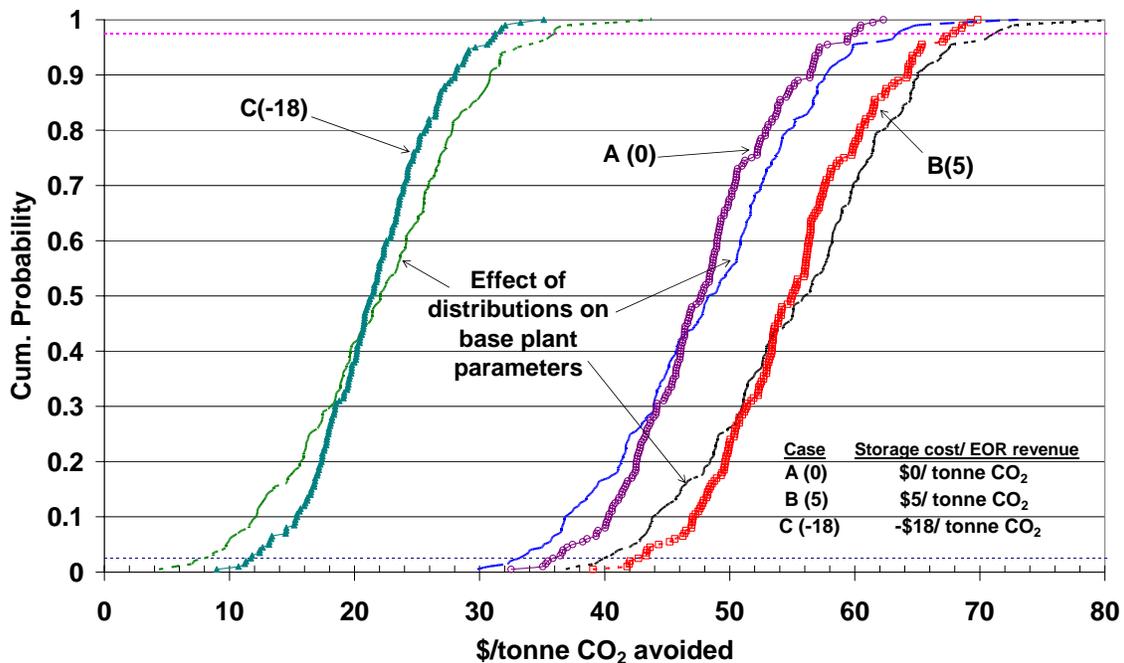
Figure 5.6. Effect of considering uncertainty in base plant parameters (viz. base plant gross cycle heat rate, plant capacity factor, fixed charge factor and coal price) on the distribution of CO$_2$ avoidance cost derived in figure 5.5.

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5.2. Applications to CO_2 Retrofits

As discussed earlier, a large number of existing coal-fired power plants may be candidates for CO_2 retrofits under a sufficiently stringent climate policy. Because of multi-pollutant interactions, the cost of CO_2 mitigation will be affected by policies for other pollutants, especially SO_2. Considering that most (~ 70%) of the current coal-based capacity in the U.S. does not have SO_2 scrubbers, several cases were modeled to analyze the impacts of adding post-combustion CO_2 capture systems to existing coal plants (see Table 5.3).

These cases are based on the analysis of the existing US coal plants database that includes details about 2294 units in as many as 85 fields including plant location, heat rate, plant capacity factor, cola characteristics, and environmental controls (NETL 2002). Two sets of power plants have been identified, *viz.* those without any sulfur controls and those with wet scrubbers (FGD systems) for SO_2 emission control. Some of the key distinguishing features of these sets have been listed in Table 5.4.

Table 5.3. Scenarios for retrofit studies of existing coal-fired power plants

Case	Base plant (PC+ESP)	SO$_2$ Capture System (FGD)	CO$_2$ Capture System (MEA)	Auxiliary NG Boiler	Sec. Steam Turbine
Reference plant without any sulfur control (R)					
R	✔	-	-	-	-
A1	✔	-	New	-	-
A2	✔	-	New	New	-
A3	✔	-	New	New	New
B1	✔	New, 99%	New	-	-
B2	✔	New, 99%	New	New	-
B3	✔	New, 99%	New	New	New
Reference plant with FGD system (Rf)					
Rf	✔	✔	-	-	-
C1	✔	✔	New	-	-
C2	✔	✔	New	New	-
C3	✔	✔	New	New	New
D1	✔	Upgrade, 99%	New	-	-
D2	✔	Upgrade, 99%	New	New	-
D3	✔	Upgrade, 99%	New	New	New

Table 5.4 Typical key features of existing coal-fired power plants in U.S.

Characteristic	Plants without any sulfur capture	Plants with FGD systems
No. of units	946	109
Total nameplate capacity	232 GW	51 GW
Plant size (gross)	540 MW	441 MW
Plant capacity factor	73%	76%
Net plant heat rate	10,150 Btu/kWh	10,530 Btu/kWh
Sulfur content in coal	~ 0.7% w/w	~ 1.1% w/w

Table 5.3 lists the various cases modeled using IECM-CS. The first set uses a typical plant without any sulfur control as the reference plant (R). It complies with federal standards for plants constructed before July 1978. Here, one needs to install a sulfur scrubber prior to the amine system during CO_2 retrofitting as per the process requirements for an amine system (B1-B3). A case where only an amine-system is added to the reference plant (A1-A3) is without any practical relevance and is included only for the sake of completeness of this analysis, and it reinstates the importance of SO_2 control as a pre-requisite for CO_2 control using amine system.

The second set of cases uses a typical plant already equipped with an FGD system that meets federal New Source Performance Standards (NSPS) (which in this case is 73% SO_2 removal) as reference (Rf). Here, the SO_2 removal efficiency of these systems are just enough to meet the existing emission standards (CFR 1999), however they do emit a substantial amount of SO_2 (about 390 ppm) that maybe undesirable for amine-based CO_2 capture system. This plant has the option of upgrading the existing SO_2 scrubber in addition to installing an MEA system. It would be preferable to upgrade the existing FGD system to 99% removal efficiency. Hence, CO_2 retrofit options here include 2 alternatives, *viz.* adding only amine system (C1-C3) and adding amine system with

upgraded FGD system (D1-D3). The reference plants (R & Rf) for these two sets of CO_2 capture plant cases are based on the characteristics presented in Table 5.4.

For each CO_2 retrofit case, there are three potential configurations in terms of the mode of supplying energy required for the amine based CO_2 capture system.

1) Base plant derating: Regeneration steam is extracted from the existing steam cycle of the base plant. Also the electrical energy for pumps, fans and compressors is obtained from the base plant. Hence, the net output of this plant is substantially lower than that of the reference plant.

2) Auxiliary NG boiler: Here, the regeneration steam is provided by adding a new NG-fired boiler so as to avoid steam extraction and subsequent derating of the existing base plant. However, the electrical energy requirement of CO_2 capture system is fulfilled through the base plant, and hence the overall output of the capture plant is lower than reference case. It maybe noted that adding a new NG-fired boiler to produce only low-pressure steam for sorbent regeneration is not a very practical approach, and is included here only for the sake of completeness of this analysis.

3) Auxiliary NG boiler plus secondary steam turbine: In this case, an auxiliary NG-fired boiler is installed to generate high pressure steam that can be passed through a turbine to generate some electricity. The exhaust from the steam turbine is then used to regenerate the sorbent. The electricity generated by this unit is often more than the electrical energy requirement of the CO_2 capture unit, and hence the overall power generation of the capture plant may exceed that of the reference plant.

In contrast to the new plant analysis presented earlier, the retrofit scenarios assume, as a bounding case, that all existing capital equipment has been fully amortized. The cost of electricity is then determined only by plant O&M costs, plus any new capital expenditures for CO_2 control. The latter includes the costs of any new sulfur removal systems that are installed to minimize the cost of CO_2 avoidance. We may also credit the

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CO$_2$ unit for the market value of additional SO$_2$ allowances that are generated and traded under the national acid rain control program. The remaining plant life is assumed to be at least 15 years. The base capital cost of the CO$_2$ capture unit is the same as that for a greenfield site, but we then consider an additional retrofit cost premium to account for site-specific retrofit difficulties. We also assume higher plant heat rates and energy penalties in providing steam to the MEA system.

5.2.1. *Results for plants without any sulfur control*

The results from the case study of CO$_2$ retrofitting of existing PC plants without any sulfur control have been summarized in Table 5.5. The coal used in this case study was Appalachian low-S bituminous coal with 0.64% sulfur content and heating value of 13,080 Btu/lb. The delivered price of this coal was assumed to be $37.1/tonne which is about $0.96/ MMBtu.

As noted earlier (Rao and Rubin 2002), we find that addition of FGD system along with MEA-based CO$_2$ capture system reduces the overall cost. Cost of electricity and that of CO$_2$ avoidance are substantially lower (almost 50% or less) in cases B1-B3 (in which FGD system was also installed) as compared to cases A1-A3 (in which only MEA-based CO$_2$ capture system was retrofitted). In cases A1 and B1 where steam and electricity required for CO$_2$ capture system is extracted from the base plant, the net plant output goes down to about 55-60% of the original plant output. This may have serious implications in terms of capacity planning and dispatch. On the other hand, option B3, which seems to be the most economic way (based on COE) to retrofit an existing PC plant with CO$_2$ capture system, leads to additional power generation and is expected to have a matching demand. Although the CO$_2$ capture system is designed to capture 90% of the CO$_2$ emissions generated by the reference plant, the options using auxiliary NG-fired boilers end up avoiding only about 60-65% of the emissions. In case of a very stringent climate policy, even the CO$_2$ emissions from the newly added NG-fired boilers may have to be captured.

Again, it must be noted that these results are sensitive to the assumptions made about the depreciation of the reference plant, retrofit factors in capital cost estimation, and the price

of natural gas. Table 5.6 shows the effect of some of these factors on the costs in cases B1 and B3.

Table 5.5. Case study results for plants with FGD systems

	R	A1	A2	A3	B1	B2	B3
Gross capacity (MW)	540	540	540	540	540	540	540
Net power generation (MW)	507.3	320.2	442.2	580.7	309.6	427.5	561.4
Net cycle heat rate (Btu/kWh)	10,150	16,010	16,810	13,790	16,560	17,210	14,090
Coal consumption (tonne/hr)	178.5	177.7	177.7	177.7	177.7	177.7	177.7
NG consumption (tonne/hr)	-	-	45.96	57.44	-	44.43	55.54
CO_2 emission (tonne/ hr)	469.3	46.7	170.4	201.3	46.9	166.5	196.3
NO_x emission (tonne/ hr)	1.34	1.32	1.58	1.65	1.32	1.57	1.64
SO_x emission (kg/ hr)	2214	11	11	11	0.13	0.13	0.13
CO_2 emission (gCO_2/ kWh)	925.1	145.9	385.3	346.7	151	389.4	349.7
NO_x emission (gNO_x/ kWh)	2.64	4.11	3.58	2.83	4.25	3.68	2.91
SO_x emission (gNO_x/ kWh)	4.36	0.03	0.025	0.019	0.0004	0.0003	0.0002
FGD waste (tonne/hr)	-	-	-	-	7.1	7.1	7.1
Spent MEA sorbent (tonne/hr)	-	17.1	17.1	17.1	1.4	1.4	1.4
CO_2 product (tonne/ hr)	-	420.6	420.6	420.6	421.9	421.9	421.9
CO_2 emissions captured (%)	-	90%	64%	57%	90%	65%	58%
CO_2 emissions avoided (%)	-	84%	58%	63%	84%	58%	62%
TCR (M$)	0*	235.5	270.8	365.6	239.2	273.4	367.8
TCR ($/ kW)	0*	735.3	612.5	629.5	772.7	639.4	655.2
COE ($/ MWh)	18.28	107.4	99.07	83.59	70.33	72.11	63.19
$/ tonne CO_2 avoided	-	114.4	149.7	112.9	67.2	100.5	78.1

*The existing plant is assumed to be completely depreciated.

In case B1, the cost of electricity and cost of avoidance both increase with either increase in reference plant capital cost (assuming less than 100% depreciation) or an increase in retrofitting capital cost requirement (assuming retrofit factor >1). The CO_2 avoidance cost may increase as high as 18% relative to the base case B1 discussed earlier depending upon the particular assumptions presented here in Table 5.6. However, these estimates

are not affected by natural gas price change, since we are not adding any auxiliary NG boiler in this case.

Table 5.6 Case study results for plants without FGD systems: sensitivity analysis

Case	R	B1		B3	
	COE ($/MWh)	COE ($/MWh)	$/tonne CO_2 av.	COE ($/MWh)	$/tonne CO_2 av.
Nominal case (100% depreciation, NG price = $4/mcf, rf=1)	18.3	70.33	67.2	63.19	78.0
Nominal case + SO_2 control credits	18.3	69.26	65.8	62.60	77.0
80% depreciation of reference plant	23.3	78.46	71.3	67.69	77.1
50% depreciation of reference plant	30.9	90.62	77.1	74.46	75.7
Retrofit factor (rf) = 1.25	18.3	75.58	74.0	67.58	85.6
Retrofit factor (rf) = 1.5	18.3	79.47	79.0	71.2	91.9
NG price = $2.5/mcf	18.3	70.33	67.2	56.28	66.0
NG price = $5.5/mcf	18.3	70.33	67.2	70.09	90.0
80% deprec., rf=1.25, NG =$2.5/mcf	23.3	83.71	78.0	65.19	72.8
80% deprec., rf=1.25, NG =$4/mcf	23.3	83.71	78.0	72.09	84.8
80% deprec., rf=1.25, NG =$5.5/mcf	23.3	83.71	78.0	78.99	96.8

In case B3, although the cost of electricity is lower than case B1, the CO_2 avoidance cost is higher owing to the additional CO_2 emissions coming from the auxiliary NG boiler. These costs are highly sensitive to natural gas prices. Changing the natural gas price from a low $2.5/mscf to $5.5/mscf leads to 25% increase in COE and 36% increase in the CO_2 avoidance cost. Only at the low natural gas price assumption is option B3 found to be cheaper than option B1 in terms of the CO_2 avoidance cost. The cost of electricity and cost of avoidance both increase with an increase in retrofitting capital cost requirement when we assume retrofit factor greater than 1. However, assumption about higher capital cost of the reference plant (assuming less than 100% depreciation) leads to increase in cost of electricity but a nominal decrease in the cost of CO_2 avoidance.   At 50% depreciation level, the CO_2 avoidance cost was estimated to be about 3% lower than the base case B3 (100% depreciation). The CO_2 avoidance cost may increase as high as 44%

relative to the base case B1 discussed earlier depending upon the particular assumptions presented in Table 5.6.

5.2.2. Results for plants with FGD systems

The results from the case study of CO_2 retrofitting of existing PC plants with FGD sulfur control have been summarized in Table 5.7. The coal used in this case study was medium-S bituminous with 1.4% sulfur content and heating value of 12,180 Btu/lb. The delivered price of this coal was assumed to be $30.6/ tonne which is about $0.78/ MMBtu.

Table 5.7. Case study results for plants with FGD systems

	R	C1	C2	C3	D1	D2	D3
Gross capacity (MW)	441	441	441	441	441	441	441
Net power generation (MW)	405.3	244.5	351.7	473.6	245.7	346.0	459.9
Net cycle heat rate (Btu/kWh)	10370	17130	17670	14190	17040	17580	14260
Coal consumption (tonne/hr)	156.6	155.9	155.9	155.9	155.9		155.9
NG consumption (tonne/hr)	-	-	40.43	50.54	-	37.79	47.25
CO_2 emission (tonne/ hr)	400.3	39.86	148.7	175.9	40.0	141.7	167.1
NO_x emission (tonne/ hr)	1.175	1.155	1.386	1.444	1.155	1.372	1.425
SO_x emission (kg/ hr)	1960	9.752	9.752	9.752	0.24	0.24	0.24
CO_2 emission (gCO_2/ kWh)	987.8	163.0	422.8	371.4	162.8	409.6	363.4
NO_x emission (gNO_x/ kWh)	2.9	4.7	3.94	3.05	4.7	3.96	3.1
SO_x emission (gNO_x/ kWh)	4.84	0.04	0.03	0.02	0.001	0.0007	0.0005
FGD waste (tonne/hr)	9.8	9.8	9.8	9.8	13.18	13.18	13.18
Spent MEA sorbent (tonne/hr)	-	14.54	14.54	14.54	1.364	1.364	1.364
CO_2 product (tonne/ hr)	-	358.8	358.8	358.8	360	360	360
CO_2 emissions captured (%)	--	90%	63%	56%	90%	65%	58%
CO_2 emissions avoided (%)	--	83%	57%	62%	84%	59%	63%
TCR (M$)	0*	181.3	213.1	296.5	153.1	182.7	263.3
TCR ($/ kW)	0*	741.7	605.9	626.1	623.0	528.0	527.4
COE ($/ MWh)	20.04	118.3	105.4	86.87	69.19	71.14	61.94
$/ tonne CO_2 avoided	-	119.1	151.1	108.4	59.6	88.4	67.1

*The existing plant is assumed to be completely depreciated.

In cases C1 and D1 where steam and electricity required for CO_2 capture system is extracted from the base plant, the net plant output goes down to about 60% of the original plant output. On the other hand, options C3 and D3 lead to about 15% increase in power generation capacity. Again we find that upgrading the FGD system (to 99% SO_2 removal efficiency) along with the addition of MEA-based CO_2 capture system does help in reducing the overall cost. Cost of electricity and that of CO_2 avoidance are substantially lower (about 30-40% and 40-50% lower, respectively) in cases D1-D3 (in which FGD system was upgraded) as compared to cases C1-C3 (in which only MEA-based CO_2 capture system was retrofitted).

Again, it must be noted that these results are sensitive to the assumptions made about the depreciation of the reference plant, retrofit factors in capital cost estimation, and the price of natural gas. Table 5.8 shows the effect of some of these factors on the costs in cases D1 and D3.

Table 5.8. Case study results for plants with FGD systems: sensitivity analysis

Case	R	D1		D3	
	COE ($/MWh)	COE ($/MWh)	$/tonne CO_2 av.	COE ($/MWh)	$/tonne CO_2 av.
Nominal case (100% depreciation, NG price = $4/mcf, rf=1)	20.04	69.19	63.5	61.94	72.8
Nominal case + SO_2 control credits	20.00	68.13	62.1	61.36	71.8
80% depreciation of reference plant	25.88	78.74	68.3	66.97	71.4
50% depreciation of reference plant	34.64	93.02	75.4	74.7	69.6
Retrofit factor (rf) = 1.25	20.04	73.04	68.5	65.35	78.7
Retrofit factor (rf) = 1.5	20.04	76.89	73.4	68.88	84.9
NG price = $2.5/mcf	20.04	69.19	63.5	54.77	60.4
NG price = $5.5/mcf	20.04	69.19	63.5	69.11	85.3
80% deprec., rf=1.25, NG =$2.5/mcf	25.88	82.59	73.3	63.33	65.1
80% deprec., rf=1.25, NG =$4/mcf	25.88	82.59	73.3	70.5	77.5
80% deprec., rf=1.25, NG =$5.5/mcf	25.88	82.59	73.3	77.67	90.0

We observe similar trends in cases C1-C3 Vs D1-D3, as we had noted earlier in cases A1-A3 Vs B1-B3. In general, derating seems to be the cheaper option than adding an auxiliary NG boiler, on the basis of CO2 avoidance cost. Adding auxiliary NG boiler may be cheaper if the reference plant is not completely depreciated or if NG is available at low price (below $2.5-3/mscf).

5.2.3. *Comparison with other technological options for CO_2 retrofitting*

It is also of interest to compare the results obtained for the amine-based CO_2 retrofit cases to other possible technological options available to the existing older pulverized coal power plants. In the recent times, IGCC (Integrated Gasification Combined Cycle) has been developed as an alternative technology for coal-based power generation. IGCC re-powering of old, less efficient and more polluting pulverized coal plants might be an attractive option that can also facilitate CO_2 capture. In a recent case study that we conducted to compare this option, we found that although IGCC re-powering with CO_2 capture requires a lot of new capital investment, it might be cheaper in terms of final cost of electricity and CO_2 mitigation cost (Chen, Rao et al. 2003). Some of the key assumptions made in this study were as follows:

1) Reference plants are fully amortized and equipped with FGD systems
2) Performance parameters for the reference base plant are based on DOE/NETL's database(NETL 2002)
3) Gross capacity: 270MW
4) Gross plant heat rate: 10,734 Btu/kWh
5) CO_2 emission rate: 1004 gCO_2/ kW
6) Plant capacity factor: 76%
7) CO_2 capture efficiency: 90%
8) Financial parameters: Plant life = 20 years, Interest rate = 9%
9) Amine-based retrofit: Adding 30% MEA-based CO_2 capture system
10) Re-powering: Texaco, O_2-blown, quench, GE MS7001F, Selexol-based CO_2 capture
11) Identical assumptions have been made about CO_2 product compression (2000 psig), transport and storage/disposal costs etc.

In case of amine-based CO_2 retrofits, two options were considered, *viz.* base plant derating (AR1) and adding an auxiliary NG-fired boiler with secondary steam turbine

(AR2). The case of IGCC re-powering was analyzed using an IGCC performance and cost model developed within the IECM modeling framework (CEES 2003). Again, two options have been considered, *viz.* Heat recovery re-powering (IR1, using existing steam turbine and land) and site re-powering (IR2, completely new IGCC on the existing land). The key results have been summarized in Table 5.9.

It may be noted that cost of electricity (and hence that of CO_2 mitigation) is a strong function of the natural gas price in case of amine-based CO_2 capture using auxiliary NG boiler and a turbine. Only at very low natural gas price can this configuration (AR2) compete with IGCC re-powering options (IR1 & IR2) in terms of the final cost of electricity. However, the later has the lowest CO_2 avoidance cost. This observation has also been supported by the probabilistic analysis we conducted for these options.

Table 5.9. Comparison of amine-based CO_2 capture and IGCC re-powering options for an existing pulverized coal plant

Parameter	Reference plant	Amine-based retrofits		IGCC Re-powering	
		AR1	AR2	IR1	IR2
Gross capacity (MW)	270				
Net capacity (MW)	248	140	282	590	599
CO_2 emission (gCO_2/ kWh)	1004	177	369	99	98
TCR ($/ kW)	0*	837	650	1493	1698
COE ($/ MWh)					
@ $4.7/mcf NG	21	83	70	62	67
@ $2.4/mcf NG			59		
CO_2 avoidance cost ($/ tonne CO_2 avoided)					
@ $4.7/mcf NG	-	75	77	46	51
@ $2.4/mcf NG			61		

*The existing plant is assumed to be completely depreciated.

The cost results (COE and cost of CO_2 avoidance) of the probabilistic analysis have been presented in Figure 5.7 and Figure 5.8. Under the set of assumptions stated before, amine-retrofit with internal derating of the base plant is clearly the most expensive option. The amine-retrofit option with auxiliary NG-fired boiler and a secondary steam

turbine may be able to compete with the more efficient IGCC repowering options, only on the basis of cost of electricity. However, since the auxiliary NG boiler generates secondary CO_2 emissions, this option is less effective in terms of avoiding CO_2 emissions. So, the IGCC repowering options, which are also more energy efficient, are cheaper than the amine-retrofits, in terms of the CO_2 avoidance costs. Even though the amine-retrofit with internal derating is much more expensive in terms of the cost of electricity generated than that with auxiliary NG boiler, the probability distribution for the CO_2 avoidance cost for these options indicate that there is a small probability that the derating option might be cheaper than adding auxiliary NG boiler, based on the assumptions of this analysis.

