Techno-Economic Study of CO_2 Capture Process for Cement Plants

by

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Abstract

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Carbon dioxide is considered to be the major source of GHG responsible for global warming; man-made CO_2 contributes approximately 63.5% to all greenhouse gases. The cement industry is responsible for approximately 5% of global anthropogenic carbon dioxide emissions emitting nearly 900 kg of CO_2 for every 1000 kg of cement produced! Amine absorption processes in particular the monoethanolamine (MEA) based process, is considered to be a viable technology for capturing CO_2 from low-pressure flue gas streams because of its fast reaction rate with CO_2 and low cost of raw materials compared to other amines. However, MEA absorption process is associated with high capital and operating costs because a significant amount of energy is required for solvent regeneration and severe operating problems such as corrosion, solvent loss and solvent degradation.

This research was motivated by the need to design size and cost analysis of CO_2 capture process from cement industry. MEA based absorption process was used as a potential technique to model CO_2 capture from cement plants. In this research four cases were considered all to reach a CO_2 purity of 98% i) the plant operates at the highest capacity ii) the plant operates at average load iii) the plant operates at minimum operating capacity and iv) switching to a lower carbon content fuel at average plant load. A comparison among four cases were performed to determine the best operating conditions for capturing CO_2 from cement plants. A sensitivity analysis of the economics to the lean loading and percent recovery were carried out as well as the different absorber and striper tray combinations.

Keywords: CO_2 capture; MEA; Aspen Plus, Icarus; Cement Plant; Fuel switching; Cost

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Dedicated to my mother

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Nomenclature

ABB	Asea Brown Boveri Ltd
AMP	2-amino-2-methyl-1-propanol
ASME	American Society of Mechanical Engineers
bar	unit of pressure
B:F	bottoms to feed ratio
^{o}C	degree Celsius
CFC’s	Chlorofluorocarbons
CKD	cement kiln dust
DEA	diethanolamine
DGA	diglycolamine
DIPA	diisopropanolamine
DS	design specification
ECBM	enhanced coal bed methane
EOR	enhanced oil recovery
FGD	flue gas desulphurization
GHGs	greenhouse gas
GWP	global warming potential
HETP	Height equivalent to a theoretical plate
HFC’s	hydrofluorocarbons
IEA	International Energy Agency
IGCC	Integrated gasification combined cycle
IPCC	Intergovernmental Panel on Climate Change
IPE	Icarus process evaluator
KEPCO	Kansai Electric Power Company Inc.
kg	kilogram
kPa	kilopascal
KP	Kansai packing
KS	Kansai solvent

LOI	loss of ignition
LP	low pressure
LTF	low temperature flash
MDEA	methyldiethanolamine
MEA	monoethanolamine
MHI	Mitsubishi Heavy Industries Ltd.
MW	megawatt
N/A	not available/not applicable
NGCC	natural gas combined cycle
NRC	National Research Council
OPG	Ontario Power Generation
PCC	pulverized coal combustion
ppb	Parts per billion
ppm	Parts per million
PRB	Power River Basin
PSA	pressure swing adsorption
RR	reflux ratio
SOFC	solid oxide fuel cell
TEA	triethanolamine
TSA	temperature swing adsorption
USD	US Dollar
VLE	vapor-liquid equilibrium

Chemical Symbols and Formulae

Ar	argon
C	carbon
$CaCO_3$	calcium carbonate
CaO	calcium oxide
CH_4	methane
C_2H_6	ethane
CO	carbon monoxide
CO_2	carbon dioxide
H	hydrogen
H_2	molecular hydrogen
H_2O	water
HCl	hydrogen chloride
HF	hydrogen fluoride
N	nitrogen
N_2	molecular nitrogen
N_2O	nitrous oxide
NaOH	sodium hydroxide
NO	nitrogen oxide
NO_2	nitrogen dioxide
NO_x	nitrogen oxides
O_2	molecular oxygen
O_3	ozone
S	sulphur
SF_6	sulphur hexafluoride
SO_2	sulphur dioxide
SO_x	sulphur oxides

Chapter 1: Introduction

Background

1.1 Greenhouse Gases and Climate Change Context

Climate change is one of the greatest and probably most challenging of environmental, social and economical threats facing the world this century. The issue of climate change is not about air quality or smog but it is about global warming. Human activities have altered the chemical composition of the atmosphere through the buildup of significant quantities of greenhouse gases (GHGs), which remain in the atmosphere for long periods of time and intensify the natural greenhouse effect. Increasing concentrations of greenhouse gases are likely to accelerate the rate of climate change. Concerns are growing about how increases in atmospheric greenhouse gases (GHGs) caused by human activities are contributing to the natural greenhouse effect and raising the earth's average temperature.

The earth’s average surface temperature has already increased by about 0.6°C during the 20th century. Evidence is getting stronger that most of the global warming that has occurred over the last 50 years is attributable to human activities [1]. The Intergovernmental Panel on Climate Change (IPCC) in its Third Assessment Report (published in 2001) projects that the global average surface temperatures will rise by about 1.4 to 5.8°C by the year 2100. This change would be larger than any climate change experienced over the last 10 000 years [2]. This global temperature increase is likely to trigger serious consequences for humanity and other life forms with significant regional variation. Evaporation will increase as the climate warms, which eventually will increase average the global precipitation. Soil moisture is likely to decline in many regions, and intense rainstorms are likely to become more frequent.

1.2 The Greenhouse Effect Process

The greenhouse effect is a naturally occurring process that aids in heating the earth's surface and atmosphere. It causes the atmosphere to trap more heat energy at the earth's surface and within the atmosphere by absorbing and re-emitting longwave energy as shown in Figure 1-1. It results from the fact that certain atmospheric gases, such as carbon dioxide, water vapor, nitrous oxide, methane, halocarbons and ozone are able to change the energy balance of the planet by being able to absorb long wave radiations from the Earth's surface. As energy from the sun passes through the atmosphere a number of phenomena take place. A portion of the energy (26% globally) is reflected back to space by clouds and particles. Clouds, gases (like ozone), and particles in the atmosphere absorb about 19% of the available energy. Of the remaining 55% of the solar energy passing through the Earth's atmosphere, 4% is reflected from the surface back to space. On average about 51% of the sun's radiation reaches the surface [3]. This energy is then used in number of processes including: the heating of the ground surface, the melting of ice and snow and the evaporation of water and plant photosynthesis. The heating of the ground by sunlight causes the Earth's surface to become a radiator of energy in the longwave band called infrared radiation. This emission of energy is generally directed to space. However, only a small portion of this energy actually makes it back to space. The majority of the outgoing infrared radiation is absorbed by a few naturally occurring atmospheric gases known as greenhouse gases. Absorption of this energy causes additional heat energy to be added to the Earth's atmospheric system. The now warmer atmospheric greenhouse gas molecules begin radiating longwave energy in all directions. Over 90% of this emission of longwave energy is directed back to the Earth's surface where it once again is absorbed by the surface. The heating of the ground by the longwave radiation causes the

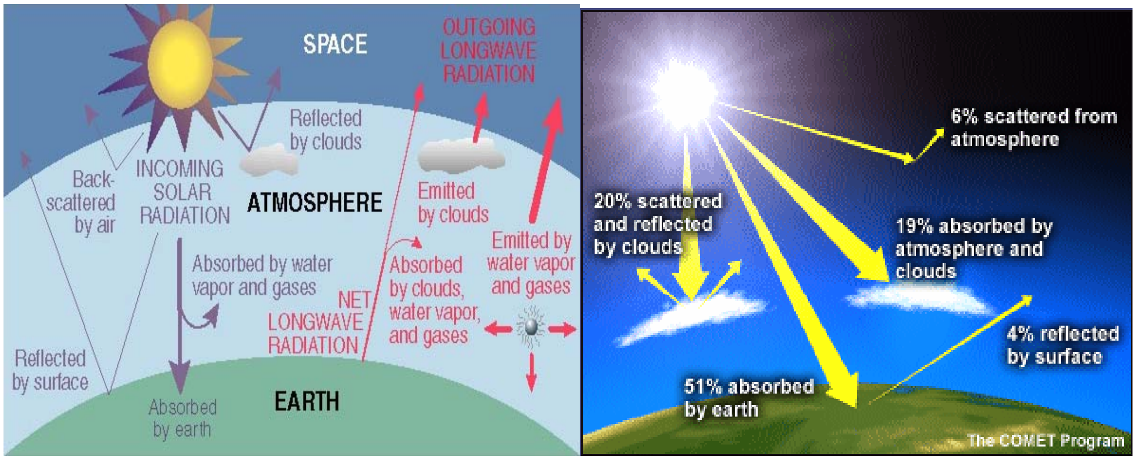

Figure 1-1: The Greenhouse Effect Process [4]

ground surface to once again radiate, repeating the cycle described above, again and again, until no more longwave is available for absorption. The net result of the greenhouse effect is the addition of extra heat energy to the earth's atmosphere and ground surface [3].

The amount of heat energy added to the atmosphere by the greenhouse effect is controlled by the concentration of greenhouse gases in the earth's atmosphere. Since the beginning of the industrial revolution (about 1700 A.D), all of the major greenhouse gases have increased in concentration, and because of these higher concentrations, scientists predict that the greenhouse effect will be enhanced and the Earth's climate will become warmer. Computer models suggest that a doubling of the concentration of the main greenhouse gas, carbon dioxide, may raise the average global temperature between 1 and 3 degrees Celsius [3].

1.3 Gases Involved in The Greenhouse Effect

We know that our atmosphere is a complex mixture of gases that trap the sun's heat near the earth's surface, similar to how the glass of a greenhouse traps the sun's warmth. A number of gases are involved in the greenhouse effect. The main greenhouse gases are:

- Carbon dioxide (CO_2)
- Water vapor (H_2O)
- Methane (CH_4)
- Nitrous oxide (N_2O)
- Tropospheric ozone (O_3) and
- Halocarbons (CFCs, HFCs, etc.)

1.3.1 Carbon dioxide (CO_2)

Carbon dioxide is the most important of the greenhouse gases released by human activities. It is the main contributor to climate change because of the quantities released. On a worldwide basis, CO_2 emissions generated from anthropogenic activities are known to be relatively small and in comparison with the gross carbon fluxes from natural systems, they represent only a fraction (~2%) of total global emissions. However, evidence suggests that they account for most of the observed accumulated CO_2 in the atmosphere [5] [6]. On the basis of global emissions information, the primary sources of CO_2 generated from anthropogenic activities are fossil fuel combustion from industry and industrial processes such as cement production, transportation, space heating, electricity generation and cooking, vegetation changes in natural prairie, woodland and forested ecosystems [7].

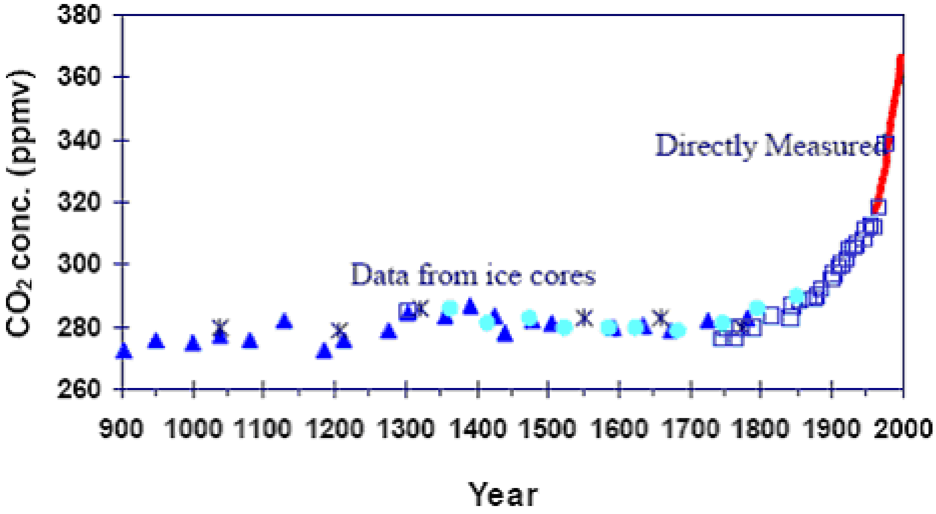

Figure 1-2: Trends in CO_2 Concentrations for past 1000 years [8]

Prior to 1700, in preindustrial times, the average concentration of CO_2 was about 290 parts per million as seen in Figure 1-2. The concentration levels of CO_2 in the atmosphere are now approaching 360 parts per million and continue to increase steadily at a rate of about 0.3-0.4% per year [9]. This increase in CO_2 in the atmosphere is primarily due to the societal changes brought about by the industrial revolution. When fossil fuels are burned, the carbon content is oxidized and released as carbon dioxide; every tonne of carbon burned produces 3.7 tonnes of carbon dioxide. The global consumption of fossil fuels is estimated to release 22 billion tonnes of carbon dioxide into the atmosphere every year – and the amounts are still climbing [9].

1.3.2 Methane (CH_4)

Global CH_4 emissions resulting from anthropogenic activities are considered to have caused an increase of about 145% in atmospheric concentrations since the mid-1700s. The current annual rate of accumulation of CH_4 is estimated to range between 40 and 60 Mt (~14–21 ppbv), or approximately 10% of total worldwide CH_4 emissions [5]. CH_4

emissions generated from human activities, amounting to ~360 Mt per year, are primarily the result of activities such as livestock and rice cultivation, biomass burning, natural gas delivery systems, domestic grazing animals, termites, landfills, and coal mining and oil and gas extraction (EPA, 1981) [10]. Anaerobic conditions associated with rice paddy flooding results in the formation of methane gas. Methane is significant because it has 21 times the heat-trapping effect of carbon dioxide [7].

1.3.3 Nitrous oxide (N_2O)

Nitrous oxide occurs naturally in the environment but human activities are increasing the quantities. The average concentration of nitrous oxide in the atmosphere is now increasing at a rate of 0.2-0.3% per year [3]. At present, it has been estimated that approximately one-third of global atmospheric N_2O is of human origin, resulting primarily from the application of nitrogen fertilizers, land-use conversion, soil cultivation, and the combustion of fossil fuels, biomass and wood. Atmospheric concentrations of N_2O have grown by about 17% since the mid-1700s [11]. Total annual emissions from all sources are estimated to be within the range of 10–17.5 Mt N_2O expressed as nitrogen (N) [12]. The other two-thirds of global atmospheric N_2O come from soil and water denitrification under anaerobic conditions.

1.3.4 Halocarbons (HFCs, PFCs, and SF_6)

Artificially created chlorofluorocarbons HFCs, PFCs, and SF_6 are the strongest greenhouse gases per molecule. However, low concentrations in the atmosphere reduce their overall importance in the enhancement of the greenhouse effect. These gases, while emitted in very small amounts, are having a lasting effect on atmospheric composition and, potentially, the climate, because they are strong absorbers of infrared radiation and have very long atmospheric lifetimes. PFCs have atmospheric lifetimes greater than 2300 years, with perfluoromethane estimated to last 50 000 years. Current measurements in the atmosphere indicate that the concentration of these chemicals may soon begin declining because of reduced emissions [7].

1.3.5 Ozone (O_3)

Ozone's role in the enhancement of the greenhouse effect has been difficult to determine. Accurate measurements of past long-term (more than 25 years in the past) levels of this gas in the atmosphere are currently unavailable. Moreover, concentrations of ozone gas are found in two different regions of the Earth's atmosphere. The majority of the ozone (about 97%) found in the atmosphere is concentrated in the stratosphere at an altitude of 15 to 55 kilometers above the Earth's surface. In recent years, the concentration of the stratospheric ozone has been decreasing because of the build-up of chlorofluorocarbons in the atmosphere. Ozone is also highly concentrated at the Earth's surface. Most of this ozone is created as a by-product of photochemical smog [3]. Table 1-1 [3] summarizes the past and present concentrations of the main greenhouse gases.

Table 1-1: Summarizes past and present concentrations of the main GHGs

Greenhouse Gas	Concentration 1750	Present Concentration	Percent Change	Natural and Anthropogenic Sources
Carbon Dioxide	280 ppm	360 ppm	29%	Organic decay Forest fires Burning fossil fuels Deforestation Land use change
Methane	0.70 ppm	1.70 ppm	143%	Wetlands Organic decay Natural gas and oil extraction Biomass burning Rice cultivation Cattle, Reuse landfills
Nitrous Oxide	280 ppb	310 ppb	11%	Forests Grasslands Oceans, Soils Soil cultivation Fertilizers Burning of fossil fuels, biomass
Chlorofluorocarbons (CFCs)	0	900 ppt	N/A	Refrigerators Aerosol spray Propellants Cleaning solvents

1.4 Greenhouse Gases and Global Warming Potential (GWP)

The Global Warming Potential (GWP) is a means of comparing the relative climate effects of the various greenhouse gases with that of an equivalent emission of CO_2 [13]. The concept of Global Warming Potential (GWP) has been developed to allow scientists and policy-makers to compare the ability of each GHG to trap heat in the atmosphere relative to another gas. By definition, a GWP is the time-integrated change in radiative forcing due to the instantaneous release of 1 kg of the gas expressed relative to the radiative forcing from the release of 1 kg of CO_2 [7]. In other words, a GWP is a relative measure of the warming effect that the emission of a radiative gas might have on the surface troposphere [7]. The GWP of a GHG takes into account both the instantaneous radiative forcing due to an incremental concentration increase and the lifetime of the gas. The indicated GWP values are calculated by integrating the effect of emissions on the climate over the next 100 years [13]. The GWP for the main greenhouse gases is shown in Table 1-2 [7] [13] [14] [15].

Molecule for molecule, CO_2 is the least effective of the major greenhouse gases. Methane, by comparison, absorbs and reradiates about 21 times as much heat energy [13]. However, the overall contribution of each greenhouse gas depends on several other factors as well, including:

- The amount of the gas released into the atmosphere each year
- The atmospheric lifetime of each gas
- The indirect effect that emissions of each gas will have on atmospheric chemistry.

Table 1-2: The Global Warming Potentials (GWP) and Atmospheric Lifetimes

GHG	Formula	Atmospheric Lifetime (Years)	GWP
Carbon dioxide	CO_2	Variable	1
Methane	CH_4	12 ± 3	21
Nitrous Oxide	NO_2	120	310
Sulphur Hexafluride	SF_6	3200	23900
Hydrofluorocarbons	**HFCs**	1 to 200+	140 to 11700

1.5 Canada’s GHGs Emissions Trend

Canadians contributed to about 731 megatonnes of CO_2 equivalent (Mt CO_2 eq.) of GHGs to the atmosphere in 2002, an increase of 2.1% over the 716 Mt recorded in the year 2001. The total emissions of all GHGs in 2002 were 20.1% above the 1990 level of 609 Mt. The average annual growth of emissions over the 1990–2002 periods was 1.7%. Emissions are now slightly above the year 2000 emissions of 725 Mt, an increase of about 1%. Approximately 74% of total GHG emissions in 2002 resulted from the combustion of fossil fuels. Another 8% were from fugitive sources, with the result that over 81% of emissions were from the Energy Sector [7]. A sectoral breakdown of Canada’s total emissions for 2002 is shown in Figure 1-3.

