A Technical, Economic and Environmental Assessment of Amine-based CO2 Capture Technology for Power Plant Greenhouse Gas Control

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Abstract

Capture and sequestration of CO_2 from fossil fuel power plants is gaining widespread interest as a potential method of controlling greenhouse gas emissions. Performance and cost models of an amine (MEA)-based CO_2 absorption system for post-combustion flue gas applications have been developed, and integrated with an existing power plant modeling framework that includes multi-pollutant control technologies for other regulated emissions. The integrated model has been applied to study the feasibility and cost of carbon capture and sequestration at both new and existing coal-burning power plants. The cost of carbon avoidance was shown to depend strongly on assumptions about the reference plant design, details of the CO_2 capture system design, interactions with other pollution control systems, and method of CO_2 storage. The CO_2 avoidance cost for retrofit systems was found to be generally higher than for new plants, mainly because of the higher energy penalty resulting from less efficient heat integration, as well as site-specific difficulties typically encountered in retrofit applications. For all cases, a small reduction in CO_2 capture cost was afforded by the SO_2 emission trading credits generated by amine-based capture systems. Efforts are underway to model a broader suite of carbon capture and sequestration technologies for more comprehensive assessments in the context of multi-pollutant environmental management.

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1. INTRODUCTION

The control of greenhouse gases is arguably the most challenging environmental policy issue facing the U.S. and other countries. An approach that is gaining widespread interest is to control CO_2 emissions by capturing and sequestering CO_2 from fossil-fuel combustion sources *(1,2)*. The key attraction of this option is that it can allow fossil fuels to continue to be used without contributing significantly to greenhouse warming. This would be a radical departure from conventional thinking about climate mitigation, which would require eliminating or severely limiting the use of fossil fuels. Given our high degree of reliance on fossil fuels (roughly 85% of commercial energy use domestically and globally), and the difficulties — technical, economic and social — of large-scale use of alternative options (like nuclear and renewables), the ability to use fossil energy while avoiding greenhouse gas emissions is a potentially attractive alternative that needs to be carefully studied. Coal-based power plants, which contribute about 30% of total U.S. CO_2 emissions, are the principal targets for this type of CO_2 control technology *(3-5)*.

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 (Figure 1), although they have not been designed for power-plant-scale operations *(6)*. They are based on different physical and chemical processes including absorption, adsorption, membranes and cryogenics *(7-11)*. The choice of a suitable technology depends on the characteristics of the flue gas stream, which depend mainly on the power plant technology. Future coal-based power plants may be designed to capture CO_2 before combustion (using coal gasification systems), or they may employ pure oxygen combustion instead of air to obtain a concentrated CO_2 stream for treatment. Figure 2 shows the variety of power plant fuels and technologies that affect the choice of CO_2 capture systems.
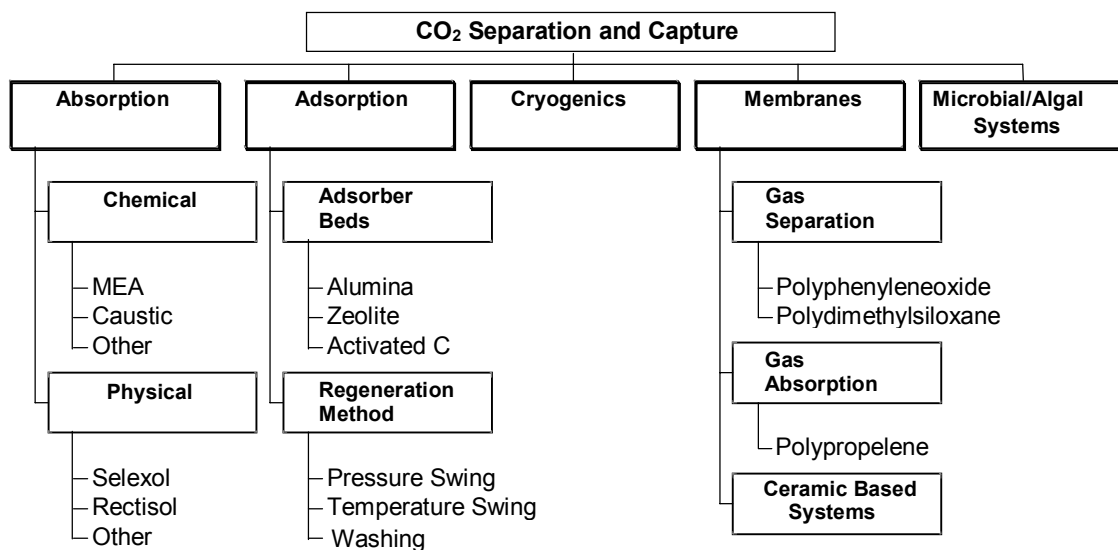
Figure 1. Technology options for CO_2 separation and capture
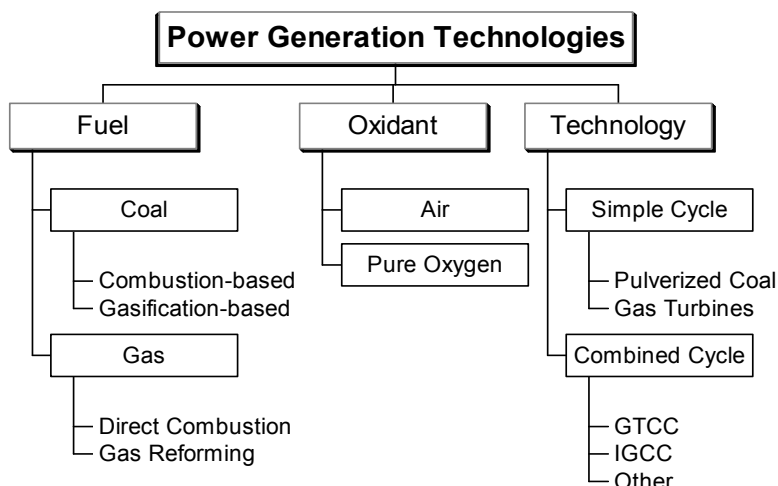
Figure 2. Technology options for fossil-fuel based power generation

1.2. Options for CO_2 Sequestration

Once the CO_2 is captured, it needs to be securely stored (sequestered). Again, there are a range of options potentially available. Geologic formations including deep saline reservoirs, depleted oil and gas wells, and unmineable coal seams are some of the potentially attractive disposal sites *(12-14)*. Ocean disposal is another option being studied *(15-16)*. 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 *(17)*. 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.

2. SCOPE OF THIS STUDY

This research is motivated by a desire to better understand the technological options for CO_2 capture and sequestration and their possible role in climate mitigation policy. Because the topic is fairly new, some of the key research questions that need to be addressed include: What kind of technologies may be used for capture and storage of CO_2? What are the key parameters that affect the performance, cost and environmental acceptability of different options? How do the alternative options compare in terms of these considerations? What are the uncertainties associated with different options? What are the benefits of R&D to reduce key uncertainties?

To begin addressing such questions, this paper focuses on current coal combustion systems. Today 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. 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 capacity compared to today *(18)*. Thus, any new policies to significantly reduce CO_2 emissions during the next two or three decades must consider not only the technology options for new power plants (which is the case typically discussed in the literature), but also the retrofitting of existing coal plants which will continue to operate for several decades to come. Such medium-term intervention to reduce CO_2 emissions has received relatively little attention to date. Hence, the present study examines the feasibility of post-combustion carbon capture at existing power plants as well as new facilities.

In this context, 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 gases, which typically contain only about 10%-12% CO_2 by volume.

- Amine-based CO_2 capture systems are a proven technology that is commercially available and in use today.

- Amine-based systems are similar to other end-of-pipe environmental control systems 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 technology advances.

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 *(19-23)*. However, detailed models of such processes are not generally available. Cost data reported in different studies also tend to be limited and often incomplete. Therefore the first objective of this study has been to develop a preliminary model of performance and cost of amine-based systems based on available information, including a characterization of key uncertainties, as reflected in the current literature.

The second objective is to apply the model to study the feasibility and cost of carbon capture and sequestration at both new and existing coal-based power plants. This analysis incorporates multi-pollutant interactions between CO_2 control, criteria air pollutants (especially SO_2 and NO_x), and air toxics (especially mercury). These aspects of CO_2 mitigation policy analysis are seldom addressed in other studies.

Finally, a third objective is to examine the impact of possible process improvements resulting from R&D. This will help to characterize the role of technological change in looking at future options and costs for CO_2 control.

3. 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 *(24,25)*. 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 *(1)*. 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 *(21)*. 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.

3.1. Process Description

A continuous scrubbing system is used to separate CO_2 from the flue gas stream. As illustrated in Figure 3, 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 original solvent is recovered. The Supplementary Information appendix contains additional details of the process chemistry and design of this system *(27-32)*.
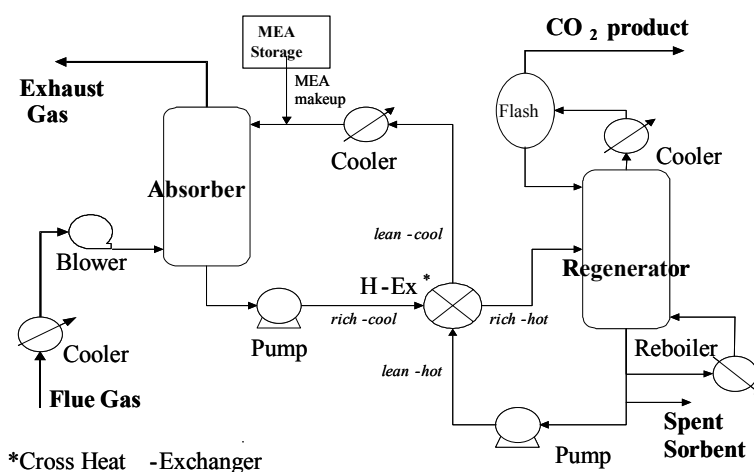
Figure 3. Flowsheet for CO_2 capture from flue gases using amine-based system

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).

4. PROCESS PERFORMANCE MODEL

The removal of CO_2 from flue gases 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. Thus, 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, CO_2 concentration, MEA concentration and absorber design). Similarly, the conditions and detailed design of the regenerator affect the energy requirements and overall performance of the system.

4.1. Performance Parameters

A mathematical model was developed to simulate the performance of a CO_2 capture and storage system based on amine (MEA) scrubbing. This CO_2 module was then added to an existing coal-based power plant simulation model, described later in this paper. Basically, there are two types of input parameters to the CO_2 performance model

Table 1. 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	-
Lean solvent CO_2 loading	mol CO_2/mol MEA	0.15-0.30	0.22	Triangular(0.17,0.22,0.25)
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	-
MEA loss (exhaust gas)	ppm	1-4	2	Uniform (1,4)
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	-
Cooling water makeup	m^3/tonne CO_2	0.5-1.8	0.8	Triangular (0.5,0.8,1.8)
Solvent pumping head	kPa	35-250	207	Triangular(150,207,250)
Pump efficiency	%	70-75	75	Uniform (70,75)
Gas-phase pressure drop	kPa	14-30	26	Triangular(14,26,30)
Fan efficiency	%	70-75	75	Uniform (70,75)
Equiv. elec. requirement	% regeneration heat	9-19	14[a]	Uniform (9,19)
CO_2 product purity	wt%	99-99.8	99.5	Uniform (99,99.8)
CO_2 product pressure	MPa	5.86-15.16	13.79	Triangular(5.86,13.79,15.16)
Compressor efficiency	%	75-85	80	Uniform (75,85)

[a] For retrofit applications, nominal value is 25.

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

- *Parameters of the CO_2 system*: The CO_2 module specifies parameters of the CO_2 capture technology, CO_2 compression system, CO_2 product transport, and CO_2 storage (sequestration) method. The basic configuration is an MEA-based absorption system with pipeline transport of liquefied CO_2 to a geologic sequestration site. Table 1 lists the model input parameters that affect overall process performance. These parameters, along with those from the base plant, are used to calculate the solvent flow rate, MEA requirement, regeneration heat requirement, and electrical energy needs of the CO_2 system. Functional relationships and default values for all model parameters were developed for a typical system design based on a detailed review of the literature, discussions with process developers and other experts, and the use of detailed process simulation models *(33,34)*. The resulting model employs fundamental mass and energy balance relationships together with empirical relationships for some key parameters. Details of the performance model are presented as an attachment *(35)*. All of the performance parameters also affect the cost of the system.

4.2. Characterization of Uncertainties

This modeling effort also incorporates a stochastic simulation capability that allows any or all model inputs to be represented by a probability distribution rather than a single deterministic value. In general, such distributions may reflect uncertainty and/or variability in system design parameters. Table 1 shows the distributions used in this paper to characterize the performance parameters of the CO_2 capture system. These distributions reflect the current literature on amine-based (MEA) systems and (in a few cases) data and judgments provided by process developers *(36)*. Detailed descriptions of these distributions are provided as an attachment *(35)*.

4.3. Model Outputs

The key outputs of the amine system performance model include

- *MEA requirement*. This depends mainly on the mass flow rate of CO_2 in the flue gas, the desired CO_2 capture efficiency, the MEA concentration, and CO_2 loadings in the solvent. Depending on the level of impurities in the flue gas, there is additional loss of solvent associated with removal of other acid gases such as SO_x and NO_2.

- *Energy requirements*. Heat requirements for solvent regeneration depend mainly on the lean sorbent loading and other system parameters. This heat is provided by low-pressure steam within the power plant, thus decreasing the net power generation efficiency. In addition, electrical energy is required for CO_2 product compression, solvent circulation, and other system requirements. The total amine system energy requirement is one of the most important model results, as it dictates the net power plant output, and hence the net cost of power generation and CO_2 avoidance.