5.3. Discussion of Results

5.3.1. Greenfield vs. retrofit applications

Case studies of new and retrofit application of amine-based CO_2 capture system at power plants have been presented in sections 5.1 and 5.2. All these cases have been summarized in Figure 5.9. Here COE for each case has been plotted against its CO_2 emission rate (CO_2/ kWh). It may be recalled that the slope of the line joining a capture plant with the respective reference plant gives the CO_2 avoidance cost for that plant. The new reference and capture plant (discussed earlier in section 5.1.1 and shown in Figure 5.2) has also been included for comparison.

It may be noted that a much higher energy penalty is expected in case of retrofit applications. This makes the cost of avoidance in these cases higher than that in the case of an efficient new plant with CO_2 capture. However, since the older plants are assumed to be completely paid of in terms of capital requirement, the retrofit applications are much cheaper than their greenfield counterparts. In the retrofit case, the cost of electricity is essentially based on the operating costs of the base plant and costs associated with the installation and operation of the CO_2 capture equipment. Hence, in terms of cost of electricity produced, the retrofit applications are much cheaper than the new pulverized coal plants with CO_2 capture.
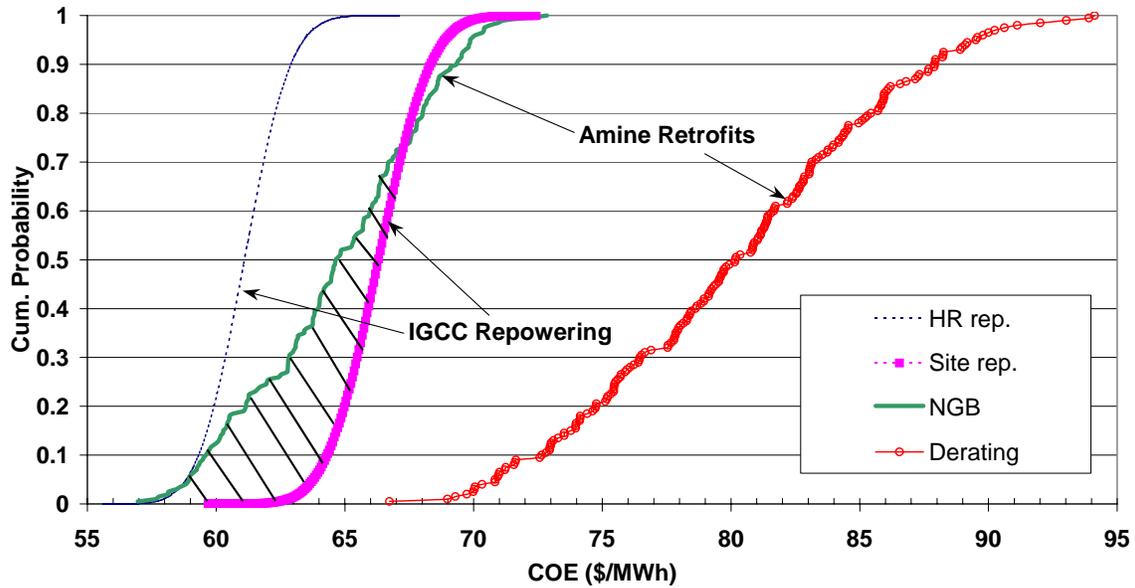
Figure 5.7. Comparison of CO₂ retrofitting options: cost of electricity
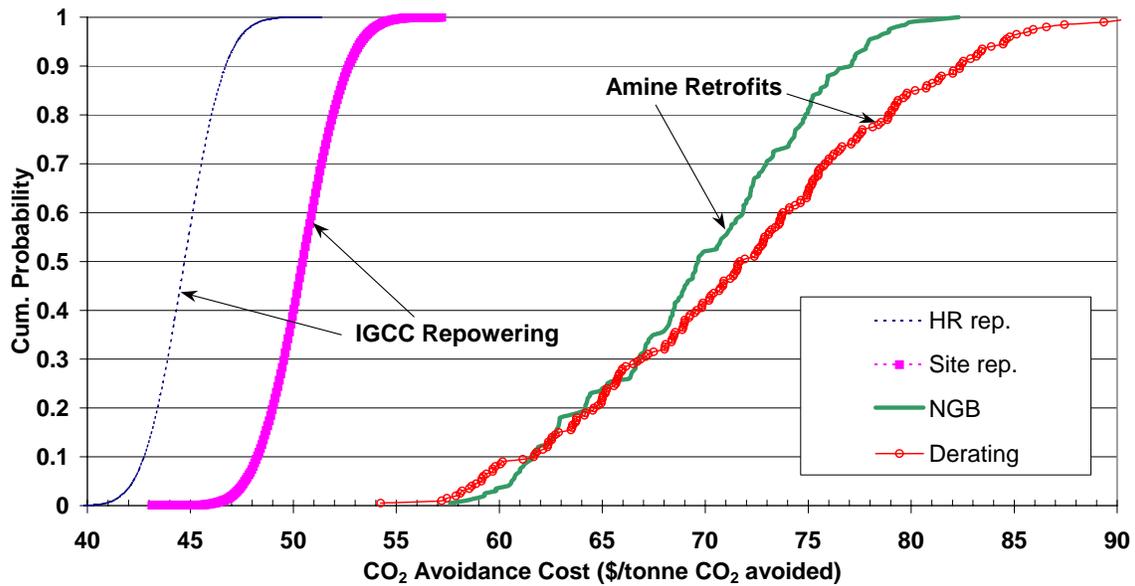
Figure 5.8. Comparison of CO₂ retrofitting options: cost of CO₂ avoidance

Given a sufficiently stringent CO_2 reduction requirement, will utilities invest in new plants with CO_2 capture, or will they prefer to spend less for retrofit applications, although it maybe a less effective measure to control CO_2 emissions?   Will the

government provide any incentives (subsidies, tax rebates *etc.*) to any of the power generation options that help in reducing CO_2 intensity of electricity sector? Answers to such questions will have a major impact on how the power industry reacts to any future regulation of CO_2 emissions and technological choices available.

Similarly, there is another set of issues related to electricity demand and supply management, dispatch dynamics, availability and price of natural gas, evolution of a domestic and/or global market for CO_2 trade, access to other (cheaper?) sources of CO_2 control credits, role of renewable energy sources (especially wind, biomass *etc.*) in the total energy mix, and other factors that will play a major role in CO_2 mitigation strategies. These factors are beyond the scope of this study, and analysis of such questions requires higher level modeling tools, such as discussed elsewhere (Johnson and Keith 2001; Johnson 2002).
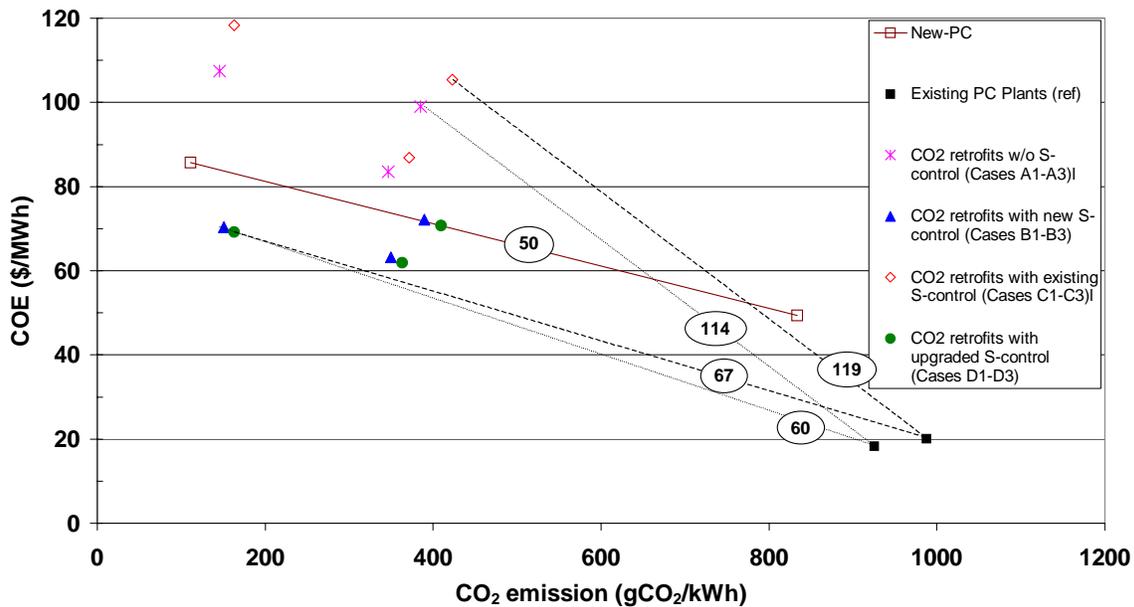
Figure 5.9. Estimation of CO_2 avoidance cost for pulverized coal plants: greenfield vs. retrofit application

5.3.2. Base plant derating vs. Auxiliary NG-fired boiler options

As we have seen in the case studies of the existing coal plants (sections 5.2.1, 5.2.2), the CO_2 capture system configurations with auxiliary NG-fired boiler and steam turbine (B3

& D3) have the lowest COEs (for natural gas price < $5.5/mscf, see Figure 5.10) than the other CO_2-retrofitting alternatives in their respective categories. Another advantage of these options is that they increase the net power generation capacity of the plant. So, they are likely to be the preferred applications of this technology. Assuming availability of enough space for this retrofit and that of natural gas fuel, this option may be looked upon as a convenient way to boost the generation capacity of existing older power plants.
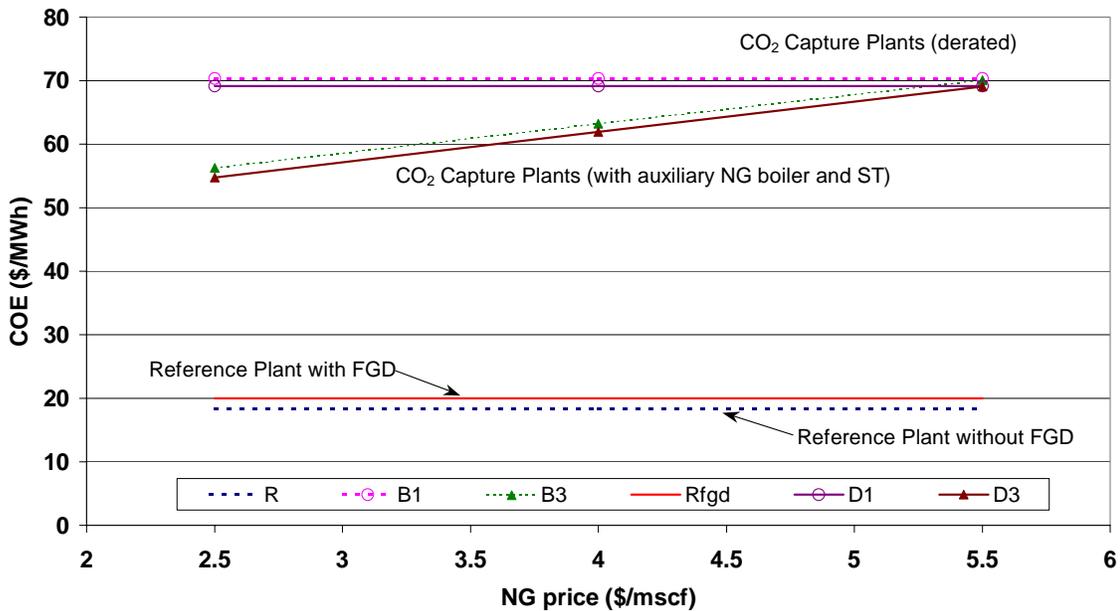
Figure 5.10. Effect of natural gas price on COE in CO_2 retrofit applications

However, their economics is very sensitive to the natural gas price. At the nominal gas price assumption of $4/mscf, their cost of CO_2 avoidance is higher than those estimated for the respective options (B1 & D1) that derate the base plant; and they can break even only at low gas prices (less than $2.5-3/mscf), as seen in

Figure 5.11. Also, assuming that CO_2 emissions from the auxiliary boiler are not controlled, the overall CO_2 emissions avoided in these cases is only about 60-65%. Whether this level of CO_2 mitigation is acceptable, depends upon the stringency of the CO_2 mitigation regulation and other market mechanisms. If new combustion sources are regulated for their emissions of criteria pollutants with concerns about the local pollution,

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then one may also have to account for NO_x-control devices in this option (not accounted for in the present analysis).
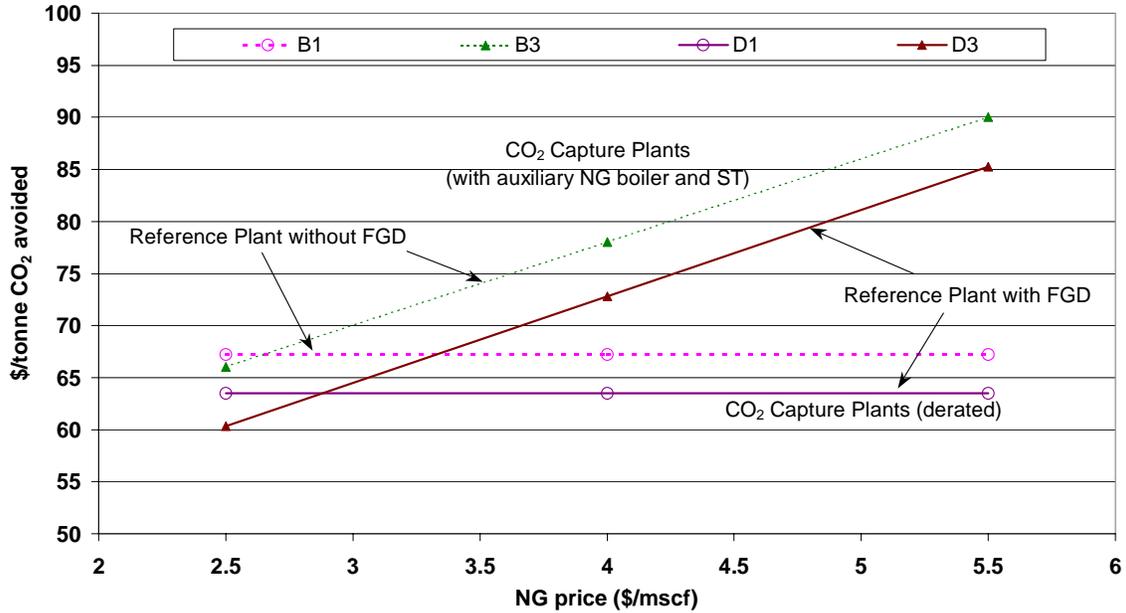
Figure 5.11. Effect of natural gas price on CO_2 avoidance cost in CO_2 retrofit applications

On the other hand, the CO_2 capture options (B1, D1) that derive their energy requirement from the base plant itself, offer higher levels of CO_2 avoidance (> 80%) and are independent of NG price fluctuations. But these options put a large energy penalty on the reference plant and the net power generation capacity of the plant may decrease by as much as 40%, depending on the site specific design of existing coal-fired units. This may have serious implications in terms of power supply capacity planning and supply management.

5.3.3. CO_2 control from coal-fired vs NG-fired power plants

The IECM-CS also has a performance and cost model of an NGCC system. Table 5. compares the results from the simulation of a new NGCC system with and without CO_2 capture with those from the case study of a coal plant presented earlier in section 5.1.

It may be noted that the capital cost of NGCC plant is quite low ($521/kW) and even after adding CO_2 capture system, it is cheaper than the reference coal plant. This lower

cost may be attributed to the higher efficiency of combined cycle technology and also to the use of cleaner fuel that cuts down the cost of environmental controls, as compared to a coal plant. NGCC plant with CO_2 capture has substantially lower CO_2 emissions and produces a smaller stream of concentrated CO_2 for storage/ disposal. In comparison with PC plant with CO_2 capture, it has much lesser (or no) environmental discharges. Although the COE from an NGCC plant (even with CO_2 capture) is substantially lower than that from a coal plant with CO_2 capture, the cost of CO_2 avoidance is much higher.

Table 5.10. Comparison of CO_2 control from a new NGCC and a new PC plant

	NGCC		PC	
	Ref.	**Capture**	**Ref.**	**Capture**
Gross capacity (MW)	518	518	500	500
Capacity factor (%)	75	75	75	75
Fuel cost ($/GJ)	3.5	3.5	1.2	1.2
Net power generation (MW)	518	445	458	341
Net cycle heat rate (Btu/kWh)	6,383	7,435	8,657	11,560
Coal consumption (tonne/hr)	-	-	215.5	214.6
NG consumption (tonne/hr)	65.9	65.9	-	-
CO_2 emission (tonne/ hr)	180.8	18.1	381.3	38.0
CO_2 emission (gCO_2/ kWh)	349.0	40.7	833.3	111.3
FGD waste (tonne/hr)	-	-	4.2	4.3
Spent MEA sorbent (tonne/hr)	-	0.39	-	1.0
CO_2 product (tonne/ hr)	-	163	-	342
CO_2 emissions captured (%)	-	90	-	90
CO_2 emissions avoided (%)	-	88	-	87
TCR (M$)	270	383	616	760
TCR ($/ kW)	521	861	1345	2228
TRR (M$/yr)	126	179	149	202
COE ($/ MWh)	36.7	61.3	49.4	89.9
$/ tonne CO_2 avoided	-	79.2	-	56.0

Figure 5.12 plots the COE and CO_2 emission rates for NGCC plant with and without CO_2 capture for various assumptions about natural gas price and plant capacity factor. As

described earlier, the slope of the line joining a reference plant and a capture plant on this plot gives the CO_2 avoidance cost. It appears that even at low gas price (\$2.5/mscf) and high capacity factor (85%), cost of CO_2 avoidance from an NGCC plant is about 30% higher than that for a PC plant of comparable net power output. However, the COE at such low gas price is even lower than the COE estimated for a PC without CO_2 control. So, even at low gas prices, controlling CO_2 from NGCC plants makes sense only if COE is the major concern rather than the cost of CO_2 mitigation.
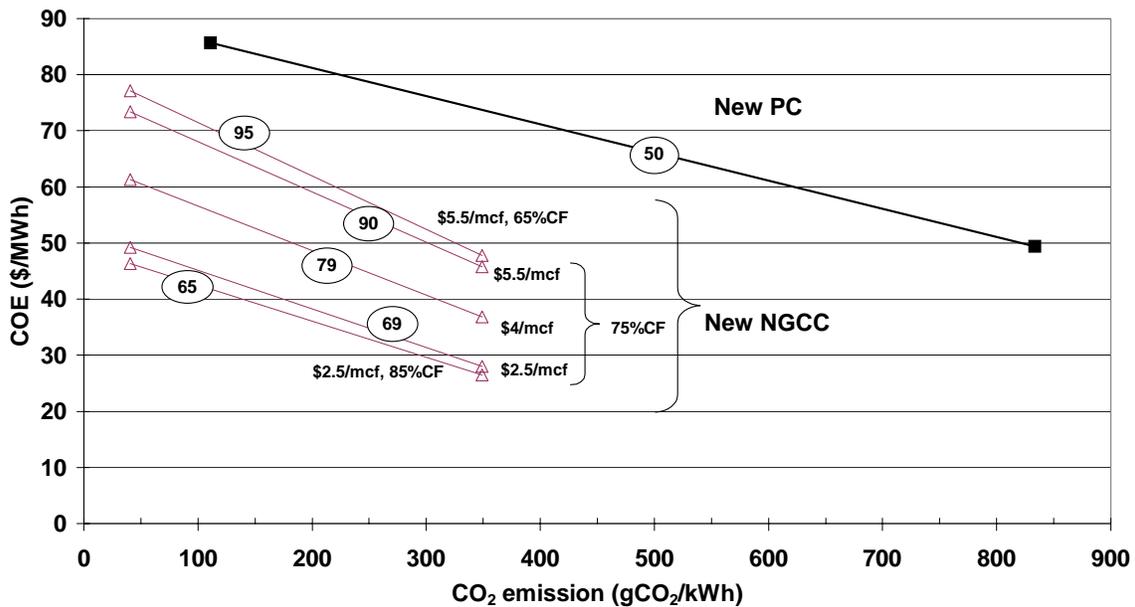
Figure 5.12. Comparison of the cost of CO_2 control from a new NGCC and a new PC plant (the numbers in the circles represent the cost of CO_2 avoidance)

Natural gas combined cycle (NGCC) is an efficient, cleaner and less capital-intensive technology. This attractive power generation technology is growing in capacity very fast and is expected to grow further in the coming decades (EIA 2000). An NGCC plant produces only about 50% CO_2 per unit of electricity as compared to a conventional coal plant. It might be desirable to capture these emissions under a stringent climate policy, and especially if it is economical (which is not the case, as seen above). Flue gas from an NGCC plant is much cleaner (due to the cleaner fuel) as compared to the flue gas from a coal plant and hence it inflicts less operational difficulties, *viz.* corrosion and disposal of spent sorbent. Today majority of the CO_2 capture plants using amine systems are

installed to scrub CO_2 from gas-fired power plants and hence there is more experience with this application. However, the flue gas from an NGCC plant is less concentrated in CO_2 and it takes more energy to regenerate the sorbent per unit of CO_2 captured. This partly explains why the cost of CO_2 avoidance from these plants was found to be higher than that from a PC plant. Also, the economics of these plants is highly sensitive to the price of natural gas. Often these plants are operated at lower capacity factors as well, which may actually lead to higher avoidance costs. So, based on this analysis, it seems that capture of CO_2 from NGCC plants, although technically feasible and favorable, is likely to be more expensive as compared to that from coal plants.

References (Chapter 5)

CEES (2003). A technical and economic assessment of Selexol-based CO2 capture technology for IGCC power plants, Annual Technical Progress Report prepared by Center for Energy and Environmental Studies, Carnegie Mellon University, Pittsburgh for U.S. Department of Energy, National Energy Technology Laboratory, Morgantown, WV. (Work performed under contract no. DOE/DE-FC26-00NT40935).

CFR (1999). New Source Performance Standards (NSPS). Code of federal regulations, Federal Register. **40CFR, Chapter 1**.

Chen, C., A. B. Rao, et al. (2003). Comparative Assessment of CO_2 Capture Options for Existing Coal-Fired Power Plants. presented at the 2nd Annual Conference on Carbon Sequestration, May 5-9, 2003, Alexandria, VA, USA.

EIA (2000). Annual Energy Outlook 2001 (with Projections to 2020), A report published by Energy Information Administration, U.S. Department of Energy, Washington, DC.

Johnson, T. L. (2002). Electricity without carbon dioxide: assessing the role of carbon capture and sequestration in US electric markets. Department of Engineering and Public Policy. Pittsburgh, PA 15213, Carnegie Mellon University**:** 247.

Johnson, T. L. and D. W. Keith (2001). "Electricity from fossil fuels without CO_2 emissions: Assessing the costs of carbon dioxide capture and sequestration in U.S. electricity markets." Journal of the Air & Waste Management Association **51**(October): 1452-1459.

NETL (2002). U.S. Coal power plants database, 2000.