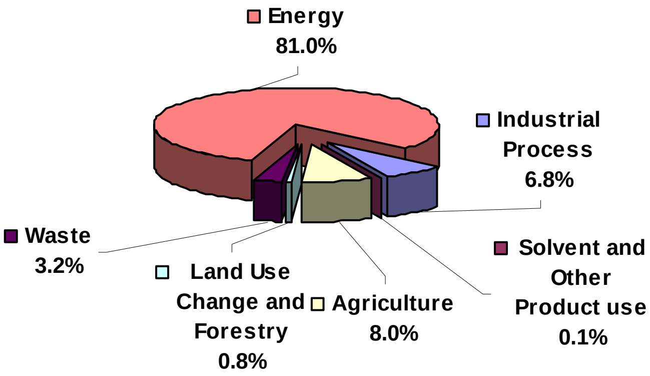

Figure 1-3: Sectoral Breakdown of Canada’s GHG Emissions, 2002

On an individual GHG basis, CO_2 had the largest share of 2002 emissions, at 78.8% (about 576 Mt), while CH_4 accounted for 12.9% (94 Mt). N_2O accounted for 7.2% of the emissions (53 Mt), PFCs contributed 0.7% (5 Mt), and SF_6 and HFCs constituted the remainder, as shown in Figure 1-4 [7].

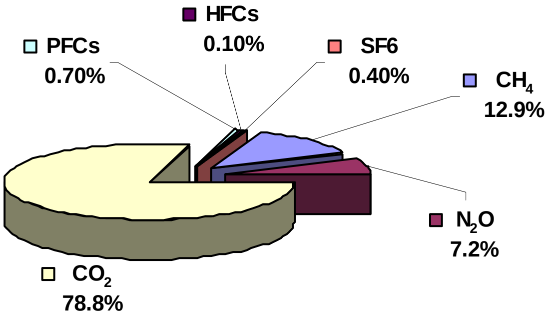

Figure 1-4: Canada's GHG Emissions by Gas, 2002

1.6 Cement Industry and Global Greenhouse Gas Emissions

The cement industry is responsible for approximately 5% of the global anthropogenic carbon dioxide emissions [16][17][18]. The cement industry emits nearly 900 kg of CO_2 per metric ton of cement produced. Because of the significant emissions per unit of cement produced, atmospheric concentrations of greenhouse gases cannot be stabilized without addressing this important emissions source. The cement industry is faced with the prospect of sharp increases in global demand over the coming decades, as well as the prospect of climate change policies that could simultaneously call for reductions in emissions of greenhouse gases. The global cement demand is projected to increase by 60% to 105% over current levels by 2020. By 2050 the cement demand will increase approximately by 225% from current levels [19]. Most of the increase in demand is in developing regions of the world, where the industry's current capital stock is relatively old and inefficient.

Cement-related greenhouse gas emissions come from fossil fuel combustion at cement manufacturing operations (about 40% of the industry's emissions); transport of raw materials (about 5%); and combustion of fossil fuel required to produce the electricity consumed by the cement manufacturing operations (about 5%). The remaining cement-related emissions (about 50%) originate from the process that converts limestone ($CaCO_3$) to calcium oxide (CaO), the primary precursor to cement [19]. It is chemically impossible to convert limestone ($CaCO_3$) to CaO and then cement clinker without generating CO_2, which is currently emitted to the atmosphere. Based on worldwide cement production and associated CO_2 emissions, the average gross unit-based emissions for the industry is approximately 0.87 kg CO_2 per kg of cement [19]. Regionally, however, unit-based emissions vary from 0.73 kg CO_2 per kg of cement (in Japan) to 0.99 kg CO_2 per kg of cement (in the United States).

1.7 GHG Emissions from Canadian Cement Industries

GHG emissions from cement production contributed an estimated 10.2 Mt (or 1.4%) to Canada's national GHG emission total in 2002. Over the 1990–2002 periods, the cement industry experienced a 14% increase in GHG emissions. Process CO_2 emissions from the production of clinker have increased by 21%, while emissions from stationary combustion increased by 3%. Over the same time frame, clinker production increased by 21% [21]. Between 2001 and 2002, GDP for the cement industry showed an increase of 2% [20] and GHG emissions increased by 3.5%. This overall increase is the result of a 4% increase in emissions from stationary combustion and a 3% increase in process emissions [7].

1.8 The Kyoto Protocol and Canada's Kyoto Challenge

The United Nations Kyoto Protocol is an international agreement aimed at addressing the issue of climate change. The protocol was agreed to in 1997. It commits developed countries to collectively reduce GHG emissions to 5.2% below 1990 levels by the period 2008-2012. Canada's share is a 6% reduction below 1990 levels. The Kyoto protocol also introduces three market mechanisms that involve transferring emissions credits to help developed countries meet their targets for reducing GHG emissions:

Canada accounts for about 2% of the world's man-made greenhouse gas emissions [7]. Canada ratified the Kyoto Protocol on December 17, 2002 and under the terms of the Kyoto Protocol, Canada is be obliged to cut greenhouse gas emissions to 6% below 1990 levels by 2010 to 2012, or earn credits through Kyoto's market mechanisms to lower the volume of emissions that it would have to reduce. Canadians face some daunting challenges in meeting Kyoto target.

The 1990–2002 data on Canada's GHG emissions demonstrate progress in reducing emissions in many areas of the economy, but also indicate areas where further efforts are needed. The average annual growth of emissions over the 1990–2002 periods was 1.7%. Total emissions of all GHGs (about 731 megatonnes of equivalent) in 2002 were 20.1% above the 1990 level of 609 Mt. In 2002, Canada's emissions increased by 15 Mt, up from the 2001 level of 716 Mt. The energy sector was responsible for most of this change, with emissions increasing over 10.4 Mt. GHG emissions associated with manufacturing in 2002 were up by 0.1 Mt over 2001, an increase of 2.2%. Between 2001 and 2002, transportation sector emissions increased in almost all modes of transportation [7].

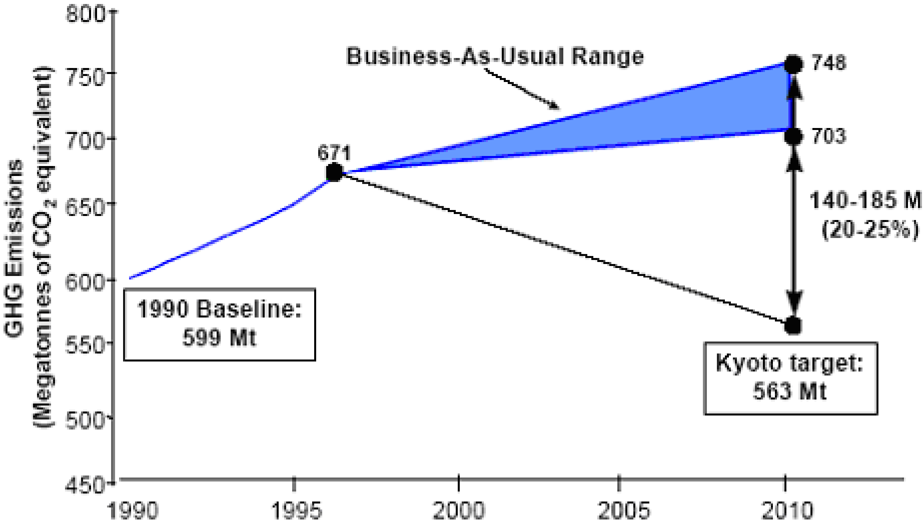

Figure 1-5: Canada's Kyoto Challenge [9]

The Canadian government estimates that total emissions of all GHGs in Canada were 20.1% higher in 2002 than in 1990. On a business-as-usual (BAU) basis, assuming that current trends in population growth, economic development, and energy consumption continue, Canada's GHG emissions will be 40% (about 240 Megatonnes) higher than our Kyoto target by 2010 as shown in Figure 1-5. To meet Canada's Kyoto target, Canadians would have to reduce GHG emissions by almost 4.5% per year between 2002 and 2010. In other words, Canada's emissions reductions would have to average out at about 1 tonne of carbon per Canadian per year beginning in 2002.

1.9 Motivation

The cement industry emits enormous amounts of carbon dioxide into the atmosphere, i.e., nearly 900 kg of CO_2 per tonne of cement produced and the industry is faced with the prospect of sharp increases in global demand over the coming decades, as well as the prospect of climate change policies that could simultaneously call for reductions in emissions of greenhouse gases. This significant emission per unit of cement produced is believed to have adverse effects on the atmosphere. To overcome the carbon dioxide emission problem capturing carbon dioxide is introduced and used in industries. This research study is motivated by a desire to capture large quantities of CO_2 emissions from cement industry which contributes to global warming and climate change; thus helping to accomplish the goals set out by the Kyoto Protocol and meets Canada's Kyoto Challenge within the stipulated period 2010 to 2012.

1.10 Objectives

Carbon dioxide capture from cement plant flue gas using amine-based CO_2 capture technology requires huge amounts of energy mostly in the form of heat. The overall objective of this thesis is to evaluate the feasibility of obtaining the heat required for amine absorption for a particular recovery of carbon dioxide for a given a set of equipment specifications and operating conditions from cement plants and to develop a model that simulates the removal of CO_2 using MEA absorption from cement plant flue gas and design a process that will minimize the cost of CO_2 capture for the cement industry. To design, model, cost and optimize CO_2 capture processes for the cement industry Aspen Plus™ and Icarus Process Evaluator will be used.

1.11 Scope of Work

The research includes design and costing of CO_2 capture from cement plant flue gas. The scope of work includes the followings:

- Focus on MEA based carbon dioxide capture process
- Design the MEA capture process using Aspen Plus™
- Size and evaluate the cost of MEA capture process using Icarus Process Evaluator
- Study the effect of the operating conditions on the cost using Aspen Plus™ and Icarus Process Evaluator

1.12 Expected Results

The expected results are:

- a flow sheet for designing a CO_2 capture process for a cement plant based on the MEA process
- a complete heat and material balance of the CO_2 capture process
- a capital and operating cost estimate for the cement plant
- an estimate for the cost of CO_2 capture

Chapter 2: Theory and Literature Review

2.1 Raw Materials for Cement Production

Dispersed and homogenized raw mixes for the manufacture of cement clinker consists basically of two generically different natural raw materials – calcium carbonate (or limestone) and aluminosilicates (or argillaceous substances). Certain additives such as bauxite, laterite, iron ore or blue dust, sand or sandstone, etc. are used to compensate the specific chemical shortfalls in the raw mix composition. A limestone with a minimum of 44% to 45% CaO and maximum 3.0% to 3.5% MgO, 0.6% R_2O, 0.6% to 0.8% SO_3, and 0.015% to 0.05% Cl is regarded as a cement grade limestone, provided its SiO_2, Al_2O_3, and Fe_2O_3 contents satisfy the ultimate modulii values of raw mixes (ratios used in calculating the chemical composition of the raw mix, i.e., lime ratio, silica ratio, molecular ratio) [22].

The proper dissociation to release reactive lime from limestone in a kiln system is the most important technological requirement in cement production. The color of limestone reflects the level and nature of associated minerals and impurities. High purity calcite limestones are white in color and carbonaceous and ferruginous materials essentially cause other shades in the less pure varieties. The ***porosity*** of the limestones varies from almost 0% to 30% and for predominantly occurring compact limestones, it is in the range of 2% to 5%. The ***water absorption*** in limestones follows the pattern of pores and level of porosity. While it may be negligible in dense varieties, it may go up to 20% in porous chalky varieties. The ***Moh's hardness*** of most limestones lies in the range of 2 to 4. The apparent density of limestones on drying at 110°C ranges from 1500 to 2600 kg/m^3. The bulk density largely depends on the apparent density of limestones, their particle size distribution, and particle shape. In general, the reported range of variation of bulk density of limestones is 1400 to 2700 kg/m^3, although typical values fall in the range of 1400 to 1600 kg/m^3 [22].

In addition to the above basic physical properties, a few characteristics like crushability and grindability as well as the physical behaviour of crushed and screened limestones are important for the process of cement manufacture.

2.2 Cement Making Process

Cement manufacturing process has witnessed several prominent changes during the last few decades. These changes have occurred in almost every operation of cement manufacturing. The major developments have been in the areas of material processing, energy conservation, instrumentation, environment, and safety issues [22]. The process flow diagram of the cement manufacturing process is shown in Figure 2-1 [27]. The basic cement manufacturing steps are discussed briefly to understand the cement manufacturing process.

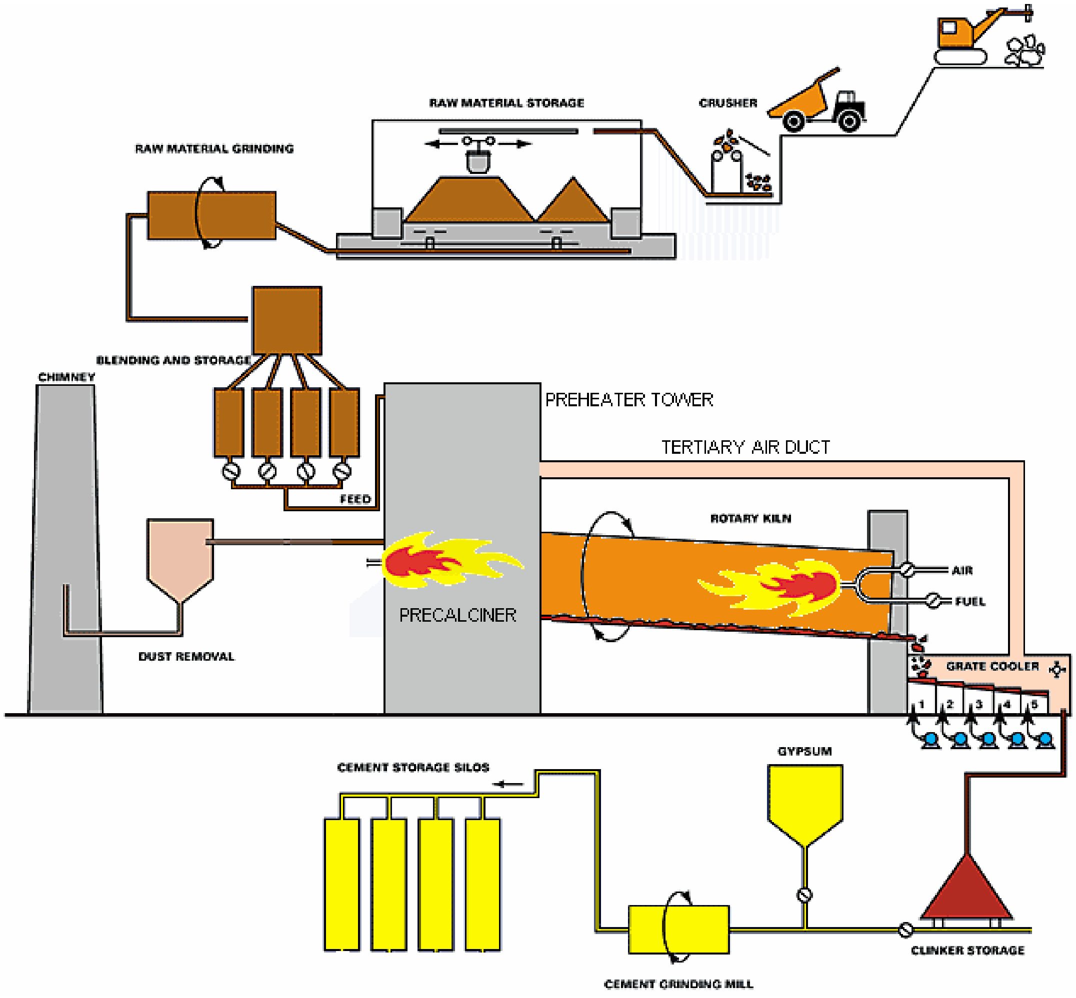

Figure 2-1: Process flow diagram of cement manufacturing process

2.2.1 Materials Preparation and Raw Milling

Limestone and other naturally occurring raw materials are usually received in the plant from the quarry in maximum sizes of 1 to 2 m. In order to convert these raw materials into clinker and cement, it is necessary to produce a raw mix with a top size of about 0.2 mm. The reduction ratio required for most raw materials is consequently 1000-2000:0.2 [22]. Such a degree of reduction cannot be achieved through a single machine or a one-stage operation. Further, following the stage of reduction, good homogeneity has to be achieved for the raw mix, as this is essential for product quality and plant efficiency. In parallel with this, it should also be borne in mind that in order to ensure continuity of the manufacturing process, stocks must be built up at various points.

The process route for raw materials preparation is multi-stage and complex and generally the following steps are:

- Crushing (one-, two- or three-stage)
- Pre-homogenization and storage of crushed materials
- Grinding (diverse milling systems)
- Proportioning, homogenization, and storage of fine material
- Quality control in raw meal preparation

2.2.1.1 Crushing

Crushing is the primary size-reduction process carried out under the action of external mechanical forces in the cement plant. Generally, three levels of crushing are recognized: **coarse** (top product size, up to 150 to 200 mm), **intermediate** (up to 40 to 70 mm), and **fine** (up to 5 to 25 mm).

Crushing is affected by compression, shear, impact, and attrition – singly or in combination. In the cement industry, the raw materials undergo primary reduction in single- or twin-rotor hammer crushers, or impact crushers. Jaw crushers, also in combination with roll crushers or gyratory crushers are used to deal with hard and abrasive materials.

2.2.1.2 Pre-homogenization and Storage of Crushed Materials

In cement production, it is often necessary to pre-homogenize the raw materials after secondary crushing, particularly when their chemical/mineralogical composition varies over a wide range. This unit operation is utilized primarily for the main raw materials, viz., the limestone and the aluminosilicate component.

The pre-homogenizing systems are also known as "blending beds". The blending effect, H, is generally determined as the ratio between the standard deviation S_{in} of one significant chemical parameter of the input to a given stockpile and the standard deviation of the same parameter S_{out} of the output from the same stockpile, as follows:

$$\text{Blending Effect, H} = \frac{S_{in}}{S_{out}}$$

In this context, it is important to note that normally it is not possible in practice to measure homogenizing effects greater that 1:10, due to the statistical inaccuracies inherent in the representative sampling and analyses of lumpy materials. In practice, however, the blending efficiency of the pre-homogenization systems lies within this range [22].