- *Environmental emissions*. The CO_2 control system generates several new waste products that are accounted for in the model, principally ammonia gas (generated by degradation of MEA) and reclaimer bottoms (a potentially hazardous solid waste generated during recovery of spent sorbent from the process). On the other hand, the CO_2 capture system also reduces emissions of particulate matter and acid gases like SO_2, HCl and NO_2. Removal of trace metals like mercury and other air toxics is not well-characterized for these systems, but is anticipated to be similar to other wet scrubber systems. A complete accounting of multi-pollutant emissions is provided by the IECM framework in which the CO_2 module is embedded, as discussed below.

4.4. The IECM Framework

The Integrated Environmental Control Model (IECM) is a power plant simulation model developed by Carnegie Mellon University for the U.S. Department of Energy (USDOE) *(37, 38)*. It includes a menu of technological options for controlling criteria air pollutants (SO_2, NO_x, and particulates), mercury (an air toxic), and solid wastes. The new CO_2 module has been integrated into the existing IECM modeling framework, allowing the analysis of alternative CO_2 capture and sequestration technologies in complex plants involving multi-pollutant emission controls. (This newly integrated version is designated IECM-CS.)

5. PROCESS COST MODEL

The CO_2 capture and sequestration system cost model is directly linked to the process performance model described earlier. The cost model follows the framework used in the IECM to ensure consistency in plant-level economic calculations *(35)*. There are four types of cost calculated by this model based on available data *(8,19,23,26,30,39-42)*.

5.1. Capital Cost

The total capital requirement (TCR) of a system is the sum of direct equipment costs (which depend on one or more performance variables that determine the size or capacity of a component), plus various indirect costs that are calculated as fractions of the total process facilities cost (PFC), following the EPRI cost estimating guidelines *(43)*. Table 2 lists the elements of capital cost, showing the nominal parameter values plus the uncertainties used later for probabilistic analyses. For the MEA system, the absorber capital cost depends mainly on the flue gas flow rate. The cost of the regenerator section and the CO_2 compressor scale mainly with the mass flow rate of CO_2 captured. Indirect cost factors are estimated based on available data for amine systems and similar technologies.

Table 2. MEA Cost Model Parameters and Nominal Values

Capital Cost Elements	Nom. Value*	O&M Cost Elements	Nom. Value*
Process Area Costs (9 areas)[a]		*5.2. Fixed O&M Costs (FOM)*	
Total Process Facilities Cost	PFC[b]	Total Maintenance Cost	2.5 % TPC[j]
Engineering and Home Office	7 % PFC[c]	Maintenance Cost	40 % of total maint. cost
General Facilities	10 % PFC[d]	Allocated to Labor	
Project Contingency	15 % PFC[e]	Admin. & Support Labor	30 % of total labor cost
Process Contingency	5 % PFC[f]	Cost	
Total Plant Cost (TPC) = sum of above		Operating Labor	2 jobs/shift[k]
Interest During Construction	calculated	*Variable O&M Costs (VOM)*	
		Reagent (MEA) Cost	$1250/tonne MEA[l]
Royalty Fees	0.5 % PFC[g]	Water Cost	$0.2/m^3
Pre-production Costs	1 month[h] VOM & FOM	CO_2 Transport Cost	$0.02/tonne CO_2/km[m]
Inventory (startup) Cost	0.5 % TPC[i]	CO_2 Storage/Disposal Cost	$5/tonne CO_2[n]
Total Capital Reqmt (TCR) = sum of above		Solid Waste Disposal Cost	$175/tonne waste[b]

*Uncertainty distributions are given below. [a]The individual process areas modeled are: flue gas blower, absorber, regenerator, solvent processing area, MEA reclaimer, steam extractor, heat exchanger, pumps, CO_2 compressor. The sum of these is the total process facilities cost (PFC). The uncertainty distributions used are: [b]Normal (1.0,0.1), [c]Triangular (5,7,15), [d]Triangular (5,10,15), [e]Triangular (10,15,20), [f]Triangular (2,5,10), [g]Triangular (0,0.5,0.5), [h]Triangular (0.5,1,1), [i]Triangular (0.4,0.5,0.6), [j]Triangular (1,2.5,5), [k]Triangular (1,2,3),[l]Uniform (1150,1300), [m]Triangular (0.004,0.02,0.06), [n]Triangular (-10,5,8)

5.3. O&M Costs

Operating and maintenance (O&M) cost elements are also listed in Table 2. Major variable cost items include the cost of sorbent (MEA) and the costs of CO_2 transport and storage. Fixed costs include the costs of maintenance and labor. Energy costs are handled internally in the model by derating the overall power plant based on the calculated energy requirement. This increases the cost per net kilowatt-hour delivered by the plant.

5.4. Cost of Electricity

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). Two key parameters in this calculation are the levelized fixed charge factor (used to amortize capital expenses), and the plant capacity factor. The fixed charge factor is based on the plant lifetime (book life) and after-tax discount rate (or interest rate, or rate of return), while the capacity factor reflects the average annual hours of plant operation. The incremental cost of electricity attributed to CO_2 control is then the difference in COE between plants with and without the CO_2 capture and storage system.

5.5. Cost of CO_2 Avoided

Analysts commonly express the cost of an environmental control system in terms of either the cost per tonne of pollutant removed or the cost per tonne "avoided." For an energy-intensive system like amine scrubbers there is a big difference between the cost per tonne CO_2 removed and the cost per tonne CO_2 avoided based on *net* plant capacity. Since the purpose of adding a capture unit is to reduce the CO_2 emissions per net kWh delivered, the cost of CO_2 avoidance (relative to a reference plant with no CO_2 control) is the economic indicator most widely used. It can be calculated as

$$\text{Cost of } CO_2 \text{ Avoided (\$/tonne)} = \frac{(\$/\text{kWh})_{\text{capture}} - (\$/\text{kWh})_{\text{reference}}}{(\text{tonne } CO_2/\text{kWh})_{\text{reference}} - (\text{tonne } CO_2/\text{kWh})_{\text{capture}}}$$
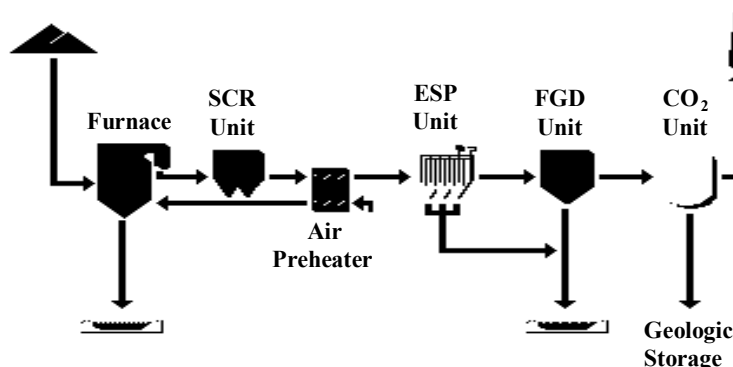
Figure 4. Schematic of an NSPS plant including CO_2 capture unit

6. CASE STUDY RESULTS FOR A NEW PLANT

The IECM-CS simulation framework models a complete coal-fired power plant with multi-pollutant environmental controls, including CO_2 capture and sequestration. The reference case is assumed to be a new 500 MW unit (gross) burning low-sulfur western U.S. coal, and meeting current federal New Source Performance Standards (NSPS) for SO_2, NO_x and particulates *(44)*. The CO_2 capture case adds an MEA scrubber. Figure 4 shows a schematic of the plant and its environmental control systems, including the CO_2 capture unit. Table 3 lists key plant design parameters and the uncertainties assumed for probabilistic analysis.

Table 3. Design Parameters for Case Study of New Pulverized Coal Plant

Parameter	Value	Parameter	Value
Gross plant size (MW)	500	Emission standards	2000 NSPS[d]
Gross plant heat rate (kJ/kWh)	9600[a]	NO_x Controls	LNB[e] +SCR[f]
Plant capacity factor (%)	75[b]	Particulate Control	ESP[g]
		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 sub-critical unit. Uncertainty case includes supercritical unit. The uncertainty distributions used are: Unc = Uniform(9230,9600); [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]See Table 1 for uncertainty. [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)

6.1. Deterministic Results

We run the model first for the reference power plant (without CO_2 capture) and then for the same plant including CO_2 capture. Table 4 quantifies the impact of the CO_2 unit on plant performance, plant discharges and plant costs. Figure 5 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 \$59/tonne for this case. Most of this cost (79%) is associated with the CO_2 capture process (including gas compression). CO_2 transport (8%) and storage (13%) account for the remainder.

Table 4. Deterministic Results for a New 500 MW Plant

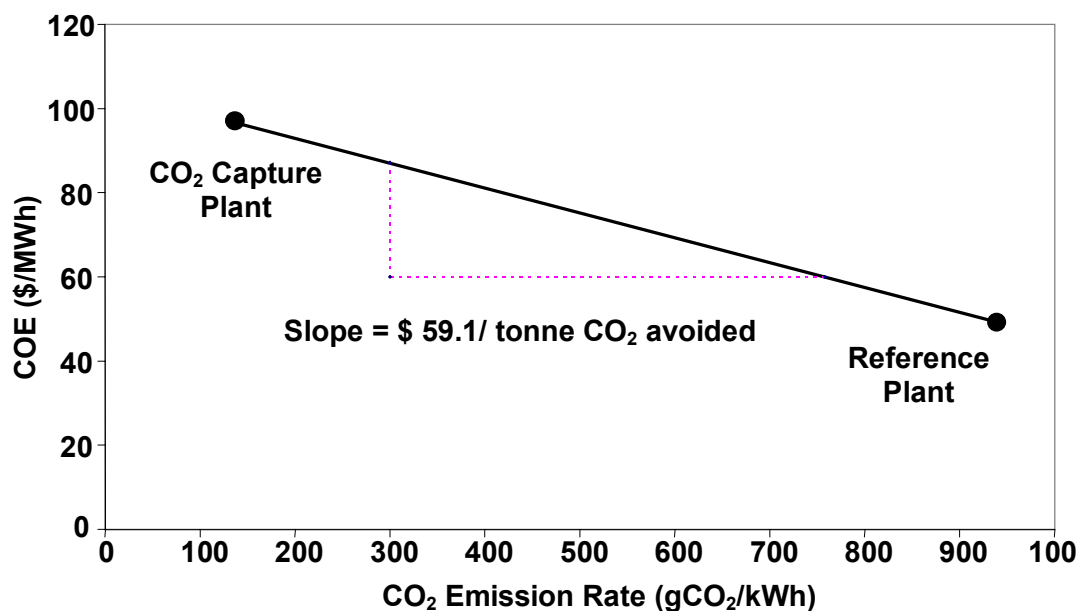| Parameter	Units	Reference Plant	w/CO_2 Control
Net plant capacity	MW (net)	462	326
CO_2 emission rate	g CO_2/kWh (net)	941	133
SO_2 emission rate	g SO_2/kWh (net)	2.45	0.0003
NO_x emission rate	g NO_x/kWh (net)	0.45	0.58
CO_2 sequestered	10^6 tonne CO_2/yr	-	2.58
Cost of electricity	\$/MWh (net)	49.2	97.0
CO_2 mitigation cost	\$/tonne CO_2 avoided	-	59.1

Figure 5. Nominal CO_2 avoidance cost for the case study coal plant
(New 500 MW, Low-S coal, 75% CF)

Addition of the CO_2 capture system increases the total plant capital cost from \$571M (for the reference plant) to \$705M. Energy requirements consume about 27% of gross plant capacity,

mostly for solvent regeneration (49% of total) and CO_2 product compression (34% of total). Solvent circulation and fan power account for the remaining share (17% of total). MEA makeup requirements contribute about 10% to the cost of CO_2 capture. The presence of SO_2 impurities can substantially increase these 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 only 70% SO_2 removal in the reference case. The costs of this upgrade were charged to the CO_2 unit.

6.2. Probabilistic Results

Different assumptions about power plant design, coal properties, plant operation and CO_2 control 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 *(35)* and (in some cases) the authors' judgment. Note that the term "uncertainty" is used loosely here to include parameter variability.