Rao, A. B. and E. S. Rubin (2002). "A Technical, Economic, and Environmental Assessment of Amine-based CO_2 Capture Technology for Power Plant Greenhouse Gas Control." Environmental Science and Technology **36**: 4467-4475.

6. R&D MANAGEMENT: POTENTIAL IMPROVEMENTS AND COST REDUCTION

Technological innovations in CO_2 capture and storage technologies are being pursued worldwide under a variety of private and government-sponsored R&D programs. While much of this R&D is directed at novel concepts and potential breakthrough technologies, there are also substantial efforts to improve CO_2 capture technologies already in use. As discussed before, amine-based CO_2 capture systems have been proposed as one of the solutions for existing as well as new conventional coal-fired power plants. This technology is commercially available today, though it is highly energy intensive and costly, as found in this study and elsewhere (Smelser, Stock et al. 1991; Hendriks 1994; Leci 1996; Simbeck 1998; Desideri and Paolucci 1999; Laboratories 1999; Mariz and al. 1999; Jeremy and Herzog 2000; Parsons Infrastructure & Technology Group 2002; Rao and Rubin 2002).

Hence, there is a major effort being made to improve amine systems for post-combustion CO_2 capture application at power plants. How far can these systems be improved in near future? Technical experts in this field can have a sense about this. As part of this research we sought to quantify the informed judgments of experts to derive uncertainty distributions for various key parameters that will affect the performance and cost of these systems in the future. This is accomplished following the "expert elicitation" protocol widely discussed in the literature and a generally accepted norm to estimate uncertainty when data is lacking (Morgan and Henrion 1990; Frey 1991; Morgan and Keith 1995).

Here we have made an effort to understand what the experts in this field think about future amine systems. Further, we have used the experts' judgments as input to our model (IECM-CS) to estimate the possible cost reductions in future systems.

6.1 Methodology

The following procedure was used to estimate the probability distributions for key performance parameters of future amine systems.

6.1.1. Identification of the parameters

As described before, a model (IECM-CS) has been developed to simulate the performance and cost of post-combustion CO_2 capture using amine-based systems at fossil fuel power plants. Analysis of the various parameters helped us identify the key parameters that have a substantial influence on the performance and cost of these systems. Consideration also was given to the clarity of parameter definition, frequency of mention in the literature, and total number of questions to be asked of experts. On this basis, the following set of parameters related to the amine-based CO_2 capture system were selected:

- Sorbent concentration (wt %)
- Sorbent regeneration heat requirement (kJ/ kg CO_2 captured)
- Sorbent loss (kg/ tonne CO_2 captured)
- Sorbent cost (US$/ tonne sorbent)

In addition to these, the following parameters were also included in the questionnaire:

- CO_2 compressor efficiency (%)
- Net plant efficiency of the power plant (with and without CO_2 capture)
- Total capital requirement (TCR, $/kW) for the power plant (with and without CO_2 capture)
- Cost of electricity (COE, cents/ kWh) (with and without CO_2 capture)

6.1.2. Identifying the experts

Professionals working in the area of amine-based CO_2 capture were identified through their authorship of technical papers and participation in International Conferences. Experts were contacted via email and requested to participate in this study. The initial response to this appeal was quite encouraging, and within the time constraints available, final responses were obtained from about a dozen experts. The experts' participation in this study was completely voluntary. There was balanced representation from industry, academia, independent research laboratories and private consultancies. Experts from all over the world participated in this study. The names of these experts have been listed in the acknowledgement section.

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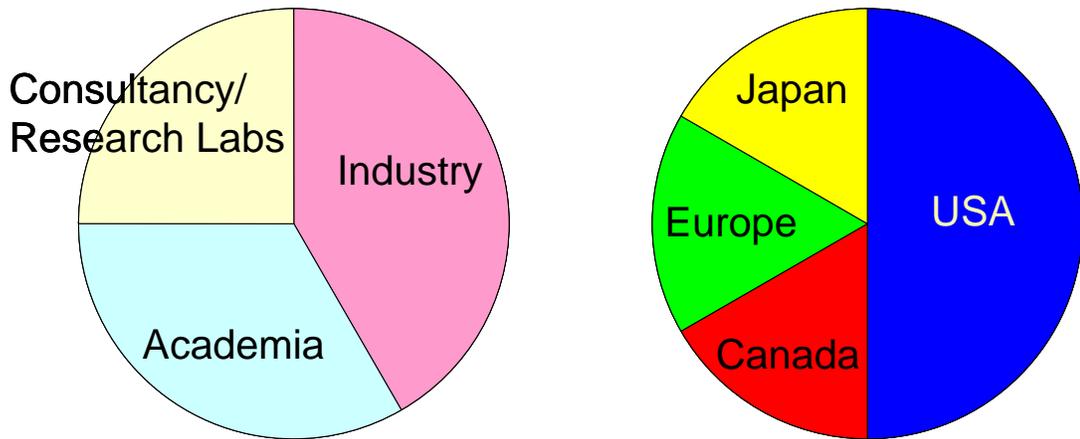
Figure 6.1. Experts' affiliations (N=12)

The initial email sent to the potential experts described the purpose of this exercise. Most of the subsequent communication was carried out via email and was supplemented by personal/ telephonic interviews.

6.1.3. The questionnaire

Once the key process parameters were identified, a detailed questionnaire was designed (see *Appendix E*). It included a brief description of the current amine-based CO_2 capture systems. This was helpful to set the stage and to ensure consistent use of terminology. Next, it included a checklist of parameters of a current amine-based system. Experts were asked to comment ("OK" or "not OK") on the typical values and ranges of these parameters and to provide replacements if the original numbers were "not OK". This part of the questionnaire helped to know the individual perceptions of the experts about the current systems (baseline). Next, there were detailed questions designed to obtain the uncertainty distributions for the parameters identified above, for a particular scenario of CO_2 capture using future amine system at power plants. Assumptions about the size, location, capture efficiency of this system at a new low-sulfur coal-fired power plant for the year 2015 were outlined. The final part of the questionnaire asked the experts to prioritize various research objectives aimed at improving the performance of amine systems and reducing the cost of CO_2 capture at power plants.

6.1.4. *The questionnaire responses*

Responses were obtained from 12 experts over a period of 3 months. Results were processed to plot the uncertainty distributions for various parameters as indicated by all the experts. These compiled results were sent back to all the respondents. This was done in order to avoid any misinterpretation of the responses during data analysis/ units conversion *etc.* and also to give the experts another chance to review their responses. The respondents were requested to reply within a stipulated timeframe if they wanted to make any changes to their original responses. Only a couple of experts changed their responses during this step. Figure 6.6, Figure 6.7, Figure 6.8, and Figure 6.9 show the final set of distributions for the four parameters derived from the experts' responses about the parameter values for future amine-systems. Some of the data points in these distributions (either min/max values or 5-percentile/95-percentile values, whichever were not provided by an expert) have been extrapolated or interpolated on the basis of the rest of the data points provided by the expert.

6.2 Expert Judgments on Current Amine-based CO_2 Capture Systems

The initial pages of the questionnaire included a checklist of parameters of a current amine-based system. The parameters included were:

- Absorber inlet flue gas pressure

- Temperature of flue gas entering the absorber

- Lean sorbent CO_2 loading (mole CO_2/ mole sorbent)

- Heat required for sorbent regeneration (kJ/ kg CO_2)

- Allowable levels of other components in flue gas

- Maximum train size (tonnes CO_2 per day)

- Energy required for CO_2 compression to 2000 psig (kWh/ tonne CO_2)

Experts were asked to comment ("OK" or "not OK") on the typical values and ranges of these parameters that were provided (based on current IECM-CS defaults) and to provide replacements if the original numbers were "not OK". This part of the questionnaire helped to know the individual perceptions of the experts about the current systems

(baseline). The last item in this list (related to CO_2 compression) was not related to the amine-based CO_2 capture system. Nonetheless, some of the experts did respond to this parameter as well. Responses to this parameter were also obtained from a separate questionnaire (containing the same question about CO_2 compression energy) that was sent to another set of experts on post-capture processing of CO_2.

The responses for these parameters have been summarized in Figure 6.2, Figure 6.3, Figure 6.4, and Figure 6.5. The experts have been represented by letters A, B, C *etc.* that have been randomly assigned to the experts. The responses to the question on the allowable levels of other components in flue gas have been summarized below:

1) SO_2 (< 10 ppmv): Most of the experts believe that a limit of 10 ppmv SO_2 is good enough, with few exceptions. One expert suggested that it should be less than 5 ppmv. Another one suggested that the number could lie in the range 2-50 ppmv, and the limit should be decided by the economics of makeup (sorbent requirement) and (spent sorbent) disposal. Lastly, there was an expert who questioned if the suggested limit of 10 ppmv was practical as he believed that even with a caustic polishing, real plants might achieve SO_2 levels of about 20 ppmv.

2) NO_2 (< 10 ppmv): Almost all of the experts agreed with this limit, except one person who questioned if this number has any practical basis, and suggested that this number may lie in the range 5-50 ppmv.

3) O_2 (~ 3.5% v/v): Most of the experts found this limit acceptable. Two experts believed that higher O_2 levels could be allowed, one of them quoting 8% as the limit, another one claiming that the system can work even with 17% O_2, while a third expert said that these systems can work at any oxygen levels although the sorbent degradation economics may worsen at higher levels.

6.3 Expert judgments on Future Amine-based CO_2 Capture Systems

The next section of the questionnaire tried to gather technically informed probabilistic judgments about some of the key parameters of a future amine-based CO_2 capture system built around the year 2015. The basic assumptions laid out included the following:

- An amine-based plant that treats the flue gas stream from a coal-combustion source which is about 12% CO_2 and which has been pre-treated for removal of SO_x and NO_x, and removes 90% of CO_2 from the flue gas stream.

- The plant has been optimized for the lowest overall cost of CO_2 avoidance ($/ tonne CO_2 avoided), considering both capital and operating costs (including energy costs) over the life of the plant.

- R&D support for this technology continues to steadily grow at a modest rate through 2015, and includes several new large-scale applications to coal-fired power plants.

The responses to these questions have been processed to obtain the probability distributions presented in Figures 6.6-6.9. It may be seen that there is considerable diversity in these responses, especially in comparison to the distributions for the case of a current commercial amine-based CO_2 capture system. Also, not all the experts responded to all the questions.

Most of these responses are optimistic, in the sense they predict an improvement in the parameter value in future systems. For example, consider the sorbent regeneration heat requirement, the parameter that is crucial to the overall energy penalty of this system. Figure 6.10 shows the "best guess" estimates for the current and future systems, as given by each expert. The future estimates are consistently lower than the current ones, implying better performance in the future.
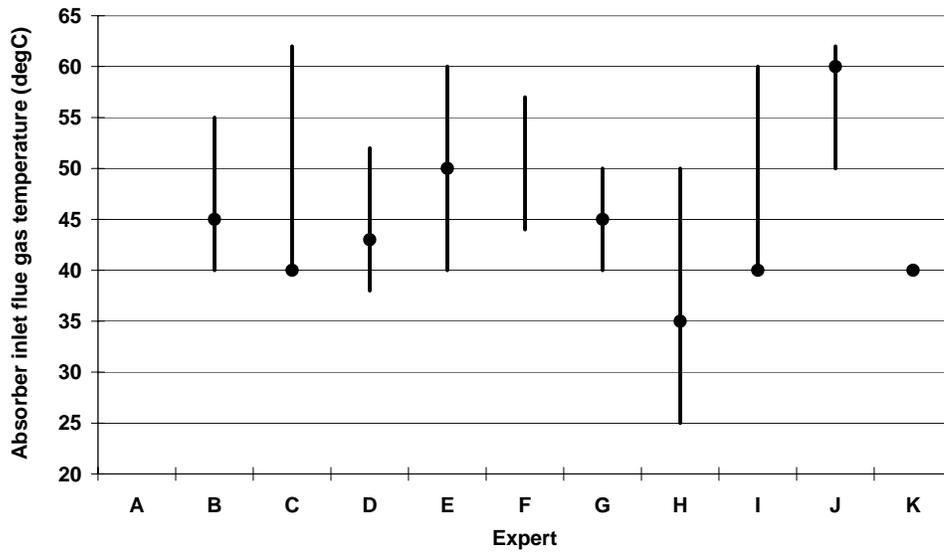
Figure 6.2. Expert judgements on current commercial amine systems for CO$_2$ capture from flue gases: absorber inlet flue gas temperature (deg C)
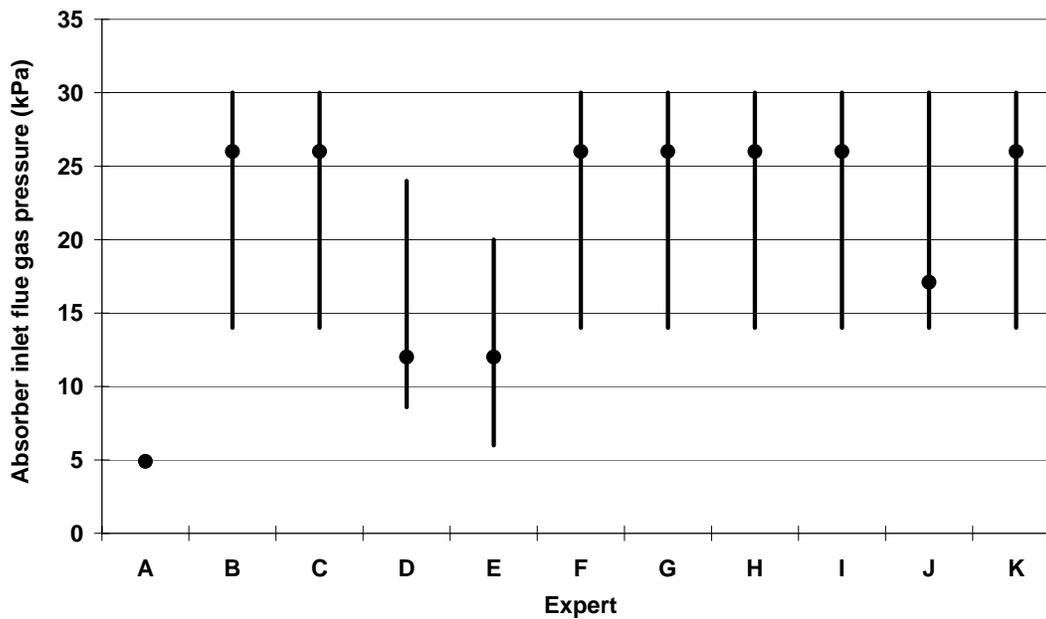
Figure 6.3. Expert judgements on current commercial amine systems for CO$_2$ capture from flue gases: absorber inlet flue gas pressure (kPa)
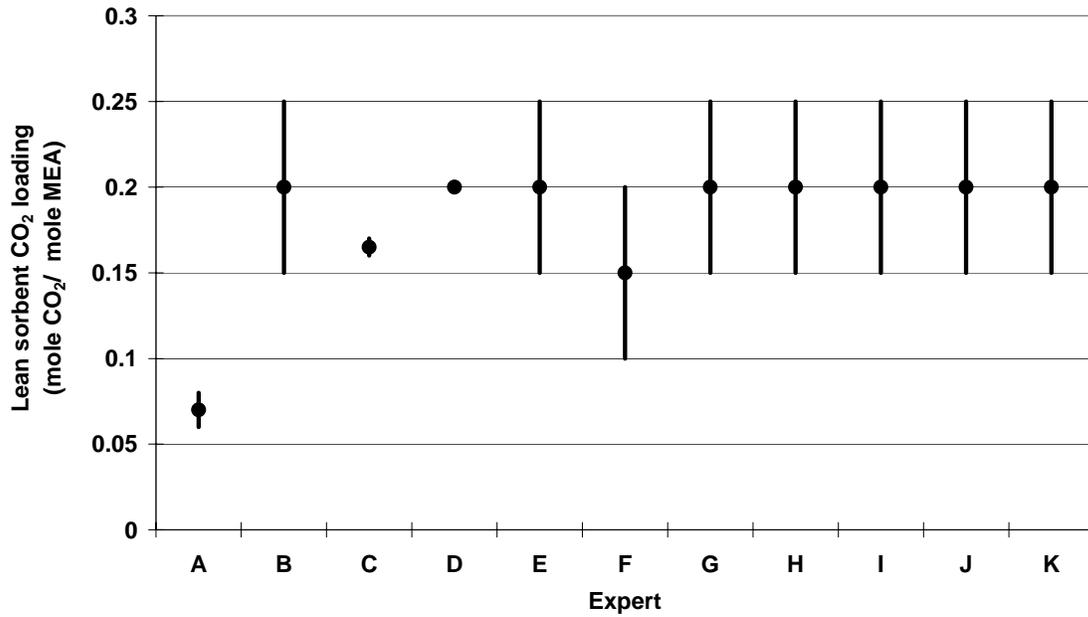
Figure 6.4. Expert on current commercial amine systems for CO_2 capture from flue gases: lean sorbent CO_2 loading (mole CO_2 / mole MEA)
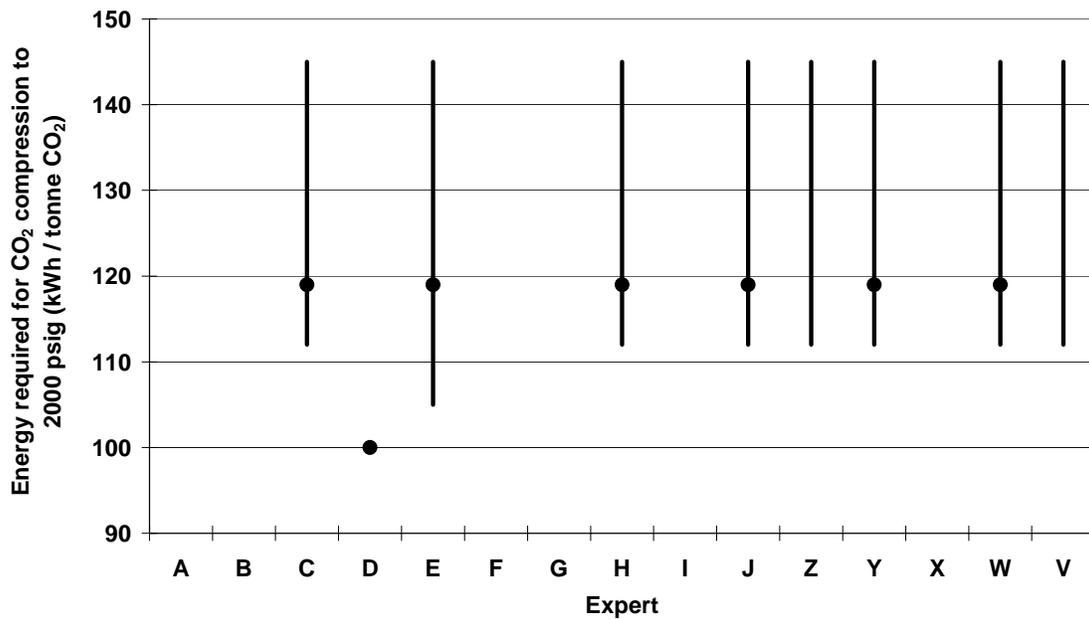
Figure 6.5. Expert on current commercial amine systems for CO_2 capture from flue gases: energy required for CO_2 compression to 2000 psig (kWh / tonne CO_2)

The basic motivation behind this study was to understand what experts in this field believe about the possible scope for improvement in the performance of amine-based CO_2 capture systems and for reduction in the cost of CO_2 capture. From the experts' responses, improvement in the various parameters relative to the current baseline was estimated. The "best guess" judgments gave a nominal/ most probable improvement, while the minimum or maximum values (depending upon the parameter) led to the "most optimistic" estimation of improvement. The relative improvement is calculated as the difference in the current and future estimate expressed as a percentage of the current baseline.

$$\text{Relative improvement in parameter Z (\%)} \;=\; (f_Z) * \frac{(Z)_{\text{future}} - (Z)_{\text{current}}}{(Z)_{\text{current}}} \times 100$$

Where, Z_{future} = Expert's judgment about the value of parameter Z in a future system

Z_{current} = Current baseline value of parameter Z

f_Z = Parameter specific multiplier that reflects the "improvement"

= (+1) for Z = Sorbent concentration

= (-1) for other three parameters viz. sorbent regeneration heat requirement, sorbent loss and sorbent cost
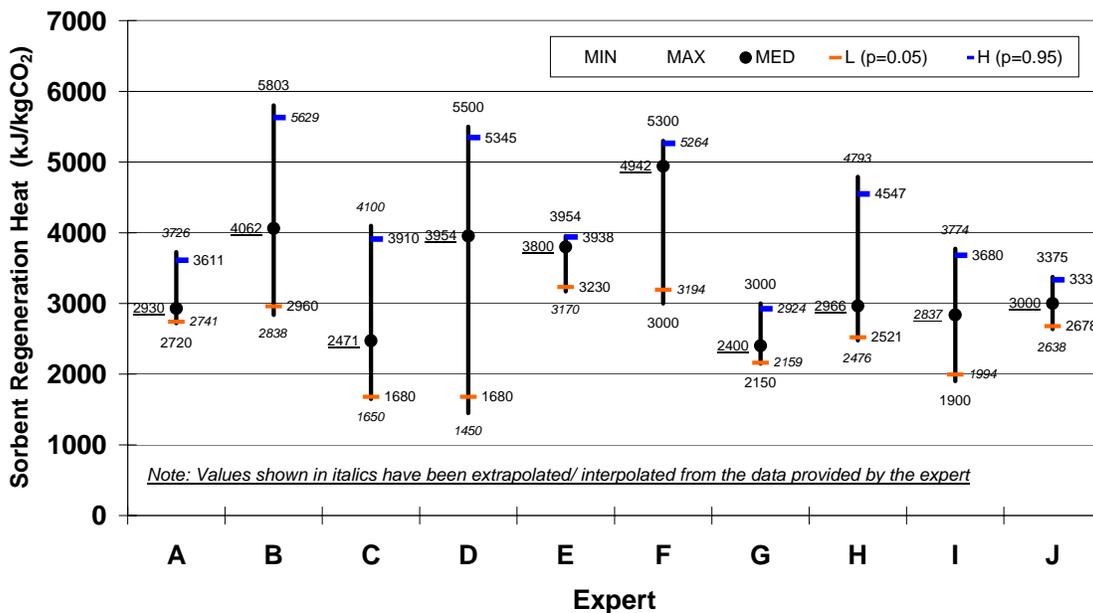
Figure 6.6. Expert judgments on future commercial amine systems: Parameter 1: sorbent regeneration heat (kJ/ kgCO₂)