2.2.1.3 Grinding

The crushed and preblended raw materials are further processed into a fine raw mix by dry or wet grinding. Dry grinding of raw materials is practiced most extensively due to escalating energy costs and progressive obsolescence of the wet process of manufacturing.

The majority of grinding in the world is performed with the help of ball mills. The material to be ground passes through the rotating cylinders that are full of steel balls, and grinding takes place by the impact of and attrition from the grinding balls tumbling inside the cylinder. Efficiency and output of the ball mills primarily depend on the optimum utilization of ball charge energy for coarse and fine grinding [22].

2.2.1.4 Homogenization of Raw Meal

The homogenization of raw meal prior to the calcination process has always been a very important step in the production of clinker. It is widely known that there are four methods of raw meal homogenization utilized in the cement industry. The first and the traditional one is the slurry mixing practiced in the wet-process plants; other methods are adopted in dry process plants and include mechanical systems, pneumatic systems, and gravity systems [22].

In slurry homogenization, the mill slurry is transported to a slurry silo, the average composition of which is tracked with the help of chemical analysis of spot samples. When the silo is about two thirds full, the slurry is homogenized with compressed air. The slurry from the silos is sent to a basin and further homogenized with the help of mechanical, pneumatic, or pneumomechanical arrangements before entering the kiln. Experience shows that a slurry basin works as the ideal blender.

The mechanical system consists of multiple storage silos, each of which is provided with regulated withdrawal facilities. Blending is achieved by an orderly withdrawal of material at variable rates from all silos. While this type of mixing consumes lower power, the system requires a great deal of material handling that increases power consumption. In addition, the required of number of silos is obviously more than in other systems. As a result of these shortcomings, this kind of homogenization system is not in wide use in the cement industry [22].

The most common homogenization system used over the past several decades is the pneumatic unit based on the air fluidization method. Air introduced through a permeable medium in the silo bottom causes the raw meal to behave almost like a fluid. This agitational method is known to provide high blending efficiency for dry material. But at the same time this method is the highest consumer of power [22].

The gravity approach to homogenization has been conceived only in the expectation of achieving reduced power consumption.

2.2.1.5 Quality Control in Raw Meal Preparation

All the above unit operations are obviously controlled with onsite and offsite equipment. Hence a control laboratory with appropriate hardware and software becomes an integral part of the raw materials preparation facility [22].

The preparation of a correct and uniform kiln feed is an insurance for the quality of the finished product and entails the monitoring of the quality of the incoming raw materials, composition of the preblending stockpile, variation of the raw mill output in terms of chemical composition and particle size distribution, and finally the efficacy of the homogenization system and the consistency of its output forming the kiln feed. The monitoring and consequential control functions are carried out with the help of sampling arrangements, use of specialized analytical instruments, and application of appropriate computerized software.

2.2.2 Kiln Operation Systems

Once raw materials have been selected and blended, and ground and homogenized into a fine and uniform kiln feed, they must then be subjected to enough heat to allow the clinkering reactions to proceed. This is the pyro-processing stage of cement manufacture, beginning with the kiln feed material extracted from storage and weighed and transported to the kiln, and finishing with the clinker from the cooler going to clinker storage. A schematic diagram pyro-processing stage is shown in Figure 2-2.

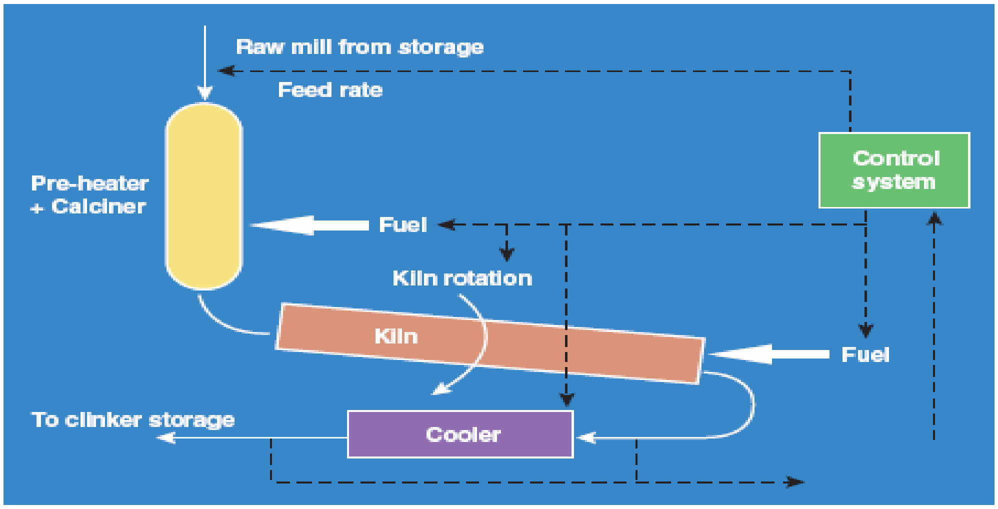

Figure 2-2: Schematic diagram of kiln operation system [22]

The main chemical reactions to produce the calcium silicates that later give cement its bonding strength occur in the kiln system. It's the heart of the cement-making process – a horizontally sloped steel cylinder, lined with firebrick, turning from about 1 to 3 rpm. Kilns are frequently as much as 3.7 m in diameter! The finely ground raw material or the slurry is fed into the kiln at the upper end. At the lower end is a roaring blast of flame, produced by precisely controlled burning of powdered coal, oil or gas under forced draft. From the upper end of the kiln the raw material slides and tumbles down the kiln through progressively hotter zones toward the flame. At the lower end of the kiln, fuels such as powdered coal and natural gas feed a flame that reaches 1870^{o}C. In the hottest part of the kiln, the raw materials reach about 1480^{o}C and become partially molten. This intense heat triggers chemical and physical changes inside the raw materials. Expressed at its simplest, the series of chemical reactions converts the calcium and silicon oxides into calcium silicates, cement's primary constituent. At the lower end of the kiln, the raw materials emerge as a new substance: red-hot particles called clinker. The heated air from the coolers is returned to the kilns, a process that saves fuel and increases burning efficiency. There is a combination of endothermic and exothermic reactions occurring in an extremely complicated chemical reaction sequence. The raw material composition, mineralogical composition and the time and temperature profile of these materials in the kiln determine the ultimate composition and mineralogy of the clinker, which in turn determines the performance of the cement produced [22].

The pyro-processing stage is generally regarded as the heart of the cement-making process. It is the stage in which most of the operating costs of cement manufacture appear, and is also therefore the stage where most of the opportunities for process improvement exist. The pyro-processing stage is also the main source of CO_2 emissions in a cement plant.

2.2.2.1 Wet Process Kilns

The long wet process kiln, with a length to diameter ratio (L/D) of up to 40, was the main clinker producing plant for most of the 20th century. It is a relatively simple process, with the main advantage of slurry preparation being the eases of milling, handling, blending, storage, pumping, and metering. It is also less prone to low level dust emission.

In wet process systems, the material preheating system is comprised of metal chains hanging in the cold end of the kiln, which absorb heat from gases and heat the material that flows over them. The chains actually provide a greater surface area for contact between hot gases and the material clinging to the chains [22].

The main problem with long wet kilns is their poor fuel efficiency, because of the water to be evaporated from the slurry. This became a severe problem only when the cost of fuel escalated during the 1970s, and only a few wet kilns have been built since that time. Another disadvantage of a wet process kiln is that it is limited in production rate because of mechanical limitations on kiln size. A 1500 tpd wet kiln is a large kiln, with 2000 tpd being an upper economic limit without encountering severe maintenance problems. On the other hand dry kiln processes with capacity of 3000 tpd are common [22].

2.2.2.2 Long Dry Kilns

Dimensionally, long dry kilns are similar to long wet kilns. These kilns became popular particularly in North America. Their main advantage over wet kilns is the improved fuel consumption because the kiln feed is dry. However, without any enhanced heat transfer fittings in the preheating zone, kiln exit temperatures of 700°C or more meant that water spray cooling was required, and, overall, very little advantage was realized over the wet process. However, kiln internals included kiln chains (similar to wet kilns), kiln metallic crosses, and ceramic heat exchangers. The crosses and ceramic heat exchangers basically split the kiln into 3 or 4 cross-sectional areas over a distance of about 15 to 20 m, splitting both the feed and gas flow, and providing improved heat transfer [22].

With these enhancements, the kiln gas exit temperatures were reduced to 350°C – 400°C, specific fuel consumption improved some 30% and output increased by 35% to 40% compared to wet kilns. Kiln production rates for long dry kilns are marginally higher than long wet kilns [22].

2.2.3 Preheaters and Precalciners

2.2.3.1 Preheaters

Ideally, the temperature differential between the fresh feedstock entering the kiln and the gas stream leaving the kiln is to be as low as is practically possible. Preheaters consist of several (up to six) stages arranged vertically in such a way that the kiln gas enters each stage from the stage below, and the material enters the stage from the stage above. Each stage consists of a mostly vertical duct and a cyclone. The duct provides the space where heat exchange takes place between the gas and solid. The pulverized state of the kiln feed offers an immense heat exchange surface. At the same time, the very small size of the solid particles minimizes the limiting effect of thermal conductivity and guarantees temperature uniformity across the particle cross section. In a properly built system, where the particles are evenly distributed across the duct, the temperatures of gas and solid equalize almost instantly, unlike in the mixing of liquids.

The solid/gas suspension then enters the cyclone where it is separated. Gas leaves the cyclone through the vertical pipe (thimble) and enters the stage above, and the solid material falls into the stage below. Each stage, therefore, operates as an ideal parallel-flow heat exchanger, but all stages are arranged in a counter flow pattern. In practice, the material temperature increases in each stage by 150°C-250°C, with a corresponding drop of the gas temperature [22].

2.2.3.2 Precalciners

The suspension preheater could eventually have become the basis for practically successful precalciners. Precalciner is beneficial for the fuel combustion, making it much more intensive than the diffusion flame in the firing zone. It is important to remember that the kiln production rate is proportional to the input of heat energy. In kilns with single-point firing, this input is limited by the amount of fuel that can be burned safely and efficiently in the burning zone, which is in essence a function of the kiln diameter. The precalciner, being effectively the second firing zone, greatly expands the system's capability [22].

Efficient and successful operation of a precalciner depends on a number of design aspects. The precalciner performance may be characterized by the combustion efficiency, calcination rate, vulnerability to build-ups and blockage, and ultimately, the production rate and fuel consumption.

2.2.4 Clinker Coolers

In cement manufacturing, formation of clinker nodules occurs at the entrance to the hottest part of the kiln with a material temperature of around 1280°C. The clinker is preferably in the form of 10 mm to 25 mm size nodules that exit from the front end of the kiln into the cooler. It is critical that cooling of the clinker is rapid to secure a phase composition and the rate of cooling can be critical to the clinker quality and performance of cement. It is equally important that the heat exchange between clinker and air is efficient to ensure proper cooling, and at the same time maximizes the recovery of heat to secondary air, tertiary air, and the related process requirement [22].

At the discharge end of the kiln, the clinker is a red hot substance and is also to some extent still reacting chemically toward creating various clinker minerals. The purpose of the clinker cooling is to recoup some of the heat in the clinker, thereby making it cool enough to handle. The other objective is to stop the chemical reactions in the clinker at the point most favourable to the cement quality [22]. The most commonly used clinker cooler used in cement industry is grate clinker cooler as shown in Figure 2-3.

Figure 2-3: Grate clinker cooler [22]

2.2.5 Grinding and Finish Milling

The finish mill system in cement manufacturing is used for reducing the size of the clinker from as large as several centimetres in diameter to a size that is a maximum of

100 micrometers across. This process is accomplished by grinding (milling) with the use of ball mill, roller mills, roll presses, or some combination of these processes.

Clinker may be harder or easier to grind depending on how it is burned, its chemical composition, and how it is cooled. One of the major causes of hard grinding is the amount of C2S (belite) in the clinker. One of the reasons for the formation of belite clusters in clinker is either the use or the inclusion of coarse silica such as quartz, in the raw mix. Long retention times in the kiln result in large C3S (alite) and/or large C2S (belite) crystals. Both of these large crystals are hard to grind, in addition to causing a dusty clinker with higher than desirable amounts of fines [22].

In cement manufacturing plant, the finish milling system is comprised of four basic components namely, 1) feeders, 2) mill, 3) elevator, and 4) separator.

2.2.6 Cement Storage, Shipping, and Packaging

Primary concerns in the handling, packaging and shipping of cement and clinker are the need to keep the products protected from moisture and to guard against impacting the environment in any negative manner. Clinker is relatively more tolerant of adverse weather conditions than cement. The handling considerations for clinker are different from the handling considerations required for the finished cement. It is undesirable to expose either clinker or cement to moist conditions as the hydration of the product results in degradation, although the clinker is somewhat more forgiving [22].

2.2.6.1 Storage of Cement

At the cement plant, cement is stored in silos. A significant portion of cement is later shipped in bulk to terminals for onward distribution to the ultimate customers. When plants are designed, there are two options available for the manufacturer, build large clinker capacity storage or large finish cement capacity storage. The cement will be discharged from the storage into a conveying system, which may include dependent upon the configuration of storage system, screws, belts, elevators and air slides into a final discharge silo for loading into various bulk containers or packaging processes [22].

2.2.6.2 Packaging of Cement

Cement is routinely bagged for convenience at the construction site and for small jobs. Bag sizes for use on construction sites vary significantly. Bagging is usually carried out with the help of rotating packing machines. The cement transferred in this way is aerated so that it flows easily. The bag must be able to fill quickly to the desired weight, yet be able to de-aerate quickly to be palletized. The exterior coating of the bag must allow easy cleaning. Advances in the two-ply construction in cement bags have gained popularity with cement manufacturers [22].

2.3 Fuels Used in the Cement Industry

The appropriate selection and use of a fuel has always been and still is a matter of great concern for the cement industry when used in the kiln for clinkering. The current fierce competition in the cement market and the high impact of the item “fuel cost” in the final price of the product is making companies look at the most economic mix to fire in their kilns. Table 2-1 shows the percentage of fuel consumption in different unit of cement plants based on process types.

Table 2-1: Fuel consumption in cement industry

Item of Consumption	Process Types	
	Dry-Process	Wet-Process
Kiln operation	80% to 100%	95% to 100%
Raw matter drying	0% to 15%	-
Solid fuel drying	0% to 4%	0% to 5%
Other heating systems	0% to 2%	0% to 2%

2.3.1 The Use of Solid Fuels

Solid pulverized fuels are most frequently used by the cement industry. Two fuels actually meet the needs of cement plants:

Coal, which in global terms is the most widely used fuel by the cement industry. Depending on the level of metamorphism coal can be classified as lignite, sub-bituminous, bituminous, and anthracite. Bituminous coal ranks first in preference, followed by sub-bituminous coal and lignite.

Petcoke’s share in the cement production energy matrix grew significantly in the last 10 years. Cement plants helped promote petcoke in the last decade from a waste fuel to a classical fuel because of the cost per unit of released energy during firing is cheaper compared to coal and fuel oil. Petcoke is a black, shiny solid presented as small granules or “needles.” It is an oil refinery byproduct that results from the thermal decomposition of heavy oils. In the composition of petcoke we find mainly carbon, although it can also show high levels of sulphur and heavy metals, such as nickel and vanadium.

Delayed petcoke, fluidcoke, flexicoke are three different type of petcoke used in cement industry. Among the three types of petcoke, flexicoke has the lowest content of volatile materials – between 2.5% and 3.5%.

Solid Fuels Other than Coal and or Petcoke

The use of residual solids by cement plants has increased noticeably. The use of waste fuel by cement plants is a common practice called co-processing, increasing in popularity because it provides reduction of cement production costs as the price for each unit of energy released while firing a waste is far below the price of a classical fuel. The cement industry burns a large number of used tires as waste fuels. The major technical difficulties that prevent a wider use of firing tires in cement kilns is mainly associated with the capacity of equipment to chop tires, the acquisition cost of tire chopping equipment, and the consumption of electric power in the tire chopping procedure. The solid residues that show better results when injected through the main burner include: saw dust, wood chips, plastic shavings (provided the plastic is chlorine-free) and a mix of paper, cardboard, textiles, and plastic.

2.3.2 The Use of Liquid Fuels

Fuel oil is widely used in the cement industry because it is very easy to fire in its liquid phase. It originates from fossils because it is an oil byproduct essentially formed by paraffinic, olefinic, naphthalenic, and aromatic hydrocarbons. Fuel oil was classified initially into six categories and later reduced to only four, based on viscosity, density, and the firing requirements of heating and nebulization (transformation of the liquid into a cloud of droplets through its injection at a high speed into a gaseous atmosphere): Four types of fuel oil are #1, #2, #4, #6 Fuel Oil [22].

#1 Fuel oil is the lightest and least viscous of all fuel oils. It is a light coloured product obtained from fractional distillation of oil. It needs no preheating and is easily nebulized for firing purposes.

#2 Fuel oil is also a product from the distillation of oil, separated under higher temperatures than those deployed for #1 fuel oil. It is a light-yellow liquid that does not need preheating and can be nebulized for firing purposes by mechanic means, i.e., under pressure injection through an appropriate nozzle.

#4 Fuel oil is typically a mix of distilled fractions and liquid residues from oil distillation. It is difficult to fire and requires some special care during nebulization. Nevertheless, it can be fired without preheating.

#6 Fuel oil is also known as bunker C oil, a residual liquid fuel from oil distillation. It is the heaviest and most viscous fuel oils. Efficient firing can be accomplished only after it has been conveniently nebulized through either high pressure or the use of an auxiliary fluid, such as steam or compressed air. Moreover, because it is very viscous, it needs to be heated, handled, and fired. #6 Fuel oil contains the highest amount of ash compared to the other four types of fuel oil because it is the heaviest fraction in the oil distillation, which concentrates the residues. On the other hand, the price per unit of energy released in firing fuel oil #6 is the lowest of all fuel oils. That is the reason why it is used in industrial applications, particularly in cement plants.

The Use of Liquid Waste Fuels

Fuel oils are among the liquid fuels most frequently used by cement plants. In the last few years the share of liquid wastes such as solvents, paint sludge, used lubricant oils, subproducts of the petrochemical industry, etc., has grown considerably in cement plants energetic matrix.

2.3.3 The Use of Gaseous Fuels

The share of natural gas in the cement plant energy matrix has grown slightly in the past few years. Natural gas is a mixture of hydrocarbons with a broad predominance of methane (CH_4) in volumetric contents varies between 70% and 99.6%. Ethane (C_2H_6) is the second most frequent component in natural gas, whereas propane (C_3H_8), butane (C_4H_{10}), carbon dioxide (CO_2), nitrogen (N_2), hydrogen (H_2), oxygen (O_2), and even heavier hydrocarbons appear depending upon the origin, in smaller amounts. Table 2-2 gives a list of typical gaseous waste fuel burned in cement plant.