Figure 6a first shows the effect of considering uncertainties and design variability only in the performance parameters of the MEA system (from Table 1). The resulting distribution for cost of CO_2 avoidance has a 95-percentile range of $43-72/tonne. The main contributors to this range are the lean solvent CO_2 loading (which determines the reboiler duty) and equivalent electrical penalty of the regeneration steam requirement. Figure 6b shows the additional uncertainty contributed by CO_2 cost model parameters, including the costs of the CO_2 capture unit, pipeline transport and geologic storage (see Table 2). Now the overall cost of CO_2 avoided shows a much wider range of $33-73/tonne (95% probability interval). The dominant factor here is the assumed cost of CO_2 storage. The high end of the distribution corresponds to a disposal cost of up to $8/tonne CO_2 stored, while the low end reflects a cost credit (revenue) of up to $10/tonne when CO_2 is used for enhanced oil recovery (EOR).
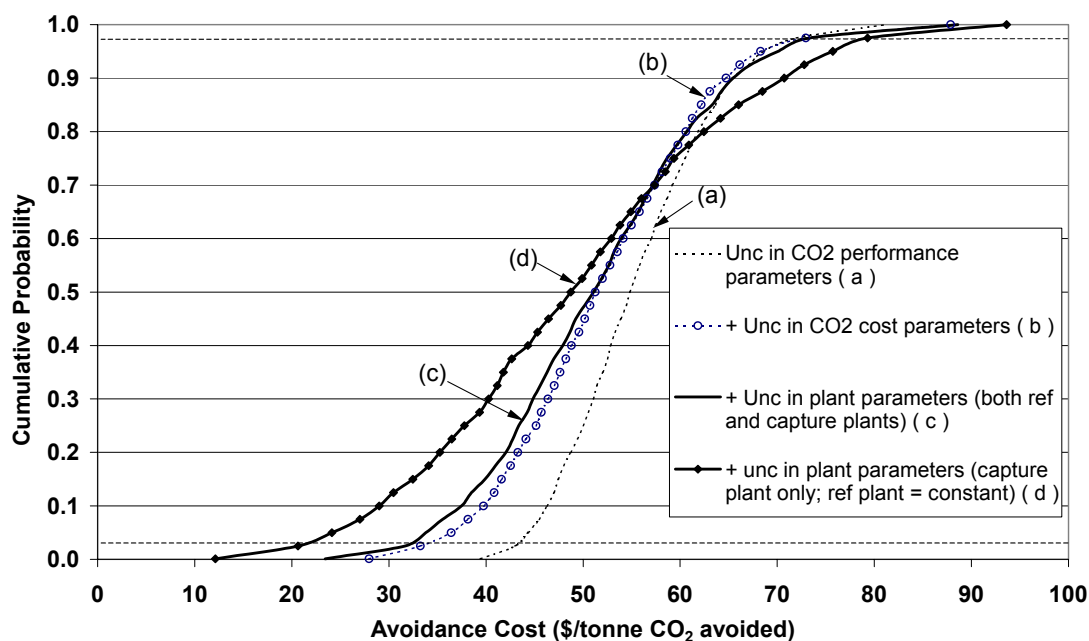
Figure 6. . Uncertainty in the cost of CO_2 avoided, decomposed into three parameter categories. The shaded areas at the top and bottom of the graph demark the 95% probability interval.

Figure 6c and Figure 6d show the effect of additional uncertainties (or variability) for the assumed power plant efficiency (heat rate), capacity factor, coal price, and fixed charge factor (see Table 3). If these parameter values are identical for the reference plant and capture plant (Figure 6c), the range in the CO_2 avoidance cost increases slightly ($32-73/tonne), although the COE of each plant is strongly affected. However, if the capture plant is different from the reference plant (Figure 6d) the CO_2 mitigation cost now broadens to a much wider range of $21-79/tonne CO_2 avoided (95% probability interval). The dominant factors here are the fixed charge rate and plant capacity factor, which strongly influence the COE for the overall plant, and thus the cost of CO_2 avoided.

Note that the tails of the overall distribution span an order of magnitude in the cost of CO_2 avoided. Note too that the distributions in Figure 6 are not symmetric about the deterministic value of avoided cost shown in Figure 5. This is a consequence of the distributions assumed for key model parameters like lean sorbent loading, CO_2 transportation and disposal cost and power plant heat rate. The median and mean cost (Figure 6c) are both $51/tonne CO_2 avoided.

7. 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).

Table 5. Scenarios for Retrofit Studies of a 500 MW Plant

CASE	A	B	C	D
Coal Type	Low-S	Low-S	Low-S	High-S
Existing SO_x control	*None*	*None*	FGD (70% removal)	FGD (~ 90% removal)
Existing NO_x control	LNB	LNB	LNB	LNB
CO_2 Retrofit Option	MEA system	MEA system plus new FGD	MEA system plus FGD upgrade	MEA system plus FGD upgrade

Cases A and B are for a plant burning low-sulfur western coal (0.48% S, as in Table 3) with no other SO_2 emission controls. It complies with federal standards for plants constructed before July 1978. To reduce CO_2 emissions the plant can either retrofit an MEA system (Case A), or install both MEA plus a new FGD system (Case B) to reduce the input of sulfur impurities to the MEA unit. In Case C the low-sulfur coal plant is already equipped with an FGD system that meets federal New Source Performance Standards (NSPS), which in this case is 70% SO_2 removal. This plant has the option of upgrading the existing SO_2 scrubber (to 99% removal) in addition to installing an MEA system. Case D is similar to Case C except that the plant burns high-sulfur (3.25% S) eastern bituminous coal. Federal standards in this case require about 90% reduction in potential SO_2 emissions, achieved with a wet FGD system.

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 also credit the 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 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 (see Table 1).

7.1. Results for Low-Sulfur Coal Plants

Figure 7 shows the deterministic results for the cost of electricity versus CO_2 emission rate for the three low-sulfur coal plant retrofit cases (A, B, C), plus the new plant case shown earlier. Without CO_2 capture, the fully amortized reference plant produces electricity at a cost that is 60-65% lower than the new plant. This lower COE reduces the cost of CO_2 capture when the energy-intensive MEA unit is added to the plant. The analysis also shows that for an existing facility without SO_2 control (Cases A and B), the addition of an SO_2 scrubber along with the CO_2 scrubber significantly reduces the cost of carbon mitigation. For a plant that already has FGD (Case C), upgrading the existing system to achieve higher sulfur removal efficiency further reduces the cost of carbon capture. These results are consistent with the design premises of other studies *(26, 39, 40)* that include SO_2 removal prior to an amine scrubber.
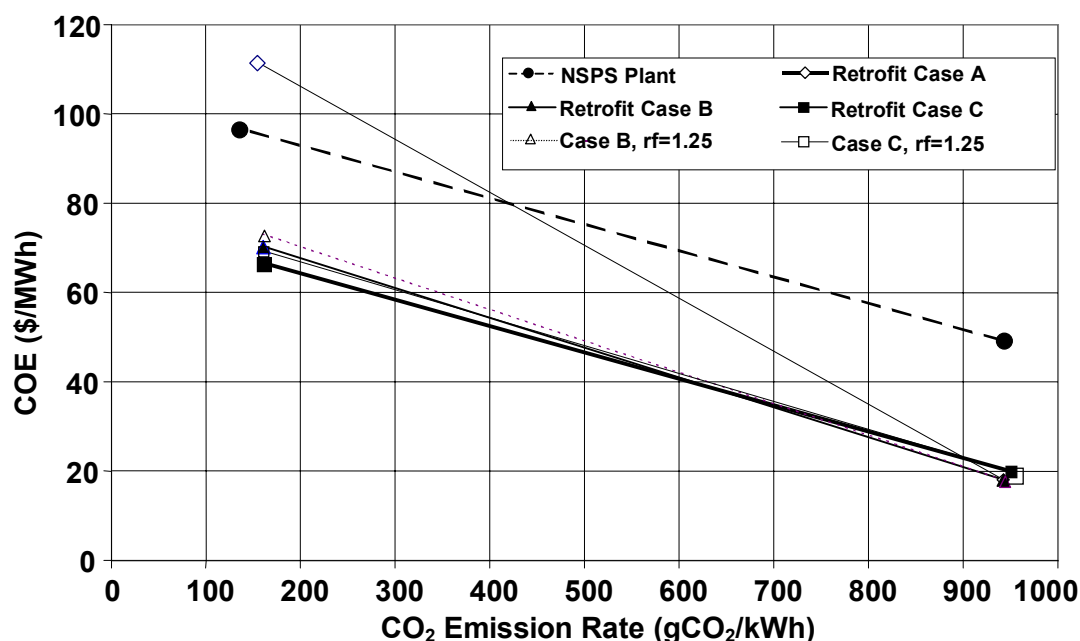
Figure 7. Estimation of CO$_2$ avoidance cost for low-S coal plants

In all cases, retrofitting the power plant with an MEA system results in substantial added reductions in SO$_2$ emissions, which is an additional environmental benefit. The interaction between SO$_2$ and CO$_2$ controls is shown in Table 6. If credits for SO$_2$ reductions at recent market prices are taken into account, the carbon mitigation cost decreases slightly, by about \$1-1.5/tonne CO$_2$ avoided. Even with SO$_2$ credits, however, the CO$_2$ mitigation cost for the retrofit plant is likely to be higher than for a new plant. This is mainly because of the much higher energy penalty resulting from limitations in efficient heat integration. Retrofit costs also are likely to be higher because of site-specific difficulties of installing new equipment at an existing plant location. Assuming a 25% capital cost premium for retrofit applications (Cases B and C with rf = 1.25 in Figure 7), adds about \$3-5 per tonne of CO$_2$ avoided. Still, the overall COE for the retrofitted plant (Cases B and C) is less than the new plant COE because of the amortized capital. Other schemes that can reduce the energy penalty costs (e.g., an auxiliary boiler system fired by low-cost natural gas) could potentially improve the economics of retrofit applications.

Table 6. Effect of SO$_2$ Controls on Carbon Mitigation Costs
for the Low-Sulfur Retrofit Scenarios

Quantity	Reference Plant (no CO$_2$ control)	Case A (MEA only)	Case B (MEA + FGD)	Case C (MEA + FGD upgrade)
Net power (MW)	470	288	275	275
COE ($/MWh)	18.0	111.8	70.4	66.7
SO$_2$ emission rate (g/kWh)	4.51	0.04	0.0004	0.0004
Total SO$_2$ emission (tonne/yr)	13,916	69	0.7	0.7
New SO$_2$ capture in FGD (tonne/yr)	0	0	13,777	7,451
New SO$_2$ capture in MEA(tonne/yr)	0	13,847	138	138

Total new SO_2 capture (tonne/yr)	0	13,847	13,915	7,589
Mitigation cost ($/tonne CO_2 avoided)				
No SO_2 credit	-	118.8	67.0	59.2
$150/tonne SO_2 credit	-	117.4	65.5	58.4
No SO_2 credit, Retrofit factor = 1.25	-	123.2	70.5	62.7

7.2. Results for High-Sulfur Coal Plants

CO_2 mitigation costs for Case D (3.25 % S coal) were again minimized by adding new FGD capacity to achieve 99% SO_2 removal. The COE for the fully amortized plant was 27 $/MWh without CO_2 control and 75 $/MWh with controls, yielding a nominal CO_2 mitigation cost of $64/tonne avoided. The effects of SO_2 credits and retrofit cost premiums were similar to those for the low sulfur coal cases in Table 6.

8. BENEFITS OF R&D

R&D programs that improve the performance of current amine-based system can reduce the future cost of CO_2 capture. The Supplementary Information includes a hypothetical example in which improvements in MEA system performance, and reductions in uncertainties, produce a substantial reduction in the cost of CO_2 avoided. New or improved sorbents, lower regeneration heat requirements, and overall process optimization are some of the potential sources of cost reduction. Government-sponsored R&D efforts, together with industrial initiatives like the CO_2 Capture Project *(45)*, are actively pursuing such goals. Our future modeling research will seek to realistically estimate potential process improvements, and apply those results to quantify potential R&D benefits and priorities.

9. DISCUSSION

The present study has established a framework for quantifying the impacts of carbon capture and sequestration technology on power plant performance, emissions and economics. For amine (MEA)-based absorption systems applied to coal-fired power plants, the cost of carbon avoidance was shown to depend strongly on assumptions about the reference plant design, details of the CO_2 capture and storage system designs, and interactions with other pollution control systems. Climate mitigation policy models that overlook or oversimplify these factors may produce misleading results regarding the feasibility and cost of CO_2 capture and sequestration options.

The presence of acid gas impurities like SO_2 and NO_2 in power plant flue gas was seen to adversely affect the performance and cost of the CO_2 removal system. Adding or upgrading an FGD unit to remove SO_2 was essential to minimize the cost of carbon mitigation. The presence of NO_x had a much smaller effect on CO_2 capture costs since most NO_x is NO, not NO_2. Because of such interactions, more stringent future regulation on SO_x, NO_x and other emissions (e.g., for control of $PM_{2.5}$, ozone, air toxics or acid rain) could make subsequent carbon capture less expensive (at the margin), thus integrating local and regional air pollution policies with the global issue of climate change.

An analysis of retrofit options found that the large energy requirements of CO_2 capture lead to a more substantial loss of plant capacity compared to a new plant affording better heat integration. Site-specific difficulties may further increase the capital cost of a retrofit installation. Thus, the overall cost of CO_2 capture is likely to be greater than that of a new plant, despite the lower cost of electricity for plants that are fully or partially amortized, or the potential credits for new SO_2 reductions accompanying CO_2 controls.

The large plant derating resulting from amine-based CO_2 controls also will significantly affect the capacity planning decisions of electric utilities if this technology is widely implemented. Analysis of these system-wide effects on future capacity requirements, technology choices, demand projections, costs, and CO_2 emissions require a more comprehensive modeling framework *(46)*, but must be part of any large-scale policy analysis of CO_2 capture and sequestration technologies.

Finally, an important feature of the study is that it takes into account the uncertainties and variability in key performance and cost parameters that influence the cost of carbon mitigation. Understanding the nature of these uncertainties, and the potential for reducing them, is crucial to projecting future costs and capabilities of new technologies for carbon capture and sequestration. A characterization of potential R&D benefits is the subject of ongoing analysis.

10. SUPPLEMENTARY INFORMATION

Additional details accompanying this paper can be found in an attachment (referenced as [45] throughout the report). The attachment is in draft form and will be resubmitted in final form when it is completed..

11. REFERENCES

1. 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.

2. Parson, E.A. and D.W. Keith (1998). "Fossil fuels without CO_2 emissions", *Science*, **282**, 1053-1054.

3. Herzog H.J., E. Drake and E. Adams (1997). "CO_2 capture, reuse and storage technologies for mitigating global climate change", A White Paper (Final report), MIT Energy Laboratory, Cambridge, MA.

4. Booras, G.S. and S.C. Smelser (1991). "An engineering and economic evaluation of CO_2 removal from fossil-fuel-fired power plants", *Energy,* **16**(11-12), 1295-1305.