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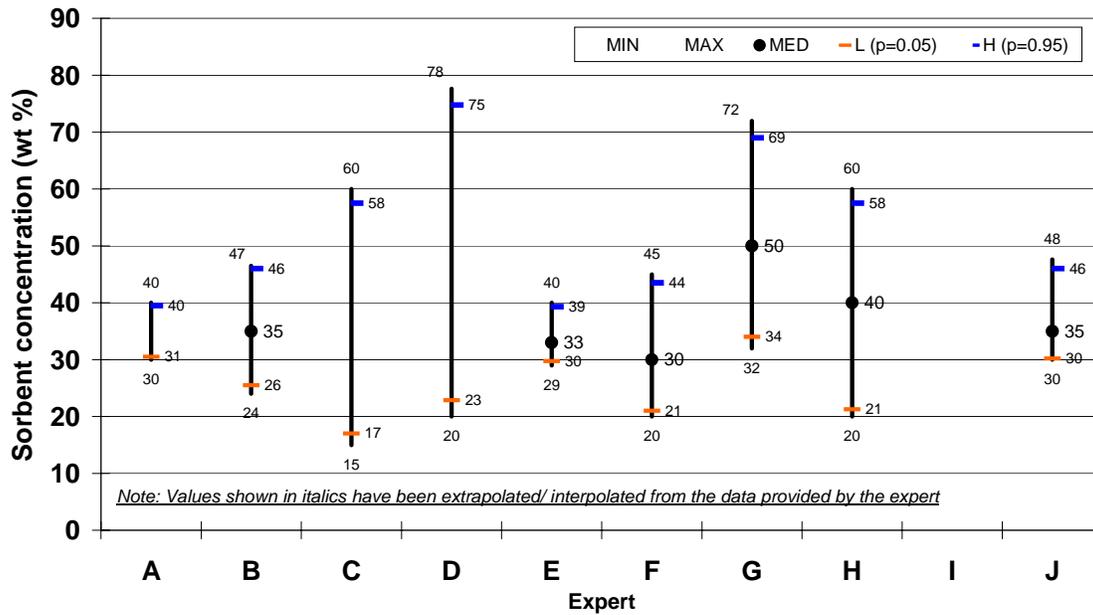
Figure 6.7. Expert judgments on future commercial amine systems: Parameter 2: sorbent concentration (% w/w)
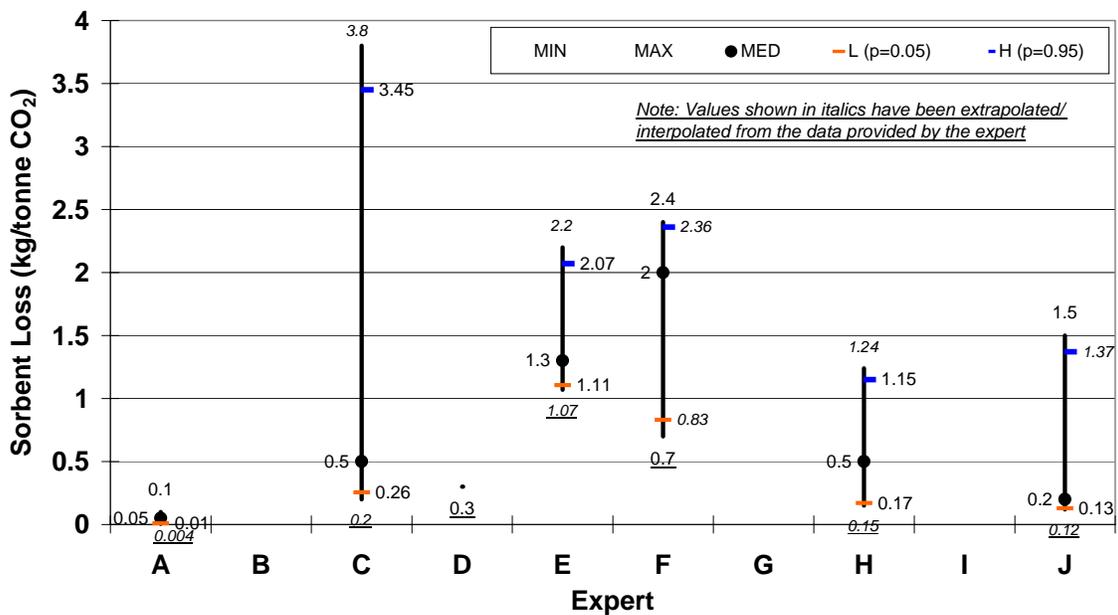
Figure 6.8. Expert judgments on future commercial amine systems: Parameter 3: sorbent loss (kg MEA/ tonne CO$_2$)
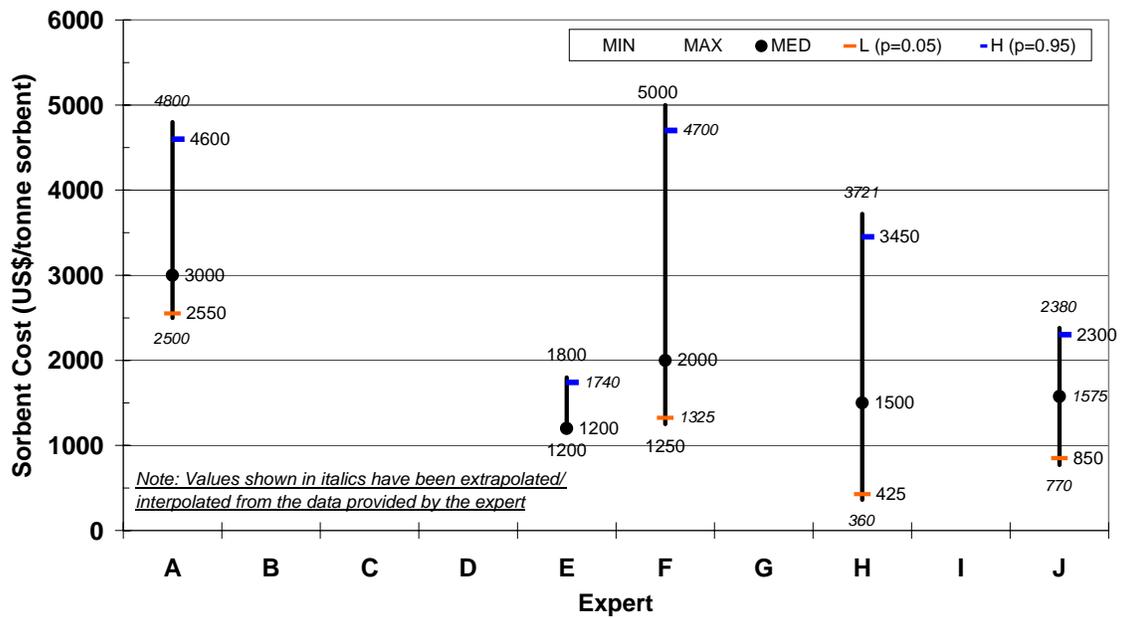
Figure 6.9. Expert judgments on future commercial amine systems: Parameter 4: sorbent loss (US$/ tonne CO₂)

Based on the "best guess" and "most optimistic" future judgments, the relative improvement is estimated and the results have been summarized in Tables Table 6.1, Table 6.2, Table 6.3, and Table 6.4.

6.3.1. Parameter 1: Sorbent Regeneration Heat Requirement

As we have seen before, this parameter determines the energy penalty of the amine system. Figure 6.10 shows the absolute values of this parameter as predicted by the various experts for a future system, compared against their own current baseline "best guess" values.

It may be noticed that all the experts are optimistic about future improvements in the performance of the amine-based CO_2 capture systems and predict that the sorbent regeneration heat requirement would be lower in the future. The best guess estimates predict an improvement in the range of 5 to 40 percent as compared to the experts' individual current estimates. The average improvement is about 23%. The most optimistic judgments are in the range of 15-73%, averaging about a 43% improvement.
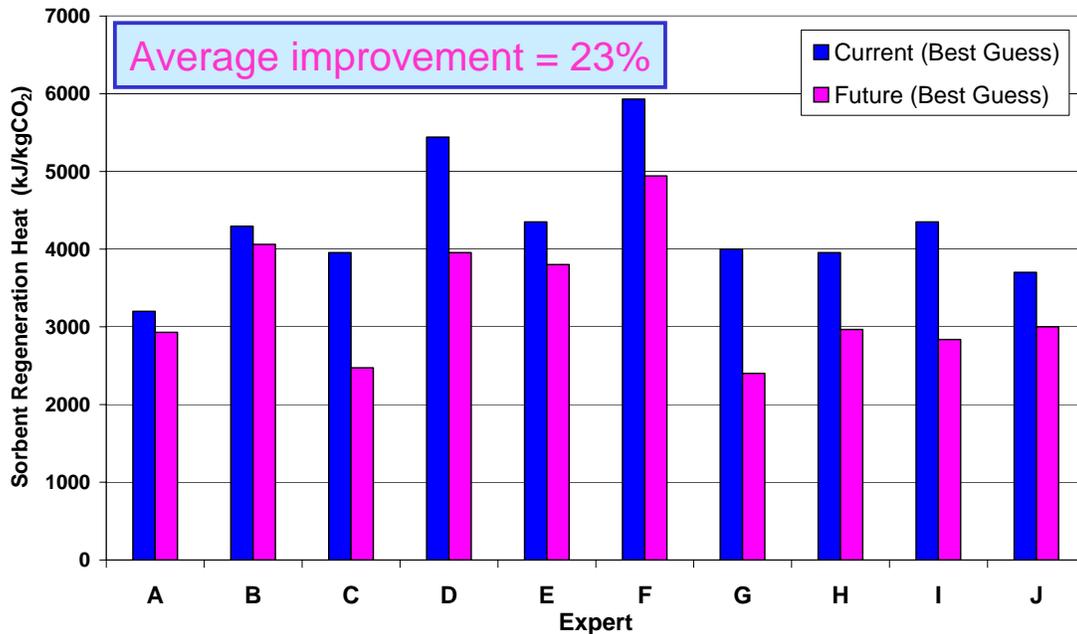
Figure 6.10. Sorbent regeneration heat requirement (kj/ kgco₂): current baseline vs. future projections

Table 6.1. Relative improvement in sorbent regeneration heat requirement

Based on	Average	Range
"Best Guess" future judgments	23%	5%-40%
"Most Optimistic" future judgments	43%	15%-73%

Experts cited advanced amines, hindered amines, special mixtures of amines, or equipment change as possible strategies to achieve these improvements. A few said they

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had "no idea" as to how the improvements would be achieved, but were optimistic nonetheless that such improvements were achievable.

6.3.2. Parameter 2: Sorbent Concentration

Most of the current commercial systems use an amine sorbent concentration of less than or equal to 30% w/w. Many experts believe that using a higher concentration of sorbents may help reduce the energy penalty of this system and improve the overall performance. At the same time, there are a few experts who believe that sorbent concentration is a non-issue.

Table 6.2. Relative improvement in sorbent concentration

Based on	Average	Range
"Best Guess" future judgments	23%	0%-67%
"Most Optimistic" future judgments	80%	33%-160%

The "best guess" value for sorbent concentration in a future system averaged about 37% w/w (which is about 23% improvement over the current baseline), while the most optimistic future judgments average to about 54% w/w (which is 80% improvement over the current baseline). The main breakthroughs that would enable the use of such higher concentrations are reported to be solving corrosion problem, special additives, corrosion inhibitors, improved metallurgy in the absorber, specially formulated amines, and lower oxygen content in flue gas.

6.3.3. Parameter 3: Sorbent Loss

The current baseline for sorbent loss is about 1.5 kg sorbent lost per tonne of CO_2 captured. Although most of the experts believe that the sorbent loss in future amine systems will be lower than that in the current systems, there are few experts who think otherwise. Hence the range of "best guess" estimates of improvement includes a few negative numbers. Across all experts, however, they average out to about a 50% improvement, *i.e.*, the sorbent losses in future systems are most likely to be about half of the current levels (*i.e.* about 0.76 kg sorbent/ tonne CO_2). The most optimistic judgments averaged across all the experts predict a much lower sorbent loss (on an average, about

0.36 kg sorbent/ tonne CO_2) in future systems, which is more than a 75% improvement over the current baseline. The key research requirements that can lead to such performance include better sorbents, better inhibitors, better design and operating conditions, and zero oxygen content in flue gas.

Table 6.3. Relative improvement in sorbent loss

Based on	Average	Range
"Best Guess" future judgments	49%	(33%)-97%
"Most Optimistic" future judgments	76%	29%-100%

6.3.4. Parameter 4: Sorbent Cost

While the current baseline cost of the amine (MEA) sorbent is about \$1.2/ kg sorbent, the expected improvements in future amine systems may come at a cost premium.

Table 6.4. Relative improvement in sorbent cost

Based on	Average	Range
"Best Guess" future judgments	(48%)	(140%)-4%
"Most Optimistic" future judgments	3%	(100%)-71%

Based on the "best guess" estimates of each expert, on average the future sorbents may cost almost 50% more. The most optimistic judgments estimated that the cost of future sorbents may be almost the same as that of the current sorbents, or even slightly cheaper. According to the various experts, there are several factors that may affect the cost of future sorbents, *viz.* natural gas (the ultimate feedstock) availability and prices, demand, and generic versus specialized application.

6.4 Estimation of Future Cost Reductions

Next, we sought to estimate the process cost reductions that would result from the improvements in future amine systems as envisaged by the various experts. For this purpose, we used the experts' responses about the four parameter values noted above as input to the IECM-CS model. All other model parameters were kept at their default

values. In these cost estimates, we assumed that the CO_2 product is compressed to 2000 psig. The cost of CO_2 transport and storage is not included.

The main cost parameters of interest are the capital cost of the CO_2 capture system, the incremental cost of electricity and the cost of CO_2 avoidance. Table 6.5, Table 6.6, and Table 6.7 give the estimated cost reductions as percentages of the estimates for current systems.

Table 6.5. Expected reduction in capital cost from improvements in four process parameters (relative to current baseline)

Based on	Average	Range
"Best Guess" future judgments	6%	(2%)-9%
"Most Optimistic" future judgments	16%	7%-19%

Table 6.6. Expected reduction in incremental COE from improvements in four process parameters (relative to current baseline)

Based on	Average	Range
"Best Guess" future judgments	18%	(8%)-29%
"Most Optimistic" future judgments	35%	20%-37%

Table 6.7. Expected reduction in cost of CO_2 avoidance from improvements in four process parameters (relative to current baseline)

Based on	Average	Range
"Best Guess" future judgments	18%	(8%)-30%
"Most Optimistic" future judgments	36%	21%-38%

It must be noted that these results reflect expert judgments on only 4 parameters of the amine system model. Even in this case, the average expected cost reductions, especially in the incremental cost of electricity and cost of CO_2 avoidance, are quite encouraging. There are many more parameters that may help to reduce overall system costs, *e.g.*, better power plant heat integration, improved column designs, shorter construction times and improvements to the base plant characteristics (lower heat rate, lower air leakage *etc.*). Thus, future amine-based CO_2 capture technology may be a more cost effective option to control greenhouse gas emissions from fossil fueled power plants in the next two decades or so. However, future improvements in this technology (and resulting cost reductions)

will depend a lot on continuing R&D investments and incentives for deployment resulting from government actions and policies (Taylor and etal 2003).

6.5 Assessing R&D Priorities

In the last section of the questionnaire, (see *Appendix E*), experts were asked to indicate their R&D priorities as to reduce the cost of CO_2 capture and the cost of CO_2 avoidance using amine-based system by year 2015.   Items (B1-B13) are related specifically to amine-based systems, while items (A1-A4) are related to the reference plant characteristics and items (C1-C3) are related to post-capture processing of CO_2 product stream.   In all, 19 research objectives were listed, and the experts were requested to classify these items into the following three categories:

- High Priority (H)

- Medium Priority (M)

- Low Priority (L)

Table 6.8 briefly summarizes the experts' responses to this question.   While there were diverse views on the relative importance of each research objective, all the experts agreed (almost unanimously) on the following items as the top priority issues:

1) To develop sorbents with lower regeneration energy requirement [B4]

2) To develop less expensive technologies for CO_2 storage/ disposal [C3]

3) To improve heat integration within the power plants (to reduce the energy penalty due to steam extraction for sorbent regeneration) [A4]

4) To develop more efficient power plants (lower heat rate) [A1]

Clearly the topmost priority R&D objective (ranked "High Priority" by 82% of the experts) is to develop sorbents having lower regeneration energy requirement. From our

Table 6.8. Summary of research priorities identified by the experts

No.	Research Objective	% of experts who believe that this item is of:		
		High Priority	Medium Priority	Low Priority
A1	To develop more efficient power plants (lower heat rate)	50%	20%	30%
A2	To improve boiler designs so that fuel can be burned with lower excess air (typically for a coal plant, ~20% excess air is used)	40%	10%	50%
A3	To develop more efficient technologies for SO_x and NO_x control so as to reduce the acidic gas impurities in the flue gas stream	27%	27%	45%
A4	To improve heat integration within the power plants to reduce the energy penalty for CO_2 capture due to steam extraction for sorbent regeneration	55%	27%	18%
B1	To develop more efficient fans for flue gas handling	0%	27%	73%
B2	To develop CO_2 absorbers that can handle higher sorbent concentrations	36%	27%	36%
B3	To develop sorbents with higher CO_2 loading capacity	45%	27%	27%
B4	To develop sorbents with lower regeneration energy requirement	82%	18%	0%
B5	To develop absorber columns offering lower pressure drops	18%	73%	9%
B6	To develop absorber columns offering higher CO_2 capture efficiencies	27%	27%	45%
B7	To develop more efficient pumps for sorbent circulation	0%	9%	91%
B8	To develop more efficient heat exchanging devices	0%	45%	55%
B9	To reduce the cost of sorbent manufacturing	9%	64%	27%
B10	To develop sorbents with lower makeup requirements (less losses)	27%	64%	9%
B11	To develop less expensive technologies for disposal of spent sorbents	18%	45%	36%
B12	To develop better instrumentation/ automation in the CO_2 capture system so as to reduce the labor requirement	0%	18%	82%
B13	To develop better construction materials so as to reduce the losses due to corrosion	27%	36%	36%
C1	To develop more efficient compressors for CO_2 compression	0%	60%	40%
C2	To develop a transport technology that can handle low-pressure CO_2 streams	9%	18%	73%
C3	To develop less expensive technologies for CO_2 storage/ disposal	64%	36%	0%

previous analysis, this parameter is crucial in determining the overall energy penalty of this system, and hence affects the overall cost of CO_2 avoidance. The other complimentary factor that helps determine the overall energy penalty is the level of heat

integration between the power plant and the amine system. Significant improvements are required on these fronts so as to make amine-based CO_2 capture economically competitive. Researchers around the world have been working in this direction, and the initial results seem to be promising (Mimura, Matsumoto et al. 2000; Iijima and Kamijo 2002; Veawab, Tontiwachwuthikul et al. 2002; Reddy and Roberts 2003).

References (Chapter 6)

Desideri, U. and A. Paolucci (1999). "Performance modelling of a carbon dioxide removal system for power plants." Energy Conversion and Management **40**: 1899-1915.

Frey, H. C. (1991). Probabilistic Modeling of Innovative Clean Coal Technologies: Implications for Technology Evaluation and Research Planning. Department of Engineering and Public Policy. Pittsburgh, PA, USA, Carnegie Mellon University**:** 605.

Hendriks, C. (1994). Carbon Dioxide Removal from Coal-fired Power Plants. The Netherlands, Kluwer Academic Publishers.

Iijima, M. and T. Kamijo (2002). Flue gas CO2 recovery and compression cost study for CO2 enhanced oil recovery. Sixth International Conference on Greenhouse Gas Control Technologies, Kyoto, Japan.

Jeremy, D. and H. J. Herzog (2000). The cost of carbon capture. presented at the Fifth International Conference on Greenhouse Gas Control Technologies, 13-16 August, Cairns, Australia.

Laboratories, A. P. P. (1999). A design study of the application of CO2/O2 combustion technology as retrofit to an existing coal fired boiler. Calgary, Alberta, TransAlta Corp.

Leci, C. L. (1996). "Financial implications on power generation costs resulting from the parasitic effect of CO2 capture using liquid scrubbing technology from power station flue gases." Energy Conversion and Management **37(6-8)**: 915-921.

Mariz, C. and e. al. (1999). Recovery of CO2 from flue gases: commercial trends (paper No. 340). the Canadian Society of Chemical Engineers Annual Meeting, Saskatchewan, Canada.

Mimura, T., K. Matsumoto, et al. (2000). Development and application of flue gas carbon dioxide recovery technology. Fifth Greenhouse Gas Control Technologies, Cairns, Australia.

Morgan, M. G. and M. Henrion (1990). Uncertainty: A Guide to Dealing with Uncertainty in Quantitative Risk and Policy Analysis. Cambridge, UK, Cambridge University Press.

Morgan, M. G. and D. W. Keith (1995). "Subjective Judgments by Climate Experts." Environmental Science & Technology **29**(10): 468A-476A.

Parsons Infrastructure & Technology Group, I. (2002). Evaluation of fossil fuel power plants with CO2 recovery. Pittsburgh, PA, USDOE/NETL/EPRI.

Rao, A. B. and E. S. Rubin (2002). "A Technical, Economic, and Environmental Assessment of Amine-based CO_2 Capture Technology for Power Plant Greenhouse Gas Control." Environmental Science and Technology **36**: 4467-4475.

Reddy, S. and C. A. Roberts (2003). ECONOMINE FGSM Plus: An Enhanced Amine Based CO_2 Capture Process. 2nd Annual Conference on Carbon Sequestration, Alexandria, VA, USA.

Simbeck, S. (1998). A portfolio selection approach for power plant CO2 capture, separation and R & D options. Fourth Greenhouse Gas Control Technologies, Switzerland.

Smelser, S. C., R. M. Stock, et al. (1991). Engineering and economic evaluation of CO2 removal from fossil-fuel-fired power plants, EPRI IE-7365, Volume 1, Project 2999-10, a research project final report prepared by Fluor Daniel Inc., for EPRI and IEA.

Taylor, M. R. and etal (2003). "The influence of government actions on innovative activities in the development of environmental technologies to control sulfur dioxide emissions from stationary sources." submitted to ES&T.

Taylor, M. R. and etal (2003). "The influence of government actions on innovative activities in the development of environmental technologies to control sulfur dioxide emissions from stationary sources." submitted to ES&T.

Veawab, A., P. Tontiwachwuthikul, et al. (2002). Performance and cost analysis for CO2 capture from flue gas streams: Absorption and regeneration aspects. Sixth International Conference on Greenhouse Gas Control Technologies, Kyoto, Japan.

7. CONCLUDING REMARKS

The basic objective of this research was to carry out a technical, environmental, and economic assessment of the amine-based CO_2 capture technology for greenhouse gas control at power plants. This chapter summarizes the key findings presented in this thesis.

7.1 Model Development

Performance and cost model of this technology were developed that served as the analytical tool for this assessment. The performance model was derived using a detailed chemical process simulator together with data obtained from process developers and through personal communication with experts. It was directly linked to the cost model, which used cost data from published literature. Cost estimates for any technology that has not been widely implemented for the application of interest (in this case, large-scale power generation) is inherently subject to uncertainty. Thus, the model has probabilistic capabilities, which allow the user to use uncertainty distributions rather than a single value for various model parameters. In the model applications discussed below, such distributions were based on multiple data sources *viz.*, published literature, technical reports, and experts' technical judgment. Finally, the amine system model was integrated with a larger fossil-fuel based power generation modeling framework (IECM) in order to study multi-pollutant interactions and to assess various CO_2 control options on a consistent basis.