Table 2-2: Typical waste fuel burned in cement kilns

Gaseous waste	Landfill gas
Liquid Waste	Cleansing solvents
	Paint sludges
	Solvent contaminated waters
	"Slope" – residual washing liquid from oil waste and oil products storage tanks
	Used cutting and machining oils
	Waste solvents from chemical industry
Solid or Pasty Waste	Farming residues (rice husk, peanut husk, etc.)
	Municipal waste
	Plastic shavings
	Residual sludge from pulp and paper production
	Rubber shavings
	Sawdust and wood chips
	Sewage treatment plant sludge
	Tannery waste
	Tars and bitumens
	Used catalyst
	Used tires

2.4 Fuel Consumption in the Canadian Cement Industry

Table 2-3 shows fuel consumption in the Canadian cement industry between 1992 and 2000. Since 1993 there has been a 20% increase in total fuel consumption in the cement industry, from 53,215 TJ in 1993 to 64,043 TJ in 2000. Coal is the dominant fuel used by the cement industry, with natural gas a distant second. Coke produced from coal has also increased over the time period. Also worth noting is the increase in consumption of wood waste and waste fuels. Wood waste increased from zero TJ in 1993 to 35 TJ in 2000. Similarly, consumption of waste fuels has increased by 46% between 1993 and 2000. Still, waste fuels make up a small percentage of total energy use in the industry, with the highest percentage – nearly nine percent of the total – in 1999 [23].

Table 2-3: Fuel consumption in Canadian cement industry, 1992 to 2000, TJ

Consumption (TJ)	1992	1993	1994	1995	1996	1997	1998	1999	2000
Coal	21244	21480	23017	23730	23071	26250	25041	28224	30192
Coke	158	294	272	464	445	288	186	441	916
Petroleum coke	7584	8931	7178	9621	9850	7095	8727	9683	8263
Natural Gas	13106	12676	12423	14673	12141	13411	14422	12286	11916
Electricity	5778	5850	6244	6518	6441	6749	6881	7219	7305
Middle dist	80	64	234	79	110	123	81	86	62
Heavy Fuel Oil	2341	1702	1484	2014	2069	1841	2389	2999	2156
LPG/Propane	23	23	37	0	0	0	0	1	1
Wood Waste	0	0	0	142	103	94	93	71	35
Waste Fuels	1167	2185	4422	3764	3767	1895	5932	6003	3197
Total Energy	51485	53215	55311	61005	58997	57746	63752	67013	64043

2.5 The Sources of Carbon Dioxide in Cement Industry

The cement industry is responsible for approximately 5% of the global anthropogenic CO_2 emissions (based on data from [17] [18] [24] [25]). Carbon dioxide is released in a large amount in the cement industry because combustion is the heart of this industry and the reaction itself in the kiln also produce huge amount of carbon dioxide. In cement manufacturing operations about 40% of the industry's emissions come from fossil fuel combustion; about 5% from transport of raw materials; and about 5% emissions comes from combustion of fossil fuel required to produce the electricity consumed by cement manufacturing operations. The remaining cement-related emissions (about 50%) originate from the process that converts limestone ($CaCO_3$) to calcium oxide (CaO), the primary precursor to cement [26].

Cement manufacturing is energy intensive involving high temperatures and a large amount of heat combustion and carbon dioxide emissions involved in cement making process. Large amount of carbon dioxide is emitted from calcining the limestone or other calcareous materials according to the following reaction (R 1):

$$CaCO_3 \xrightarrow{\Delta} CaO + CO_2 \qquad \textbf{(R 1)}$$

It is seen from reaction R1 that it is chemically impossible to convert limestone ($CaCO_3$) to CaO and then cement clinker without generating CO_2, which is currently emitted into the atmosphere. The calcination normally takes place at 800-900˚C. This reaction thermally decomposes $CaCO_3$ into CaO and CO_2. Typically, cement contains the equivalent of about 64.4 percent CaO. Consequently about 1.135 units of $CaCO_3$ are required to produce 1 unit of cement or 1.6 tonnes of raw materials is needed for 1 tonne of clinker production. Approximately 50 percent by weight of $CaCO_3$ is lost as carbon dioxide during the production. Roughly the industry emits nearly 900 kg of CO_2 for every 1000 kg of cement produced. The carbon dioxide in the flue gas from cement production normally ranges from 22 to 28% on a molar basis.

Table 2-4 shows data from [17] [18] [24] [25], representing the total estimated emissions, cement demand, unit emissions, energy intensity, and clinker factor (kg of clinker per kg of cement) in 2000 for the global cement industry. The gross unit-based emissions for the industry were approximately 0.87 kg CO_2 per kg of cement as shown in the Table 2-4. Unit-based emissions vary globally from 0.73 to 0.99 kg CO_2 per kg of cement. There is similar variation in energy intensity and clinker factor. Two of the important factors that drive unit-based CO_2 emissions are the energy intensity and clinker factor. Lowering the energy intensity lowers the fossil fuel combustion during production. Lowering the clinker factor directly reduces both the process emissions and the associated fuel-related emissions.

Table 2-4: Cement Demand, Total and Unit CO_2 Emissions, Energy Intensity and Clinker Factor in Major World Regions for the Year 2000 [26]

Region	Total Cement Demand (Mt/year)	Unit Emissions (Mt CO_2/ Mt Cement)	Total CO_2 Emissions (Mt/year)	Energy Intensity (MJ/kg Cement)	Clinker Factor (kg/kg)
1. USA	90	0.99	90	5.50	0.88
2. Canada	9	0.91	8	5.20	0.88
3. W. Europe	220	0.84	186	4.04	0.81
4. Japan	82	0.73	60	3.10	0.80
5. Aus & NZ	8	0.79	6	4.08	0.84
6. China	500	0.90	449	4.71	0.83
7. SE. Asia	123	0.92	112	4.65	0.91
8. Rep of Korea	44	0.90	40	4.05	0.96
9. India	69	0.93	64	4.71	0.89
10. FSU	88	0.81	71	5.52	0.83
11. Other E. Europe	37	0.89	33	5.20	0.83
12.S&L. America	134	0.82	109	4.48	0.84
13. Africa	87	0.85	74	4.75	0.87
14. Middle East	80	0.85	68	4.92	0.89
Total	1571	0.87	1371	-	-

To calculate the total carbon dioxide emission, different emission factors are used as shown in Table 2-5 related to the cement production [27]. Energy-related emission factors were calculated from the Environment Canada publication “Canada’s GHG Inventory 1997 Emissions and Removal Trends”. For emissions from calcination, the method provided in the U.S. Climate Wise Program was applied and it is consistent with the Intergovernmental Panel on Climate Change (IPCC) Guidelines. The amount of carbon dioxide released during the calcination step is based on the lime content of the clinker or cement.

Table 2-5: Emission factors for different fuels used in cement industry

Fuel or Process	Emission Factor
Coal	2.52 kg CO_2 / kg of fuel
Coke	2.48 kg CO_2 / kg of fuel
Natural Gas	1.88 kg CO_2 / m^3 of fuel
Distillate Oil	2.83 kg CO_2 eq / L of fuel
Diesel	2.73 kg CO_2 eq / L of fuel
Gasoline	2.36 kg CO_2 eq / L of fuel
Propane	1.53 kg CO_2 eq / L of fuel
Calcination	0.52 - 0.54 kg CO_2 eq / kg of clinker

2.6 Greenhouse Gas Emissions in Canadian Cement Industry

Table 2-6 shows greenhouse gas emissions associated with the cement industry in Canada [23]. Overall, the cement industries contributed 7.3 percent of all greenhouse gas emissions from manufacturing and mining industries in Canada. Table 2-6 shows both greenhouse gas emissions attributable to the burning of fuels, as well as that attributable to the cement making process itself (calcination). In fact, process-related greenhouse gases are roughly equivalent to combustion related greenhouse gases. In Canada, the cement industry experienced a 20% increase in fuel consumption between 1993 and 2000, and the figures below indicate that between 1993 and 2000, the cement industry realized a 21% increase in greenhouse gas emissions. The largest increases are the result of emissions from coal and coke. It is important to note that the estimates of greenhouse gas emissions are calculated by emissions factors, which means that actual greenhouse gas emissions could vary according to the exact make-up and quality of the fuels used. Coke from coal, waste fuels, coal and petroleum coke have relatively high greenhouse gas emissions factors – all in the range of 0.085 per terajoule - while natural gas has an emissions factor of roughly 0.05.

Table 2-6: Total Greenhouse Gas Emissions by Fuel Source in Canadian Cement Industry, 1992 - 2000, Thousand Metric Tons

FUEL	1992	1993	1994	1995	1996	1997	1998	1999	2000
Coal	1714	1732	1861	1919	1865	2123	2005	2260	2423
Coke	13	25	23	39	38	24	16	37	78
Petroleum Coke	636	749	602	807	826	595	701	779	665
Natural Gas	652	629	608	727	601	663	713	608	591
Middle Dist.	5	4	17	5	8	9	5	6	4
Heavy Fuel Oil	173	126	110	149	153	136	174	218	157
LPG/Propane	1	1	2	0	0	0	0	0	0
Wood Waste	0	0	0	12	8	8	8	6	3
Waste Fuels	100	187	381	323	323	162	509	515	274
GHG Emissions (Combustion)	3294	3453	3604	3981	3822	3720	4131	4429	4195
GHG Emissions (Process CO2)	4440	4525	5332	6035	5722	6156	6198	6474	6679
Total GHG Emissions	7734	7978	8936	10016	9544	9876	10329	10903	10874

2.7 The Case Study for This Thesis

The St. Marys cement plant was chosen as the case study in this research. St Marys Cement Plant was founded in 1912 and the plant is located in Ontario, Canada. It is the largest employer in the town of St. Marys, Ontario and one of the Canada's leading cement manufacturers. It employs 157 people on 1,500-acre site. The plant produces an average of 2,400 tons of clinker per day and with a name plate capacity of 680,000 tons of cement per year. Table 2-7 shows the data for cement production, fuel consumption, energy consumption, and raw material consumption from the St. Marys cement plant for the year 2002 to 2004. The rock composition used in St. Marys plant contained the material as Loss of Ignition (LOI)% 42.882, CaO% 51.53, SiO_2% 1.796, Al_2O_3% 0.478, Fe_2O_3% 0.383, MgO% 2.146, SO_3% 0.539, K_2O% 0.094, Na_2O% 0.051, P_2O_5% 0.041,

Mn_2O_3% 0.034, and TiO_2% 0.026. St. Marys cement used diesel, coal, coal fines, and petcoke for fuels. The heating value for petcoke is 32.563 MJ/kg and the composition is C% 87.6, S% 5.8, N% 1.9, H% 1.7, volatiles% (including moisture) 3.2, and ash% 0.3. The heating value for coal is 26.967 MJ/kg and the composition is C% 50, S% 2.35 on a wet basis. The coal fines and coal are used 1: 3 ratio as fuel in kiln burner. The clinker composition is SiO_2% 21.75, Fe_2O_3% 3.11, Al_2O_3% 4.46, CaO% 65.98, MgO% 3.07, SO_3% 1.41, K_2O% 0.57, Na_2O% 0.22, TiO_2% 0.23, P_2O_5% 0.14, and Mn_2O_3% 0.08. The flue gas flow rate and composition from St. Mary cement plant is shown in Table 3-1. The research was performed on the basis of four cases below:

- **Case I:** the plant operates at the highest capacity
- **Case II:** the plant operates at average load
- **Case III:** the plant operates at minimum operating capacity
- **Case IV:** Switching to a lower carbon content fuel at average plant load

Table 2-7: Production data, Fuel, Energy, and Raw Material consumption report for St. Marys Cement for the year 2002 - 2004

	2002	2003	2004
Production Report			
Clinker (Tonnes)		567,926	636,839
Total Cement Produced (Tonnes)	591,746	613,721	609,110
Fuel Consumption			
Diesel in Quarry (Litres)	646,971	303,754	604,574
Coal (Tonnes)	40,034	28,401	20,760
Petcoke (Tonnes)	16,419	19,488	37,788
Energy Consumption			
MJ	1,747,839,672	1,645,956,255	2,074,213,668
millions BTU	1,655	1,559	1,964
Total kWh			94,373,070
Runtime Hours – Kiln	6,811	6,366	6,964
Raw Material Consumption			
Rock	889,905	750,795	855,713
Alumina	1,843	1,636	525
Silica Flour	62,600	75,844	80,194
Silica Fume	914	1,092	648

2.8 How Can CO_2 be captured?

Developing economically feasible technology to reduce CO_2 emission from fossil fuel burning process is becoming more and more essential. Several alternative strategies have been proposed to reduce CO_2 emissions from fossil fuel combustion and these strategies include switching of fuels from higher carbon content to low carbon content fuels with lower CO_2 emission, fuel balancing, and improving plant efficiencies through utilization of advanced fossil fuel technologies and CO_2 capture and sequestration. Among the options, greater reduction of CO_2 from flue gas is expected to be technically possible using CO_2 capture and sequestration. Three main techniques for CO_2 capture are:

- Post-combustion capture
- Pre-combustion capture
- Oxy-fuel combustion

Post combustion capture: A variety of techniques can be used to capture CO_2 from the flue gases for post combustion capture process. The most commonly thought method to capture CO_2 from a coal burnt flue gas is amine scrubbing, using MEA as the solvent. The amine from the scrubber is heated by steam to release high purity CO_2 and the CO_2 free amine is then reused in the scrubber. However, the major shortcoming for the application of the MEA CO_2 capture process is energy intensive for solvent regeneration, and necessitates high capital and operating costs.

Pre-combustion capture: The low concentration of CO_2 in flue gas means that a large volume of gas has to be handled, which results in large equipment sizes and high capital costs. A further disadvantage of the low CO_2 concentration is that powerful chemical solvents have to be used to capture CO_2 and regeneration of the solvents to release the CO_2 requires a large amount of energy. If the CO_2 concentration and pressure could be increased, the CO_2 capture equipment would be much smaller and different physical solvents could be used, with lower energy penalties for regeneration. This can be achieved by pre-combustion capture. The fuel is reacted with oxygen or air, and in some cases steam, to give mainly carbon monoxide and hydrogen. The carbon monoxide is reacted with steam in a catalytic reactor, called a shift converter, to give CO_2 and more hydrogen. The CO_2 is separated and the hydrogen is used as fuel.

Oxyfuel combustion: The concentration of CO_2 in flue gas can be increased greatly by using concentrated oxygen instead of air for combustion. The oxygen would be produced by cryogenic air separation. If fuel is burnt in pure oxygen, the flame temperature is excessively high, so some CO_2 rich flue gas would be recycled to the combustor to make the flame temperature similar to that in a normal air-blown combustor. The advantage of oxygen-blown combustion is that the flue gas has a CO_2 concentration of around or above 80%, compared to 4-14% for air blown combustion, so only simple CO_2 purification is required. It may be possible to omit some of the flue gas cleaning equipment such as flue gas desulphurisation, which would reduce the net cost of CO_2 capture. The disadvantage of oxyfuel combustion is that a large quantity of oxygen is required, which is expensive, both in terms of capital cost and energy consumption.

2.9 Types of CO_2 Capture Technology

Several technologies have been developed to capture carbon dioxide from flue gases. They are:

- Chemical solvent scrubbing
- Physical solvent scrubbing
- Adsorption
- Membranes
- Cryogenics

2.9.1 Chemical solvent scrubbing

Chemical solvent scrubbing is currently the most favoured method to capture CO_2 from low-pressure flue gases. Chemical solvent scrubbing involves the reaction between CO_2 and the absorbents. The chemical absorbents used are aqueous solutions of mono, di or tri ethanol amines, diisopropanol amine, sodium carbonate, sodium hydroxide, and potassium carbonate [28]. The most common solvent used for CO_2 removal from low-pressure flue gas is monothanolamine (MEA). The flue gas is cooled down and particulates and other impurities are removed as much as possible before CO_2 removal. The flue gas is then contacted counter-currently with the chemical solvent in an absorption vessel equipped with packings or plates where much of the CO_2 is absorbed by chemically reacting with the chemical solvent and forms a loosely bound intermediate compound. The CO_2-rich solvent is then passed into a stripper column where with the addition of heat the reverse reaction takes place. The bonds are broken with the application of heat thus regenerating the original solvent and producing a carbon dioxide stream. The CO_2-free solvent is recycled to the absorption vessel and the process runs continuously. CO_2 released in the stripper is compressed for transport and storage. CO_2 recovery of 98% with product purity in excess of 99% can be achieved using the amine scrubbing technology [29].

Amine scrubbing technology has been established for over 60 years in the chemical and oil industries, for removal of hydrogen sulphide and CO_2 from gas streams. DEA (diethanolamine) has also been used in addition to other chemicals such as potassium carbonate, but MEA is the most commonly used for flue gas applications [30] [31] [32] [33].

The disadvantages of this process are the limited CO_2 loadings and significant energy requirements as well as corrosion in presence of O_2 and high solvent degradation rates in the presence of SO_2 and NO_2. The process also requires extensive equipment for circulating large volumes of liquid absorbents, high solvent consumption and heat exchange and large energy losses. New or improved solvents with higher CO_2 absorption capacities, faster CO_2 absorption rates, high degradation resistance and low corrosiveness and energy use for regeneration are needed to reduce equipment sizes and capital operating cost.

2.9.2 Physical solvent scrubbing

Physical absorption is usually used when the flue gas contains very high partial pressures of CO_2, typical for systems operating much above atmospheric pressure. The low concentration of CO_2 results in a lower partial pressure, which reduces significantly the efficiency of absorption because the solubility of CO_2 is directly proportional to its partial pressure. The technology development needs for physical solvents are similar in principle to those for chemical solvents. The CO_2-bearing gas is typically contacted counter-currently with the absorbent in a packed or plate tower and the advantage of such solvents is that the regeneration is performed by pressure reduction resulting in much lower energy consumption that causes the absorbed CO_2 to flash into the vapour phase. The main physical solvents that could be used for CO_2 capture are typically cold methanol (Rectisol process), N-methyl-2-pyrrolidone, polyethylene glycol dimethylether (Selexol process), propylene carbonate (Fluor process) and sulfolane. The physical solvents have relatively low affinity for NOx and O_2; however, SO_2 is 100 times more soluble than CO_2 and thus the pre-removal of SO_2 is still required. Therefore, SO_2 levels must be reduced to 1 ppm before a physical solvent is used to minimize the amount of make-up solvent.