5. Jacco, C.M.F., Hendriks, C.A. and K. Blok (1995). "Carbon dioxide recovery from industrial processes", *Climatic Change*, **29**(4), 439-461.

6. Desideri U. and R. Corbelli (1998). "CO_2 capture in small size cogeneration plants: Technical and economical considerations", *Energy Conversion and Management*, **39**(9), 857-867.

7. Riemer, P., Audus, H. and A. Smith. (1993), "Carbon dioxide capture from power stations", IEA Greenhouse Gas R&D Programme, Cheltenham, United Kingdom.

8. Hendriks, C. (1994). *Carbon Dioxide Removal from Coal-fired Power Plants*, 14-223, Kluwer Academic Publishers, The Netherlands.

9. Mimura, T., Satsumi, S., Iijima, M. and S. Mitsuoka (1999). "Developments on energy saving technology for flue gas carbon dioxide recovery by the chemical absorption method and steam system in power plant", in *Greenhouse Gas Control Technologies (ed. by Eliasson B., Riemer P. and A. Wokaun), Proceedings of the 4th International Conference on Greenhouse Gas Control Technologies,* 30 August – 2 September 1998, Interlaken, Switzerland, Elsevier Science Ltd.

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

11. Audus, H. (2000). "Leading options for the capture of CO_2 at power stations", presented at the *Fifth International Conference on Greenhouse Gas Control Technologies,* 13 – 16 August 2000, Cairns, Australia.

12. Adams, D., Ormerod, W., Riemer, P. and A. Smith (1993). "Carbon dioxide disposal form power stations", IEA Greenhouse Gas R&D Programme, Stoke Orchard, Cheltenham, Gloucestershire, GL52 7RZ, UK.

13. Byrer, C.W., and H.D. Guthrie (1998). "Coal deposits: potential geological sink for sequestering carbon dioxide emissions from power plants", *Proceedings of the AWMA's Second International Specialty Conference,* Oct 13-15, 1998, Washington, DC, Air & Waste Management Association, Pittsburgh, PA.

14. Lindeberg, E. and S. Holloway (1999). "The next steps in geo-storage of carbon dioxide", in *Greenhouse Gas Control Technologies (ed. by Eliasson B., Riemer P. and A. Wokaun), Proceedings of the 4th International Conference on Greenhouse Gas Control Technologies,* 30 August – 2 September 1998, Interlaken, Switzerland, Elsevier Science Ltd.

15. Herzog, H.J. (1998). "Ocean sequestration of CO_2 – an overview", *Proceedings of the AWMA's Second International Specialty Conference, Oct 13-15, 1998, Washington, DC*, Air & Waste Management Association, Pittsburgh, PA.

16. Spencer, D.F. (1999). "Integration of an advanced CO_2 separation process with methods for disposing of CO_2 in oceans and terrestrial deep aquifers", in *Greenhouse Gas Control Technologies (ed. by Eliasson B., Riemer P. and A. Wokaun), Proceedings of the 4th International Conference on Greenhouse Gas Control Technologies,* 30 August – 2 September 1998, Interlaken, Switzerland, Elsevier Science Ltd.

17. 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.

18. EIA (2000). *Annual Energy Outlook 2001 (with Projections to 2020),* Energy Information Administration, U.S. Department of Energy, Washington, DC.

19. Smelser, S.C., Stock, R.M. and G.J. McCleary (1991). "Engineering and economic evaluation of CO_2 removal from fossil-fuel-fired power plants", EPRI IE-7365, Vol. 2, Project 2999-10, prepared by Fluor Daniel Inc., for EPRI and IEA, EPRI, Palo Alto, CA.

20. Kohl, A.L. and R.B. Nielsen (1997). *Gas Purification*, 40-277, 5th ed., Gulf Publishing Company, Houston, TX.

21. Herzog, H.J. (1999). "The economics of CO_2 capture", in *Greenhouse Gas Control Technologies (ed. by Eliasson B., Riemer P. and A. Wokaun), Proceedings of the 4th International Conference on Greenhouse Gas Control Technologies,* 30 August – 2 September 1998, Interlaken, Switzerland, Elsevier Science Ltd., 1999, 101-106.

22. Chakma A. and P. Tontiwachwuthikul (1999). "Designer solvents for energy efficient CO_2 separation from flue gas streams", in *Greenhouse Gas Control Technologies (ed. by Eliasson B., Riemer P. and A. Wokaun), Proceedings of the 4th International Conference on Greenhouse Gas Control Technologies,* 30 August – 2 September 1998, Interlaken, Switzerland, Elsevier Science Ltd.

23. Simbeck, D (1999). "A portfolio selection approach for power plant CO_2 capture, separation and R&D options", in *Greenhouse Gas Control Technologies (ed. by Eliasson B., Riemer P. and A. Wokaun), Proceedings of the 4th International Conference on Greenhouse Gas Control Technologies,* 30 August – 2 September 1998, Interlaken, Switzerland, Elsevier Science Ltd.

24. Kaplan, L.J. (1982). "Cost-saving process recovers CO_2 from power-plant flue gas", *Chemical Engineering*, **89**(24), 30-31.

25. Pauley, C.R., Simiskey, P.L. and S. Haigh (1984). "N-Ren recovers CO_2 from flue gas economically", *Oil & Gas Journal*, **82**(20), 87-92

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

27. Sartori, G., Ho, W.A., Thaler, W.A., Chludzinski, G.R. and J.C. Wilbur (1994). "Sterically hindered amines for acid gas absorption", in *Carbon Dioxide Chemistry: Environmental Issues, ed. by J. Paul and C. Pradier*, The Royal Society of Chemistry, Cambridge, UK.

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

29. Suda, T., Fujii, M., Yoshida, K., Iijima, M., Seto, T. and S. Mitsuoka (1992). "Development of flue gas carbon dioxide recovery technology", *Energy Conversion and Management*, **33**(5-8), 317-324.

30. Leci, C.L. (1996). "Financial implications on power generation costs resulting from the parasitic effect of CO_2 capture using liquid scrubbing technology for power station flue gases", *Energy Conversion and Management*, **37**(6-8), 915-921.

31. Yagi, T., Shibuya, H. and T. Sasaki (1992). "Application of chemical absorption process to CO_2 recovery from flue gas generated in power plants", *Energy Conversion and Management*, **33**(5-8), 349-355.

32. Barchas, R. and R. Davis (1992). "The Kerr-McGee/ABB Lummus Crest technology for the recovery of CO_2 from stack gases", *Energy Conversion and Management*, **33**(5-8), 333-340.

33. ProTreat ™ (2002). Optimized Gas Treating, Inc., Houston, TX.

34. Aspen Plus ® 11.1 (2002). Aspen Technology, Inc., Cambridge, MA.

35. Rao, A.B. (2002). Performance and Cost Models of an Amine-Based System for CO_2 Capture and Sequestration, Report to DOE/NETL, from Center for Energy and Environmental Studies, Carnegie Mellon University, Pittsburgh, PA, (a draft is included with this document as an attachment).

36. Mariz C., Fluor-Daniel, Aliso Viejo, CA, Private Communications, 2000-01.

37. Rubin, E.S., Kalagnanam, J.R., Frey, H.C. and M.B. Berkenpas (1997). "Integrated environmental control modeling of coal-fired power systems", *Journal Air & Waste Management Assn.,* **47,** 1180-1188.

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

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

40. Simbeck, D. R. 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 2000, Cairns, Australia.

41. Chapel, D., Ernst, J. and C. Mariz (1999). "Recovery of CO_2 from flue gases: Commercial trends", presented at the Canadian Society of Chemical Engineers Annual Meeting, 4-6 October, 1999, Saskatoon, Saskatchewan, Canada.

42. EPRI (2002). Evaluation of Innovative Fossil Fuel Power Plants with CO_2 Removal. Technical Report 1000316, EPRI, Palo Alto, CA, U.S. Department of Energy/NETL, Pittsburgh, PA.

43. TAG (1993). *Technical Assessment Guide,* EPRI TR 102276, EPRI, Palo Alto, CA.

44. CFR (1999). Code of federal regulations, *Federal Register*, July 1, 1999, 40CFR, Ch.I.

45. CO_2 Capture Project (2001). Accessed at www.co2captureproject.com.

46. 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," *Proceedings of First National Conference on Carbon Sequestration*, U.S. Department of Energy/NETL, Pittsburgh, PA.

Details of
A Technical, Economic and Environmental Assessment of Amine-based CO$_2$ Capture Technology for Power Plant Greenhouse Gas Control

Appendix to Annual Technical Progress Report

Reporting Period October 2000 – October 2001

Anand B. Rao

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1. INTRODUCTION

1.1. Technology Options for CO₂ 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). They are based on different physical and chemical processes including absorption, adsorption, membranes and cryogenics. Figure 1 and Table 1 briefly summarizes the salient features of these technology options (Riemer, et al., 1993; Hendriks, 1994; Mimura et al., 1999; Jeremy, 2000; Audus, 2000). The choice of a suitable technology (which mainly depends on the power plant technology) depends upon the characteristics of the gas stream from which CO_2 needs to be separated. Future power plants may be designed so as to separate out CO_2 from coal before combustion (using coal-gasification systems), or they may employ pure oxygen combustion instead of air so as to obtain a concentrated CO_2 stream for treatment. Figure 2 shows the variety of power plant fuels and technologies that affect the choice of CO_2 capture systems. In this report, post-combustion capture of CO_2 from flue gas streams of conventional power plant using amine-based absorption systems has been considered.
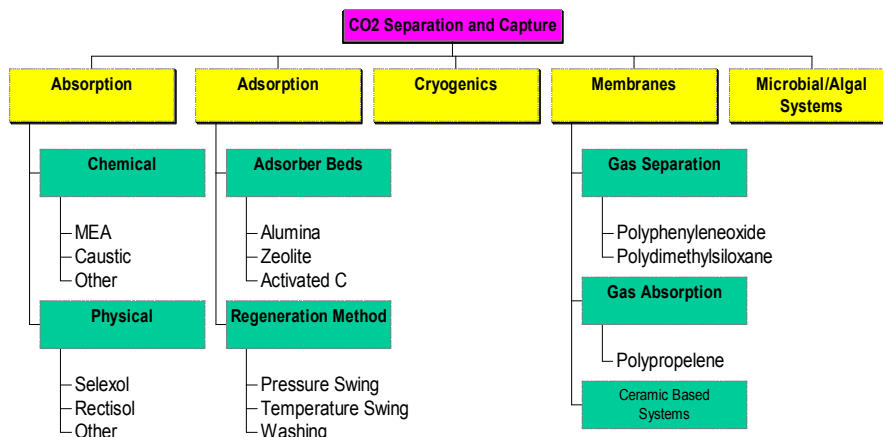
Figure 1. Technology Options for CO₂ Separation and Capture
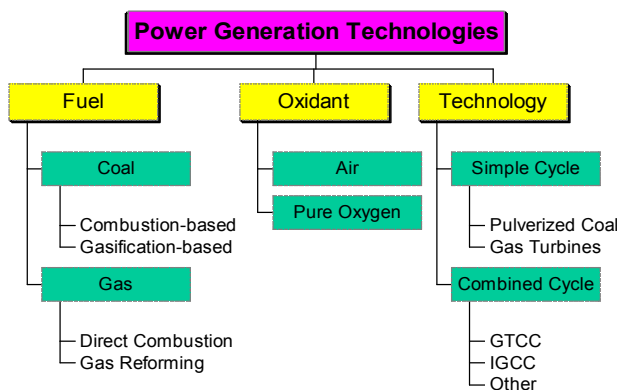
Figure 2. Technology Options for Fossil-Fuel based Power Generation

Table 1. Comparison of Technology Options for CO2 Separation and Capture

Technology Option	System Requirements	1.2. Advantages	Problems/ Drawbacks
Absorption (Chemical)	Absorber and stripper sections Chemical solvent (*e.g.* MEA, HPC)	Suitable for dilute CO_2 streams (typical flue gas from power plants) Operates at ordinary T & P Commercially available, proven technology	The heat of solvent regeneration is very high Significant solvent losses due to acidic impurities in the gas stream
Absorption (Physical)	Absorber and stripper sections Physical solvent (*e.g.* Selexol)	Less energy required Solvents are less susceptible to the impurities in the gas stream	Requires high operating pressure Works better with gas streams having high CO_2 content
Adsorption	Adsorber bed(s)	Very high CO_2 removal is possible	Requires very high operating pressures Costly
Membranes	Membrane filter(s)	Upcoming, promising technology Space efficient	Requires very high operating pressures May require multiple units and recycling due to lower product purity Very costly

1.3. Post-combustion amine-based absorption of CO_2 from flue gases

Today 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. 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 the absolute amount compared to today, according to the latest DOE projections [20]. Natural gas use is projected to account for 31% of power generation in 2020. Thus, any serious policies to reduce CO_2 emissions during the next two decades must consider not only the technology options for new power plants (which is the case commonly discussed in the literature), but also the retrofitting of existing coal and natural gas plants which will continue to operate for several decades to come. This medium-term intervention to reduce CO_2-emissions has received very little attention to date.

In current systems which use air for combustion, post-combustion CO_2 separation from the flue gas stream has to be carried out. 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%-12 % CO_2 by volume.

- Amine-based CO_2 capture systems are a proven technology that are commercially available today.

- Amine-based systems are similar to other end-of-the-pipe 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.

1.3. Model Configuration Options

For post-combustion CO2 capture from flue gas, the amine-based CO_2 capture system, which is the current commercially available technology, has been chosen for this model. There is a major R&D effort going on worldwide to improve this technology – mainly to reduce the high energy penalty of this technology. A substantial part of the energy requirement consists of heat or steam requirement for sorbent regeneration. Depending upon how this steam is supplied, there are three configuration options available. These are shown graphically in Figure 3 and described below.