7.2 Model Applications

Today there is a growing consensus in the global scientific community on the need to control CO_2 emissions to mitigate the problem of global climate change problem. Capture and sequestration of CO_2 is being proposed as a medium term solution that can allow continued use of conventional fossil-fuel resources over the next century without significantly contributing to global carbon emissions from electric power generation (and, potentially, other industrial sources). Electric utilities, being one of the largest stationary sources of CO_2 emissions, are likely to be the early targets in this effort. Thus, the performance and cost model developed in this research were used to study the option of post-combustion CO_2 capture from a new PC plant using amine-based system. This can

serve as a reference against which other CO_2 mitigation options could be compared. Also, as we have discussed before, there exists a large existing fleet of coal-fired capacity that may have to be retrofitted under a stringent greenhouse gas control policy. Thus, retrofit applications as well as new plant applications are of interest.

7.2.1. New Plants

Using the model, we conducted case studies of a new coal-fired power plant with an amine-based CO_2 capture plant. It was concluded that in order to capture CO_2 from a pulverized coal plant, a much bigger plant has to be built in order to provide the same net power output as a plant without CO_2 capture. This is because approximately 22% of the gross power generation is consumed by the CO_2 capture and compression systems. It was found that a capture plant designed to capture 90% of the CO_2 mass emissions effectively "avoids" about 87% of the emissions per net kWh generated because the CO_2 capture system itself consumes substantial energy and contributes to CO_2 emissions. An amine system also reduces the total emissions of SO_x and particulates to very low levels. However, the capture plant has higher emission rates of NO_x and ammonia (per kWh). The capture plant uses 33% more coal to deliver the same amount of electricity, and therefore also has larger amounts of bottom ash, fly ash and FGD waste to be disposed. The capture plant also has to dispose of the spent sorbent (from the amine system) and the CO_2 product stream.

The addition of CO_2 capture and storage increases the capital requirement as well as the cost of electricity generated. Our case studies showed the capital cost of the new plant to increase by 56% to 66% (based on $/kWh), while the COE increased by about 73% to 82% for a plant with 90% capture. Cost results were found to be sensitive to assumptions about the plant capacity factor and fixed charge factor.

The CO_2 avoidance cost was estimated to be about $50/ tonne CO_2 avoided. As mentioned earlier, it is higher than the CO_2 capture cost ($36/ tonne CO_2 delivered). This estimate includes about $8/ tonne CO_2 captured for transport and storage/ disposal of the concentrated CO_2 product stream. From the probabilistic analysis, the 95% confidence interval was estimated as $39-71/ tonne CO_2 avoided. It may be noted that these results

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are within reasonable agreement with some of recent estimates published by other studies (Simbeck and McDonald 2000; Stork Engg Consultancy B.V. 2000; ALSTOM, AEP et al. 2001; Parsons Infrastructure & Technology Group 2002; Singh, Croiset et al. 2002).

7.2.2. Retrofit Applications

Retrofit application of this technology to existing coal-fired plants is another important scenario to be considered. Several cases have been modeled and presented in section 5.2. If the existing plant already has a sulfur-control system, it helps in reducing the cost of CO_2 capture using amine-based systems. Optimization studies showed that it is further helpful if the existing sulfur-control system is upgraded to the maximum achievable removal efficiency to minimize SO_2 removal in the amine system. It is necessary to add a new, highly efficient sulfur scrubber upstream of amine-based system if one does not already exist at the reference plant. If all the energy required for CO_2 capture and compression system is extracted from the base plant ("derating" case), the net output of the plant falls substantially (by almost 40%), much more so that at a new plant, because of generally poorer heat integration potential at older plants. Accordingly, the cost of electricity may go up by 150-300% as compared to that of reference plant, for various cases analyzed. Adding an auxiliary natural-gas-fired boiler (with a secondary steam turbine) may be a lower-cost option for providing auxiliary energy if lower levels of CO_2 avoidance (~ 60-65%) are acceptable and if natural gas is available at low price. The flue gas from the auxiliary natural gas boiler is more dilute in CO_2 content, so capturing this CO_2 is not cost effective. This additional CO_2 emission from the auxiliary natural gas boiler reduces overall CO_2 avoidance level of the system. The cost of electricity (and that of CO_2 avoidance) in this case may be lower than that of the derating case; however, it is very sensitive to the cost and availability of natural gas and assumptions about the retrofit factors. This option might be attractive if the gas price is below $2.5/mscf.

7.2.3. Capacity Factor Assumption

Typically, in a plant-level analysis like this, it is assumed that the CO_2 capture plant has the same capacity factor as that assumed for the reference plant (Jeremy and Herzog 2000; Simbeck and McDonald 2000; Stork Engg Consultancy B.V. 2000; ALSTOM, AEP et al. 2001; Parsons Infrastructure & Technology Group 2002; Singh, E.Croiset et

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al. 2002). The resulting cost difference with CO_2 capture then reflects the added cost under identical operating conditions. In practice, however, any number of parameters affecting overall plant costs – such as the capacity factor – may differ in plants with and without a CO_2 capture system. Based on dispatch models for two regions of the U.S., recent studies by Johnson and Keith found that the CO_2 capture plants would be operated at higher capacity factors (Johnson 2002; Johnson and Keith 2004). However, there are other factors that could discourage the higher utilization of these plants. For example, plants using amine-based CO_2 capture system were shown earlier to have NO_x emission rates that are substantially higher than that for a similar reference plant without CO_2 capture (see Table 5.2). This could reduce their use under regional NO_x caps. In general, the capacity factor of a plant will be determined by a number of factors related to the overall generation mix and the constraints under which plants in a utility system operate. While such factors are clearly important, they are beyond the scope of this thesis. However, the models developed here can be readily used to explore the cost implications of alternative assumptions regarding the capacity factor and other base plant parameters.

7.3 Expert Elicitations

The technical judgment of various experts in this field was elicited to explore the potential for reducing the cost of current amine systems. The experts belonged to academia, industries, research laboratories and private consultancies.

7.3.1. Scope for improvement

The experts' responses indicate that future amine-based systems can perform significantly better as compared to their current counterparts, given a substantial program of R&D. The average relative improvement in various parameters as predicted by the experts is reported below in Table 7.1. The numbers in the parentheses represent the range of estimates provided by various experts.

Various possible strategies are cited to achieve these improvements. Most experts emphasize development of better sorbents (advanced amines or special mixtures of amines). Other items on the list of suggestions included equipment change, special additives, corrosion inhibitors, improved metallurgy for absorbers, better design and

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operating conditions, and reducing the oxygen levels in the flue gas to zero. All these improvements in performance parameters will help in reducing the future cost of CO_2 avoidance.

Table 7.1. Relative improvement in amine-system parameters

Model Parameter	Improvement estimates based on	
	"Best Guess" judgments	"Most Optimistic" judgments
Sorbent Regeneration Heat Requirement (kJ/ kg CO_2)	23% (5%-40%)	43% (15%-73%)
Sorbent Concentration (wt %)	23% (0%-67%)	80% (33%-160%)
Sorbent Loss (kg MEA/ tonne CO_2)	49% ((33%)-97%)	76% (29%-100%)
Sorbent Cost ($/ tonne MEA)	(48%) ((140%)-4%)	3% ((100%)-71%)

However, the cost of improved sorbents is expected to be substantially higher than current prices, or at best remain at the current level under the most optimistic scenario. Higher sorbent costs were also attributed to natural gas (the ultimate feedstock) availability and prices, demand, and generic versus specialized application.

7.3.2. Possible future cost reductions

The experts' responses about the four parameter values for future amine systems (Table 7.1) were used as input to the IECM-CS model along with the other parameter defaults. Table 6.2 gives the results obtained using the average improvement numbers reported in Table 7.1 (in bold), with the range obtained using individual responses given in the parentheses.

It may be noted that all the four parameters under consideration, *viz.* sorbent regeneration heat requirement, sorbent concentration, sorbent loss and sorbent cost, directly affect the total energy penalty and makeup sorbent requirement cost (the most important variable cost items of this technology). However these parameters only indirectly affect the

capital cost of this system and hence have less impact on the same. This fact is reflected in the cost reductions that are expected to be lower in case of capital requirement as compared to those in cost of electricity.

Table 7.2. Expected reduction in future cost of CO_2 avoidance

Cost Parameter	Expected cost reduction based on	
	"Best Guess" judgments	**"Most Optimistic" judgments**
Total Capital Cost (M$)	**6%** ((2%)-9%)	**16%** (7%-19%)
Cost of Electricity ($/MWh)	**18%** ((8%)-29%)	**35%** (20%-37%)
Cost of CO_2 avoidance ($/ tonne CO_2 av.)	**18%** ((8%)-30%)	**36%** (21%-38%)

7.3.3. *Probabilistic distribution of current and future costs of CO_2 avoidance*

In case of probabilistic runs, most of the experts seem to be optimistic about the potential for substantial cost reductions in future.

Figure 7.1 shows the cumulative probability distributions for avoidance cost based on the four parameter distributions for current and future systems. The average of the experts' responses have been used as input to the model to generate these curves. The distribution for future costs shows substantially lower values than the current cost distribution. The cost difference (cost reduction) maybe accredited to the R&D inputs being invested to improve this technology. In other words, this cost difference or cost saving maybe looked upon as the benefit of R&D investments in this technology. From Figure 7.1, the difference in the median values of mitigation cost in current versus future scenario is about $8/ tonne CO_2 avoided, or about 20% of the current value; and the difference between the average value of each distribution is about $9/ tonne CO_2 avoided, or about 21% of the current value. When the distributions are derived for each individual expert, the difference in the median as well as the average values of mitigation cost is found to

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have a range of \$2-9/ tonne CO_2 avoided, or about 3-24% of the current value. Based on these measures of the benefit of R&D, the wider the application of this technology, the larger the number of tonnes of CO_2 avoided and the bigger the overall benefit.
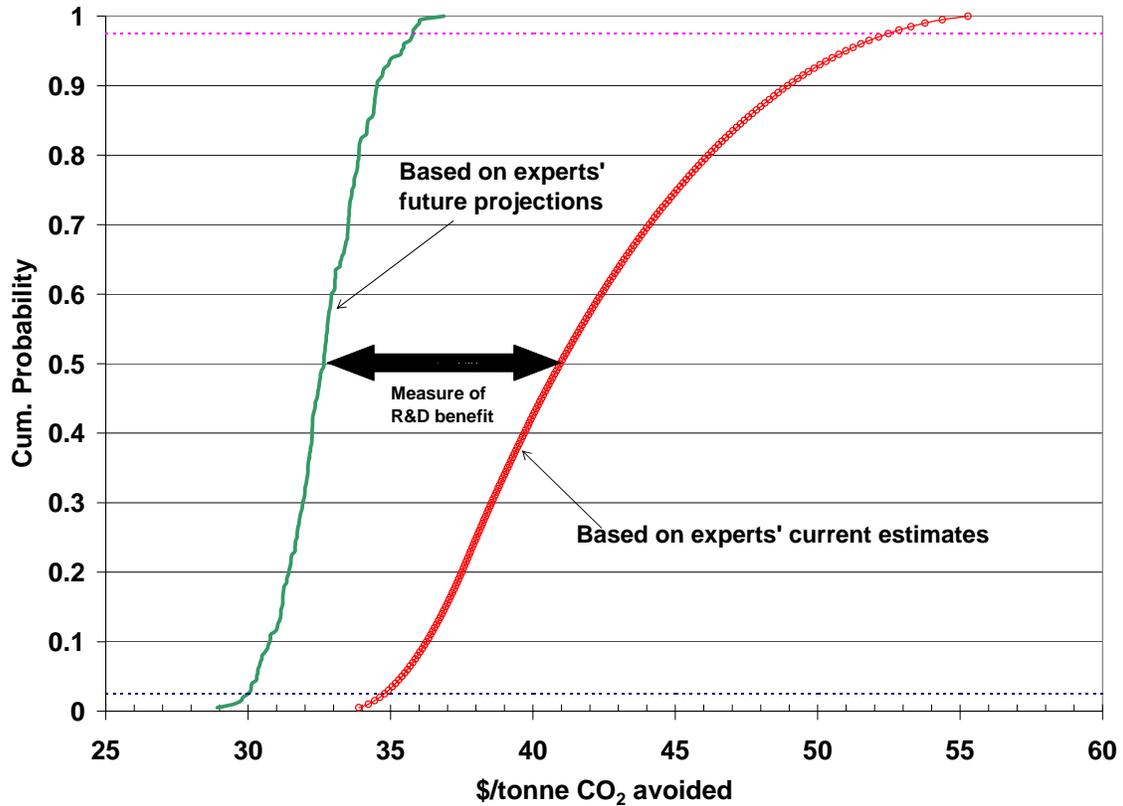
Figure 7.1. Probabilistic estimate of future cost reduction

7.3.4. R&D priorities

Although there were diverse views on the relative importance of various research objectives, all the experts seem to agree (almost unanimously) on the following items as the top priority issues:

5) To develop sorbents with lower regeneration energy requirement [B4]

6) To develop less expensive technologies for CO_2 storage/ disposal [C3]

7) To improve heat integration within the power plants (to reduce the energy penalty due to steam extraction for sorbent regeneration) [A4]

8) To develop more efficient power plants (lower heat rate) [A1]

So, the topmost priority R&D objective is believed to be to develop sorbents having lower regeneration energy requirement. Together with more effective heat integration schemes, this may be the key to the success of this technology for CO_2 capture applications in future. Various researchers around the world have been working in this direction, and the initial results seem to be promising

7.4 Directions for Future Work

There are several directions in which this work may be furthered. First, the post-capture CO_2 processing modules (for transport, storage or usage of CO_2 product) could be developed in more detail. Then, there are also a few areas in the current performance and cost models for amine system that could be improved with the availability of additional data.   For example, it is likely that there are correlations between some of the performance and cost parameters (e.g. sorbent concentration, lean sorbent loading and maintenance costs) which are not established in the current model. Also, default values for various parameters for CO_2 absorption systems using sorbents other than MEA could be included in the present model.

It would also be useful to elicit appropriate group of experts on other parameters of interest, such as equivalent energy penalty factor (F_{HE}), direct and indirect capital costs, *etc*. A more elaborate (and time consuming) protocol for eliciting expert judgments, including greater reliance on face-to-face interviews, also may help in better understanding of their views.

Another direction of future work might be applying the models developed here to address climate policy issues more extensively. For example, a power plant database could be used with the model to assess the feasibility of retrofitting amine-based systems to existing coal plants, the emission reductions that realistically could be achieved through such an effort, and the overall cost of such a program. Another application might be in the form of a case study of a prospective power plant sited in a developing country with site-specific cost data and regulatory framework. This work might also be extended in the direction of modeling other technological options for CO_2 capture at power plants,

including advanced combustion systems using pure oxygen. IECM-CS can provide a consistent framework for comprehensive comparisons of various technological options for carbon sequestration.

References (Chapter 7)

ALSTOM, AEP, et al. (2001). Engineering feasibility and economics of CO2 capture on an existing coal-fired power plant, Final report prepared by ALSTOM Power Inc., ABB Lummus Global Inc., ALSTOM Power Environmental Systems and American Electric Power (AEP). report no. (PPL-01-CT-09) submitted to Ohio Department of Development, Columbus, OH and US Department of Energy/ NETL, Pittsburgh, PA**:** 163.

Jeremy, D. and H. J. Herzog (2000). The cost of carbon capture. presented at the Fifth International Conference on Greenhouse Gas Control Technologies, 13-16 August, Cairns, Australia.

Johnson, T. L. (2002). Electricity without carbon dioxide: assessing the role of carbon capture and sequestration in US electric markets. Department of Engineering and Public Policy. Pittsburgh, PA 15213, Carnegie Mellon University**:** 247.

Johnson, T. L. and D. W. Keith (2004). "Fossil electricity and CO2 sequestration: how natural gas prices, initial conditions and retrofits determine the cost of controlling CO2 emissions." Energy Policy **32**: 367-382.

Parsons Infrastructure & Technology Group, I. (2002). Evaluation of fossil fuel power plants with CO2 recovery. Pittsburgh, PA, USDOE/NETL/EPRI.

Simbeck, D. and M. McDonald (2000). Existing coal power plant retrofit CO_2 control options analysis. presented at the Fifth International Conference on Greenhouse Gas Control Technologies, 13-16 Augus, Cairns, Australia.

Singh, D., E. Croiset, et al. (2002). Techno-economic study of CO2 capture from an existing coal-fired power plant: MEA scrubbing vs. O2/CO2 recycle combustion. Waterloo, Ontario, Canada**:** 27.

Singh, D., E.Croiset, et al. (2002). Economics of CO2 capture from a coal-fired power plant - a sensitivity analysis. Sixth Greenhouse Gas Control Technologies, Kyoto, Japan.

Stork Engg Consultancy B.V., A. (2000). Leading options for the capture of CO2 emissions at power stations. UK, IEA GHGRDP.

APPENDIX A

POTENTIAL ENVIRONMENTAL IMPACTS OF AMINE SYSTEMS

Amine system is an "end-of-pipe" treatment. It captures CO_2 generated during combustion of a fossil fuel in a power plant. Although an end-of-pipe solution is less desirable than pollution prevention approaches, it is a common practice to reduce environmental impact of industrial activities (Rubin 2001). Often it is observed that the environmental control technologies, especially those based on chemical processes, transfer the pollutants (or their derivatives) from one phase (or part of the ecosystem) to another phase (or part of the ecosystem). Many times, this could result in "pollution transfer" rather than "pollution control." For example, a flue gas desulfurization (FGD) system that controls SO_2 emission from a power plant flue gas produces "FGD waste," a solid waste product that needs to be managed later. Many of the wastewater treatment plants separate out contaminants from the water in the form of a solid residue, which again has to be taken care of. In some cases, as in the case of FGD waste, it is possible to make use of this solid product in the manufacture of construction materials. This is the most benign strategy for the environmental protection, and is a good example of industrial ecology. However, this may not be always feasible and some of the solid wastes may have to be disposed of. Depending upon the chemical nature of such solid waste, it is either stored in a landfill or incinerated. Precautions must be taken while disposing the solid waste to avoid creating a new problem in the form of soil pollution, groundwater contamination or air pollution.

In case of MEA-based CO_2 capture systems, environmental problems may arise from the spent sorbent slurry discharged from the MEA reclaimer (also called *reclaimer bottoms*) and the emissions of MEA and ammonia (NH_3) carried by the treated flue gas. Here is a brief description of these issues. It may be noted that the large concentrated CO_2 stream generated as a "byproduct" of this process may have its own set of concerns, and is beyond the scope of this study.

A. Spent Sorbent or Reclaimer Bottoms

As mentioned before, MEA degrades under the real operating conditions in a plant. Typically, the degradation products are separated in a reclaimer and disposed of as hazardous waste (Barchas and Davis 1992). The waste is often picked up by a contactor and is incinerated at a permitted facility. This method of waste disposal is expensive and the CO_2 capture plant may have to spend substantial amount depending upon the waste quantity.

Recent study undertaken at the National Energy Technology Laboratory (NETL), Pittsburgh, has identified various chemical species that are found in the reclaimer waste samples from an operational CO_2 capture plant (Strazisar, Anderson et al. 2003). The major organic compounds which were found in these samples include monoethanolamine (C_2H_7NO, MEA), 3-hydroxyethylamino-N-hydroxy-ethyl propanamide ($C_7H_{16}N_2O_3$), 4-hydroxyethyl-2-piperizinone ($C_6H_{12}N_2O_2$), 2-hydroxyethylamino-N-hydroxyethyl acetamide ($C_6H_{14}N_2O_3$), Ammonia (NH_3), and N-acetylethanolamine ($C_4H_9NO_2$) with traces of several other compounds.

The analysis of inorganic species revealed the presence of seven metal cations at a concentration of 0.2 ppm or more. Since sodium carbonate is added to the reclaimer to regenerate MEA, sodium ions dominate all other metal ions (> 800 ppm). Other metals (potassium, selenium, arsenic, calcium, iron, aluminum, and zinc) are believed to originate mostly from the coal. Additionally, mercury was found to be present in the reclaimer waste at a concentration of 1 ppb. Anion concentrations (chloride, nitrate, fluoride, sulfate, phosphate, bromide) are found to be relatively higher (80-49000 ppm). The amounts of halide anions match the typical halogen concentrations found in coal. This might be the result of "heat-stable-salts" formed by reaction between MEA and mineral acids (HCl, HF, HBr) in the flue gas.

This analysis helps to understand why the spent sorbent (reclaimer bottoms) is treated as hazardous waste. From economic as well as environmental perspective, it is preferable for the CO_2 capture plant to cleanse the flue gas upstream, before entering the amine

system, so as to minimize sorbent degradation losses, corrosion problems, makeup sorbent requirement and cost of disposing the spent sorbent.

B. Environmental (Gaseous) Emissions

a) MEA entrainment with treated flue gas: As mentioned before, MEA is a primary amine with high pH. It is completely soluble in water and is readily biodegradable (Dow 2000). It is a reactive compound and is classified as a corrosive material. So, proper care needs to be taken while handling this compound. Toxic gases and vapors (such as oxides of nitrogen and carbon monoxide) may be released in a fire involving MEA. It can cause irritation and be harmful to gastrointestinal tract, nausea, vomiting and diarrhea when swallowed. It is an irritant to eyes and continued contamination of the eyes could cause permanent injury. MEA is also known to cause mild irritation to skin and repeated or prolonged contact may cause dermatitis.   Inhalation of mist or aerosols can produce severe respiratory irritation and cause sore throat, coughing, headache, nausea and vomiting. In terms of chronic effects, lungs may be affected with repeated or prolonged exposure. MEA may have effects on the central nervous system, gastrointestinal tract and kidneys (Deltrex 2000). Several animal studies were conducted by exposing rats and rabbits to MEA vapors of 12-102 ppm for several weeks. Skin was found to be the significant target tissue at relatively high concentration of MEA vapor; other sensitive target tissues include the kidneys and livers (Knaak, Leung et al. 1997).   Based on various laboratory tests, it is found to be lacking in genotoxicity.

MEA is not listed as a hazardous air pollutant under the Clean Air Act of 1990 (CAA 1990).   The National Institute for Occupational Safety and Health (NIOSH) has established a recommended exposure limit (REL) for MEA of 3 ppm (8 mg/m3) as a time-weighted average for a normal 10-hour workday and a 40-hour workweek and a short-term exposure limit (STEL) of 6 ppm (15 mg/m3) for periods not to exceed 15 minutes (OSHA 1996). These limits are based on the risk of skin, eye and respiratory irritation and narcotic effects. However, no signs or symptoms of chronic exposure to ethanolamine have been reported in humans. MEA has been regarded as hazardous according to the criteria of Worksafe Australia (Deltrex 2000). Based on the LD_{50} value of 505 mg/ kg body weight (in rabbits exposed by dermal route), MEA is classified as a

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"dangerous" chemical that falls within the European Risk Category 3 (Chemrest 2003). The European Toxicity Risk Code for this chemical is "XI = Irritant" which matches with the hazard category "Harmful/Irritant" assigned by Worksafe Australia (Deltrex 2000). It may be noted that MEA is widely used in various applications such as a dispersant for agricultural products; as a softening agent for hides; as an accelerator in the production of antibiotics, polishes and waving solutions for hair; as a rubber accelerator; an intermediate in the production of emulsifiers, soaps and detergents; and in some hair-care products. In case of MEA-based CO_2 capture systems, trace quantities (1-4 ppm) of MEA may be found in the exhaust flue gas at the stack. Under well-mixed atmospheric conditions, it is less likely to pose a risk of direct exposure to human beings. Also, only a small amount of MEA is expected to partition to the atmosphere. It may be removed by reactions with photochemically generated hydroxyl radicals or by precipitation. Thus, relatively low levels of MEA are expected to be present in the atmosphere; and owing to its short half-life, it is unlikely to adversely impact air quality (Davis and Carpenter 1997). However, MEA's complete solubility in water may facilitate its entry into surface and ground water.