CO_2 capture literature for physical solvent scrubbing has been reported only for Selexol process. IGCC power plants operate at a high pressure and the partial pressure of CO_2 is much higher than the conventional power plants. In IGCC power plant the flue gas CO_2 concentration is about 35-40% and the total pressure is at least 20 bar. The use of Selexol has been reported for IGCC power plants. The physical absorption process for oxy fuel combustion scheme can be a very efficient approach for capturing CO_2 because the CO_2 concentration in the flue gas varies from 80% to 95% on a dry basis. A high concentration of CO_2 produces a very high partial pressure of CO_2 that makes physical solvent scrubbing a suitable option [34] [35] [36].

2.9.3 Adsorption

Adsorption is due to the significant intermolecular forces between gases and the surfaces of certain solid materials. Some solid materials with high surface areas, such as activated carbon and molecular sieves (zeolites) can be used to separate CO_2 from gas mixtures by adsorption. These adsorbents are normally arranged as packed beds of spherical particles. The process operates with the basic adsorption and regeneration steps on a repeated cycle. Gas is fed into a bed of solids in the adsorption step that adsorbs CO_2 and allows other gases to pass through the bed. When the bed becomes completely saturated with CO_2, the feed gas is switched to another clean adsorption bed and the completely saturated bed is regenerated to remove the CO_2. Adsorption of a single gas or multiple layers of gases can be carried out depending on the temperature, partial pressures, surface forces and adsorbent pore sizes [28].

In pressure swing adsorption (PSA), the gas mixture flows through the beds at elevated pressures and low temperatures until the adsorption of the desired constituent approaches equilibrium conditions at the bed exit and the adsorbent is regenerated by reducing pressure. Once regenerated, the beds are ready for another adsorption cycle. In temperature swing adsorption (TSA), hot inert gas or external heat is used to regenerate adsorbents by raising their temperature. In electric swing adsorption (ESA) regeneration takes place by passing a low-voltage electric current through the adsorbent. PSA and TSA are used commercially for gas separation and are used to some extent in hydrogen production and in the removal of CO_2 from natural gas. It has been found that PSA is superior to TSA in all cases due to the high-energy requirements and low speed of regeneration in TSA. ESA is not yet commercially available but it is said to offer the prospect of lower energy consumption than the other processes. Adsorption is not yet a highly attractive approach for the large-scale industrial separation of from flue gases because the capacity and CO_2-selectivity of available adsorbents is low. However, adsorption may be successful when combined with another capture technology [37].

2.9.4 Membranes

The differences in physical or chemical interactions between gases and membrane material cause one component to pass through the membrane faster than another component. Multiple stages and/or recycle of one of the streams are necessary to achieve high degrees of separation with the membrane process. This leads to increased complexity, energy consumption and costs. When expanded to use with large-scale plants, the cost will likely be more expensive compared to other capture techniques. Several membranes with different characteristics may be required to separate high-purity CO_2. Various types of membranes are currently available, including porous inorganic membranes, palladium membranes, polymeric membranes and zeolites.

Gas separation membranes and gas absorption membranes are two types of membrane processes. In gas separation membranes, the porous structure of the solid membrane permits the preferential permeation of the mixture constituents. The gas mixture containing CO_2 is introduced into the membrane separator at elevated pressure and CO_2 passes preferentially through the membranes and is recovered on the shell side of the separator at reduced pressure. Gas permeability and membrane selectivity are the main parameters considered for designing and operating a separation process. Membranes for CO_2 separation from flue gases have not been widely explored because of the low pressure of the flue gases. The atmospheric flue gas needs to be pressurized to achieve the desired separation, the costs are almost double than those of conventional amine capture process because of the compression energy.

On the other hand, gas absorption membranes are microporous solids that are used as contacting devices between a gas and a liquid. The CO_2 diffuses through the membrane and is removed by an absorption liquid such as amine, which selectively removes certain components. It is more compact than conventional membrane separators. In contrast to gas separation membranes, it is the absorption liquid, not the membrane that gives the process its selectivity.

2.9.5 Cryogenics

Cryogenics is a low-temperature separation technique. Cryogenic separation of gas mixture involves compressing and cooling down the CO_2 bearing gas mixture in several stages to induce phase changes in CO_2 and invariably other mixture components in the case of flue gases. A major disadvantage of cryogenic separation of CO_2 is its inherent energy intensiveness to provide the refrigeration necessary for the process, particularly for dilute gas streams. Another disadvantage is the possible plugging of solid CO_2 clusters and ice that comes from the water vapour freezing in the CO_2 feed mixture. Cryogenic separation has the advantage that it enables direct production of liquid CO_2, which could be easily transported and shipped to the storage.

Cryogenic separation is widely used commercially for purification of CO_2 from streams that already have high CO_2 concentrations (typically >90%). It is not normally used for more dilute CO_2 streams, although it has recently been claimed that CO_2 can be captured (by freezing it as a solid) from atmospheric pressure flue gases. The most promising applications for cryogenics are expected to be for separation of CO_2 from high-pressure gases, such as in pre-combustion capture processes, or oxyfuel combustion in which the input gas contains a high concentration of CO_2 [37].

2.9.6 Other techniques

Chemical looping combustion, in which, direct contact between the fuel and the combustion air is avoided by using a metal oxide to transfer oxygen to the fuel in a two-stage process [38]. In the reduction reactor, the fuel is oxidised by reacting with a metal oxide, which it converted to a lower oxidation state. It is then transported to a second reactor, the oxidation reactor, where it is re-oxidised by reacting with O_2 in the air. The major development issue associated with chemical looping combustion is development of a metal oxide material that is able to withstand long-term chemical cycling and is resistant to physical and chemical degradation from impurities generated from fuel combustion [37].

2.10 CO_2 Storage and Usage

2.10.1 Industrial Usage of CO_2

Captured CO_2 can either be utilized or stored. Carbon dioxide captured from process streams are used for many purposed including:

- food processing and carbonation
- synthesis of chemicals such as urea, methanol, organic and inorganic carbonates and as a solvent
- direct biofixation through growth of biomass/algae for fuel

On a scale of interest for CO_2 capture from cement plant, CO_2 can be used in CO_2 enhanced oil recovery (CO_2 EOR) and enhanced coal bed methane recovery (ECBM). Figure 2-4 gives the breakdown of CO_2 utilization in the USA in 1989.

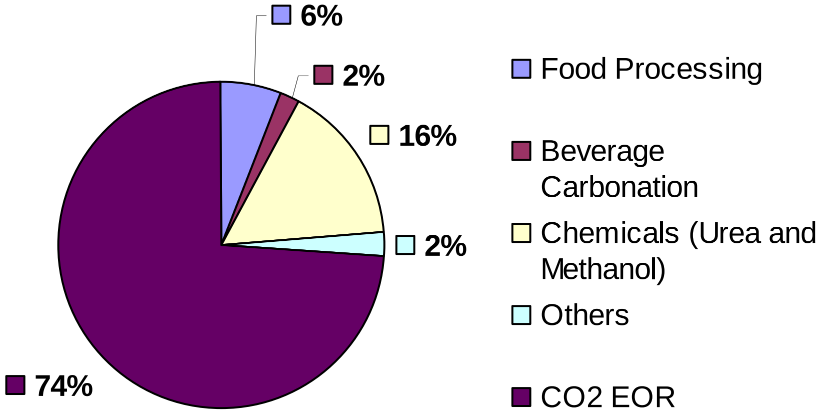

Figure 2-4: Annual US CO_2 utilization in 1989

In 1989, a total of the 4.7 Mt was consumed in the form of liquid/solid CO_2 in the USA of which the food processing industry consumed around 2.7 Mt/year. The cooling of food, especially ice cream, meat products, and frozen foods, was the main user for both solid and liquid CO_2. In the food processing industry, approximately 0.9 Mt/year of CO_2 is used for beverage carbonation in the USA, with soft drinks and beer production consuming the largest quantity. Urea, methanol and other chemical manufacturing consumed around 6.5 Mt/year of CO_2 [39]. Other uses such as the production of dimethyl carbonate (DMC), CO_2 based alkylene polycarbonate, metal processing, etc. consumed around 0.7 Mt/year of CO_2. More recent estimates report that global consumption for chemical synthesis is 89 Mt/year [40]. The use of carbon dioxide for EOR and ECBM is more attractive. In CO_2 enhanced oil recovery (CO_2 EOR), CO_2 is used as a miscible solvent. There are 74 enhanced oil recovery projects operating in the U.S and around 28 Mt/year of CO_2 was consumed in 1998 in the USA for CO_2 EOR. Most of this carbon dioxide is extracted from natural CO_2 reservoirs. But CO_2 comes from anthropogenic sources. In ECBM, carbon dioxide can be injected into suitable coal beds where it would be adsorbed on the coal, locking it up permanently provided the coal is never mined. Moreover, it preferentially displaces methane that exists along with the coal and this methane could be sold as fuel [37].

2.10.2 Long-Term Utilization/Storage of CO_2

2.10.2.1 CO_2 Enhanced Oil Recovery

The use of CO_2 EOR can increase the oil recovery in a reservoir by 10-15% [40]. CO_2 is already used in more than 74 EOR projects in the USA and elsewhere and about 30 Mt CO_2/year is consumed. CO_2 is injected into the oil reservoir and comes up with the produced oil and is usually separated from the oil and re-injected back into the reservoir to minimize operating costs. The current source of CO_2 used in CO_2 EOR projects is mainly from natural underground reservoirs of CO_2. However, it is possible to use CO_2 captured from fossil fired power plants if the cost of the product delivered on site is competitive with CO_2 recovered from natural reservoirs.

The Weyburn CO_2 EOR project in Saskatchewan, Canada, is an example of a scheme where all of the CO_2 is supplied from a North Dakota (USA) coal gasification plant located approximately 300 km away from the injection site. Worldwide, EOR projects have the potential to sequester 120 Gt of CO_2 [37].

2.10.2.2 CO_2 Storage – Mineral Carbonation

Many methods of CO_2 storage are currently being researched. These methods include storage in saline aquifer and mineral carbonation. Mineral carbonation is advantageous because it provides a means for permanent disposal with no requirements for monitoring, or worries for re-emission. Mineral ores containing magnesium or calcium can be chemically combined with CO_2 to form chemically more stable carbonates. Mineral carbonation involves the reaction of carbon dioxide with minerals such as fosterite and serpentine. The overall reactions are shown below.

$$Mg_2SiO_4 + 2CO_2 \rightarrow 2MgCO_3 + SiO_2 \text{(Fosterite)} \qquad \textbf{(R 2)}$$

$$Mg_3Si_2O_5(OH)_4 + 3CO_2 \rightarrow 3MgCO_3 + 2SiO_2 + 2H_2O \text{ (Serpentine)} \qquad \textbf{(R 3)}$$

Both reactions are exothermic and thermodynamically favourable at low temperatures. The mineral carbonation process first may require processing of the minerals to reduce the particle size, removal of trace metals, and heat treatment to speed up the reaction rate to acceptable levels.

2.11 Overview of MEA-CO_2 Capture Process

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 [67]. Carbon dioxide capture processes using monoethanolamine (MEA) based solvent is an organic chemical absorption process. It was developed over 60 years ago as a general, nonselective solvent to remove acidic gas impurities (e.g., H_2S, CO_2) from natural gas streams. The process was then adapted to treat flue gas streams for CO_2 capture. Fluor Daniel Inc., Dow Chemical Co., Kerr-McGee Chemical Corp., and ABB Lummus Crest Inc. were some of the initial developers of the MEA-based technology for CO_2 capture. Typically, about 75-90% of the CO_2 is captured using this technology with a purity of product CO_2 stream greater than 99%.

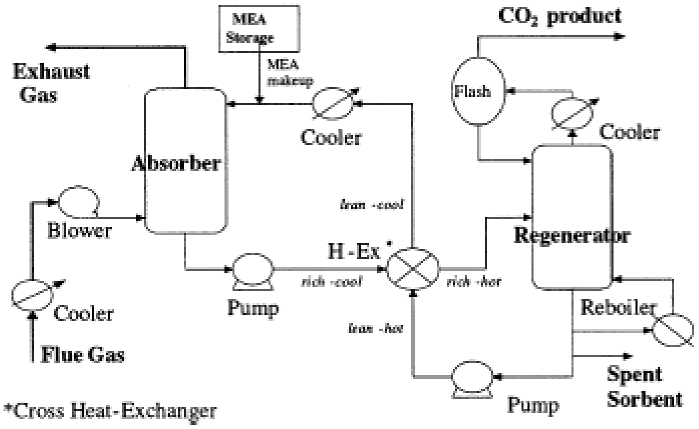

Figure 2-5: Flowsheet for CO_2 Capture process from flue gases using MEA solvent

The amine process has two main elements, which are the absorption and stripping columns as illustrated in Figure 2-5 [67]. Amines in the water solution react with CO_2 in the absorption column, forming chemical compounds that separate CO_2 from the gas mixtures at a higher rate than the natural CO_2 absorption in pure water. The monoethanolamine (MEA), which is a primary amine, produces the carbamate ion when it reacts with CO_2. The main advantages in using MEA/water solutions are its high CO_2 reactivity, its high limit load (0.5 mole of CO_2 per mole MEA) and its low molecular weight. The limit of 30% by wt of MEA in water was set to avoid corrosion. Its main drawback is the stability of the carbamate ion that makes the regeneration more heat-demanding to regenerate the solvent.

Acid gases such as SO_2 and NO_2 react with MEA to form heat-stable salts that reduce the CO_2 absorption capacity of the solvent and raise the MEA make-up to cover additional losses. Thus, very low concentrations of these gases (typically 10 ppm) are desirable to avoid excessive loss of solvent. The problem is especially acute for SO_2 because its concentration in flue gases is typically 700- 2500 ppm for 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.

The flue gas entering the absorption column flows up through the vessel, countercurrently to the aqueous solution. The carbon dioxide in the flue gas reacts chemically with the solvent while the purified gas is vented to the atmosphere, and the solvent enriched by CO_2 is pumped from the absorption tower to a heat exchanger. The rich solvent is preheated in the lean/rich exchanger by the hot lean solution returning from the stripper on its way back to the absorber. The rich solvent solution enters the top of the regenerator where it flows down through the vessel countercurrently to the stripping steam generated in the reboiler. Steam and solvent vapours move up the regenerator column and condenses as CO_2 is liberated. Uncondensed steam and carbon dioxide leave the top of the regenerator and then enter the reflux condenser. The condensate is returned to the system while the carbon dioxide is removed. The lean solvent solution is pumped from the bottom of the regenerator directly to the lean/rich heat exchanger. The solvent leaves the lean/rich heat exchanger after giving up heat to the rich solution and then enters a cooler, where its temperature is further lowered before being returned to the absorber.

2.11.1 MEA-CO_2 Absorption Reaction Mechanisms

Carbon dioxide capture with MEA solution involves exposing a gas stream to an aqueous amine solution which reacts with carbon dioixide in the gas to form a soluble carbonate salt by an acid-base neutralization reaction. Aqueous solutions of amines absorb CO_2 primarily by chemical reaction; the reactions between the amines and CO_2 are quite complex. However, when carbon dioxide is absorbed into aqueous MEA (monoethanolamine) solutions, the following two overall reactions should be considered [41] [42].

$$CO_2 + 2RNH_2 \Leftrightarrow RNH_3^+ + RNHCOO^- \qquad \textbf{(R 4)}$$

$$CO_2 + RNHCOO^- + 2H_2O \Leftrightarrow RNH_3^+ + 2HCO_3^- \qquad \textbf{(R 5)}$$

Where, R refers to $HOCH_2CH_2$. At very short exposure times of the liquid to the gas encountered in industrial absorbers, the effect of reaction (R 5) can be neglected, and only reaction (R 4) affects the absorption rate of carbon dioxide [43].

Reaction (R 4) takes place in two steps:

$$CO_2 + RNH_2 \Leftrightarrow H^+ + RNHCOO^- \qquad \textbf{(R 6)}$$

$$H^+ + RNH_2 \Leftrightarrow RNH_3^+ \qquad \textbf{(R 7)}$$

The equilibrium constants of reactions (R 6) and (R 7) are in the order of 10^{-5} and 10^{10} l/g-mole, respectively [42]. The second step reaction (R 7) is ionic and virtually instantaneous, whereas the first-step reaction (R 6) is second order, that is, first order with respect to both carbon dioxide and MEA, and is rate controlling.

The equilibrium of reaction (R 4) lies to the right under high pressure and low temperature (typically < 40°C) conditions and it shifts to the left under low pressure and high temperature conditions. The forward reaction of (R 4), which is favoured under high pressure and low temperature, represents the absorption step and the reverse reaction, which is favoured under low pressure and high temperature, represents the regeneration step. Therefore, CO_2 can be absorbed under high pressure and low temperature conditions and the CO_2 loaded solvent can be regenerated under low pressure and high temperature conditions in a stripping column [52]. Carbon dioixide can be separated by heating in a separate stripping column. The major advantage of this technique is that, in the ideal case, the amine is not consumed and is continuously recycled back during the process.

The capacity of the solution for CO_2 for reaction (R 6) is limited to approximately 0.5 mole of CO_2 per mole of amine, even at relatively high partial pressures of CO_2 because of the high stability of the carbamate and its low rate of hydrolysis to bicarbamate.

2.11.2 Corrosion in MEA-CO_2 Capture Process

Carbon dioxide (CO_2) capture unit using reactive amine solvents is constantly subject to excessive corrosion problem. In MEA absorption process the main operational problem results from corrosion, which has a great impact on the plant's economy as it can result in unplanned downtime, production losses, and reduced equipment life as well as limiting the operational ranges of process parameters.

Corrosion mechanism in an aqueous amine-CO_2 system is not well-understood. Several mechanisms have been used to postulate the corrosion phenomena. Riesenfeld and Blohm [44] suggested that the corrosion is associated with an evolution of CO_2 from the rich amine solution. The evolved CO_2 then reacts directly with carbon steel to form iron carbonate ($FeCO_3$). However, in most cases, mechanisms of iron dissolution in a CO_2-water system are used to represent the corrosion mechanism in the aqueous amine-CO_2 system. Three different types of iron dissolution reactions have been suggested. First, the dissolution reaction involves a reduction of hydrogen ion (H^+) as shown is reaction (R 8). Second, bicarbonate ion (HCO_3^-) in the solution functions as an oxidizing agent in the reduction in reaction (R 9). Third, the reaction is governed by undissociated carbonic acid (H_2CO_3) reaction (R 10).

$$Fe + 2H^+ \rightarrow Fe^{2+} + H_2 \qquad \textbf{(R 8)}$$

$$Fe + 2HCO_3^- \rightarrow FeCO_3 + CO_3^{2-} + H_2 \qquad \textbf{(R 9)}$$

$$Fe + H_2CO_3 \rightarrow FeCO_3 + H_2 \qquad \textbf{(R 10)}$$

Recently, a mechanistic corrosion model was established to identify the oxidizing agents responsible for corrosion reactions in an aqueous amine-CO_2 system [45].