Base plant de-rating: 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 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 amine system, optimal heat integration may not be achievable, and is likely to lead to 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 by adding an auxiliary NG-fired boiler. Often it would be combined with a steam turbine which could generate some additional power (mainly used to supply electrical energy demand of CO_2 capture unit), and 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 NG fuel. Also, the combustion of NG leads to additional CO_2 emissions (and NO_x emissions). So, there are at least two possible sub-options available

1. The fluegas from the auxiliary boiler is cooled down to acceptable exhaust temperature and then directly vented to the atmosphere. Here, the net CO_2 capture efficiency of the system is substantially lowered because of the additional CO_2 emissions from NG boiler. Also, the total NO_x emissions may exceed the allowable levels of emission. So, the flue gas from the auxiliary NG boiler may have to be treated for NO_x removal before venting.

2. The CO_2 capture system maybe designed so as t capture CO_2 from the additional flue gas as well. In this case, the secondary flue gas stream (after cooling and NO_x removal, if required) maybe merged with main flue gas stream, before it enters the CO_2 capture system. If the NG fuel contains H_2S, the secondary flue gas may have to be treated for SO_x removal as well. The basic purpose of the auxiliary NG boiler is to provide the

steam required for srbent regeneration. With higher amount of flue gas to be treated (and more CO_2 to be captured), the amine-system would require more steam and thus a bigger auxiliary NG boiler would be required (which means more secondary flue gas!). An optimal size of auxiliary NG boiler maybe 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 to the originally desired level, but it often requires substantially big auxiliary NG boiler facility. This may not be always practically feasible (space constraints for retrofit applications, fuel availability, etc.) and economically viable (higher capital cost of building a bigger CO_2 capture system as well as an auxiliary boiler, higher O&M costs etc.). In the present version of IECM, this configuration option is not included.

Figure 3. CO_2 Capture Plant Configuration Options

a)   Reference base plant

b)   Capture plant with internal derating

c)   **Capture plant with an auxiliary boiler and a secondary steam turbine**
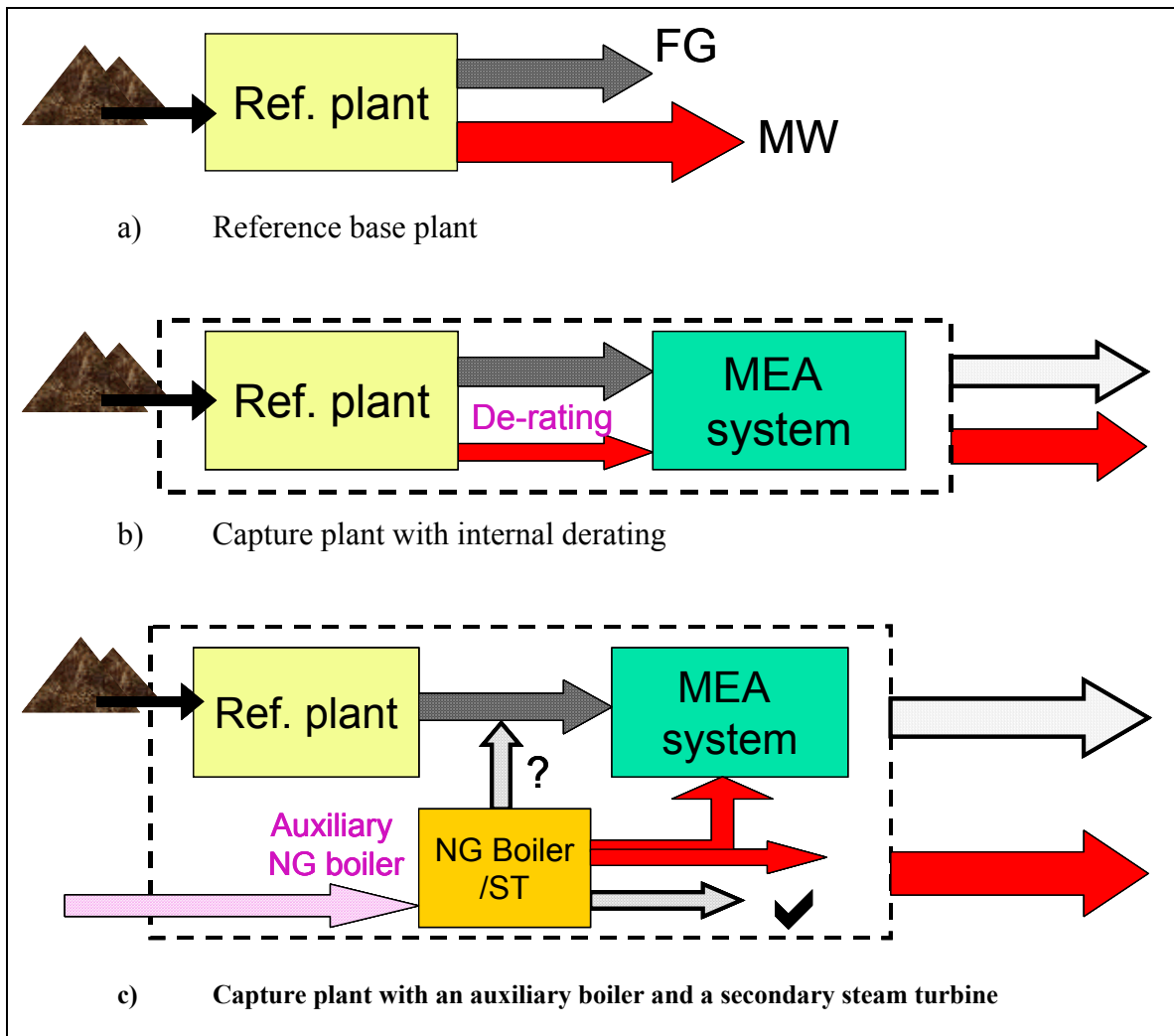
In terms of the configuration of the CO_2 capture system shown in Figure 3, the user can make the following choices as well

Direct contact cooler: The default setting in IECM chooses to include a DCC to cool the flue gas before it enters the amine system. The temperature of the flue gas affects the absorption reaction (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 (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 about 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 a must.

Choice of sorbent: At this time, MEA is the default sorbent used in the system and the nominal values of various parameters are based on a process simulation model that uses MEA. As always, the users can overwrite the nominal values of these parameters if they wish to use a different sorbent (and have the relevant data). In future, the model can adopt a different sorbent by providing the appropriate values for the key parameters.

CO_2 transportation: The default mode of CO_2 transportation is via pipelines. 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 maybe expanded in future to include detailed parameters about pipeline transport and also other transport options.

CO_2 storage/ disposal: The default option for CO_2 disposal is underground geological storage. A nominal cost of \$5/ tonne CO_2 has been suggested, which can be changed 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, maybe expanded in future to include details about the various storage/ disposal options.

2. AMINE-BASED CO_2 CAPTURE SYSTEMS

2.1. Historical Developments

Combustion of fossil fuels in air leads to a gaseous product stream that mainly contains nitrogen, carbon dioxide, water vapor and 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 an obvious source of CO_2 available at no cost. The whole idea of separating CO_2 from flue gas streams started in 1970's, not with concern about the greenhouse effect, but as a possibly economic source of CO_2, mainly for enhanced oil recovery (EOR) operations. Even today, about 80% of CO_2 production is used for EOR (Chapel et al., 1999). Several commercial CO_2 capture plants were constructed in the US in the late 1970's and early 1980's (Kaplan, 1982; Pauley, et al., 1984). 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 plants are still in operation today. But all these plants are much smaller (in terms of tonnage of CO_2 handled) than a typical power plant. Figure 4 gives a rough idea about the various industrial applications of CO_2 capture technologies and their relative magnitude of operations. 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 (USDOE, 1999; Statoil, 2001). The international research community is closely monitoring this facility.

■ mtonCO2/day	EOR	Fertilizer	Soda Ash	Food-grade	Sleipner West Gasfield, Norway	500MW Coal-based Powerplant
mtonCO2/day	1200	150	800	320	3000	9000
Year	'80s	?	1978	?	1996	2010?
Process	Dow MEA	Dow MEA	Kerr-McGee	Dow MEA	MEA based	MEA based
Location	TX & NM	IGFC	NACC	NEEA, MA	Norway	?
Status	shut down	operational	operational	operational	operational	prospective
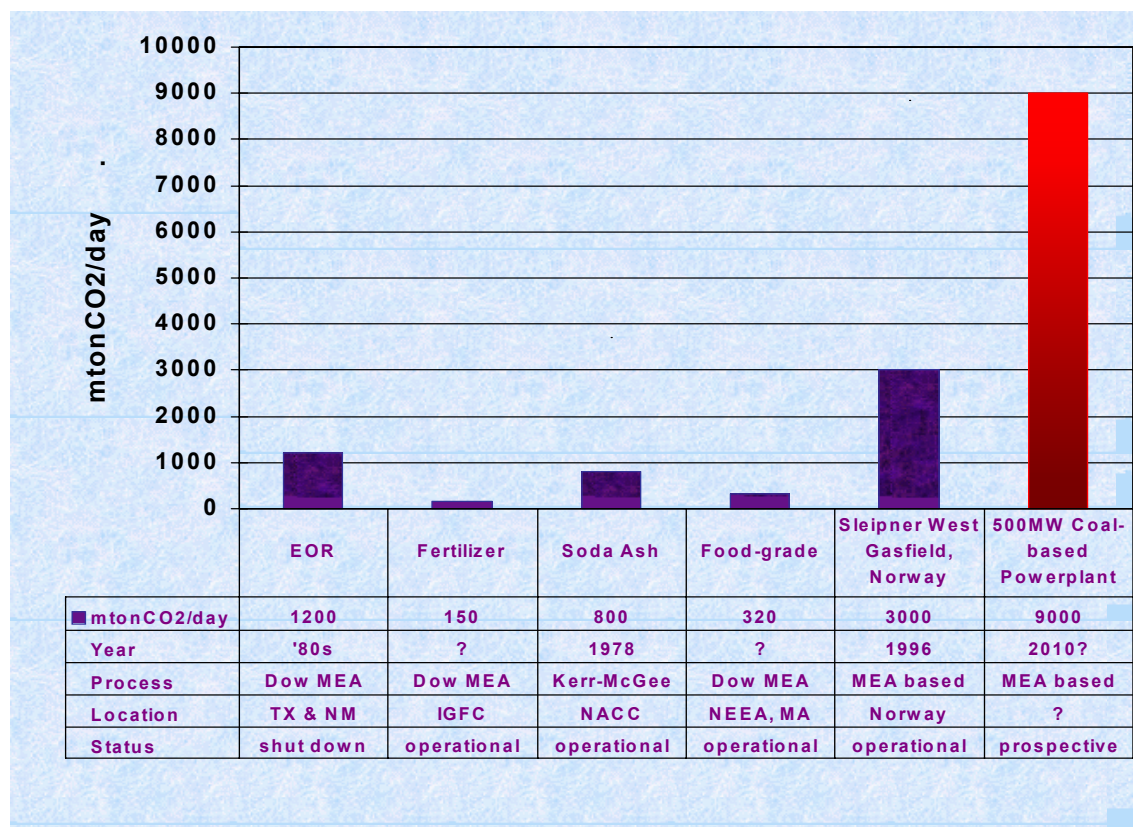
Figure 4. Major industrial applications of CO₂ capture systems

All these plants capture CO_2 with processes based on chemical absorption using a monoethanolamine (MEA) based sorbent. 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 sorbent to remove acidic gas impurities (e.g. H_2S, CO_2) from natural gas streams (Herzog, 1999). 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., were some of the initial developers of MEA-based technology for CO_2 capture. About 75%-95% 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 et al., 2000). Data for the Econamine FG process are more readily available. So, the performance and cost model is based on this 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 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. On the other hand, the ABB Lummus Crest process uses a 15%-20% w/w MEA solution without any inhibitor (Marion 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).

2.2. Process Description

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. Chemical absorption systems tend to be more efficient than the other systems shown in Appendix A, as the process is accompanied by a chemical reaction that enhances the overall mass transfer from gas phase to liquid phase.