According to the Dow Chemical Company (one of the leading manufacturers of MEA), MEA is "practically non-toxic" to aquatic organisms (Dow 1998). It is claimed that as MEA exhibits low (or no) aquatic toxicity, is biodegradable in soil, surface waters and wastewater treatment plants and does not show any tendency to bioaccumulate, the routine manufacture, use and disposal of this compound is not expected to adversely impact the environment (Davis and Carpenter 1997). However, U.S. Environmental Protection Agency has classified MEA as having "moderate" ecological toxicity, based on the effects on aquatic organisms, such as fish and algae (USEPA 2002). This aspect might need further careful investigation before MEA-based CO_2 capture systems are installed in large numbers.

It may be noted that the market cost of MEA sorbent lost via leakage or entrainment is likely to be trivial in the overall cost of running the plant. However the externalities and potential risks posed by the same might be significantly higher. In view of the above information, considering the potential health and environmental risks posed by MEA, it is

important to take special care to avoid MEA losses through entrainment in the absorber and leakage in the circulation piping.

b) Nitrosamines: Another potential problem that may arise from amine-based systems is in the form of "nitrosamines", which are known carcinogens that are formed by a reaction between an amine and a nitrogen oxide. These compounds were found in the samples of the lean sorbent (2.91 μmole/ ml). However, detectable amount of nitrosamines was not found in the reclaimer waste, which may be explained in terms of their low boiling point (Strazisar, Anderson et al. 2003). The trace quantities of MEA entrained by the flue gas may combine with nitrogen oxides in the atmosphere (or present in the flue gas itself) and may constitute to health hazards in the surrounding areas.

c) Ammonia: Ammonia (NH_3) is generated as a dissociation product of MEA. Loss of MEA (and emission of NH_3) via dissociation route is unavoidable under the normal operating conditions in the amine system. As seen in the case study of a new plant (Section 5.1 & *Appendix D*), the emission rate of NH_3 from a CO_2 capture plant is substantially higher than that from a power plant without CO_2 capture system. Emission of ammonia is not regulated under the current air pollution control programs, however, it is a known source of air (and water) pollution (Ashman, Sutton et al. 1998; Howarth, Sharpley et al. 2002). Especially in the polluted urban areas, changes in ammonia emissions have the most significant impact on the total particulate matter levels (Ansari and Pandis 1998). The treated flue gas coming from a capture plant is almost free of SO_2, but it contains a significant amount of NO_x. Gaseous emissions of ammonia, SO_2 and NO_x form secondary particulates (ammonium nitrate (NH_4NO_3) and ammonium sulfates (NH_4HSO_4 and $(NH_4)_2SO_4$)). These compounds are significant part of fine particulate matter ($PM_{2.5}$) in air and can play an important role in visibility impairment and regional haze (Battye, Aneja et al. 2003). Hence, the multi-fold increase in NH_3 emissions from a power plant owing to the amine-based CO_2 capture system has to be considered carefully in the light of local and regional air pollution issues. The contribution from all the industrial sources combined makes only a small fraction (0.2%-9.6%) of total ammonia emission inventory that is dominated by emissions from livestock waste and fertilizer application (Battye, Aneja et al. 2003). However, these

point sources of ammonia, especially in the vicinity of urban centers, can have substantial impact on the particulate matter levels and the air quality in general. So, the severity of the impacts of the higher ammonia emissions from the capture plant will significantly depend on the land-use pattern around the power plant.

References

Ansari, A. S. and S. N. Pandis (1998). "Response of inorganic PM to precursor concentrations." Environmental Science and Technology **32**(18): 2706-2714.

Ashman, W. A. H., M. A. Sutton, et al. (1998). "Ammonia: emission, atmospheric transport and deposition." New Phytol. **139**: 27-48.

Barchas, R. and R. Davis (1992). "The Kerr-McGee/ ABB Lummus Crest technology for the recovery of CO2 from stack gases." Energy Conversion and Management **33**(5-8): 333-340.

Battye, W., V. P. Aneja, et al. (2003). "Evaluation and improvement of ammonia emissions inventories." Atmospheric Environment **37**: 3873-3883.

CAA (1990). Clean Air Act Amendments. Public Law 101-549, Section 112, Nov. 15, 1990.

Chemrest (2003). Chemical toxicity and risk information: Ethanolamine (http://www.chemrest.com/Toxicity%20and%20Risk%2020Codes/Ethanolamine.htm).

Davis, J. W. and C. L. Carpenter (1997). "Environmental assessment of the alkanolamines." Reviews of Environmental Contamination and Toxicology **149**: 87-137.

Deltrex (2000). Monoethanolamine 99% (http://www.deltex.com.au/products/s133.htm), Deltrex Chemicals, a Division of Campbell Brothers Ltd., Australia).

Dow (1998). Product information: Ethanolamines - Toxicology Overview [Form No. 111-01338-1198AMS, CH 132-007-E-1198AMS] (wysiwyg://221/http://www.dow.com/alkanolamines/eth/mea.html), The Dow Chemical Company.

Dow (2000). "Specialty Alkanolamines": Product information available at the website of The Dow Chemical Company (www.dow.com/alkanolamines/).

Howarth, R. W., A. Sharpley, et al. (2002). "Sources of nutrient pollution to coastal waters in the United States: Implications for achieving coastal water quality goals." Estuaries **25**(4b): 656-676.

Knaak, J. B., H.-W. Leung, et al. (1997). "Toxicology of Mono-, Di-, and Triethanolamine." Reviews of Environmental Contamination and Toxicology **149**: 1-86.

OSHA (1996). Occupational safety and health guideline for ethanolamine (http://www.osha-slc.gov/SLTC/healthguidelines/ethanolamine/recognition.html), OSHA (Occupational Safety and Health Administration), U.S. Department of Labor.

Rubin, E. S. (2001). Introduction to engineering and the environment. New York, NY, The McGraw-Hill Companies, Inc.

Strazisar, B. R., R. R. Anderson, et al. (2003). "Degradation pathways for monoethanolamine in a CO_2 capture facility." Energy & Fuels **17**(4): 1034-1039.

USEPA (2002). Aqueous and semi-aqueous solvent chemicals: Environmentally preferable choices (http://www.epa.gov/opptintr/solvents/finlchr4.htm), USEPA (U.S. Environmental Protection Agency).

APPENDIX B

CHARACTERIZATION OF UNCERTAINTY AND VARIABILITY

The term "uncertainty" is a vast umbrella that covers a variety of concepts. Uncertainty often arises from various sources including lack of complete information, conflicting sources of information, variability in a process or an object, linguistic imprecision, and approximations incorporated in a model to simplify the real life situations (Morgan and Henrion 1990). Uncertainty in an empirical quantity is usually expressed using a probability distribution.

Any techno-economic analysis, and especially that of new energy and environmental control technologies that are still in the research phase, involves uncertainties regarding the performance and costs. These uncertainties come from incomplete data and numerous assumptions and approximations built into simulations. Some parameters, especially the cost parameters are influenced by a larger set of factors outside the scope of the particular study, and fluctuations in these quantities may be seen as "inherent randomness" when viewed within this limited focus area. In addition, there may be significant variability in plant or process design assumptions across different studies or organizations.

Uncertainty and variability are often ignored or treated in a limited way using sensitivity analysis. However sensitivity analysis suffers from shortcomings resulting from the difficulty in evaluating the effect of simultaneous variations in several parameters and the lack of insight into the likelihood of obtaining any particular result.

A more robust approach is to represent uncertainties and/or variability in model parameters using probability distributions. Using probabilistic simulations techniques, simultaneous uncertainties in any number of input parameters can be propagated through a model to determine their combined effect on model outputs. The result of a probabilistic simulation includes both the possible range of values for model output parameters and information about the likelihood of obtaining various results. The development of ranges and probability distributions for model input parameters can be

based on information available in published studies, statistical data analysis and/or the judgments of process engineers with relevant expertise.

One of the distinguishing features of this modeling effort is a probabilistic capability that allows model inputs to be represented by probability distributions rather than single deterministic values. Depending upon the parameter, these distributions reflect the ranges of values reported in the literature, modeling approximations, the evolving nature of the technology, practical considerations in running such plants and variety of plant or process design assumptions. Some of the distributions are also based on an expert elicitation exercise (explained in Chapter 6) in which experts were asked about the nominal values and possible ranges for several important parameters characterizing the performance of current commercial MEA-based systems capturing 90% CO_2 from the flue gas of a typical coal-fired power plant.

While designing an amine-based CO_2 capture system for a given flue gas (from a power plant application), there are certain parameters that could be specified independent of others. For example, the sorbent concentration, CO_2 capture efficiency target, lean sorbent loading, temperature and pressure of sorbent regeneration steam, CO_2 product pressure, *etc*. Distributions for such parameters reported in different studies essentially represent variability. Experts' technical judgments were useful in defining these distributions.

Probability distributions for parameters such as the efficiencies of fans, pumps or compressors represent the possibility of encountering fluctuations in the performance due to inherent characteristics of these devices in an operating plant. Uncertainties arising from real plant operating conditions and approximations in process simulations are also reflected in the distributions for parameters such as MEA makeup rate, removal efficiencies for other acid gases (SO_2, NO_2 and HCl), and equivalent energy penalty from extraction of sorbent regeneration steam from the steam cycle of the base plant.

In case of cost parameters such as capital costs, O&M costs, cost of reagents (including MEA) and cost of disposal, the distributions represent both variability and uncertainty

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arising from disagreement in data sources and inherent characteristics of market mechanism.

Then there are certain process parameters that are interdependent. For example, the CO_2 capture efficiency (η) is a function of various operating variables of the system for a given column design.

$$\eta_{CO2} = f(y_{CO2}, C, L/G, \phi_{lean}, T_{fg},)$$

where, y_{CO2} = CO_2 concentration in the flue gas,

                  C = sorbent concentration,

                  L/G = sorbent circulation,

                  ϕ_{lean} = lean sorbent loading, and

                  T_{fg} = temperatures of the flue gas

In the process model that we have developed here, the CO_2 capture efficiency (η) is a user-defined parameter (treated as an independent parameter) and sorbent circulation (L/G) is estimated as:

$$(L/G) = f(y_{CO2}, C, \eta_{CO2}, \phi_{lean}, T_{fg})$$

Similarly, sorbent regeneration heat requirement (Q/L) is estimated as:

$$(Q/L) = f(L/G, \phi_{lean}, C)$$

Probability distributions for these calculated parameters are not explicitly defined. Instead, the probability distributions on the independent parameters are propagated to these parameters throughout the model.

Table 3.1 (in Chapter 3) lists the probability distributions developed for performance model parameters. Similarly, Table 4.1 (in Chapter 4) lists the probability distributions developed for some of the cost model parameters. These distributions reflect both uncertainty and variability in system designs. These tables have been reproduced here for convenience. *Appendix C* discusses the basis for these distributions.

Table 3.3. Amine System Performance Model Parameters and Uncertainties

Performance Parameter	Units	Data (Range)	Nominal Value	Unc. Representation (Distribution Function)
CO_2 removal efficiency	%	Mostly 90	90	-
SO_2 removal efficiency	%	Almost 100	99.5	Uniform(99,100)
NO_2 removal efficiency	%	20-30	25	Uniform(20,30)
HCl removal efficiency	%	90-95	95	Uniform(90,95)
Particulate removal eff.	%	50	50	Uniform(40,60)
MEA concentration	wt%	15-50	30	Triangular (20,30,40)
Lean sorbent CO_2 loading	mol CO_2/mol MEA	0.15-0.30	0.2	Triangular(0.1,0.2,0.25)
Nominal MEA make-up	kg MEA/tonne CO_2	0.5-3.1	1.5	Triangular(0.5,1.5,3.1)
MEA loss (SO_2)	mol MEA/mol SO_2	2	2	-
MEA loss (NO_2)	mol MEA/mol NO_2	2	2	-
MEA loss (HCl)	mol MEA/mol HCl	1	1	-
NH_3 generation	mol NH_3/mol MEA oxidized	1	1	-
Caustic consumption in MEA reclaimer	kg NaOH/tonneCO_2	0.13	0.13	-
Activated carbon use	kg C/tonne CO_2	0.075	0.075	-
Gas-phase pressure drop	kPa	4-37	14	Triangular(4,14,37)
Fan efficiency	%	70-80	75	Uniform (70,80)
Sorbent pumping head	kPa	200	200	Triangular(150,200,250)
Pump efficiency	%	70-80	75	Uniform (70,80)
Equiv. elec. requirement	% regeneration heat	9-19	14[a]	Uniform (9,19)
CO_2 product pressure	MPa	7.58-15.16	13.79	Triangular(7.58,13.79,15.16)
Compressor efficiency	%	75-88	80	Uniform (75,88)

[a] for retrofit applications, nominal value is 22.

Table 4.4. Amine System Cost Model Parameters and Uncertainties

Cost Parameter	Units	Nominal Value	Unc. Representation (Distribution Function)
Process facilities	M$	calc	*Normal(1.0,0.1)
Engineering and home office	% PFC	7	Triangular (5,7,10)
General facilities	% PFC	10	Triangular (5,10,15)
Project contingency	% PFC	15	Triangular (10,15,20)
Process contingency	% PFC	5	Triangular (2,5,10)
Royalty fees	% PFC	0.5	Triangular (0,0.5,0.5)
Startup cost	Month of TOM	1	Triangular(0.5,1,1)
Inventory cost	% TPC	0.5	Triangular(0.4,0.5,0.6)
Total maintenance cost	% TPC	2.5	Triangular(1,2.5,5)
Operating labor	Jobs/shift	2	Triangular(1,2,3)
MEA cost	$/ tonne MEA	1200	Uniform (1100,1300)
Waste disposal cost	$/ tonne waste	175	*Normal(1.0,0.1)
CO_2 transport cost	$/ tonne CO_2/ km	0.02	Triangular (0.004,0.02,0.08)
CO_2 storage/disposal cost	$/ tonne CO_2	5	Triangular (0,5,10)
Revenue from CO_2 usage in EOR or ECBM[b]	$/ tonne CO_2	-10[a]	Triangular (-20,-10,0)

References

Morgan, M. G. and M. Henrion (1990). Uncertainty - A guide to dealing with uncertainty in quantitative risk and policy analysis. Cambridge, UK, Cambridge University Press.

PROBABILITY DISTRIBUTIONS FOR AMINE SYSTEM MODEL PARAMETERS

Here we briefly explain the basis for the probability distributions for various model parameters (Table 3.3 and Table 4.3) used in the probabilistic analysis presented in Chapter 5. As mentioned in *Appendix B*, these distributions take into account the data reported in literature, modeling approximations, and experts' technical judgments (especially when sufficient data is not available).

1. **CO_2 capture efficiency (%)**

Almost all the studies report CO_2 capture efficiency of 90%. This seems to be the optimum capture level in the MEA-based CO_2 capture systems, especially for Fluor Daniel's process using 30% MEA sorbent. Few studies mention desired capture efficiency of 85% or less. Only the Kerr-McGee/ ABB Lummus amine process using 20% MEA sorbent is reported to capture as high as 96% CO_2. Here we use 90% as the default value for this parameter and do not specify any distribution.

CO_2 capture efficiency (%)	Reference(s)
80	[1]
85	[2], [3]
85-90	[4]
90	[5], [6], [7], [8], [9], [10], [11], [12], [13], [14],[15], [16], [17], [18], [19], [20], [21], [22], [23]
85-95	[24], [25]
93	[26]
96	[27], [28]

2. **Acid gas removal efficiency of MEA system (%)**

MEA system designed to remove CO_2 from the flue gas also removes other acid gases (SO_2, NO_2 and HCl) if they are present. Just like other wet scrubber technologies, it also removes sizable portion of the particulate matter in the flue gas. Unfortunately, no data points are available regarding the removal efficiencies of these species in the literature.

Hence these distributions are solely based on personal communication with technical experts working in this field for a long time [29-31].

- SO_2 removal efficiency of MEA system (%): Uniform(99,100)

- NO_2 removal efficiency of MEA system (%): Uniform(20,30)

- HCl removal efficiency of MEA system (%): Uniform(90,95)

- Particulate removal efficiency of MEA system (%): Uniform(40,60)

3. **MEA concentration (% w/w)**

Today the commercially available MEA-based technology supplied by Fluor Daniel uses 30% MEA sorbent with some inhibitors. Other suppliers who do not use this inhibitor, prefer to use lower MEA concentration in the range of 15-20%. The highest MEA concentration is reported to be 50% (an optimization study, not in a real plant). Since the model is based on Fluor Daniel's Econamine FG process, the nominal value of 30% for MEA concentration is used. Given the higher energy penalty of dilute sorbent systems, it is unlikely that systems with sorbent concentrations less than 20% will be built today. At the same time, in spite of ongoing R&D efforts, an optimized amine system with 50% MEA is far from reality today. Hence a triangular distribution (20,30,40) is used to convey the variability in current commercial system designs.

MEA concentration (%)	Reference(s)
12, 18	[24], [25]
15	[2]
15-20	[32], [19]
20	[27], [28], [33]
20-30	[1], [34], [35]
30	[6], [8], [9], [12], [14], [33], [13], [4], [36], [37], [17], [24], [26]
50	[17]

4. **Lean sorbent CO_2 loading (mole CO_2 / mole MEA)**

The regenerated (or lean) sorbent contains some "left-over" CO_2. The level of lean sorbent CO_2 loading mainly depends upon the initial CO_2 loading in the sorbent and the amount of regeneration heat supplied, or alternatively, the regeneration heat requirement

depends on the allowable level of lean sorbent loading. The range of values reported in the literature corresponds to variety of designs and operating conditions. Although a higher value translates into lower regeneration heat requirement (and hence lower energy penalty), experts in this technology believe that operating amine system at such conditions will lead to severe corrosion and maintenance problems [29, 30]. Similarly, a very low value like 0.05 implies very high regeneration energy requirement, which may not be economically viable due to the large energy penalty. So, the first (0.05) and the last (0.3) data points have been omitted while characterizing the distribution for this parameter. Here we use a nominal value of 0.2 and a triangular distribution (0.1, 0.2, 0.25) that is based on literature survey as well as experts' opinions.

Lean sorbent CO_2 loading (mole CO_2 / mole MEA)	Reference(s)
0.05-.2	[33]
0.1	[34]
0.16	[38]
0.2	[39], [40]
0.21	[38]
0.215	[27], [28]
0.24	[17]
0.25	[17]
0.3	[2]

5. **Nominal MEA makeup (kg MEA/ tonne CO_2)**

During the normal operating conditions, MEA is lost through polymerization and oxidation reactions. In order to keep the system continuously running, sorbent makeup is required. Typically this value is reported by various studies in the units of kg MEA per tonnes of CO_2 captured. The range of values indicate both variability due to differences in system design and operating conditions as well as uncertainty due to complex chemistry involved. Here we use a nominal value of 1.5 kg MEA per tonne CO_2 captured, with a triangular distribution (0.5, 1.5, 3.1).

6.

MEA makeup rate (kg MEA/ tonne CO_2 captured)	Reference(s)
0.5-2	[20]
0.7	[28], [14]
0.9	[41]
1.5	[38], [9], [17]
1.6	[42]
1.8	[8], [11], [41]
3.1	[36]

7. **Blower pressure head (kPa)**

Blower pressure head (ΔP_{fg}) is the pressure head that needs to be provided to the flue gas in the blower so that it can overcome the pressure drop in the absorber tower. Typically, it would depend on the absorber design. We find a range of values reported in the literature, as well as indicated by the experts. Based on these numbers, the nominal value for this parameter is taken as 14. The variability is represented by a triangular distribution covering the range (4, 14, 37) of all these values.

Blower pressure head (kPa)	Reference(s)
4	[2]
5	[3], [17]
12	[28]
14	[25], [19]
15	[26]
17	[6]
20	[42], [9]
37	[33]

8. **Sorbent pumping head (kPa)**

Sorbent pumping head ($\Delta P_{sorbent}$) is the pressure head to be provided to the sorbent using sorbent circulation pumps, so that it can overcome the pressure losses in the absorber and regenerator columns. A typical value is about 200 kPa [17], and a modest range around the same has been assumed to represent the uncertainty coming from lack of sufficient data, triangular(150,200,250).

9. **Blower and pump efficiency (%)**

Flue gas blower and sorbent circulation pump are common mechanical devices used in industrial operations. These devices have performance ratings that may vary depending upon the make/type and the operating conditions. Typically, the efficiencies would be around 70-80%. Hence we use a uniform distribution (uniform (70, 80)).

10. **Equivalent electricity generation penalty (% of regeneration heat energy)**

As explained before, this factor (F_{HE}) gives the equivalent loss of power generation capacity due to extraction of steam from the power plant steam cycle to meet the sorbent regeneration heat requirement. This factor mainly depends on the design issues such as the quality of steam used for sorbent regeneration, and the level of heat integration between the base plant and the amine system. Recent developments suggest that the energy penalty can be substantially reduced by better heat integration and in-house practices [43, 44]. Here we use a nominal value of 14 with a uniform distribution (9,19) for a new plant. Retrofit applications are likely to have some constraints on the level of heat integration, and hence have a higher value for this factor. The values obtained from the literature are in the range (20,25), hence we use a nominal value of 22.

Equivalent electricity loss (%)	Reference(s)
9	[17]
16	[14], [17]
17	[45]
19	[6]
20-25	[28], [27]
22	[2], [26]
25	[3]

11. **CO_2 product pressure (psig)**

Most of the data sources report the value of CO_2 product pressure to be around 2000 psig. It is believed that pressure of this magnitude would be sufficient to overcome pressure losses incurred during the transport of CO_2 product over a reasonable distance (100 miles or so) without any need for recompression stations to keep the CO_2 product in liquefied state. Depending upon the assumptions about the transportation distance and other details, this value may be different. The CO_2 product pressure has to be above 1070 psig,

the critical pressure of CO_2. Also, the product pressure of 2200 psig is sufficiently high for transport of CO_2 via pipelines. So, a triangular distribution has been adopted to reflect this range 1100-2200 with a nominal value of 2000.

CO_2 product pressure (psig)	Reference(s)
1164	[17]
1207	[6]
1451	[3], [46]
1595	[2]
2000	[47], [27], [28], [48], [16], [49]
2035	[8], [11], [9]
2180	[42], [26]
3200	[5]

12. **Compressor efficiency**

This is the effective efficiency of the compressors used to compress CO_2 product to the desirable pressure. As seen from the reported data, the compression efficiencies lie in the range of about 75% to 88%. Hence the nominal value of compressor efficiency (η_{comp}) is taken as 80% with a triangular distribution to cover the data range.

CO_2 compression efficiency (%)	Reference(s)
75	[50]
75-85	[3]
85	[17]
88	[28]

It may be noted that probability distributions have not been assigned to some performance parameters such as MEA losses due to acid gases (SO_2, NO_2, HCl), ammonia generation and consumption of caustic and activated carbon. These are the parameters that reflect stoichiometry (balance among chemical species).

Next we have discussion about the probability distributions for some cost parameters.