It is observed that the level of corrosion in amine treating plants is influenced by a number of factors including [46]:

- the type of amine used
- the presence of contaminants in the solution
- CO_2 loading, amine concentration

- operating pressure and temperature
- solution velocity
- oxygen content in the system, and degradation products

Generally, the primary amine (MEA) is found to be more corrosive than the secondary amine (DEA) and tertiary amine (MDEA).

The carbon dioxide loading in the solvent strongly affects the corrosion rate and an increase CO_2 loading intensifies the system's aggressiveness with the corrosion process. The recommended maximum CO_2 loadings in the rich amine solution are in the range of 0.25-0.40 mol/mol for MEA. The rich amine solution is generally more corrosive than the lean amine.

An increase in amine concentration also causes a higher corrosion rate, because at high amine solution concentration the reaction rate would be high. To limit the plant corrosion within acceptable and manageable levels, most amine plants are operated in compliance with the recommended maximum amine concentrations of 20 to 30 % by weight.

Temperature has a significant impact on corrosion. If the amine solution temperature is too high, the reaction rate of each reaction would increase including corrosion reaction in order to maintain the equilibrium of the system.

Solution velocity can cause severe erosion and corrosion because the protective film developed by inhibitors can be removed or damaged by the shear force of a high velocity fluid stream.

Aqueous amine solutions can be degraded in the presence of carbon dioxide and oxygen. Solvent degradation leads to corrosion in MEA plants. Oxygen exists in the flue gas and reacts with other substances in the system. The source of oxygen in the flue gas is from the excess air in the combustion step. The corrosion reaction occurred from oxygen are shown in Equation (R 11) and (R 12).

$$2Fe + 2H_2O + O_2 \rightarrow 2Fe(OH)_2 \qquad \textbf{(R 11)}$$

$$2Fe(OH)_2 + 2H_2O + \frac{1}{2}O_2 \rightarrow 2Fe(OH)_3 \qquad \textbf{(R 12)}$$

The most susceptible areas include the bottom portion of the absorber, flash drum, rich-lean heat exchanger, regenerator and reboiler. The corrosion detected in these areas occurs in many forms, i.e. uniform, pitting, erosion, galvanic, stress corrosion cracking and inter-granular.

Corrosion can be reduced by a number of approaches including [46]:

- using of proper equipment design
- using of corrosion resistant materials instead of carbon steel
- maintaining low temperatures of the solution and the steam in the reboiler
- removing suspended solids and amine degradation products continuously
- adding caustic to the circulating amine solution
- maintaining the lowest possible pressure on stripping columns and reboilers
- using of corrosion inhibitors

Of these alternatives, the use of corrosion inhibitors is considered to be the most economical technique for corrosion control.

2.11.3 Solvent Degradation in MEA-CO_2 Capture Process

If the loaded amine solution contacts with oxygen-containing gas, the amine is subject to oxidative degradation. The exact mechanism for the degradation is still unclear, but several studies have shown that the degradation products are oxidized fragments of the amine including NH_3, formate, acetate, and peroxides [47]. There are several reaction mechanisms possible in the degradation of alkanolamines. One of these is the reaction of H_2S (captured from the flue gas) with O_2 to form free sulphur, which upon heating, reacts with the amine and forms dithiocarbamates, thioureas and other decomposition products. These products cannot be reconverted to the free amine by application of heat. In addition, thiosulfuric acid may be formed which combines with the amine to form a heat-stable salt that cannot be recovered for gas absorption by heating. The major heat-stable salt consists of sulphate and oxalate. Flue gases contain some gases such as sulphur dioxide, nitrous oxide, and oxygen which can react with the solvent to form heat-stable salt which cause the solvent degradation and reduce the efficiency of the absorption process. These compounds react with amines and cannot be regenerated, which subsequently increase MEA use to cover additional losses. There are many factors that affect the solvent degradation. The flue gas temperature affects the degradation rate. Hot flue gas can increase the degradation rate and reduce the absorption efficiency. The flue gas should be cooled down before entering to the MEA absorption system. Another way to reduce the degradation problem is to remove these contaminate gases responsible for solvent degradation before being entered into the absorption system. Flue gas desulphurization (FGD) is used to remove sulphur content gas in the flue gas. Reclaimers are used in MEA absorption processes for the removal of solvent degradation. Around 1-10% of the total solvent circulation rate would be treated in the reclaimer.

2.12 Economic Analysis of Capture Process

The economic analysis for MEA-CO_2 absorption process consists of two parts; capital investments, and total product cost. The details of each part are described below [48].

2.12.1 Capital Investment

Capital invesment can be divided into two categories namely direct and indirect costs.

2.12.1.1 Direct Costs

Direct costs include the following costs:

Purchased equipment

The cost of purchased equipment is the basis of estimating capital investment. This includes all equipment listed on a complete flowsheet, spare parts and non-installed equipment spares, surplus equipment, supplies, and equipment allowance, inflation cost allowance, freight charges, taxes, insurance, duties, allowance for modification during startup. Sources of equipment prices, methods of adjusting equipment prices for capacity, and methods of estimating auxiliary process equipment are therefore essential to the estimator in making reliable cost estimates.

Purchased-equipment installation

The installation of equipment involves costs for labour, foundations, structural supports, platforms, construction expenses, insulation, paint and other factors; directly related to the erection of purchased equipment. Depending upon the complexity of the equipment and type of the plant in which the equipment is installed, the installation costs for equipment are estimated from 25 to 55 percent of the purchased equipment cost.

Instrumentation and controls

This cost includes the costs for instrument costs, installation-labour costs, calibration, and expenses for auxiliary equipment and materials required for instrumentation. Total instrumentation costs depend on the amount of control required and may amount to 6 to 30 percent of the purchased costs for all equipment. Depending on the complexity of the instruments and the service, additional charges for installation and accessories may amount to 50 to 70 percent of the purchased cost, with the installation charges being approximately equal to the cost for accessories.

Piping

The cost for piping covers the process pipe, labour, valves, pipe hangers, fittings, pipe, supports, insulation for piping and other items involved in the complete erection of all piping used directly in the process. This kind of cost can be varied depending on the type of the chemical processes, which can be divided into solid process, solid-fluid process, and fluid process. The process plant piping can run as high as 80 percent of purchased equipment cost or 20 percent of fixed capital investment. Material and labour for pipe insulation is estimated to vary from 15 to 25 percent of the total installed cost of the piping.

Electrical installations

The electrical installation consists of four major components, namely, power wiring, lighting, transformation and services, and instrument and control wiring. The cost for electrical installations consists primarily of electrical equipment, materials and labour. In ordinary chemical plants, electrical installations cost amounts to 10 to 15 percent of the value of all purchased equipment. The electrical installation cost is generally estimated between 3 to 10 percent of the fixed-capital investment.

Building (including services)

This cost consists of labour, materials, and supplies involved in all building connected to the plant.

Yard improvements

Yard improvement cost includes costs for fencing, site clearing, grading, road, walkways, railroad, fences, parking area, wharves and piers, and landscaping. Yard improvement cost for chemical process is approximately 10 to 20 percent of the purchased equipment cost or 2 to 5 percent of the fixed capital investment.

Service facilities

Utilities for supplying steam, water, power, compressed air, and fuel are part of the service facilities of an industrial plant. Waste disposal, fire protection, and miscellaneous service items, such as shop, first aid, and cafeteria equipment and facilities, require capital investment, which are included under the general heading of service facilities cost. The total cost for service facilities in chemical plants generally ranges from 30 to 80 percent of the purchased equipment cost.

Land

The land cost for the plants is approximately 4 to 8 percent of the purchased equipment cost or 1 to 2 percent of the total capital investment.

2.12.1.2 Indirect Costs

Engineering and supervision

The cost for engineering and supervision includes the costs for construction design and engineering, drafting, purchasing, accounting, construction and cost engineering, travel, reproductions, communications, and home office expense. Generally and it is approximately 30 percent of the purchased equipment cost or 8 percent of the total direct costs of the process plant.

Construction expenses

The construction or field expense includes temporary construction and operation, construction tools and rentals, home office personnel located at the construction site, construction payroll, travel and living, taxes and insurance, and other construction overhead. For ordinary chemical plant, the construction expense is approximately 10 percent of the total direct cost of the plant.

Contractor’s fee

The contractor’s fee varies for different situations, but it can be estimated to be 2 to 8 percent of the direct plant cost or 1.5 to 6 percent of the fixed capital investment.

Contingency

Contingency factor is usually included in an estimate of capital investment to compensate for unpredictable events, such as storms, floods, strikes, price changes, small design changes, errors in estimation, and other unforeseen expenses. Contingency factor ranging from 5 10 15 percent of the direct and indirect plant costs are commonly used.

2.12.2 Estimation of Total Product Cost

Capital investment is only one part of complete cost estimate. Another equally important part is the estimation of costs for operating the plant and selling the products which can be grouped under the heading of total product cost. This cost can be divided into two categories of manufacturing costs and general expenses. Manufacturing costs are also known as operating or production costs.

2.12.2.1 Manufacturing Costs

The manufacturing costs associated with all expenses directly connected with the manufacturing operation or the physical equipment of the process plant. These expenses can be divided into three classifications as given below.

Direct production cost

This cost includes the expenses directly associated with the manufacturing operation. These types of cost involve expenditures for:

- Raw materials
- Operating labour
- Operating supervision
- Power and Utilities (Steam, electricity, fuel, refrigeration, and water)
- Maintenance and repairs
- Operating supplies
- Laboratory charges

Fixed charges

Fixed charges are the expenses which do not vary with the change in production rate. They involve:

- Depreciation
- Property taxes
- Insurance
- Rent

Plant overhead costs

These costs are similar to the fixed charges in that they do not vary with the change of the production rate. They consist of:

- Medical expenses
- Safety and protection
- General plant overhead
- Payroll overhead
- Packaging
- Restaurant
- Recreation
- Salvage
- Control laboratories
- Plant superintendence
- Storage facilities

2.12.2.2 General Expenses

In addition to the manufacturing cost, the general expenses are involved in every plant's operation. The general expenses may be classified as:

- **Administrative expenses** include costs for executive and clerical wages, office supplies, engineering and legal expenses, upkeep on office buildings, and general communications.

- **Distribution and marketing expenses** are costs incurred in the process of selling and distributing the various products. These costs include expenditure for materials handling, containers, shipping, sales offices, sales man, technical sale services and advertising.
- **Research and development expenses** are the costs for any progressive concern which wishes to remain in a competitive industrial position.

Financing (interest) expenses include the extra costs involved in procuring the money necessary for the capital investments.

2.13 Literature Review of CO_2 Capture From Flue Gases

The literature review is performed on the basis of process flowsheet evaluation, equipment design criterion such as types of column, tray type, type of packing, height of packing, the mass transfer region, solvent selection, solvent circulation rate, solvent degradation, corrosion and other process parameters to optimize the reboiler duty, as well as cost of captured CO_2. Some of the works are summarized below:

Desideri and Paolucci [31] modeled an amine based carbon dioxide removal and liquefaction system using Aspen Plus simulation software for flue gases of conventional natural gas and coal fired 320 MWe power plants. The simulation was performed for absorption/stripping columns together with a liquefaction unit. The heat required for the rich solution regeneration process was provided to the kettle reboiler by steam extracted from the power plant connected to the CO_2 removal plant. The study was performed with an aim to minimize the energy requirements for the process to remove CO_2 when it was integrated in a fossil-fuel-fired power plant. The capital and variable costs of the CO_2 removal and liquefaction system was estimated and their influences on the cost of electricity generation were calculated.

The model described by authors allows the estimation of optimal absorber and stripper performance parameters for low energy consumption. The authors concluded that the goal of 90% CO_2 emissions reduction was, however, penalized by the very high capital costs of the removal plant, which make this solution not feasible at the current price of electricity.

Freguia and Rochelle [49] studied the details of MEA based CO_2 removal model from flue gases with Aspen RateFrac with a purpose to understand how the design variables affect each other at the level of the whole process, not of the single absorber and stripper. Aspen Plus RateFrac module was used to integrate the heat transfer and multi component mass-transfer relationships for a packed absorber and stripper column. The rigorous Electrolyte-NRTL thermodynamics property method was inserted to predict enthalpies, equilibrium vapour pressures of CO_2 and H_2O, and solution speciation. The solvent they used in the model contains about 30 wt. % aqueous monoethanolamine (MEA). The model was adjusted with the laboratory wetted wall column data and field data from a commercial plant. The solvent loop was not closed. It was kept open at the absorber inlet. In this way the lean solvent becomes an input to the model. Sensitivity analyses were performed on solvent rate, column heights, CO_2 removal, heat-stable salt loading and concentration, stripper pressure, and absorber temperature on process variables to find operating conditions at low steam requirement.

Freguia and Rochelle mentioned that the solvent circulation rate affects the steam requirements heavily and recommended operating the process at a solvent rate slightly higher than the optimum**.** The steam consumption was reduced by a greater height of packing in the absorber. The effect is significant with less packing, whereas with a large amount of packing the effect disappears and the reboiler duty reaches an asymptote. For the stripper steam consumption was only reduced slightly by increasing the packing height. The stripper height required is much less than that of the absorber because, due to higher temperature, the reaction rate is not limiting. An increase in CO_2 content in the flue gas and the reduction of CO_2 removal are accompanied by a small reduction in the specific steam consumption.

Freguia and Rochelle showed that he addition of a strong acid to the solvent forms heat stable salts, which can slightly reduce the energy requirement by making the equilibrium relationship more linear. They claim that at a constant solvent rate, the addition of HSS can change steam consumption by as much as 40%.

The author mentioned that the stripper pressure effect was not very significant; therefore, it is possible to operate stripper at any convenient pressure but they found that the optimum pressure is approximately 1.5 to 2 atm.

Singh et al. [50] presented a techno-economic comparison of the performance of two different technologies of flue gas CO_2 scrubbing using MEA and O_2/CO_2 recycle combustion combined with a low temperature flash (LTF) unit to the retrofit of a typical 400 MW_e coal fired power plant. Commercially available process simulation packages such as Hysys and Aspen Plus were used to develop the model. The simulation was conducted with a goal to capture 90% CO_2 from the flue gas of pulverised coal fired boiler that burns western sub-bituminous coal producing 400 $MW_{e.}$ Both capital and operating costs were evaluated for the two technologies investigated using a variety of sources such as vendor input, published and personal sources, and an engineering sizing and costing software package such as Icarus Process Evaluator.

Singh et al. considered in their study that the power output to the grid must remain fixed and a considerable amount of supplementary energy was supplied to either CO_2 separation processes generated by gas turbine combined cycles (GTCCs), gas turbines and steam boilers The flue gas was treated in four separate absorption/regeneration column trains to eliminate the structural. The author used the following key operating conditions of the MEA capture process:

Absorber pressure 1.2 bar, absorber feed stream temperature 40^oC, stripper pressure 2.0 bar, stripper condenser temperature 40^oC.

Singh et al. concluded that both processes were expensive; however, O_2/CO_2 appeared to be a more attractive retrofit option. The CO_2 capture cost for the amine case was $55/ton of CO_2 avoided (3.3 ¢/kWh) and that for the O_2/CO_2 case was lower at $35/ton of CO_2 avoided, (2.4 ¢/kWh). This represented an increase of 20–30% in current electricity prices. This study showed that the O_2/CO_2 case was less expensive than MEA scrubbing.

Alie et al. [51] studied the amine capture process for a 500 MW power plant by simulating this process using Aspen Plus simulation program. The simulation converged using a decomposition method presented in Alie's paper. The specification for this work was the 85% carbon dioxide recovery and 98% mole purity of carbon dioxide product stream. The reboiler duty of the regenerator was considered because it has the most impact on the operating cost of this process. The decomposition method was divided into three phases. The first was to simulate the stand-alone absorber and vary α_{lean} (mole ratio of carbon dioxide to MEA in lean-solvent stream). Then the conditions of the solvent stream were investigated. The second phase was to simulate the stand alone regenerator using the results from phase I as the input data. The reboiler duty was compared between each case to find the best condition that give the minimum reboiler duty. The results from decomposition method were compared with the coupled process flowsheet in the third phase.

Chakma et al. [52] studied the CO_2 separation for low pressure flue gases from coal fired power plants. The study was carried out using HYSIM process simulator for the MEA capture process for a $1x10^7$ m^3 STP/day of flue gas containing 15% of CO_2. They studied the sensitivity of the effect of inlet gas composition, feed gas pressure, reboiler energy inputs, number of stages and type of solvent, solvent concentration on the cost of capturing CO_2. A consistent set of restrictions were imposed for the simulation. The simulation was designed for a recovery of 95% carbon dioxide. The CO_2/amine loading was also restricted to reduce corrosion potentials. The maximum reboiler temperature was set at 122^oC to prevent thermal degradation of solvent. The top stage temperature of the stripper was set at 50^oC to minimize the solvent losses due to vaporization. To minimize the pressure drop in the absorber and stripper, valve trays with venturi openings were selected for both columns. The steam for the reboiler and pumps was assumed to be provided by the boilers in the power plants.

The authors examined the effect of absorber pressure on the cost of CO_2 removal by varying the absorber pressure from 1.15 to 2 atm and reported that there is no cost benefit in operating the absorber column at higher pressure. They recommended that the absorber should be operated at as low a pressure as possible.

The authors studied the effect of the solvent flow rate on the cost of separation and reported that the cost of CO_2 separation increases almost exponentially with solvent flow rate and recommended that the solvent flow rate should be kept at a minimum. On the point of MEA concentration, the authors mentioned that an increase in MEA concentration decreases the cost of separation substantially but the possibilities of corrosion and solvent degradation must be taken into consideration in selecting the concentration level of the solvent.

Chakma et al. studied the effect of the number of absorber and stripper trays on the cost of separation and mentioned that an increase in the number of trays in absorber as well as in stripper reduces the overall cost of CO_2 capture. In the absorber, the higher number of trays allows lower solution circulation rates thus reducing the overall costs. In the case of stripper, an increase in the number of trays lowers the reflux ratio for a given separation duty and thereby in the reduction of separation costs.

The authors carried out the simulation runs with different solvents like Monoethanolamine (MEA), Diisopropanolamine (DIPA), Diethanolamine (DEA) Diglycolamine (DGA), Methyldiethanolamine (MDEA) and Trieethanolamine (TEA) to compare their cost effectiveness in CO_2 removal from power plan flue gases and concluded that from a cost effectiveness point of view, none of the commonly available gas treating alkanoamine solvents can compete with 20 or 30 wt% MEA.