In a power plant application (Figure 5) cooled flue gases flow vertically upwards through the absorber countercurrent to the absorbent (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 solution 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, then sent back to the absorber. Some fresh MEA is added make up for losses incurred in the process.
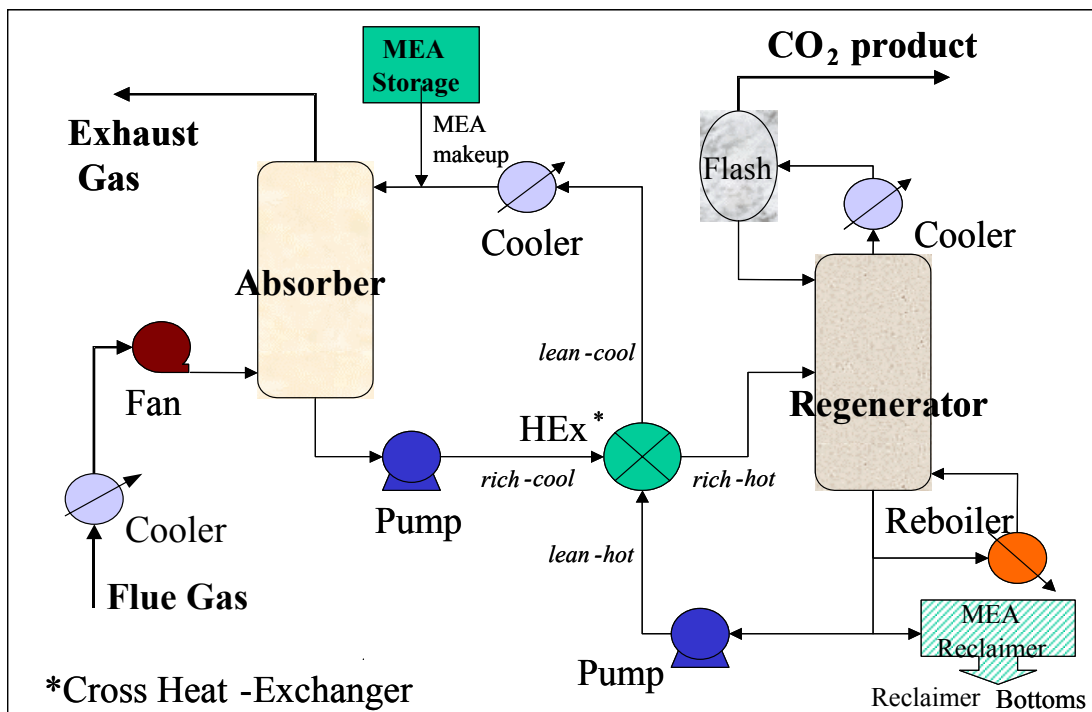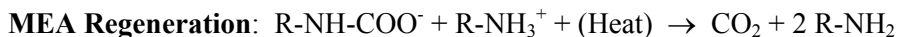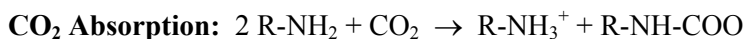

Figure 5. Flow Sheet for CO_2 Capture from Flue Gases using Amine-based System

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 to long distances to the designated storage or disposal facility.
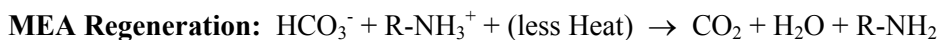
2.3. Process Chemistry

The process chemistry is complex, but the main reactions taking place are [26]

$$\textbf{CO}_2 \textbf{ Absorption: } 2 \text{ R-NH}_2 + CO_2 \rightarrow \text{R-NH}_3^+ + \text{R-NH-COO}$$

$$\textbf{MEA Regeneration}: \text{R-NH-COO}^- + \text{R-NH}_3^+ + (\text{Heat}) \rightarrow CO_2 + 2 \text{ R-NH}_2$$

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 lot of heat energy to break the bonds and to regenerate the sorbent.

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 (Hezorg et al., 1997; Sartori, 1994).

$$\textbf{CO}_2 \textbf{ Absorption: } \text{R-NH}_2 + CO_2 + H_2O \rightarrow \text{R-NH}_3^+ + HCO_3$$

$$\textbf{MEA Regeneration: } HCO_3^- + \text{R-NH}_3^+ + (\text{less Heat}) \rightarrow CO_2 + H_2O + \text{R-NH}_2$$

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 Equipment

The CO_2 capture and separation system consists of the following capital 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 (the absorption being an exothermic process is favored by low temperatures), to minimize sorbent losses (higher temperature may lead to sorbent losses due to evaporation and degradation), and to avoid excessive loss of moisture with the exhaust gases. 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. 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 has to overcome a substantial pressure drop as it passes through a very tall absorber column, countercurrent to the sorbent flow. Hence the cooled flue gas has to 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 some of the CO_2 from the flue gas gets dissolved in the sorbent. The column may be plate-

type or a packed one. 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 needs to 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 has to be cooled down before it could 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 with the application of heat and CO_2 gets separated from the sorbent to leave reusable sorbent behind. In case of unhindered amines like MEA, the carbamate formed is stable and it takes large amount 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 give a fairly rich CO_2 stream.

Reboiler: The regenerator is connected with a reboiler which 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/IP 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: 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 caustic helps in freeing of some of the MEA. The recovered MEA is taken back to the sorbent stream while the bottom sludge (reclaimer waste) is sent for proper disposal.

Sorbent processing area: The regenerated sorbent has to be further cooled down even after passing through the rich/lean cross heat exchanger using a cooler, so that the sorbent temperature is brought back to acceptable level (about 40 deg C). Also, in order to make up for the sorbent losses, a small quantity of fresh MEA sorbent has to be added to the sorbent stream. So, 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.

CO_2 drying and compression unit: The CO_2 product may have to be carried to very long distances via pipelines. Hence it is desirable that it does not contain any moisture in order to avoid corrosion in the pipelines. 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 moisture and other impurities (e.g. N_2) at acceptable levels.

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 seems to be the obvious mode of transportation. 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 be considered. 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 (PanCanadian, 2001).

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 Fig. A-3 and Table A-2 in Appendix A). 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 [14-16]. Ocean disposal and terrestrial sinks are additional options being studied [17-18]. 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 [19]. 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 for capture of CO_2 from power plant flue gases, it has its own drawbacks. The main problems maybe enlisted as follows

Energy Penalty: The stable carbamate ion requires substantial energy to break the bonds. So, 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 needs to be built in order to achieve the same "net" power generation capacity, as it would have been without CO_2 capture.

Loss of Sorbent: Some of the sorbent is lost during the process because of a variety of reasons including mechanical, entrainment, vaporization and degradation (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 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 electrodialysis are also being proposed for this purpose (Yagi et al., 1992).

Corrosion: Corrosion control is very important in amine systems processing oxygen-containing gases. In order to reduce corrosion rates, corrosion inhibitors, lower concentrations of MEA, appropriate materials of construction and mild operating conditions are required (Barchas and Davis, 1992).

3. PERFORMANCE MODEL DEVELOPMENT

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 (Smelster, 1991; Kohl and Nielsen, 1997; Chapel et al., 1999; Herzog, 1999; Chakma and Tontiwachwuthikul, 1999; Simbeck, 1999; Marion 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 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 solvents. Similarly, the design of and conditions in the regenerator affect the energy requirement and the overall performance of the system.

3.1. Process Simulation Tool

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. The *ProTreat* package makes exclusive use of a column model that treats the separation as a mass transfer rate process.

- *ASPEN-Plus* is a powerful process engineering tool for the design and steady-state simulation and optimization of process plants.

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 compressor, cooler, absorber, heat exchangers, regenerator, sorbent circulation pumps etc. Many parameters were varied in the *ProTreat* model. Table 2 summarizes the parameters that were changed and the ranges for each parameter.

Table 2. 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	Input	-	0.73-5.56
7	Reboiler heat duty (Q)	Input	GJ/hr	95-664
8	Q/L	Input	MJ/kmole	2.4-22.5
9	CO_2 capture efficiency	Output	%	41.2-99.9
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	40.4-71.6

The following set of parameters related to the design/configuration of the CO_2 capture system were held constant:

- Absorber height: 40 ft

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

- Inlet flue gas pressure: 3 psi

- Solvent 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 CO_2 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°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 using SAS to derive performance equations

The IECM uses response-surface models to characterize the performance of various technologies. Simple algebraic equations are derived from the process simulation runs and used as performance equations rather than having a detailed process simulation module inside IECM. The key performance output variables were regressed against all the input variables to obtain linear/ logarithmic relationship among them. The data collected from the process simulation runs was used to carry out these multivariate linear regressions using a statistical package called SAS. Only those variables with significance value greater than 0.9995 were retained in the performance equations.

3.3. Performance Parameters

A preliminary model was developed to simulate the performance of a CO_2 capture system based on amine (MEA) scrubbing. This CO_2 module was then added to an existing coal-based power plant simulation model (called IECM), described later in this section. Basically, 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 shown in Appendix A are still under construction.

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

These parameters are used to calculate the solvent 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 several 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.

3.3.1. 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 (after env'l. 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 is G, and the molar fraction of CO_2 in the flue gas is y_{CO2}.

  > Temperature of flue gas = T_{fg}
  >
  > Plant capacity factor = PCF (%)
  >
  > Annual hours of operation = HPY = (PCF/100)*365*24 hrs/yr

3.3.2. Parameters to configure the CO_2 system

These are the choices the user can make in order to configure the CO_2 capture system.

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

- Sorbent regeneration steam supply: Steam extraction from the base plant (default, internal derating) or Steam generated from an auxiliary NG boiler (w/ ST)

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

- Mode of CO2 storage/ disposal: Underground geologic reservoir (default) or EOR or ECBM or Depleted oil/gas wells or Ocean

3.3.3. Parameters controlling the performance of the CO_2 system

Parameters controlling the performance of the CO_2 system: The numerical values to the input parameters are specified by the user. The intermediate and final output parameters are then derived using the performance equations. 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.

CO_2 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.

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

Most of 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 nominal value, but the user can specify the desired level of CO_2 capture efficiency.

MEA concentration (C_{MEA})
The solvent used for CO_2 absorption is a mixture of monoethanolamine (MEA) with water. 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 solvent 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 solvent concentration and the user may choose any value between 15-40%.

Lean solvent CO_2 loading (ϕ_{min})
Ideally, the solvent 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) solvent contains some "left-over" CO_2. The level of lean solvent CO_2 loading mainly depends upon the initial CO_2 loading in the solvent 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 (0.05-0.3).

Liquid to gas ratio (L/G)
The liquid to gas ration 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 one of the parameters derived by the process simulation model.

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.

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

Removal efficiency ($\eta_{acid\ gas}$) and stoichiometric MEA loss ($n_{MEA,acidgas}$)
As discussed before, MEA is an alkaline solvent 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. Also, these gases form heat stable salts (HSS) with MEA that can not be regenerated even after application of heat. So, they cause a (permanent) loss of MEA solvent that may be estimated according the stoichiometry of their reactions with MEA. The typical removal efficiencies of these gases in the absorber using MEA solvent designed for 90% removal of CO_2 are given in Table 3.

Table 3. Removal Efficiency of Acid Gases Due to MEA Solvent
(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	$\eta_{HCl} = 95\%$	$n_{MEA,\ HCl} = 1$

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. Else, this is same as that obtained from the base plant.

The temperature of the flue gas affects the absorption reaction (absorption of CO_2 in MEA solvent 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 (direct contact cooler, flue gas blower, absorber), is directly related to the flue gas temperature.

Nominal MEA loss ($\dot{m}_{MEA, nom}$)
MEA is a reactive solvent. 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 kgMEA/ mton CO_2.

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

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

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

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

NH_3 Generation (n_{NH_3})
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).

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

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. [Each mole of MEA lost in oxidation takes up additional mole of MEA in HSS formation.]

$$n_{MEA, organics} = \left(1 \frac{\text{mole MEA}}{\text{mole org. acids}} \right)$$

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 about 0.13 kg NaOH/ mton CO_2.

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 Na salt of organic acid to the reclaimer bottoms.

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

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

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

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

$$= \dot{m}_{NaOH} \, / \, 40 \tag{7}$$

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% (which may be a function of particle size distribution).

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

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

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

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

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

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

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 have to be controlled. The acceptable level of purity of CO_2 product for most of the applications is about 99.8%.

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 derived form the process simulation model.

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) which is derived from the process simulation model, by the total sorbent circulation molar flow rate (MEA sorbent plus dilution water) in the CO_2 capture system.

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

Unit heat of sorbent regeneration (q_{regen})

This is the amount of heat required for the regeneration of the MEA solvent (loaded with CO_2) in the stripper/ regenerator section. It is expressed as 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 825 Btu/ lb CO_2 (i.e. about 1900 kJ/ kg CO_2). The actual amount of heat required for regeneration of the solvent is much higher (about 2-3 times higher than this theoretical minimum), mainly because of the large amount of latent heat taken up by the dilution water in the solvent. A wide range of numbers have been reported for the regeneration heat requirement of MEA system. Majority of the sources report a heat requirement of about 3800-4000 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 captued (m_{CO2}).

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

Enthalpy of regenerating steam (q_{steam})

The regeneration heat is provided in the form of LP 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 solvent 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.

Heat to electricity equivalence factor (F_{HE})

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

From the data obtained from the available studies (Smelster et al., 1991; Mimura et al., 1997; Bolland and Undrum, 1999; Marion et al., 2001; Hendriks, 1994), this factor has been found to lie in the range (9, 22) for a new plant and (22, 30) for retrofit cases. So, the nominal value (for this new plant application) has been taken as 14%.

$$F_{HE} = 14\% \text{ i.e. } F_{HE} = 0.14 \frac{\text{(kW.s) Electric}}{\text{(kJ) Heat}}$$

So, if 10,000 kJ is the regeneration heat requirement for CO_2 capture operation, then the corresponding loss in power generation capacity of the power plant is estimated as 14% of 10,000 kJ i.e. 1400 kW.s, or (1400/3600 =) 0.39 kWh. It may be noted that, in case of retrofit applications, the energy penalty might be significantly higher, and F_{HE} may be around 25%.

Blower pressure head (ΔP_{fg})

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 is about 26 kPa (~3.8 psi).

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%.

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

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

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 maybe delivered to the required destination in liquid form and (as far as possible) without recompression facilities en route. The critical pressure for CO_2 is about 1070 psig. The typically reported value of final pressure to which the product CO_2 stream has to be pressurized using compressors, before it is transported is about 2000 psig.

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 takes lot of energy, and is a principle contributor to the overall energy penalty of a CO_2 capture unit in a power plant.

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 maybe 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.

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.

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

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.

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 basically a function of the total heat requirement for sorbent regeneration in the amine system.

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).

Secondary steam turbine power generation efficiency (η$_{ST2}$)
This is the efficiency of the secondary steam turbine added along 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.