13. **Sorbent cost ($/ tonne sorbent)**

The cost of the sorbent depends upon various market forces including the cost of natural gas (which is the ultimate feedstock for its manufacture). Hence it is likely to be

fluctuating. Based on the data and experts' judgments, the nominal value was taken as $1200/ tonne sorbent with a uniform distribution (1000, 1300).

Sorbent cost ($/ tonne sorbent)	Reference(s)
970	[42]
1189	[11]
1300	[28]
1340	[17, 51]

14. **Total maintenance cost (% TPC)**

The total maintenance cost is generally specified as a fraction of total plant cost. This parameter depends on some design parameters (*e.g.* sorbent concentration, column design) as well as the operating conditions (*e.g.* flue gas characteristics, CO_2 loading levels). Hence there is likely to be some variability as well as performance uncertainty reflected in the range of data. Based on the data as well as experts' judgments, 2.5% was taken as the nominal value with a triangular distribution (1, 2.5, 5).

Total maintenance cost (% TPC)	Reference(s)
1.5	[42]
2	[11]
2.2	[28]
4	[2]
4	[17]

15. **CO_2 transport cost**

There is a wide range of numbers reported for the cost of transporting the concentrated stream of CO_2 product. Most of the variability may be explained in terms of assumptions about the location of the pipeline and design parameters (safety factors, spare capacity *etc.*). Few representative numbers are listed here. The nominal value is taken as $0.02/ km per tonne CO_2, with a triangular distribution (0.004,0.02,0.08).

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CO$_2$ transport cost ($/ km per tonne CO$_2$)	Reference(s)
0.004-0.03	[52]
0.01-0.03	[53]
0.01-0.05	[54]
0.012	[55]
0.024-0.061	[34]
0.03-0.08	[56]

The distributions for other cost parameters (mainly the indirect capital costs) are based on those for other environmental control technologies.

References

1.   Langeland, K. and K. Wilhelmsen, *A study of the costs and energy reuirement for carbon dioxide disposal.* Energy Conversion and Management, 1993. **34(9-11)**: p. 807-814.

2.   Stork Engg Consultancy B.V., A., *Leading options for the capture of CO2 emissions at power stations*. 2000, IEA GHGRDP: UK.

3.   Bolland, O. and H. Undrum. *Removal of CO2 from gas turbine power plants: Evaluation of pre- and postcombustion methods*. in *Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September*. 1998. Interlaken, Switzerland: Elsevier Science Ltd.

4.   Rao, A.D. and W.H. Day, *Mitigation of greenhouse gases from gas turbine power plants.* Energy Conversion and Management, 1996. **37(6-8)**: p. 909-914.

5.   Slater, M., E. West, and C. Mariz. *Carbon dioxide capture from multiple flue gas sources*. in *Sixth International Conference on Greenhouse Gas Control Technologies*. 2002. Kyoto, Japan.

6.   Parsons Infrastructure & Technology Group, I., *Evaluation of fossil fuel power plants with CO2 recovery*. 2002, USDOE/NETL/EPRI: Pittsburgh, PA.

7.   Jeremy, D. and H.J. Herzog. *The cost of carbon capture*. in *presented at the Fifth International Conference on Greenhouse Gas Control Technologies, 13-16 August*. 2000. Cairns, Australia.

8.   Chapel, D., J. Ernst, and C. Mariz. *Recovery of CO2 from flue gases: commercial trends (paper no. 340). in presented at the Canadian Society of Chemical Engineers Annual Meeting, 4-6 October.* 1999. Saskatoon, Saskatchewan, Canada.

9.   Desideri, U. and A. Paolucci, *Performance modelling of a carbon dioxide removal system for power plants.* Energy Conversion and Management, 1999. **40**: p. 1899-1915.

10.   Simbeck, D. *A portfolio selection approach for power plant CO2 capture, separation and R & D options.* in *Fourth International Conference on Greenhouse Gas Control Technologies, 30 August - 2 September.* 1998. Interlaken, Switzerland: Elsevier Science Ltd.

11.   Fluor, *Evaluation of technologies for the removal of CO2 from coal fired power plant flue gas.* 1998, a report prepared by Fluor Daniel Canada Inc. for TransAlta, November 1998.

12.   Desideri, U. and R. Corbelli, *CO2 capture in small size cogeneration plants: Technical and economical considerations.* Energy Conversion and Management, 1998. **39(9)**: p. 857-867.

13.   Leci, C.L., *Development requirements for absorption processes for effective CO2 capture from power plants.* Energy Conversion and Management, 1997. **38(Suppl)**: p. S45-S50.

14.   Mimura, T., et al., *Development of energy saving technology for flue gas carbon dioxide recovery in power plants by chemical absorption method and steam system.* Energy Conversion and Management, 1997. **38(Suppl)**: p. S57-S62.

15.   Leci, C.L., *Financial implications on power generation costs resulting from the parasitic effect of CO2 capture using liquid scrubbing technology from power station flue gases.* Energy Conversion and Management, 1996. **37(6-8)**: p. 915-921.

16.   Tontiwachwuthikul, P.T., W. Kritpiphat, and D. Gelowitz, *Carbon dioxide production from co-generation for enhanced oil recovery: An economic evaluation.* The J. of Can. Petro. Tech., 1996. **35(6)**: p. 27-33.

17.   Hendriks, C., *Carbon Dioxide Removal from Coal-fired Power Plants.* 1st ed. 1994, The Netherlands: Kluwer Academic Publishers. 14-223.

18.   Riemer, P., H. Audus, and A. Smith, *Carbon Dioxide Capture from Power Stations.* 1994, a report published by IEA Greenhouse Gas R&D Programme: UK.

19.   Barchas, R. and R. Davis, *The Kerr-McGee/ ABB Lummus Crest technology for the recovery of CO2 from stack gases.* Energy Conversion and Management, 1992. **33**(5-8): p. 333-340.

20.   Suda, T., et al., *Development of flue gas carbon dioxide recovery technology.* Energy Conversion and Management, 1992. **33**(5-8): p. 317-324.

21.   Yagi, T., H. Shibuya, and T. Sasaki, *Application of chemical absorption process to CO2 recovery from flue gas generated in power plants.* Energy Conversion and Management, 1992. **33**(5-8): p. 349-355.

22.   Nakayama, S., et al., *Pulverized coal combustion in O2/CO2 mixtures on a power plant for CO2 recovery.* Energy Conversion and Management, 1992. **33(5-8)**: p. 379-386.

23.   Horn, F.L. and M. Steinburg, *An improved carbon dioxide power plant.* Energy Progress, 1982. **2(3)**: p. 154-159.

24.   Sander, M.T. and C.L. Mariz, *The Fluor Daniel "Econamine FG" process: Past experience and present day focus.* Energy Conversion and Management, 1992. **33(5-8)**: p. 341-348.

25.   Wiggins, W.R.J. and R.L. Bixler, *Sources, recovery and tranportation of CO2.* Energy Progress, 1983. **3(3)**: p. 132-135.

26.   Smelser, S.C., R.M. Stock, and G.J. McCleary, *Engineering and economic evaluation of CO2 removal from fossil-fuel-fired power plants.* 1991, EPRI IE-7365, Volume 1, Project 2999-10, a research project final report prepared by Fluor Daniel Inc., for EPRI and IEA.

27.   Marion, J., et al. *Engineering feasibility of CO2 capture on an existing US coal-fired power plant.* in *Twenty Sixth International conf. On coal utilization and fuel systems.* 2001. Clearwater, FL.

28.   ALSTOM, et al., *Engineering feasibility and economics of CO2 capture on an existing coal-fired power plant.* 2001, Final report prepared by ALSTOM Power Inc., ABB Lummus Global Inc., ALSTOM Power Environmental Systems and American Electric Power (AEP). report no. (PPL-01-CT-09) submitted to Ohio Department of Development, Columbus, OH and US Department of Energy/ NETL, Pittsburgh, PA. p. 163.

29.   Dingman, J.D., *personal communication.* 2002.

30.   Mariz, C., *personal communication.* 2000-2003.

31.   Gupta, J.C., *personal communication.* 2002.

32. DuPart, M.S., T.R. Bacon, and D.J. Edwards, *Understanding corrosion in alkanolamone gas treating plants.* Hydrocarbon Processing, 1993. **May**: p. 89-94.

33. Kohl, A.L. and R.B. Nielsen, *Gas Purification.* 5th ed. 1997, Houston, TX: Gulf Publishing Company. 40-277.

34. Anada, H.R., et al., *Economics of by-product CO2 recovery and transportation for EOR.* Energy Progress, 1983. **3(4)**: p. 233-243.

35. Stokes, K.J., *Choosing an ammonia plant CO2 removal system for today's conditions.* Nitrogen, 1981. **131**: p. 35-38.

36. Chakma, A., *An energy efficient mixed solvent for the separation of CO2.* Energy Conversion and Management, 1995. **36(6-9)**: p. 427-430.

37. Jou, F., A.E. Mather, and F.D. Otto, *The solubility of CO2 in a 30 mass percent monoethanolamine solution.* The Canadian Journal of Chemical Engineering, 1995. **73 (Feb)**: p. 140-147.

38. Guezebroek, F.H., L.H.J.M. Schneider, and G.J.C. Kraaijveld. *Exergy analysis of amine-based CO2 removal technology.* in *Sixth International Conference on Greenhouse Gas Control Technologies.* 2002. Kyoto, Japan.

39. Veawab, A., et al. *Performance and cost analysis for CO2 capture from flue gas streams: Absorption and regeneration aspects.* in *Sixth International Conference on Greenhouse Gas Control Technologies.* 2002. Kyoto, Japan.

40. Alie, C., et al., *Simulating CO2 capture from pulverized coal fired power plants: Assessment of MEA-CO2 capture processes using AspenPlus.* 2002, International test network for CO2 capture: Report on 3rd workshop (May 16-17, 2002), Apeldoorn, The Netherlands (Report number PH4/11).

41. Pauley, C.R., P.L. Simiskey, and S. Haigh, *N-Ren recovers CO2 from flue gas economically.* Oil and Gas Journal, 1984. **82**(20): p. 87-92.

42. Singh, D., et al., *Techno-economic study of CO2 capture from an existing coal-fired power plant: MEA scrubbing vs. O2/CO2 recycle combustion.* 2002: Waterloo, Ontario, Canada. p. 27.

43. Reddy, S. and C.A. Roberts. *ECONOMINE FGSM Plus: An Enhanced Amine Based CO₂ Capture Process.* in *presented at the 2nd Annual Conference on Carbon Sequestration, May 5-9, 2003.* 2003. Alexandria, VA, USA.

44. Gibbins, J.R. and R.I. Crane, *Scope for reductions in the cost of CO2 capture using flue gas scrubbing with amine solvents,* in *Draft paper.* 2003. p. 27.

45.   Alie, C., et al., *Simulatiion of CO2 capture from coal-basd power plants: Supplementary energy management*. 2003, International test network for CO2 capture: Report on 5th workshop (June 2-3, 2002), Carnegie Mellon University, Pittsburgh (Report number PH4/22).

46.   Bolland, O. and P. Mathieu, *Comparison of two CO2 removal options in combined cycle power plants.* Energy Conversion and Management, 1998. **39(16-18)**: p. 1653-1663.

47.   Iijima, M. and T. Kamijo. *Flue gas CO2 recovery and compression cost study for CO2 enhanced oil recovery*. in *Sixth International Conference on Greenhouse Gas Control Technologies*. 2002. Kyoto, Japan.

48.   Simbeck, D. and M. McDonald. *Existing coal power plant retrofit CO_2 control options analysis*. in *presented at the Fifth International Conference on Greenhouse Gas Control Technologies, 13-16 Augus*. 2000. Cairns, Australia.

49.   Clair, J.H. and W.F. Simister, *Process to recover CO2 from flue gas gets first large-scale tryout in Texas.* Oil and Gas J., 1983. **Feb**: p. 109-113.

50.   Price, B.C., *Processing high CO2 gas.* Energy Progress, 1984. **4(3)**: p. 169-174.

51.   Dow, *Personal communication with the Marketing Executive, Dow Chemicals*. 2003.

52.   Herzog, H.J., *A Cost model for transport of carbon dioxide*, in *Draft paper*. 2001, MIT.

53.   Wallace, D., *Capture and storage of CO2: What needs to be done?* 2000, IEA.AIE: presented at the COP6, The Hague. p. 9.

54.   NOVEM, *Potential for CO2 sequestration and enhanced coalbed methane production in the Netherlands*. 2001, NOVEM (The Netherlands Agency for Energy and the Environment): Utrecht. p. 105.

55.   Bergman, P.D., E.M. Winter, and Z.-Y. Chen, *Disposal of power plant CO2 in depleted oil and gas reservoirs in Texas.* Energy Conversion and Management, 1997. **38(Suppl)**: p. S211-S216.

56.   Battelle, *Carbon dioxide sequestration in saline formations - Engineering and economic assessment final technical report*. 2001, NETL, USDOE: Morgantown, WV. p. 93.

APPENDIX D

CASE STUDY RESULTS FOR NEW COAL PLANTS

<u>Assumptions</u>:

- Base plant (Super critical PC):

    o Coal: Wyoming powder river basin, 0.48% S, 8340 Btu/lb, 23.19/tonne

    o Heat Rate: Steam cycle HR 6750 Btu/kWh = Gross cycle HR 7923 Btu/kWh

- SO_2 control: FGD system using limestone

- CO_2 control:

    o MEA system using 30% w/w MEA-based sorbent

    o CO_2 product compression to 2000 psig

    o cost for CO_2 transport = $7/ tonne CO_2 (100 miles @ $0.0426/tonne CO_2 per km)

    o cost for CO_2 disposal = $5/ tonne CO_2

- Other details as mentioned in Tables 3.3, 4.2, 4.3 and 5.1.

Parameter	Reference Plant	Capture Plant (A)	% change	Capture Plant (B)	% change
Model Inputs:					
Gross capacity (MW)	500	500	0%	670	+34%
FGD efficiency (%)	95	99		99	
CO_2 capture efficiency (%)	-	90		90	
Energy Penalty:					
Base plant, PC (MW)	29.3	29.3	0%	39.2	+34%
FGD System (MW)	9.5	13.7	+43.4%	18.3	+92%
ESP (MW)	0.9	0.9	0%	0.9	0%
SCR (MW)	2.7	2.7	-0.4%	3.6	+33%
MEA: Regen. heat req. (equiv.)	-	60.8		81.5	
MEA: fans (MW)	-	10.4		14.0	
MEA: pumps (MW)	-	0.9		1.1	
MEA: CO_2 compression (MW)	-	40.3		54.0	

Parameter	Reference Plant	Capture Plant (A)	% change	Capture Plant (B)	% change
MEA: total (MW)	-	112.4		150.6	
Total energy penalty (MW)	42.4	158.4	+275%	212.6	+402%
7.5 Plant Performance					
Total energy penalty (%MW$_g$)	8.5%	31.8%	+275%	31.7%	+275%
Net power generation (MW)	457.6	341.2	-26%	457.6	0%
Net cycle heat rate (Btu/kWh)	8,657	11,560	+34%	11,550	+33%
Sorbent regen. heat (kJ/kgCO$_2$)	-	4575		4575	
Plant Consumption:					
Coal consumption (tonne/hr)	215.5	214.6	-0.4%	287.5	+33%
Limestone consumed (tonne/hr)	2.3	2.4	+4%	3.2	+39%
MEA consumed (kg/hr)	-	789		1057	
Emissions/ Discharges:					
CO$_2$ emission (tonne/ hr)	381.3	38.0	-90%	50.9	-87%
NO$_x$ emission (kg/ hr)	269.5	265.0	-2%	355.2	+32%
SO$_x$ emission (kg/ hr)	76.4	0.11	-99.9%	0.14	-99.8%
NH$_3$ emission (kg/ hr)	5.4	29.8	+457%	40.0	+648%
CO$_2$ emission (gCO$_2$/ kWh)	833.3	111.3	-87%	111.2	-87%
NO$_x$ emission (gNO$_x$/ kWh)	0.59	0.78	+32%	0.78	+32%
SO$_x$ emission (gNO$_x$/ kWh)	0.17	0.0003	-99.8%	0.0003	-99.8%
Bottom ash (tonne/hr)	3.5	3.5	-0.4%	4.6	+33%
Fly ash (tonne/hr)	8.0	8.0	-0.4%	10.7	+34%
FGD waste (tonne/hr)	4.2	4.3	+4%	5.8	+39%
Spent MEA sorbent (tonne/hr)	-	1.0		1.4	
CO$_2$ product (tonne/ hr)	-	342		458	
Capital costs (TCR, M$)					
PC	516.2	513.6	-0.5%	646.0	+25%
ESP	23.5	22.1	-6%	27.8	+18%
FGD	48.3	50.1	+4%	57.6	19%
SCR	27.6	27.7	+0.2%	34.8	26%

Parameter	Reference Plant	Capture Plant (A)	% change	Capture Plant (B)	% change
MEA	-	146.5		196.6	
Total - TCR (M$)	615.7	760.1	+24%	962.8	+56%
Normalized TCR ($/ kW)	1345	2228	+66%	2104	+56%
O&M costs (M$/yr)					
MEA – FOM	-	4.1		5.3	
MEA – reagent	-	6.5		8.7	
MEA – power	-	13.5		17.2	
MEA – steam (equiv.)	-	15.9		20.2	
MEA – CO_2 transport	-	7.2		9.7	
MEA – CO_2 disposal	-	11.2		15.1	
TOM – MEA	-	61.3		79.9	
TOM – PC	44.4	13.3	-70%	17.9	-60%
TOM – ESP	1.6	1.6	-1.4%	1.8	+16%
TOM – FGD	7.5	8.6	+16%	10.2	+36%
TOM – SCR	4.2	4.2	-0.5%	5.5	+30%
8. OVERALL PLANT COSTS					
Fuel cost	32.9	32.7	-0.4%	43.8	+34%
TOM – whole plant	57.6	89.1	+55%	115.2	+100%
TRR (M$/ yr), whole plant	148.7	201.5	+36%	257.7	+73%
COE ($/ MWh)	49.4	89.9	+82%	85.7	+73%
$/ tonne CO_2 avoided	-	56.0		50.2	
$/ tonne CO_2 delivered	-	40.3		36.2	

QUESTIONNAIRE SENT TO EXPERTS ON AMINE-BASED CO$_2$ CAPTURE TECHNOLOGY

Thank you for agreeing to assist us in our work on characterizing the uncertainty and variability associated with the performance of amine-based CO$_2$ capture technology. We plan to use your answers, and those of other experts as inputs to engineering-economic models we have developed to characterize the performance and cost of these systems. While we will acknowledge the experts who have assisted us in this effort, we will not identify any expert with any specific response. If you have questions or concerns please contact me:

> By E-mail: abr@cmu.edu, or

> By Phone: (412) 268-6826 or (412) 268-6115, or

> By Fax: (412) 268-1089 (attn: Anand B. Rao)

Please also send your responses by email or fax, or by regular mail to:

> Anand B. Rao
> Department of Engineering & Public Policy
> Carnegie Mellon University
> Pittsburgh, PA 15213, USA.

Alternatively, if you prefer, I can give you a call and note down your responses.

In the pages that follows we:

1. provide some background on the problem

2. list the key assumptions we are using in our model and ask you to comment on their appropriateness *for current systems*

3. ask you to make judgments about the likely values of a number of key coefficients *for future systems*

4. ask you to provide a ranking of various research objectives to reduce the cost of CO$_2$ capture and sequestration from flue gas streams.

We estimate that you may need approximately 40-50 minutes to complete this questionnaire, and hope you will find that acceptable. We really appreciate your time and help, and will be happy to share our findings with you at the end of this study.

Part 1: Background

Development of improved technology to capture and sequester the CO_2 emitted by power plants using fossil fuels — especially coal — is the subject of major research efforts worldwide. The attraction of this option is that it would allow abundant world resources of fossil fuels to be used for power generation and other applications without contributing significantly to atmospheric emissions of greenhouse gases. The two key barriers to carbon capture and sequestration (CCS), however, are the high cost of current CO_2 capture technologies, and uncertainties regarding the technical, economic and political feasibility of CO_2 storage options.

As part of the USDOE's Carbon Sequestration Program, we have developed an integrated modeling framework to evaluate the performance and costs of alternative CO_2 capture and sequestration technologies for fossil-fueled power plants, in the context of multi-pollutant control requirements. This model (called the IECM-CS) allows for explicit characterization of the uncertainty or variability in any or all model input parameters. One of the purposes of this model is to improve the understanding in both the technical and policy communities about the magnitude of CCS cost and the various factors that affect it. We also want to explore the potential for reducing costs through targeted R&D.

At this stage, many of the model parameter values and uncertainty distributions have been based on information gathered from the literature. This approach has its limitations owing to the limited availability of data and possibility of inconsistent assumptions across different studies. An alternative method that can allow a more robust analysis of uncertainty and variability involves the development of expert judgments. Here, we want to understand how experts in this field would characterize some of the key parameters that affect the performance and cost of a particular CO_2 capture technology, namely, amine-based absorption of CO_2 from flue gas. We are interested both in current system designs, and improved future designs that you might envision.

Please briefly review the *Supplementary Information* on the next two pages to make sure that we use a consistent terminology during this exercise.

Overview of amine-based CO_2 capture systems

The idea of separating CO_2 from flue gas streams started in the 1970s, not with concern about the greenhouse effect, but as a potentially economic source of CO_2, mainly for enhanced oil recovery (EOR) operations. Several commercial CO_2 capture plants were constructed in the U.S. in the late 1970s and early 1980s. CO_2 was also produced for other industrial applications such as carbonation of brine and production of products like dry ice, urea and beverages. Some of these CO_2 capture plants are still in operation today, but all these plants are much smaller than a typical power plant. The first commercial CO_2 sequestration facility started in Norway in September 1996 in response to a Norwegian carbon tax. Since then, Statoil has been storing CO_2 from the Sleipner West gas field in a sandstone aquifer 1000 m beneath the North Sea. The international research community is closely monitoring this facility.

All these plants capture CO_2 with processes based on chemical absorption using a monoethanolamine (MEA)-based solvent. MEA is an organic chemical belonging to the family of compounds known as amines. It was developed over 60 years ago as a general, non-selective solvent to remove acidic gas impurities (e.g. H_2S, CO_2) from natural gas streams. The process was then adapted to treat flue gas streams for CO_2 capture. Fluor Daniel Inc., Dow Chemical Co., Kerr-McGee Chemical Corp. and ABB Lummus Crest Inc., were some of the initial developers of MEA-based technology for CO_2 capture. Typically, about 75% to 90% of the CO_2 is captured using this technology, producing a nearly pure (>99%) CO_2 product stream.