Barchas, R and Davis, R [53] studied the recovery of carbon dioxide from oxygen-containing gases and discussed the design aspects of a recovery carbon dioxide unit. The main feature of the Kerr-McGee/Lummus Crest technology is its capability to operate with fuels ranging from natural gas to high-sulfur coal or coke. The technology is also suitable for flue gas that contains oxygen or a small amount of sulfur dioxide. Figure 2-6 shows the basic flow scheme of the Kerr-McGee/Lummus Crest carbon dioxide recovery technology. The feed gas is treated from a desulfurization system, water removal, and cooling system before being sent to an amine absorber where it is scrubbed with a 15-20 weight percent of MEA solution. Approximately 90 percent of carbon dioxide in the feed gas is recovered. A new energy saving unit can be installed as shown in Figure 2-7. An additional flash tank is installed. Once heated in a solution exchanger to a higher temperature, the rich solution is then flashed in a flash tank. From this, equipment sizes are smaller except for the solution exchanger. And the steam consumption can be reduced, including the production cost.

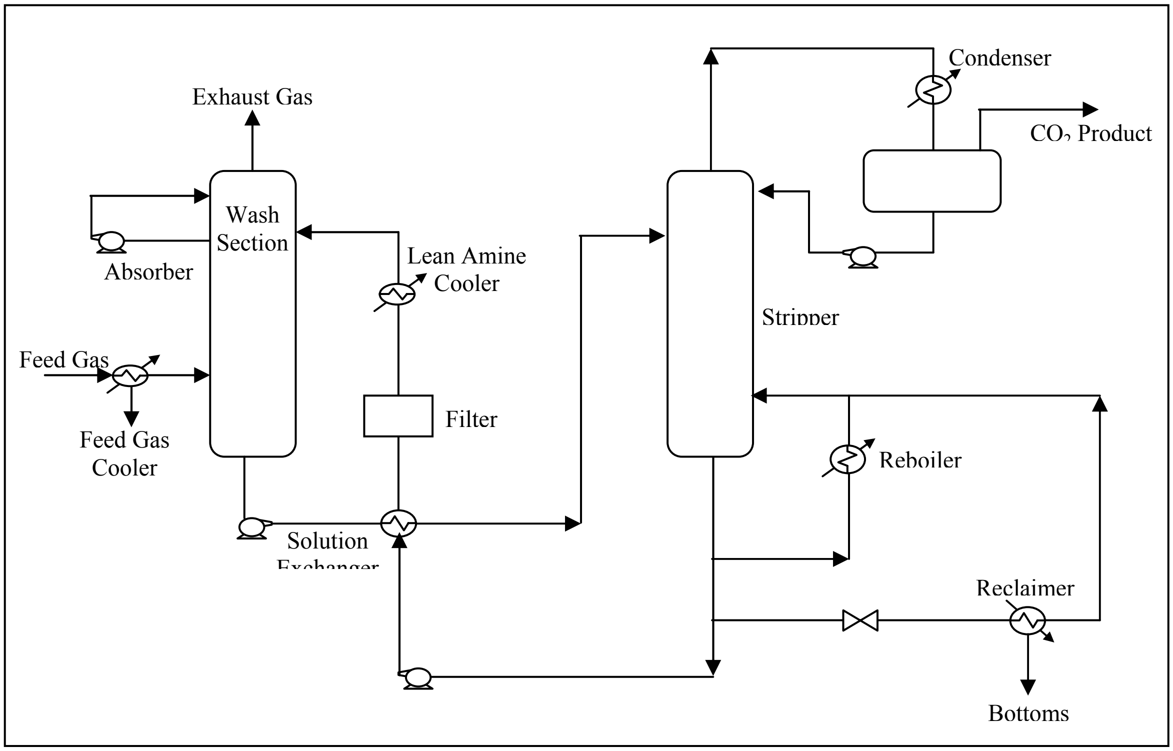

Figure 2-6: Kerr McGee/ABB Lummus Crest CO_2 Recovery Technology Basic Flow

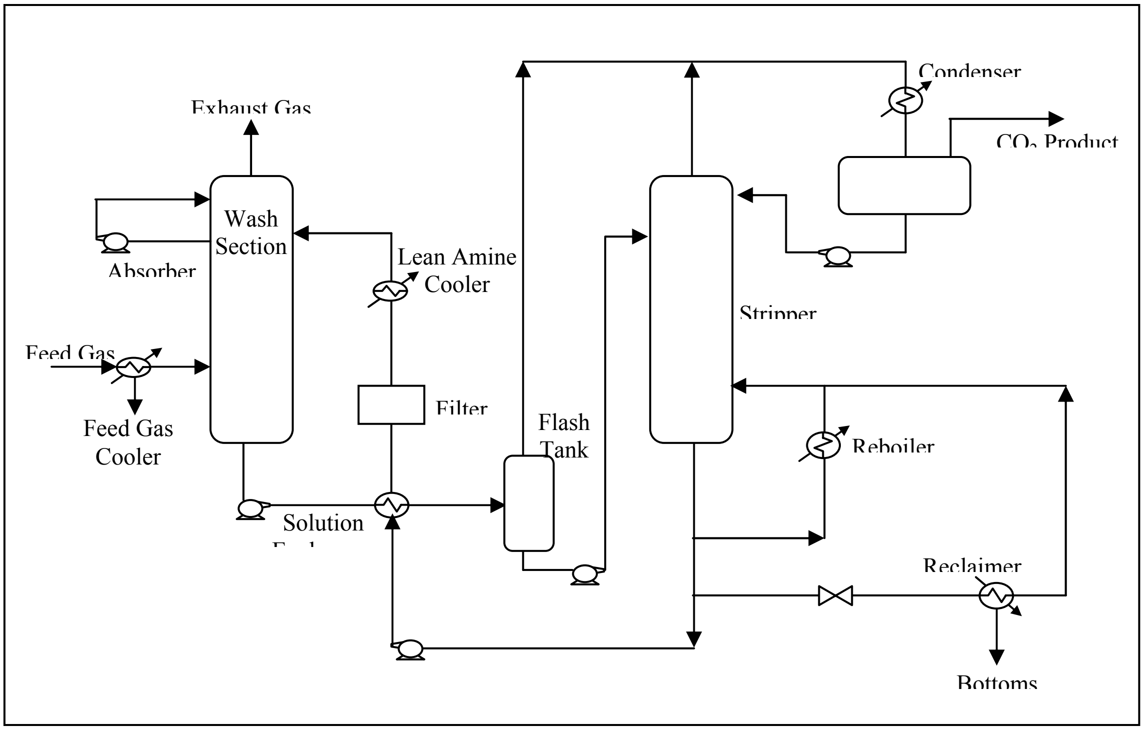

Figure 2-7: Kerr McGee/ABB Lummus Crest CO_2 Recovery Technology Energy Saving Design

Jou et al. [54] quantified the solubility of CO_2 in 30 mass % monoethanolamine (MEA) solutions at eight different temperatures of 0, 25, 40, 60, 80, 100, 120, and 150°C at CO_2 partial pressures of ranging from 0.0001 to 20,000 kPa. This work was undertaken to rationalize the data available and to produce a body of data over a wide range of temperatures and pressures, which could serve as a standard for use in the modeling of vapour-liquid equilibrium in MEA solutions.

Wilson et al. [55] provided details for the MEA based CO_2 extraction pilot plant test facilities adjacent to SaskPower's Boundary Dam coal fired power station (BDPS) and discussed the results obtained in terms of the absorption performance, or mass-transfer efficiency of the process, energy consumption for solvent regeneration, operational problems such as solvent degradation, levels of heat stable salts as well as corrosion in the CO_2 plant. The pilot plant has a capacity to handle $14x10^3$ m^3/day flue gases and captures up to 4 tonnes of CO_2/day. The amine extraction unit consisting essentially of 18" absorber and 16" regenerator treats the pre-conditioned flue gas from the Anderson 2000 unit. The process operation is based on the Fluor's Econamine FG^{SM} technology.

Wilson et al. presented some valuable results both in design and operational aspects of absorption performance and solvent regeneration. The authors in their paper reported that CO_2 absorption occurs mostly at the lower portion of the absorber. The lean CO_2 loading and absorption efficiency tend to stabilize at a certain value of heat duty, which could be considered to be the optimal operating condition. They identified three major heat stable salts (i.e., sulphate, oxalate and thiocyanate) in the context of solvent degradation.

Veawab et al. [56] focused on the evaluation of absorption performance and energy requirement for solvent regeneration of three common alkanolamines including monoethanolamine (MEA), diethanolamine (DEA), and methyldiethanolamine (MDEA). The primary amine, MEA, is the most attractive solvent for CO_2 capture because it reacts with CO_2 at a faster rate and the cost of raw materials is relatively low compared to the secondary and tertiary amines. However, the associated high capital and operating costs of absorption process using MEA guided to identify the alternative solvents that lead to a reduced cost compared to the cost of MEA process.

Veawab et al. reported that, based on the CO_2 capture efficiency, the absorption performance of the test solutions was in the following order: MEA > DEA > MDEA. MEA required the shortest column height while the MDEA required the tallest. The energy requirement for solvent regeneration of different amines was reported in the following order: MEA > DEA > MDEA. MEA is more difficult to regenerate, thus resulting in higher residual CO_2 loading of lean solution.

Tanthapanichakoon and Veawab [57] investigated the effect of heat-stable salts on corrosion in the amine treating units using an aqueous solution of 5 $kmol/m^3$ monoethanolamine (MEA) at 80^oC and under atmospheric pressure and CO_2 loading of 0.20 mol/mol. Carbon steel 1018 was used as a representative of the typical construction material of process equipment.

Heat-stable salts in amine treating units are of great concern for corrosion because the presence of heat-stable salts in MEA solution led to increased corrosion. The authors reported that the heat-stable salts, which were tested in this study, were acetate, chloride, formate, glycolate, malonate, oxalate, succinate and sulfate. The experiments showed that heat stable salts were the major contributor to the increased corrosion of carbon steel 1018 and oxalate. None of the test heat stable salts showed pitting tendency on either carbon steel or stainless steel.

Veawab [58] discussed the impacts of operating parameters on corrosion and corrosion controls in the CO_2 capture process using reactive amine solvents. The author reported that an increase in CO_2 loading intensifies the system's aggressiveness and accelerate the corrosion process. The rich amine solution is generally more corrosive than the lean amine and the recommended maximum CO_2 loadings in the rich amine solution are in the range of 0.25-0.40 mol/mol for MEA.

The author mentioned that higher temperature tends to increase the corrosion rate. The most susceptible process areas to corrosion are the rich-lean heat exchanger, the regenerator and the reboiler where the amine solutions are heated to an elevated temperature. An increase in amine concentration also causes a higher corrosion rate although high amine concentration is desirable for energy saving purposes. Solution velocity can cause severe erosion corrosion, especially in the presence of solid contaminants.

The author discussed that corrosion can be inhibited through the addition of corrosion inhibitors. Inorganic corrosion inhibitors are in practice more favoured than the organic compounds because of their superior inhibition performance. Among these, vanadium compounds, particularly sodium metavanadate ($NaVO_3$), are the most extensively and successfully used in the amine treating plants. However, there is a concern over the use of toxic inorganic inhibitors as it makes disposal of the resulting industrial waste difficult and costly.

Wilson, M. A. et al. [59] studied the MEA-based CO_2 extraction pilot plant performance of a pulverized coal power plant in Saskatchewan, Canada, to test the effectiveness of pre-conditioning the flue gases for sulphur oxide removal to reduce amine degradation. Dow Chemicals U.S.A. and Union Carbide Corporation supplied the patented amines used in the pilot plant for CO_2 extraction. In addition, the effectiveness of the Andersen 2000 sulphur dioxide scrubbing process was also measured.

The Amine Extraction Unit was used to test the Union Carbide "Flue Guard" technology and the Dow Chemical "FT" technology. Union Carbide "Flue Guard" technology provided the stabilizer, for increasing the oxygen tolerance of the amine, and corrosion inhibitor separately. Corrosion in the amine absorber and regenerator towers was generally minimal, indicating the effectiveness of the corrosion inhibitors. The stabilizer also worked effectively under adverse conditions to keep the loss of amine to oxygen degradation very low. On the other hand, the Dow Chemical FT solvent arrived with the corrosion inhibitor and stabilizer pre-mixed. The operating conditions were kept more or less the same as during the Union Carbide test to allow direct comparison of the two chemicals. The results from using the Dow solvent over the test period were much the same as for the Union Carbide technology, with similar CO_2 capture, thermal, and product purity efficiencies.

The author reported that the most significant differences noted between the Union Carbide and Dow solvents were in the levels of corrosion and in amine degradation due to oxygen. The Dow Chemical solvent exhibited slightly higher rates of corrosion, although corrosion was generally minimal for both technologies, and higher degradation rates. Since the Dow Chemicals amine came premixed, maintaining appropriate concentrations of the corrosion inhibitors and stabilizers was more problematic.

Mimura et al. [60] studied the flue gas CO_2 recovery technology and focused on the reduction of the initial and operating costs. The joint research done by Kansai Electric Power Co (KEPCO) Inc. and Mitsubishi Heavy Industries (MHI), Ltd emphasized on increase of flue gas velocity in the CO_2 absorber, reduction of the KP-1 packing weight, and reduction of amine consumption. An energy efficient solvent was developed and commercialized from the results of this research and development work.

Mimura et al. reported that KEPCO and MHI jointly developed very low-pressure loss and very high efficient gas and liquid contact KP-1 packing for atmospheric flue gas conditions. KP-1 was designed with a packing diameter of 320 mm and a flue gas velocity of 1.92 m/s. Under normal operating conditions, a pressure loss of 40 mm H_2O was found through the CO_2 absorption part (KP-1 packing part). The authors reported that the KP-1 packing was able to operate without flooding at 950 m^3/h with a flue gas velocity of 3.29 m/s.

Mimura et al. reported that KEPCO and MHI developed the KS-1 solvent and its amine consumption was around 0.35 kg/tonne CO_2 recovered, which was about 1/6 of the MEA solvent consumption. Degradation of MEA can be as high as 2 kg/tonne of CO_2 recovered. Detailed investigations identified that a combination of decomposition, vapour loss and mist loss were responsible for the amine losses. However, if vapour loss and mist loss were reduced, KS-1 loss could be reduced to between 0.1~0.2 kg/tonne CO_2 recovered.

Mimura et al. [61] in their follow up paper discussed the pilot plant performance they developed for the new energy efficient solvent KS-2. KS-2 and KS-1 have similar energy efficiency and both require 20% less energy than MEA but KS-2 is more stable than KS-1 and is a more efficient absorbent for low CO_2 content flue gas.

Mimura et al. selected promising absorbents from 80 alkanolamine specimens that have been tested in the laboratory. Subsequent tests were done in a pilot plant and showed excellent characteristics for KS-1. KS-2 was developed after KS-1, and the molecular structure of KS-2 has a higher degree of hindrance and high CO_2 absorption capacity under flue gas condition and very low CO_2 loading under regeneration condition compared to MEA and KS-1. The regeneration energy required by KS-2 absorbent during 1000 hours continuous running tests is about 700 kcal/kg-CO_2 recovered which is about 20% less than the MEA solution. Based on the heat stable salt analysis, KS-2 is very stable and may not be degraded by itself. Power plant power reduction is about 5~6% if KS-2 absorbent and KP-1 packing are used, with optimum steam in natural gas-fired power plant.

Aroonwilas et al. [62] evaluated the performance of the structured packing in terms of the volumetric overall mass-transfer coefficient (K_Ga_v) as a function of the process operating parameters including gas load, CO_2 partial pressure, liquid load, liquid temperature, solvent concentration, solvent type, and structural packing types. Three types of experiments consisting of laboratory-scale, pilot-scale, and industrial-scale trials were carried out and in each experiment different types of structured packings were used. Comparisons of CO_2 absorption performance between the tested structured and common random packings were performed. Structured packings showed excellent mass transfer performance with a lower pressure drop in comparison with random packings for the absorber. This result is due to the fact that structured packings possess a considerably higher geometric surface area per unit volume in comparison with those of random packings. Despite the superior performance of the structured packings, the cost of structured packing is generally higher than that of random ones.

Goff and Rochelle [63] studied the oxidative degradation of monoethanolamine (MEA) at conditions typical of an absorber to capture CO_2 from flue gas. Experiments were conducted in a heated, sparged reactor by bubbling air, containing varying concentrations of CO_2, through the amine solution. Degradation rates were measured as a function of iron and copper concentrations at 55°C using 7-M MEA solutions with CO_2 loadings of 0.4 and 0.15, corresponding to the conditions at the top and bottom of the absorber. Solutions with lower CO_2 loadings degraded faster than those with high loadings. The oxidative degradation was catalyzed by the presence of various multivalent cations such as iron, copper, nickel, and chromium. This study quantified the effect of varying iron and copper concentrations on the degradation rates. Dissolved iron is always present in these systems as a corrosion product; copper is a known corrosion inhibitor in MEA CO_2 removal systems, and is often added to the solutions as a copper (II) salt. Finally, this study also looked at the effect of various chelating agents and corrosion inhibitors as possible inhibitors to the oxidative degradation rates.

The authors reported that the degradation rate of solutions with high CO_2 loading increased as the concentration of dissolved iron was increased. The addition of copper further catalyzed the degradation rates of the solution. Several inhibitors and chelating agents, including (ethylenedinitrilo)-tetraacetic acid (EDTA), N-dihydroxyethyl glycine (bicine), methyldiethanolamine and phosphate were tested as possible degradation inhibitors.

Both iron and copper catalyzed the oxidative degradation of MEA, but the rate was less than first order with respect to the metal concentrations. Adding copper to a system that already contains iron will result in an increase in the degradation rate regardless of the CO_2 loading. Degradation catalyzed by either copper or iron appeared to be of the same order of magnitude. EDTA was more effective at inhibiting the degradation catalyzed by copper than iron.

Marion et al. [64] conducted a comprehensive study evaluating the technical feasibility and economics of alternate CO_2 capture for an existing coal fired 450 MW_e power plant in Conesville, Ohio. Three technology concepts were being evaluated, namely:

- **Concept A:** Coal combustion in air, followed by Kerr-McGee/ABB Lummus Global commercial MEA-based CO_2 capture process
- **Concept B:** O_2/CO_2 recycle combustion
- **Concept C:** Coal combustion in air with catalytic O_2 removal and CO_2 capture by tertiary amines MEA/MDEA process

Marion et al. reported that in Concept A, a flue gas desulfurization unit was installed downstream and integrated into the power plant to strip SO_2 from the 15% CO_2 effluent gas stream. SO_2 content was reduced to about 10 ppmv as required by the downstream amine system. Flue gases were cooled in a direct contact cooler and introduced to the MEA system where more than 96% of CO_2 was removed. The researchers concluded that

5 acres of land would be needed for the capture plant. The captured CO_2 liquid product characteristics were reported CO_2 = 99.95 vol. %; N_2 = 0.05 vol. %; temperature = 28^oC and pressure = 2000 psig.