3.4. Performance Equations

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

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

$(Q/L) = \exp(-2.4452 - 0.0037*y_{CO2} - 6.2743*\phi_{lean} + 0.0254*C)$

[adj. $R^2 = 0.96$]

$(T_{fg,out}) = 41.15 + 0.062*T_{fg,in} + 1.307*y_{CO2} - 18.872*\phi_{lean} + 0.270*C)$

[adj. $R^2 = 0.92$]

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

[adj. $R^2 = 0.95$]

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

[adj. $R^2 > 0.99$]

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 (MJ/ 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)

3.5. Model Outputs

The model has been built in Analytica, which specializes in propagation of uncertainties. The key outputs of the amine system performance model include:

- **MEA 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 solvent. Depending on the level of impurities in the flue gas, there is some loss of solvent. 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 solvent 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, solvent 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

- 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,

$\dot{n}_{CO_2}$ = Total moles of CO2 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

- Net loss of MEA = MEA makeup requirement = $\dot{m}_{MEA,makeup}$

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

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

Estimation of total sorbent circulation flow rate: From the performance equations, we find

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

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

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

- Waste generated from reclaimer:

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

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

$$M_{waste,total} = m_{waste}/f_{w,waste} \text{ kg/hr}$$

Typically, the reclaimer waste contains about 40% water.

- Activated carbon consumption

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

- Caustic Consumption in Reclaimer

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

- 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) (Smelster et al., 1991). Therefore, the process water requirement is:

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

- 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) = ΔT_w

Drop in flue gas temperature = ΔT_{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} = $m_{fg}*(\Delta T_{fg} / \Delta T_w)*(SH_{fg} / SH_w)$ tonne/hr

Therefore, the total water requirement is:

(M_w) = Process water (M_{pw}) + Cooling water (M_{cw})

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,

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

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.

Total energy penalty of CO_2 capture system is:

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

where,

$$E_{regen} = \text{as explained in (9)}$$

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

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

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

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

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

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

3.6. Characterization of Uncertainties

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. Uncertainties in these parameters reflect the ranges of values reported in the literature, the evolving nature of the technology, and practical considerations in running such plants. Table 4 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.

Table 4. 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	-
Lean solvent CO_2 loading	mol CO_2/mol MEA	0.15-0.30	0.22	Triangular(0.17,0.22,0.25)
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	-

MEA loss (exhaust gas)	ppm	1-4	2	Uniform (1,4)
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	-
Cooling water makeup	m^3/tonne CO$_2$	0.5-1.8	0.8	Triangular (0.5,0.8,1.8)
Solvent pumping head	kPa	35-250	207	Triangular(150,207,250)
Pump efficiency	%	70-75	75	Uniform (70,75)
Gas-phase pressure drop	kPa	14-30	26	Triangular(14,26,30)
Fan efficiency	%	70-75	75	Uniform (70,75)
Equiv. elec. requirement	% regeneration heat	9-19	14[a]	Uniform (9,19)
CO$_2$ product purity	wt%	99-99.8	99.5	Uniform (99,99.8)
CO$_2$ product pressure	MPa	5.86-15.16	13.79	Triangular(5.86,13.79,15.16)
Compressor efficiency	%	75-85	80	Uniform (75,85)

[a] For retrofit applications, nominal value is 25.

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 to ensure consistency in economic calculations. There are four types of cost calculated by this model based on the available data (Smelster et al., 1991; Hendriks, 1994; Leci, 1996; Chapel et al., 1999; Simbeck, 1999; Desideri and Paolucci, 1999; Jeremy and Herzog, 2000).

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).

The capital cost model is based on the cost and flow rate information obtained from Fluor Daniel Inc (Fluor Daniel, 1998). It is assumed that there are multiple trains installed to perform the CO$_2$ capture operation. Based on the same source, the maximum train size has been assumed to be 5000 tonnes per day of CO$_2$. Based on the actual CO2 capture rate ($\dot{m}_{CO_2}$) the minimum number of trains required to be installed (N$_{min}$) is determined. Different equipments have different maximum capacity limits. So, (E$_{n,i}$) defines the number of equipments required per train.

E$_{n,i}$: Each train consists of the following pieces of equipment:

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

Pumps - 2 per each train

Reboilers - 4 per each train

Special cases:

1. Each train need not have a separate installation of the solvent processing area, CO_2 transport facility and CO_2 disposal facility, and they will be installed for the whole CO_2 capture unit. Hence "E_n" in this case, may be considered as $(1/N_t)$ per train, where N_t is the total number of trains installed.

2. In case of CO_2 compressors, which have higher capacity ($\sim$ 7200 tpd CO_2), the number of compressors required is calculated accordingly. If N_c is the total number of CO_2 compressors installed, then the number of compressors installed per train may be stated as (N_c/N_t).

Different components of this system (Absorber, Regenerator, Flue gas blower etc.) are scaled, based on the flow ate of the material being handled by that particular device, using 0.6 power law e.g., the cost of absorber and flue gas blower is scaled on the basis of flue gas flow ate entering the CO_2 system. The data obtained from the Fluor Daniel report serve as reference numbers for this scaling exercise.

Actual value of scaling parameter per train (X) is calculated by dividing the magnitude of the scaling parameter (obtained from the performance model) by the minimum number of equipments required (i.e. product of minimum number of trains required and minimum number of equipments per train). e.g. if V is the value of a parameter, then X is given as

$$X = \frac{V}{E_{n,i} \bullet N_{min}}$$

So, different process areas using the same scaling parameter may have different value of X, depending upon the value of E_n.

Each process area has a reference cost (C_{ref}) based on the source sited before, and the corresponding value of the scaling parameter (X_{ref}). The cost of the equipment is calculated using the reference values and the actual value of scaling parameter (X), based on the $6/10^{th}$ rule which is commonly used in chemical engineering costing. For example, in case of a particular process area (say, area 10), we have the following cost:

$C_{10, ref}$ = Cost of equipment (area 10)

Scaling parameter = $X_{10,ref}$

From the performance model, we have:the total quantity of the scaling parameter, Y. Now, as discussed above,

N_{min} = Minimum number of trains

$E_{n,10}$ = Number of equipment (10) per train

Minimum number of equipment installed, $Z_{10,min} = N_{min} \times E_{n,10}$

Total number of equipment installed, $Z_{10} = N_t \times E_{n,10}$

where,

N_t is the actual number of trains installed (including spares)

So, the actual flue gas flow rate per train,

$$X_{10} = Y / Z_{10,min}$$

Therefore, the actual capital cost of absorber in this case may be estimated as

$$C_{10} = C_{10,ref} \cdot \left(\frac{X_{10}}{X_{10,ref}} \right)^{0.6}$$

Once the cost of a particular equipment is calculated (C_{10}), it needs to be multiplied by the total number of equipments installed (Z_{10}) in order to get the total cost of installation for that process area (10).

Similarly, in case of other process areas some physical quantity can be identified (e.g., flue gas flow rate, solvent flow rate, CO_2 product flow rate, CO_2 compression energy requirement, steam flow rate, makeup MEA flow rate etc.) that may be used for scaling of the capital cost.

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

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 essentially a large vessel where the incoming hot flue gas is made to contact with the cooling water. 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. The actual cost of the unit is estimated on the basis of the cost information available for a particular reference case study using 0.6 power law for scaling purposes.

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

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 maybe estimated using as above

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

Absorber: This is the vessel where the flue gas is made to contact with the MEA-based solvent, and some of the CO_2 from the flue gas gets dissolved in the solvent. 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 enters this vessel. The actual cost of the unit is estimated on the basis of the cost information available for a particular reference case study using 0.6 power law for scaling purposes.

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

Rich/lean cross heat exchanger: The rich (CO_2-loaded) and lean (regenerated) solvent streams are passed through this cross heat exchanger, where the rich solvent gets heated and the lean solvent gets cooled. So, the size (and cost) of this unit is mainly a function of the volumetric solvent flow rate in the absorber. It is assumed that this volumetric flow rate is fairly constant in the range of temperature and pressure conditions found in this system. The actual cost of the unit is estimated on the basis of the cost information available for a particular reference case study using 0.6 power law for scaling purposes.

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

Regenerator: This is the column where the CO_2-loaded solvent is regenerated with the application of heat. Solvent 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 solvent concentration, desired CO_2 capture efficiency, etc.). So, the cost maybe estimated using 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 basically a heat exchanger where low-pressure steam extracted from the power plant is used to heat the loaded solvent. So, the size (and cost) of this unit is a function of mainly the flow rate of the solvent as well as the flow rate of steam. The actual cost of the unit is estimated on the basis of the cost information available for a particular reference case study using 0.6 power law for scaling purposes.

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

It maybe 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 almost identical.

Steam extractor: Steam extractors are installed to take out LP/IPsteam 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 maybe used, and the next two cost items (8 and 9) are included in its place.

Auxiliary boiler: The cost of the NG boiler is estimated on the basis of the (no reheat) steam flow rate generated from the boiler. The following cost estimation formula was reported by Dale Simbeck

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

Since the steam flow rate (m_{steam}) was estimated as tonnes/hr, the following expression maybe obtained after accounting for the unit conversions

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

Secondary steam turbine: The cost of the secondary steam turbine is estimated on the basis of the electrical power generated from this new turbine. The following cost estimation formula was reported by Dale Simbeck

$$C_{ST2} = \$300*(\text{power generation expressed in kWe})$$

Since the power generation (E_{ST2}) was estimated as MWe, the following expression maybe obtained after accounting for the unit conversions

$$C_{ST2} = \$300000*(E_{ST2})$$

MEA reclaimer: In order to avoid accumulation of the heat stable salts in the solvent stream and to recover some of the lost MEA solvent, a part of the solvent stream is periodically distilled in this vessel. Addition of caustic helps in freeing of some of the MEA. The amount of MEA makeup required, maybe taken as an indicative of the amount of heat stable salts formed and the quantity of solvent 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, based on a reference study.

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

Solvent processing area: The solvent processing area primarily consists of solvent 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 solvent stream. So, the size (and cost) of this unit (together) will be a function of the total solvent flow rate, and maybe 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 a final CO_2 product at the specified pressure (about 2000 psig) that contains moisture and other impurities (e.g. N_2) at acceptable levels. Obviously, the size (and cost) of this unit will be a function of the CO_2 product flow rate, and maybe 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 cost (PFC). The various indirect costs are then estimated as fractions of the total direct cost (PFC) following the EPRI cost estimating guidelines (TAG, 1993).

Table 5 lists the elements of total capital cost. Because of data limitations some of the indirect cost factors are estimated based on other technologies.

Table 5. 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 Cost (PFC)	ΣA_i
C	Engineering and Home Office	10% PFC
D	General Facilities	10% PFC
E	Project Contingency	15% PFC
F	Process Contingency	2% PFC
G	Total Plant Cost (TPC) = sum of above	B+C+D+E+F
H	Interest Costs During Constr.	Calculated
I	Royalty Fees	0.5% PFC
J	Pre-production (Fixed O&M)	1 month
K	Pre-production (Variable O&M Cost)	1 month
L	Inventory (startup) Cost	0.5% TPC
M	Total Capital Requirement (TCR)[a]	G+H+I+J+K+L

4.2 O&M Cost

The major operating and maintenance (O&M) cost consists of some fixed costs and some variable cost elements as listed in Table 6.

Table 6. 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	$1250/ mton
Water Cost	$0.8/ 1000 gallon
CO_2 Transport Cost	$0.04/ mton CO_2 km

CO$_2$ Storage/Disposal Cost	\$5/ mton CO$_2$
Solid Waste Disposal Cost	\$175/ mton waste

The *fixed O&M* (FOM) costs include the costs of maintenance (materials and labor) and labor (operating labor, administrative and support labor). These are estimated on the annual basis (\$M/yr). The mathematical model for the fixed cost is as follows

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

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

$$FOM_{maint} = \Sigma_i (f_{maint})_i \times TPC_i \quad \text{where i = process area}$$

$$FOM_{admin} = f_{admin} \times (FOM_{labor} + f_{maintlab} \times FOM_{maint})$$

The *variable O&M* (VOM) costs include:

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 user controlled cost input variable.

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

where, UC$_{MEA}$ is the unit cost of MEA.

Cost of inhibitor (VOM$_{inhibitor}$): Addition of inhibitor makes it possible to use higher concentrations of MEA solvent 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 proper disposal of the spent sorbent i.e. the reclaimer waste, again the quantity estimated in the performance model.

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

where, UC$_{waste}$ is the unit cost of waste disposal for the reclaimer waste.

Cost of CO$_2$ 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 mton CO_2), and CO_2 product flow rate (calculated result from performance model).

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

Cost of CO_2 disposal ($VOM_{disposal}$): Depending upon the method of CO2 disposal or storage, either there may be some revenue generated (Enhanced Oil Recovery, Coal Bed Methane) which may be treated as a "negative cost", or 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, the energy costs are handled internally in the model by de-rating the overall power plant based on the calculated power requirement. This increases the cost per net kilowatt-hour delivered by the plant. The CO_2 capture unit is charged for the total electricity production foregone (energy penalty) because of capture and compression of CO_2 from the flue gas, and the base plant is credited for the same. The unit cost of electricity (COE_{noctl}) is estimated by the base plant module, or maybe overridden by a user-specified value when this energy is supplied from an external source (in that case, no credit given to the base plant). Since energy cost is one of the biggest O&M cost items for CO_2 unit, the way in which it is accounted for (internal de-rating or external provision) becomes very crucial while calculating the mitigation cost.

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

Alternatively, when regeneration steam is provided by an auxiliary NG boiler, the cost of energy maybe estimated from the total annualized cost of the new boiler and secondary steam turbine, which takes into account their capital cost requirement and cost of natural gas fuel.