Process Description

A continuous scrubbing system is used to separate CO_2 from the flue gas stream. The system consists of two main elements: an absorber where CO_2 is removed, and a regenerator (or stripper) where CO_2 is released (in concentrated form) and the sorbent is recovered.
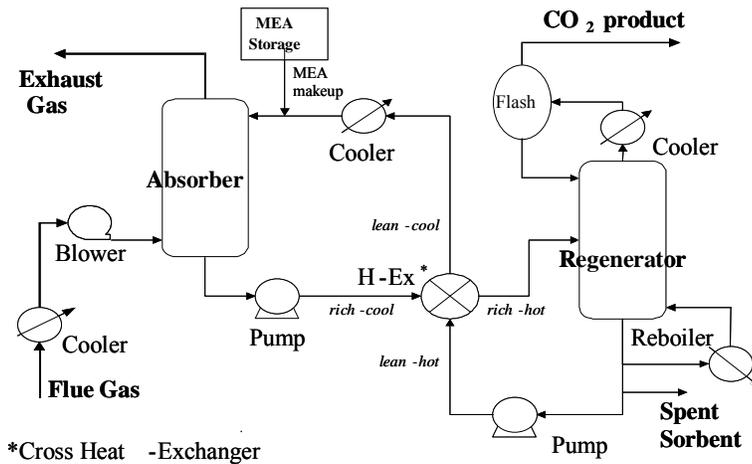
In a power plant application, cooled flue gases flow vertically upwards through the absorber countercurrent to the absorbent (consisting of MEA in a water solution, often with some additives to inhibit corrosion). The MEA reacts chemically with the CO_2 in the flue gases to form a weakly bonded compound called carbamate. The scrubbed gases are then washed and vented to the atmosphere. The CO_2-rich solution leaves the absorber and passes through a heat exchanger, then is further heated in a reboiler using low-pressure steam. The carbamate formed during absorption is broken down by the application of heat, regenerating the solvent, and producing a concentrated CO_2 stream. The hot CO_2-lean solvent is then returned to the heat exchanger, where it is cooled, then sent back to the absorber. Some fresh MEA is added to make up for losses incurred in the process. The process chemistry is complex, but the main reactions taking place are:

CO_2 Absorption: $2 \, R\text{-}NH_2 \, (MEA) + CO_2 \rightarrow R\text{-}NH_3^+ + R\text{-}NH\text{-}COO^-$ (Carbamate)   (1)

MEA Regeneration: $R\text{-}NH\text{-}COO^- + R\text{-}NH_3^+ + Heat \rightarrow CO_2 + 2 \, R\text{-}NH_2$ (2)
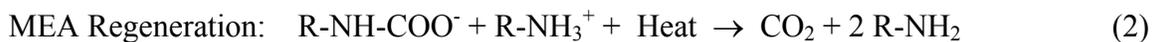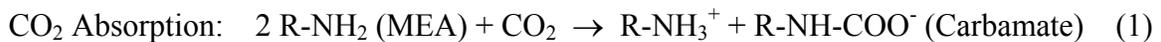
Pure MEA (with $R = HO\text{-}CH_2CH_2$) is an "unhindered" amine that forms stable carbamate; hence, only half a mole of CO_2 is absorbed per mole of amine, as shown in Reaction (1). For other "hindered" amines (where R is a bulky group), the carbamate formed is not stable, and an alternate reaction leads to a higher theoretical capacity of one mole of CO_2 per mole of amine. But the CO_2 uptake rate of hindered amines is very low. Efforts are underway to formulate better solvents by combining favorable properties of these two groups of amines.

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Some of the solvent also is lost during the process because of physical losses, entrainment, vaporization and chemical degradation. All the solvent entering the stripper does not get regenerated. Flue gas impurities, especially oxygen, sulfur dioxide and nitrogen dioxide react with MEA to form heat-stable salts, thus reducing the CO_2-absorption capacity of the solvent. However, proprietary inhibitors are available that can render the solvent tolerant to oxygen. Flue gas NO_x is not a major problem since the predominant form (~ 95%) of total NO_x in the flue gas is nitric oxide (NO), which does not react with inhibited amines. But SO_2 and NO_2 do react to degenerate MEA solvent, so very low inlet concentrations of these gases (on the order of 10 ppm) are desirable to avoid excessive loss of solvent. Since the untreated flue gases of coal-fired power plants typically contain about 700 to 2500 ppm SO_2 (plus roughly 10-40 ppm NO_2), the interaction of SO_2 with the CO_2 control system is particularly important. The heat-stable salts that are formed in these reactions may be treated in a side-stream MEA reclaimer, which can regenerate some of the MEA. Technologies such as electrodialysis are also being proposed for this purpose.

Finally, corrosion control is very important in amine systems, especially when processing oxygen-containing gases. In order to reduce corrosion rates, corrosion inhibitors, low concentrations of MEA, appropriate materials of construction and mild operating conditions are required.

A key feature of amine systems is the large amount of heat required to regenerate the solvent. This heat is typically drawn from the steam cycle and significantly reduces the net efficiency of the power plant. Substantial electrical energy also is needed to compress the captured CO_2 for pipeline transport to a storage site. As shown later in this paper, the overall energy penalty of this process has a major impact on system performance as well as cost.

From a multi-pollutant perspective, there are also important interactions between the CO_2 capture system and the control of other air pollutants, especially SO_2 and NO_x emissions. Acid gases like SO_2 and NO_2 react with MEA to form heat-stable salts that reduce the CO_2 absorption capacity of the solvent. Thus, very low concentrations of these gases (on

the order of 10 ppm) are desirable to avoid excessive loss of (costly) solvent. The problem is especially acute for SO_2 because its concentration in flue gases is typically 700 to 2500 ppm at coal-fired plants. NO_x is less of a problem because most of the NO_x is nitric oxide (NO), whereas only NO_2 (typically about 5% of total NO_x) is reactive (see Supplementary Information for additional details).

Part 2: Current Amine Systems

To begin with, please consider a CO_2 capture system which:
- uses 30% (by weight) MEA-based sorbent,
- treats the flue gas stream from a coal-combustion source which has a CO_2 concentration of about 12% (by volume) and which has been pre-treated for removal of SO_x and NO_x and cooled to a desired absorber inlet temperature,
- and removes 90% of CO_2 from the flue gas stream.

In the table that follows we have summarized the base case key parameters that we ask you to comment upon. These parameters are intended to characterize the performance of a current commercial system as system described above and in the supplemental information given earlier.

Please indicate whether you find each of these nominal values and ranges to be reasonable. If not, please indicate the value you would prefer and provide us with a brief explanation. If you prefer to use different units than the ones shown here, please indicate clearly the measure that you prefer.

Parameter	Accept-ability	If not OK, New Value	Explanation of change
Absorber inlet flue gas pressure			
Nominal value = 26 kPa gauge = 3.8 psig	o OK o not OK		
Range = 14 - 30 kPa gauge = 2 - 4.4 psig	o OK o not OK		
Temperature of flue gas entering the absorber			
Nominal value = 60 deg C = 140 deg F	o OK o not OK		
Range = 50 - 62 deg C = 122 - 144 deg F	o OK o not OK		
Lean sorbent CO_2 loading (mole CO_2/ mole sorbent)			
Nominal value = 0.2 = 7.3 x10^{-3} SCF CO_2/ gal soln	o OK o not OK		
Range = 0.15 - 0.25 = 5.5 – 9.1 x10^{-3} SCF CO_2/ gal	o OK o not OK		

Parameter	Accept-ability	If not OK, New Value	Explanation of change
Heat required for sorbent regeneration (kJ / kg CO$_2$)			
Nominal value = 4350 = 2.2 lb steam/ lb CO$_2$	o OK o not OK		
Range = 3500 - 6000 = 1.75 – 3.00 lb steam/ lb CO$_2$	o OK o not OK		
Allowable levels of other components in flue gas			
SO$_2$ < 10 ppmv	o OK o not OK		
NO$_2$ < 10 ppmv	o OK o not OK		
O$_2$ ~ 3.5 % by volume	o OK o not OK		
Maximum train size (tonnes CO$_2$ per day)			
Nominal value = 5000	o OK o not OK		
Energy required for CO$_2$ compression to 2000 psig (kWh / tonne CO$_2$)			
Nominal value = 119	o OK o not OK		
Range = (112, 145)	o OK o not OK		

Part 3: Judgments About Future Amine Systems

In this section we would like to obtain your technically informed probabilistic judgments about several key parameters of a future amine-based CO_2 capture system built around the year 2015. We would like you to consider all of the parameters listed, but feel free to skip any parts that you are not comfortable with. <u>Also, if you prefer to use some different unit of measurement in your answer, please mention it clearly</u>.

In producing your answers please assume that:

- We are still talking about an amine-based plant that treats the flue gas stream from a coal-combustion source, which is about 12%, CO_2 and which has been pre-treated for removal of SO_x and NO_x, and removes 90% of CO_2 from the flue gas stream.

- The plant has been optimized for the lowest overall cost of CO_2 avoidance ($/ tonne CO_2 avoided), considering both capital and operating costs (including energy costs) over the life of the plant.

- R&D support for this technology continues to steadily grow at a modest rate through 2015, and includes several new large-scale applications to coal-fired power plants.

Parameter 1: Amine sorbent concentration (wt %)

First please give us your <u>highest</u> estimate of the value that you think that amine sorbent concentration might be in state-of-the-art operating plants by 2015.

wt%

Suppose that we have a clairvoyant who can look ahead to 2015 and tells you that the actual number is 1.15 times this value. Can you think of any plausible story of how that might have happened?

o No o Yes

If you answered yes, briefly what is the story:

Is there at least a 5% probability that this story could turn out to be true?

o No o Yes

Next we'd like your <u>lowest</u> estimate of the value that you think that sorbent concentration might be in state-of-the-art operating plants by 2015.

wt%

Again, suppose that we have a clairvoyant who can look ahead to 2015 and tells you that the actual number is 0.85 times this value. Can you think of any plausible story of how that might have happened?

o No o Yes

If you answered yes, briefly what is the story:

Is there at least a 5% probability that this story could turn out to be true?

o No o Yes

Finally we'd like your <u>best</u> estimate of the value that you think that average sorbent concentration might be in state-of-the-art operating plants by 2015.

wt%

Parameter 2: Unit heat requirement for sorbent regeneration

First please give us your <u>lowest</u> estimate of the value that you think that sorbent regeneration heat requirement might be in state-of-the-art operating plants by 2015.

kJ/ kg CO$_2$ captured (or   lb steam/ lb CO$_2$ captured)

Suppose that we have a clairvoyant who can look ahead to 2015 and tells you that the actual number is 0.85 times this value. Can you think of any plausible story of how that might have happened?

o No o Yes

If you answered yes, briefly what is the story:

_____
_____
_____

Is there at least a 5% probability that this story could turn out to be true?

o No o Yes

Next we'd like your <u>highest</u> estimate of the value that you think that sorbent regeneration heat requirement might be in state-of-the-art operating plants by 2015.

| kJ/ kg CO_2 captured (or   lb steam/ lb CO_2 captured) |

Again, suppose that we have a clairvoyant who can look ahead to 2015 and tells you that the actual number is 1.15 times this value. Can you think of any plausible story of how that might have happened?

o No o Yes

If you answered yes, briefly what is the story:

_____
_____
_____

Is there at least a 5% probability that this story could turn out to be true?

o No o Yes

Finally we'd like your <u>best</u> estimate of the value that you think that sorbent regeneration heat requirement might be in state-of-the-art operating plants by 2015.

| kJ/ kg CO_2 captured (or   lb steam/ lb CO_2 captured) |

Parameter 3: Nominal sorbent loss (kg sorbent / tonne CO_2 captured)

First please give us your <u>lowest</u> estimate of the value that you think that nominal sorbent loss might be in state-of-the-art operating plants by 2015.

| kg/ tonne CO_2 captured |

Suppose that we have a clairvoyant who can look ahead to 2015 and tells you that the actual number is 0.85 times this value. Can you think of any plausible story of how that might have happened?

o No o Yes

If you answered yes, briefly what is the story:

_____
_____
_____

Is there at least a 5% probability that this story could turn out to be true?

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o No o Yes

Next we'd like your <u>highest</u> estimate of the value that you think that nominal sorbent loss might be in state-of-the-art operating plants by 2015.

kg/ tonne CO$_2$ captured

Again, suppose that we have a clairvoyant who can look ahead to 2015 and tells you that the actual number is 1.15 times this value. Can you think of any plausible story of how that might have happened?

o No o Yes

If you answered yes, briefly what is the story:

Is there at least a 5% probability that this story could turn out to be true?

o No o Yes

Finally we'd like your <u>best</u> estimate of the value that you think that average nominal sorbent loss might be in state-of-the-art operating plants by 2015.

kg/ tonne CO$_2$ captured

Parameter 4: Cost of sorbent (US$ / tonne sorbent)

First please give us your <u>lowest</u> estimate of the value that you think that unit cost of sorbent might be in state-of-the-art operating plants by 2015.

US$ / tonne sorbent

Suppose that we have a clairvoyant who can look ahead to 2015 and tells you that the actual number is 0.85 times this value. Can you think of any plausible story of how that might have happened?

o No o Yes

If you answered yes, briefly what is the story:

Is there at least a 5% probability that this story could turn out to be true?

o No o Yes

Next we'd like your <u>highest</u> estimate of the value that you think that unit cost of sorbent might be in state-of-the-art operating plants by 2015.

US$ / tonne sorbent

Again, suppose that we have a clairvoyant who can look ahead to 2015 and tells you that the actual number is 1.15 times this value. Can you think of any plausible story of how that might have happened?

o No o Yes

If you answered yes, briefly what is the story:

_____

_____

_____

Is there at least a 5% probability that this story could turn out to be true?

o No o Yes

Finally we'd like your <u>best</u> estimate of the value that you think that average unit cost of sorbent might be in state-of-the-art operating plants by 2015.

US$ / tonne sorbent

Parameter 5: Other Sorbents

The preceding questions have all focused on amine-based sorbents. Today there is much research being directed at developing alternative sorbents for post-combustion CO_2 capture from flue gas. How likely do you think it is that by 2015 there will be some other commercially available sorbent (other than advanced amines) that significantly out-performs amine-based sorbents in CO_2 capture from flue gas and which is economically competitive? (please mark an <u>X</u> on the line)

```
0.0     0.2     0.4     0.6     0.8     1.0
 |-------|-------|-------|-------|-------|
```

If your answer was less than 0.5, do you think that some other commercially available and economically competitive sorbent will become available some time after 2015?

o No

o Yes in approximately the year 20____

The next page addresses a parameter related to CO_2 compression. Please address this too if you can. You may assume that the product CO_2 flow rate is about 400 tonne CO_2 / hr (with 99.8% purity).

Parameter 6: Energy required for CO₂ Compression

First please give us your <u>highest</u> estimate of the value that you think that energy requirement for CO_2 compression might be in state-of-the-art operating plants by 2015.

<div style="border:1px solid black;padding:4px;">

kWh/ tonne CO₂, or **compressor efficiency =** **%**

</div>

Suppose that we have a clairvoyant who can look ahead to 2015 and tells you that the actual number is 1.15 times this value. Can you think of any plausible story of how that might have happened?

o No o Yes

If you answered yes, briefly what is the story:

_____

_____

_____

Is there at least a 5% probability that this story could turn out to be true?
o No o Yes

Next we'd like your <u>lowest</u> estimate of the value that you think that energy requirement for CO_2 compression might be in state-of-the-art operating plants by 2015.

<div style="border:1px solid black;padding:4px;">

kWh/ tonne CO₂, or **compressor efficiency =** **%**

</div>

Again, suppose that we have a clairvoyant who can look ahead to 2015 and tells you that the actual number is 0.85 times this value. Can you think of any plausible story of how that might have happened?

o No o Yes

If you answered yes, briefly what is the story:

_____

_____

_____

Is there at least a 5% probability that this story could turn out to be true?
o No o Yes

Finally we'd like your <u>best</u> estimate of the value that you think that average energy requirement for CO_2 compression might be in state-of-the-art operating plants by 2015.

<div style="border:1px solid black;padding:4px;">

kWh/ tonne CO₂, or **compressor efficiency =** **%**

</div>

Part 4: Judgments About Overall Plant Performance and Cost

In the following questions we want your comparative judgments about the overall performance and cost of a state-of-the-art operating coal-fired power plant in the year 2015, with and without amine-based CO₂ capture system. Please keep in mind that the terms "reference plant" and "capture plant" refer to the following:

Reference Plant:

- **New low-S coal-fired power plant with environmental control units (ESP, FGD and SCR)**
- Net generation = 500 MW

Capture Plant:

- **New low-S coal-fired power plant (similar to the reference plant) with environmental control units (ESP, FGD and SCR)**
- Amine-based CO_2 capture system to capture 90% CO_2 from the power plant flue gas stream
- The product CO_2 is compressed to 2000 psig (for transport to a disposal site)
- Net power generation = 500 MW
- Costs of CO_2 transport and storage are **not** included

Note: In questions about plant cost, you should also assume:

- **A mid-western US plant location**

- **Total capital requirement (TCR) includes interest during construction and other owner's costs**

- **The cost of fuel is US$1.2/GJ (HHV basis)**

- **Currency is current US dollars (US$)**

Again, feel free to skip any question you're not comfortable expressing a judgment about.

Parameter 7: Net Plant Efficiency

Consider the **reference plant** as described on the previous page.

First please give us your <u>highest</u> estimate of the value that you think that the net efficiency (HHV basis) of this reference plant might be in 2015:

%

Next we'd like your <u>lowest</u> estimate of the value that you think that the net efficiency (HHV basis) of this reference plant might be in 2015:

%

Finally we'd like your <u>best</u> estimate of the value that you think that average net efficiency (HHV basis) of this reference plant might be in 2015.

%

Now consider the **capture plant** as described on the previous page.

First please give us your <u>highest</u> estimate of the value that you think that the net efficiency (HHV basis) of this CO_2 capture plant might be in 2015:

% efficiency points lost, or   **% net efficiency**

Next we'd like your <u>lowest</u> estimate of the value that you think that the net efficiency (HHV basis) of this CO_2 capture plant might be in 2015:

% efficiency points lost, or   **% net efficiency**

Finally we'd like your <u>best</u> estimate of the value that you think that average net efficiency (HHV basis) of this CO_2 capture plant might be in 2015.

% efficiency points lost, or   **% net efficiency**

Parameter 8: Total Capital Requirement

Consider the **<u>reference plant</u>** as described on the previous page.

First please give us your <u>highest</u> estimate of the value that you think that the total capital requirement of this reference plant might be in 2015:

M$ (or	**$/kW$_{net}$)**

Next we'd like your <u>lowest</u> estimate of the value that you think that the total capital requirement of this reference plant might be in 2015:

M$ (or	**$/kW$_{net}$)**

Finally we'd like your <u>best</u> estimate of the value that you think that average total capital requirement of this reference plant might be in 2015.

M$ (or	**$/kW$_{net}$)**

Now consider the **<u>capture plant</u>** as described on the previous page.

First please give us your <u>highest</u> estimate of the value that you think that the total capital requirement of this CO_2 capture plant might be in 2015:

M$ (or	**$/kW$_{net}$)**

Next we'd like your <u>lowest</u> estimate of the value that you think that the total capital requirement of this CO_2 capture plant might be in 2015:

M$ (or	**$/kW$_{net}$)**

Finally we'd like your <u>best</u> estimate of the value that you think that total capital requirement of this CO_2 capture plant might be in 2015.

M$ (or	**$/kW$_{net}$)**

Parameter 9: Cost of Electricity

Consider the **<u>reference plant</u>** as described on the previous page.

First please give us your <u>highest</u> estimate of the value that you think that the cost of electricity generated from this reference plant might be in 2015:

$/MWh (or	**cents/kWh)**	

Next we'd like your <u>lowest</u> estimate of the value that you think that the cost of electricity generated from this reference plant might be in 2015:

$/MWh (or	**cents/kWh)**	

Finally we'd like your <u>best</u> estimate of the value that you think that average cost of electricity generated from this reference plant might be in 2015.

$/MWh (or	**cents/kWh)**	

Did you have a particular capacity factor (load factor) and capital charge factor (fixed charge factor) in mind for these estimates? If so, what are they:

Capacity factor	% (or hours/yr)
Capital charge factor	% (or plant life = yr & i = %)

Now consider the **<u>capture plant</u>** as described on the previous page.

First please give us your <u>highest</u> estimate of the value that you think that the cost of electricity generated from this plant might be in 2015:

$/MWh (or	**cents/kWh)**	

Next we'd like your <u>lowest</u> estimate of the value that you think that the cost of electricity generated from this CO_2 capture plant might be in 2015:

$/MWh (or	**cents/kWh)**	

Finally we'd like your <u>best</u> estimate of the value that you think that cost of electricity generated from this CO_2 capture plant might be in 2015.

$/MWh (or	**cents/kWh)**	

Did you have a particular capacity factor (load factor) and capital charge factor (fixed charge factor) in mind for these estimates? If so, what are they:

Capacity factor	% (or hours/yr)
Capital charge factor	% (or plant life = yr & i = %)

Part 5: Research Priorities

Today there is an ongoing effort worldwide to bring down the cost of CO_2 capture and sequestration from flue gas streams. The factors affecting this cost may be categorized as:

   a)   factors related to the "base" power plant (including other environmental control units)

   b)   factors related to the amine-based CO_2 capture technology

   c)   factors related to the post-capture processing of the concentrated CO_2 stream

On the following pages you will find a list of research objectives that may help reduce the overall cost of CO_2 capture and sequestration from flue gas streams. We would like to know how you would prioritize R&D efforts in these areas.

You may also add more items to this list (items D1, D2 etc.), and take them into consideration while going through the ranking exercise.

Initially, please indicate your priorities as:

       H: High priority   **M**: Medium priority   **L**: Low priority

Finally, please further prioritize your preferences on the same page, by putting a rank/number against each **High Priority** item, taking into account the potential effectiveness of these factors. (You may give the same rank/number to two or more items if you think they are equally preferred). Please note that rank 1 indicates the highest priority.

  [H / M / L]   (RANK)

A1: [   ] To develop more efficient power plants (lower heat rate)   (   )

A2: [   ] To improve boiler designs so that fuel can be burned with lower excess air
       (typically for a coal plant, ~20% excess air is used)   (   )

A3: [   ] To develop more efficient technologies for SO_x and NO_x control so as to
       reduce the acidic gas impurities in the flue gas stream   (   )

A4: [   ] To improve heat integration within the power plants to reduce the energy
       penalty for CO_2 capture due to steam extraction for sorbent regeneration   (   )

B1: [   ] To develop more efficient fans for flue gas handling   (   )

B2: [   ] To develop CO_2 absorbers that can handle higher sorbent concentrations   (   )

B3: [   ] To develop sorbents with higher CO_2 loading capacity   (   )

B4: [   ] To develop sorbents with lower regeneration energy requirement   (   )

B5: [   ] To develop absorber columns offering lower pressure drops
(   )

B6: [   ] To develop absorber columns offering higher CO_2 capture efficiencies   (   )

B7: [   ] To develop more efficient pumps for sorbent circulation   (   )

B8: [   ] To develop more efficient heat exchanging devices
(   )

B9: [   ] To reduce the cost of sorbent manufacturing   (   )

B10:[   ] To develop sorbents with lower makeup requirements (less losses)   (   )

B11:[   ] To develop less expensive technologies for disposal of spent sorbents   (   )

B12:[   ] To develop better instrumentation/ automation in the CO_2 capture system
     so as to reduce the labor requirement   (   )

B13:[   ] To develop better construction materials so as to reduce the losses due to
     corrosion   (   )

C1: [   ] To develop more efficient compressors for CO_2 compression   (   )

C2: [   ] To develop a transport technology that can handle low-pressure
     CO_2 streams   (   )

C3: [   ] To develop less expensive technologies for CO_2 storage/ disposal   (   )

D1: [   ]   (   )

D2: [   ]   (   )

D3: [   ]   (   )

D4: [   ]   (   )

D5: [   ]   (   )