Cost of water (VOM_{water}): Water is mainly required for 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. Also, it maybe noted that the unit cost of water (UC_{water}) may vary depending upon the location of the power plant.

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

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

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

Finally, the total annual O&M cost (TOM, \$/yr) maybe obtained as

$$TOM = FOM + VOM$$

4.3 Incremental Cost of Electricity

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

$$\text{Total annualized cost, TRR (\$/yr)} = TCR \times CRF + TOM$$

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

CRF = Capital recovery factor (%)

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). Two key parameters are the capital recovery factor (to amortize capital expenses), and the plant capacity factor (which determines the effective annual hours of operation at full load).

Cost of electricity, COE ($/MWh) $= \text{TRR} / (\text{MW}_{net}*\text{HPY})$

Where,

TRR = Total annualized cost ($/yr)

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

HPY = Annual hours of operation (hrs/yr)

So, by running two scenarios of the power plant model, one without CO_2 capture unit (reference plant) and one with CO_2 capture unit (CO_2 capture plant), we obtain the respective capital costs, O&M costs to give the annualized costs (TRR) and finally the cost of electricity (COE) with and without 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 Avoided

Many analysts like to express the cost of an environmental control system in terms of the cost per ton of pollutant removed or avoided. For energy-intensive CO_2 controls there is a big difference between the cost per ton CO_2 removed and the cost per ton "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:

$$\text{Cost of } CO_2 \text{ Avoided (\$/mton)} \ = \ \frac{(\$/kWh)_{after} \ - \ (\$/kWh)_{before}}{(t\ CO_2/kWh)_{before} \ - \ (t\ CO_2/kWh)_{after}}$$

For power plants with multi-pollutant controls the desire to quantify costs for a single pollutant sometime 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 cost of CO_2 avoidance has another interpretation in terms of the carbon-tax scenarios. Consider a scenario where every power plant is made to pay a fixed amount of tax (C-tax) that is proportional to their CO_2 emissions. Now let's have a reference plant (one that does not control its CO_2 emissions) and a CO_2 capture plant (one that captures, say 90% of its CO_2 emissions). The reference plant will pay a much higher C-tax (almost 10 times that paid by the capture plant). So, the COE for the reference plant 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 maybe reached where COE for both the plants are same (see Figure 6).

It means that at this C-tax level, the power plant might be indifferent between paying C-tax for its entire CO_2 emissions or incurring the cost of the CO2 capture unit. Above this particular C-tax level, the COE for the reference plant will be higher than that for the capture plant, as it is evident

from the figure. So, cost of CO_2 avoidance is this C-tax level, where the COE for the reference plant and capture plant become equal.
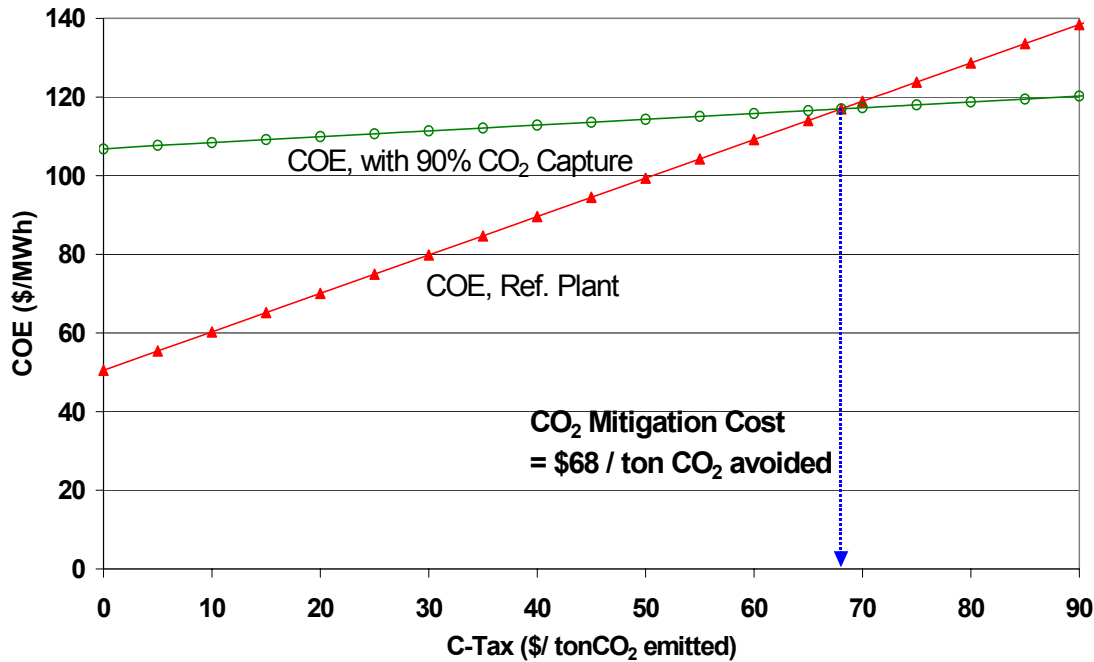

Figure 6. Cost of electricity (COE) as a function of carbon-tax

5. UNCERTAINTY DISTRIBUTION BASED ON DATA FOR COMMERCIAL SYSTEMS (WORK IN PROGRESS)

6. REFERENCES

1. Marion, J., Nsakala, N., Bozzuto, C., Liljedahl, G., Palkes, M., Vogel, D., Gupta, J.C., Guha, M., Johnson, H., and S. Plasynski (2001). "Engineering feasibility of CO2 capture on an existing US cola-fired power plant", in *the Proceedings of the 26th International Conference on Coal Utilization and Fuel Systems, March 5-8, 2001, Clearwater, FL, USA*, 941-952.

2. Herzog, H.J. et al (2001). "A cost model for transport of carbon dioxide", *a draft paper.*

3. IEA GHGRDP (2000). "International test network for CO2 capture: report on a workshop", Report No. PH3/33, December 2000.

4. IEA GHGRDP (2000). "Leading options for the capture of CO2 emissions at power stations", Report No. PH3/14, *prepared by Stork Engg Consultancy B.V., Amsterdam*, February 2000.

5. TransAlta Corp., Calgary, Alberta (1999). "A design study of the application of CO2/O2 combustion technology as retrofit to an existing coal fired boiler", *a report prepared by ABB Power Plant Laboratories*, July 1999.

6. Simbeck, D (1999). "A portfolio selection approach for power plant CO_2 capture, separation and R&D options", in *Greenhouse Gas Control Technologies (ed. by Eliasson B., Riemer P. and A. Wokaun), Proceedings of the 4th International Conference on Greenhouse Gas Control Technologies, 30 August – 2 September 1998, Interlaken, Switzerland,* Elsevier Science Ltd., 119-124.

7. Bolland, O. and H. Undrum (1999). "Removal of CO2 from gas turbine power plants: Evaluation of pre- and postcombustion methods", in *Greenhouse Gas Control Technologies (ed. by Eliasson B., Riemer P. and A. Wokaun), Proceedings of the 4th International Conference on Greenhouse Gas*

Control Technologies, 30 August – 2 September 1998, Interlaken, Switzerland, Elsevier Science Ltd., 125-130.

8. Chapel, D., Ernst, J. and C. Mariz (1999). "Recovery of CO2 from flue gases: commercial trends (paper No. 340)", *presented at the Canadian Society of Chemical Engineers annual meeting, Saskatoon, Saskatchewan, Canada*, October 4-6, 1999.

9. Appl, M. (1999). *Ammonia: Principles and Industrial Practice*, Chapter 4, Wiley-VCH, Weinheim, FRG.

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

11. Tontiwachwuthikul, P., Chan, C.W., Kritpiphat, W., Demontigny, D., Skoropad, D., Gelowitz, D., Aroonwilas, A., Mouritas, F., Wilson, M., and L. Ward (1998). "Large scale carbon dioxide production from coal-fired power stations for enhanced oil recovery: A new economic feasibility study", *The J. of Can. Petro. Tech.*, **37**(11), 48-55.

12. Desideri U. and R. Corbelli (1998). "CO_2 capture in small size cogeneration plants: Technical and economical considerations", *Energy Conversion and Management*, **39**(9), 857-867.

13. Bolland, O. and P. Mathieu (1998). "Comparison of two CO_2 removal options in combined cycle power plants", *Energy Conversion and Management*, **39**(16-18), 1653-1663.

14. Kohl, A.L. and R.B. Nielsen (1997). *Gas Purification*, 40-277, 5th ed., Gulf Publishing Company, Houston, TX.

15. Mimura, T., Simayoshi, H., Suda, T., Iijima, M., and Mituoka, S. (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.

16. Leci, C.L. (1996). "Development requirements for absorption processes for effective CO_2 capture from power plants", *Energy Conversion and Management*, **38**(Suppl.), S45-S50.

17. Tontiwachwuthikul, P.T., Kritpiphat, W., and D. Gelowitz (1996). "Carbon dioxide production from co-generation for enhanced oil recovery: An economic evaluation", *The J. of Can. Petro. Tech.*, **35**(6), 27-33.

18. Leci, C.L. (1996). "Financial implications on power generation costs resulting from the parasitic effect of CO_2 capture using liquid scrubbing technology form power station flue gases", *Energy Conversion and Management*, **37**(6-8), 915-921.

19. Rao, A.D., and W.H. Day (1996). "Mitigation of greenhouse gases from gas turbine power plants", *Energy Conversion and Management*, **37**(6-8), 909-914.

20. A. Chakma (1995). "An energy efficient mixed solvent for the separation of CO_2", *Energy Conversion and Management*, **36**(6-9), 427-430.

21. Jou, F., Mather, A.E., and F.D. Otto (1995). "The solubility of CO_2 in a 30 mass percent monoethanolamine solution", *The Canadian Journal of Chemical Engineering*, **73** (Feb), 140-147.

22. Hendriks, C. (1994). *Carbon Dioxide Removal from Coal-fired Power Plants*, 14-223, Kluwer Academic Publishers, The Netherlands.

23. Riemer, P., Audus, H., and A. Smith (1994). "Carbon Dioxide Capture from Power Stations", a report prepared for IEA Greenhouse Gas R&D Programme, UK.

24. Dupart, M.S., Bacon, T.R., and D.J. Edwards (1993). "Understanding corrosion in alkanolamone gas treating plants", *Hydrocarbon Processing*, **May** 1993, 89-94.

25. Langeland, K. and K. Wilhelmsen (1993). "A study of the costs and energy reuirement for carbon dioxide disposal", *Energy Conversion and Management*, **34** (9-11), 807-814.

26. Sander, M.T., and C.L. Mariz (1992). "The Fluor Daniel 'Econamine FG" process: past experience and present day focus", *Energy Conversion and Management*, **33**(5-8), 341-348.

27. Barchas, R. and R. Davis (1992). "The Kerr-McGee/ ABB Lummus Crest technology for the recovery of CO_2 from stack gases", *Energy Conversion and Management*, **33**(5-8), 333-340.

28. Suda, T., Fujii, M., Yoshida, K., Iijima, M., Seto, T. and S. Mitsuoka (1992). "Development of flue gas carbon dioxide recovery technology", *Energy Conversion and Management*, **33**(5-8), 317-324.

29. Yagi, T., Shibuya, H. and T. Sasaki (1992). "Application of chemical absorption process to CO_2 recovery from flue gas generated in power plants", *Energy Conversion and Management*, **33**(5-8), 349-355.

30. Smelster, S.C., Stock, R.M. and G.J. McCleary (1991). "Engineering and economic evaluation of CO_2 removal from fossil-fuel-fired power plants", EPRI IE-7365, Vol 2, Project 2999-10, *a research project final report prepared by Fluor Daniel Inc., for EPRI and IEA*, June 1991.

31. Pauley, C.R., Simiskey, P.L. and S. Haigh (1984). "N-Ren recovers CO_2 from flue gas economically", *Oil & Gas Journal*, **82**(20), 87-92

32. Johnson, J.E. (1984). "SELEXOL Solvent process reduces lean, high-CO2 natural gas treating costs", *Energy Progress*, **4**(4), 241-248.

33. Wiggins, W.R. and R.L. Bixler (1983). "Sources, recovery and tranportation of CO_2", *Energy Progress*, **3**(3), 132-135.

34. Anada, H.R., Fraser, M.D., King, D.F., Seskus, A.P., and J.T. Sears (1983). "Economics of by-product CO_2 recovery and transportation for EOR", *Energy Progress*, **3**(4), 233-243.

35. St. Clair, J.H. and W.F. Simister (1983). "Process to recover CO2 from flue gas gets first large-scale tryout in Texas", *Oil & Gas J.*, **Feb** 1983, 109-113.

36. Stokes, K.J. (1981). "Choosing an ammonia plant CO_2 removal system for today's conditions", *Nitrogen*, **131**, 35-38.

37. Vaz, R.N., Mains, G.J., and R.N. Maddox (1981). "Process to recover CO2 from flue gas gets first large-scale tryout in Texas", *Hydrocarbon Processing*, **Feb** 1981, 139-142.

38. Nakayama, S., Noguchi, Y., Kiga, T., Miyamae, S., Maeda, U., Kawai, M., Tanaka, T., Koyata, K. and H. Makino (1992). "Pulverized coal combustion in O2/CO2 mixtures on a power plant for CO_2 recovery", *Energy Conversion and Management*, **33**(5-8), 379-386.

39. Price, B.C. (1984). "Processing high CO_2 gas", *Energy Progress*, **4**(3), 169-174.

40. Horn, F.L. and M. Steinburg (1982). "An improved carbon dioxide power plant", *Energy Progress*, **2**(3), 154-159.