Marion et al. mentioned that for Concept B, the flue gas stream leaving the flue gas desulfurization unit was cooled down to 38^oC in a direct contact cooler and the flue gas leaving the cooler was split into two streams with about two third recycled back to the boiler and the remaining one third feeding the CO_2 compression and liquefaction system. The plot plan required about 3 acres of land for the CO_2 liquefaction and compression and direct contact cooling systems. The captured CO_2 liquid product characteristics were reported CO_2 = 97.8 vol. %; N_2 = 1.2 vol. %; SO_2 = 215 ppmv; O_2 = 9300 ppmv; temperature = 28^oC and pressure = 2000 psig.

Marion et al. described in Concept C an ABB designed process comprised of an optimized mixture of MEA and MDEA; installed downstream of the flue gas desulfurization unit and integrated into the power plant to strip CO_2 from the effluent gas stream. A De-Oxy catalyst was used upstream of the solvent contractor. This system recovers about 91% of the CO_2. The plot plan required for this was about 7 acres of land. The captured CO_2 liquid product characteristics were reported CO_2 = 99.97 vol. %; N_2 = 0.03 vol. %; temperature = 28^oC and pressure = 2000 psig.

Nobuo Imai [65] from Mitsubishi Heavy Industries compares the performance of KS-1 to MEA. He showed that KS-1 has improved performance relative to MEA in terms of heat of dissociation of rich-MEA, higher CO_2 solubility, lower corrosiveness, and lower degradation rate. He also discussed the cost estimates for the process.

Chapel D. et al. [66] studied carbon dioxide sources for industrial uses and considered carbon dioxide from flue gas as an important source. So the economics of carbon dioxide capture process from flue gas was studied. The capital and operating cost for carbon dioxide capture plant was determined using cost factors. They reported the main cost factors for a 1000 ton/day of carbon dioxide. From the calculation, the operating cost for carbon dioxide capture process is between 10.83 to 17.85 US $ / ton which depend on the flue gas sources. Moreover they reported that the capital cost for a large single-train plant would be lower compared to that with a multi-trains plant. They also reported that the 4600 ton/day of carbon dioxide single train plants are possible.

Rao and Rubin [67] studied the impacts of carbon capture and sequestration technology on power plant performance, emissions, and economics. Their objective was to develop a preliminary model of performance and cost of amine-based CO_2 capture process based on available information by taking into account the uncertainties and variability in key performance and cost parameters and apply that model to study the feasibility and cost of carbon capture and sequestration at existing coal-based power plants as well as new facilities.

They developed a mathematical model to simulate the performance of CO_2 capture and storage system based on amine (MEA) scrubbing and incorporated that CO_2 module to an existing coal-based power plant simulation model. They considered two types of input parameters to the CO_2 performance model:

***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 a MEA-based absorption system with pipeline transport of liquefied CO_2 to a geologic sequestration site. 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.

The authors also developed the *Process Cost Model* that directly linked the process performance model to the model used to calculate four types of cost based on available data such as capital cost, operating and maintenance costs, cost of electricity, and cost of CO_2 Avoided.

The authors showed that for amine (MEA)-based absorption systems applied to coal-fired power plants, the cost of carbon avoidance depended 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. The presence of acid gas impurities such as 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.

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.

Slater et al. [68] studied the cost of separating CO_2 from multiple flue gas sources at one of their petrochemical complexes, using Fluor's proprietary Econamine FG^{SM} amine capture technology. The recovery plant capacity was 3820 tonnes/day, representing most of the CO_2 emissions from the complex. The authors discussed in their paper the cost penalties associated with capturing CO_2 from a large number of flue gas sources and also compared the cost of the capture facility with facilities designed with either a single flue gas source or only a few selected sources.

Slater et al. mentioned some useful design points in their study. The recovery plant was designed to recover about 90% of the CO_2 from the flue gases. The product CO_2 was dehydrated and compressed to 220 bara. The reboilers only required LP steam (nominal 3.4 barg (50 psig)). MP steam (62barg) was generated in utility boilers and used to provide energy for the CO_2 capture plant. The required steam for compressors and blowers was used directly. The remaining steam was expanded to 3.4 barg through a backpressure turbine that drives an electric generator. CO_2 compression was achieved through a single steam turbine driven three-body five-stage compressor. Dehydration was accomplished through a propriety glycerol-based Shell Global Solutions. The plant described above resulted in an overall plant installed cost of $160 million.

Slater et al. reported that only a small cost penalty associated with multiple flue gas sources for a CO_2 recovery plant. The largest factor affecting plant cost with multiple flue gas sources was the wide separation between groups of sources that required separate flue gas processing trains.

Simmonds et al. [69] studied the cost and practicability of capturing CO_2 from the site using today's best available technology. The study provided a firm basis for comparison with other technologies and designs a feasible process and cost estimation and discussed the environmental impacts, the benefits and the challenges as well. The authors examined the issues of retrofitting a very large scale; post combustion, amine based capture facility using Econamine FG^{SM} process. The capture plant was designed to produce a high purity, high-pressure (220 bar) stream of 6,000 tonnes per day CO_2 (equivalent to 2.0 million tonnes per annum), suitable for use with Enhanced Oil recovery.

Simmonds et al. reported that the Econamine FG^{SM} required SO_2 and NO_x at or below 10 ppm and 20 ppm respectively to avoid excessive degradation of the solvent. An optimal plant-ducting layout would require 2 km of ducting for the flue gas. 25 MW was required for the flue gas blower of which 15 MW of this power was required to overcome the pressure drop of this ducting. The authors reported that their design called for four absorbers of 10.3 m in diameter and a stripper of 10.4 m in diameter. They reported that there were no technical restrictions on building columns of this size. The authors anticipated a cost of CO_2 capture of between US$ 50 and US$ 60 per tonne of CO_2.

Chapter 3: Process Simulation

3.1 Flue Gas Analysis for Process Simulation

The Aspen Plus™ [72] program requires the input data of the flue gas stream which are temperature, pressure, flow rate, and flue gas composition. To simulate the MEA-based CO_2 absorption process flue gas data from St. Marys Cement plant were used. The objective was to capture 85% of the carbon dioxide at a purity of 98%. St. Marys Cement plant burns mixtures of pet coke, coal and coal fines as fuel in the kiln furnace for combustion. The flue gas in cement plant comes from the combustion process as well as from the kiln reaction itself. The flue gas flow rate, composition, temperature and pressure for the process simulation are analysed on the basis of three cases based on the daily production data from St. Marys Cement Plant:

- **Case I:** the plant operates at the highest capacity
- **Case II:** the plant operates at average load
- **Case III:** the plant operates at minimum operating capacity

Table 3-1: Flue Gas Analysis from St. Marys Cement Plant for Process Simulation

	Case I	Case II	Case III
Temperature, ^oC	160	160	160
Pressure, bar	1.013	1.013	1.013
Mole Flow, kmol/hr	9851	8162	5629
Mass Flow , kg/hr	304996	252711	174283
Volume Flow, m^3/hr	350×10^3	290×10^3	200×10^3
Mass Flow, kg/hr			
H_2O	12831	10631	7332
CO_2	96956	80335	55404
N_2	187824	155626	107328
O_2	7385	6119	4220
Mass Fraction			
H_2O	0.042	0.042	0.042
CO_2	0.318	0.318	0.318
N_2	0.616	0.616	0.616
O_2	0.024	0.024	0.024
Mole Fraction			
H_2O	0.072	0.072	0.072
CO_2	0.224	0.224	0.224
N_2	0.681	0.681	0.681
O_2	0.023	0.023	0.023

3.2 Assumptions for The Simulation Process

The following assumptions are made prior to the development of the MEA based CO_2 absorption simulation process:

- To reduce the complexity of the Aspen Plus™ simulation process, the baghouse to screen out the particulates and the flue gas desulfurization to scrub sulphur dioxide are not included in the simulation process. The inlet flue gas entering the blower is considered free from all contaminants and consists primarily of CO_2, O_2, N_2, and H_2O. No NO_x and SO_2 exist in the flue gas.
- MEA is guarded with additives to tolerate the presence of O_2 in the flue gas.
- Corrosion is not taken into account in the simulation.
- MEA solvent concentration is 30 percent by weight.
- Overall carbon dioxide recovery considered ranges from 55% to 95%.
- The product carbon dioxide stream purity is specified at 98% by mole and compressed to 150 bar pressure for discharge.
- For simplicity of the simulation dryer and transportation are not included in the Aspen simulation process.

3.3 Development of An Aspen Process Flowsheet

The Aspen Plus™ process flowsheet developed using the assumptions for the simulation is shown in Figure 3-1. Detailed flowsheet explanation is described next:

3.3.1 Block Specifications

The flowsheet consists of the following blocks:

SEP

SEP separates the feed into two outlet streams, using rigorous vapour-liquid or vapour-liquid-liquid equilibrium. It was modeled with **FLASH2** unit operation model in Aspen Plus™ [72] simulation. It is used in conjunction with the unit COOLER, described later on.

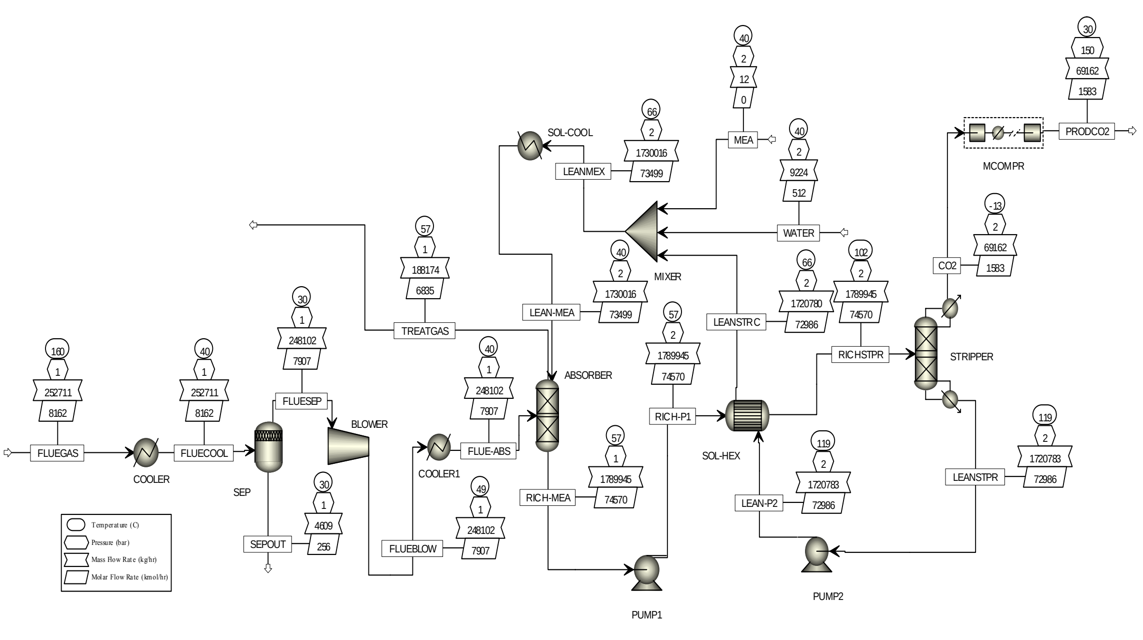

Figure 3-1: Aspen Plus™ Process flowsheet diagram for CO_2 capture process

BLOWER

The absorption column operates at pressures slightly above the atmospheric pressure (1.2 bar) and the flue gas needs to be compressed using a blower. The BLOWER is modeled using the **COMPR** unit operation model in Aspen Plus™ [72]. The BLOWER changes the stream pressure to compensate for pressure drop in the absorber. It represented as a single stage compressor. Because of the extremely large flow rate of flue gas, the compressor duty is high and sensitive to the pressure to be compressed at. It is well-known that the absorption rate increases with partial pressure of CO_2. Therefore, increase in feed pressure should increase the mass transfer rate in the absorber. But there is a penalty in the form of blower power associated with this approach.

COOLER / COOLER1 / SOL-COOL

Determines thermal and phase conditions of outlet stream and is modeled with the **HEATER** unit operation model in Aspen Plus™ [72]. **COOLER** is used to cool down the flue gas temperature to 40°C before entering the blower and some water is removed from the flue gas using the SEP block. By reducing the flue gas temperature to 40°C, the blower power consumption can be reduced significantly. **COOLER1** is used to cool down the stream leaving the blower to 40°C before it enters the absorber because the absorption reaction between MEA and carbon dioxide is favoured at lower temperatures (typically < 40°C). Lower temperatures also decrease solvent degradation and corrosion rate [52]. The **SOL-COOL** unit cools down the lean MEA solvent to the desired absorber inlet temperature of 40°C. It was observed that from 20°C to 37°C, increasing the temperature increased CO_2 absorption due to an increase in the rate of reaction between CO_2 and MEA but from 40°C to 65°C increasing the temperature decreased CO_2 absorption because the Henry's constant increases with increasing temperature [62].

PUMP1 and PUMP2

Both PUMP1 and PUMP2 are modelled with the **PUMP** unit operation model in Aspen Plus™ [72] and change stream pressure to maintain the desired pressure when any pressure drop is observed. Due to low pressure of the absorber, PUMP1 is needed to transfer the rich solution to the STRIPPER (1.9 bar). PUMP2 is used to compensate for the pressure drop in the unit SOL-HEX, MIXER, SOL-COOL, and ABSORBER for the LEANSTPR stream. By default, PUMP calculates power requirement using efficiency curves for water in a centrifugal pump.

SOL-HEX

SOL-HEX is implemented in Aspen Plus™ [72] using the **HEATX** unit operation model and it exchanges heat between the hot stream from the bottom of the stripper and the cold stream from the bottom of the absorber. SOL-HEX acts as a counter current heat exchanger where the LEANSTPR stream is used to heat up the RICH-P1 stream before going to the stripper. SOL-HEX increases the cold stream RICH-P1 temperature to a maximum of 45°C by using the heat from hot stream LEANSTPR.

MCOMPR

MCOMPR represents a multistage compressor and is used to compress the final CO_2 product. MCOMPR is modelled with the **MCOMPR** unit operation model in Aspen Plus™ [72] and changes stream pressure across multiple stages with intercoolers. It is, therefore, composed of a series of COMPR blocks with inter-cooling stages. The CO_2 compressor is required to compress the product CO_2 for transportation via pipeline. This block requires that the outlet pressure, compression ratio, isentropic efficiency, and interstage temperatures be specified. The outlet pressure is determined by the length of the pipeline, the location of the booster compressors and the ultimate end use of the product CO_2.

MIXER

MIXER combines multiple streams into one single stream and is implemented with the **MIXER** unit operation model in Aspen Plus™ [72]. Make up MEA and WATER is mixed with LEANSTRC stream to compensate for MEA and WATER losses.

ABSORBER and STRIPPER

ABSORBER and STRIPPER are modeled with the **RateFrac™** unit operation model in Aspen Plus™ [72] that performs rigorous rating and design for single and multiple columns. **RateFrac™** is a rate-based non-equilibrium model for simulating all types of multistage vapour-liquid fractionation operations by simulating actual tray and packed columns, rather than idealized representation of equilibrium stages. **RateFrac™** explicitly accounts for the underlying inter-phase mass and heat transfer processes to determine the degree of separation. It does not use empirical factors such as efficiencies and the Height Equivalent to a Theoretical Plate (HETP). The use of **RateFrac™** completely avoids the need for efficiencies in tray columns or HETPs in packed columns. **RateFrac™** [72] directly includes mass and heat transfer rate processes in the system of equations representing the operation of separation process units. It has far greater predictive capabilities than the conventional equilibrium model.

There is no condenser and reboiler in the ABSORBER column. A significant amount of CO_2 is recycled through the process and the most important factor is the amount of flue gas and solvent that flows through the column.

The purpose of modeling the STRIPPER is to minimize the reboiler heat duty. Thermal degradation of the MEA solvent due to high reboiler temperature is one of the major limitations in the stripper column. Another important operating parameter for the STRIPPER is its CO_2 recovery. A low CO_2 recovery causes a huge amount of CO_2 recirculation throughout the columns and subsequently increases the equipment size and the reboiler duty due to a large amount of material to heat. On the other hand, with a high

CO_2 recovery much less material is recirculated, but the reboiler duty increases to achieve high CO_2 separation. There is therefore an optimized CO_2 recovery in the stripper that would minimize the reboiler duty.

ABSORBER and STRIPPER need to be specified on the basis of column configurations, column type, internal geometry, and column pressure.

In ABSORBER the inlets and outlets are connected to the top and bottom of the column. The flue gas enters at the bottom of the column and the LEAN-MEA enters at the top of the column. The LEAN-MEA flows back from the stripping column after the amine regeneration process. The STRIPPER is composed of partial vapour condenser and a kettle reboiler. The feed enters the column above the mass transfer region. Inside the stripper, two design specifications are specified. The first one is to achieve the desired mass flow of CO_2 in the distillate (commonly 85% of the CO_2 in the flue gas) by varying the bottoms to feed ratio at the bottom of the stripper. The other specification is the mole purity of carbon dioxide (typically 98% CO_2 purity) in the product stream by varying the molar reflux ratio at the top of the stripper.

RateFrac™ has built-in routines for bubble cap and valve tray. The diameter of the Absorber and Stripper column needs not to be specified because it is an output of the model. The total number of trays of the column needs to be specified as well as the column diameter estimate. The default value for entrainment flooding 80%, tray spacing 24 inches, and weir height 2 inches are used to completely specify the tray geometry.

The ABSORBER and STRIPPER are specified with constant pressures through out the columns.

3.3.2 Stream Specifications

FLUE-GAS

The flue gas for CO_2 capture process is received at atmospheric pressure and a temperature of 160°C. The flue gas composition for the CO_2 capture process is composed of CO_2, N_2, H_2O, and O_2. The CO_2 molar composition in the flue gas is 22.4 percent. Ar, NO, CO, SO_2, and H_2 are not included in the flue gas for MEA based CO_2 absorption process. One of the drawbacks of an amine plant is the MEA's low tolerance to SO_2 in the flue gas. The flue gas is cooled down to 40°C before it enters the blower for pressurization. The water from the flue gas must be removed before entering the blower.

MEA and WATER

Make up MEA and WATER is added to the process to compensate for the loss from the top of the Absorber and from the Stripper distillate. The flow rates of MEA and water are calculated immediately prior to Mixer execution, therefore any initial values suffice. Make up MEA and WATER is added to the mixer at a temperature of 40°C and a pressure of 2.1 bar.

3.3.3 Property Specifications

The ELECNRTL property method is used for developing the Aspen Plus™ simulation model with the user interface because this feature assists in adding any missing ionic components, defining the solution chemistry, retrieving binary interaction parameters, and inputting parameters for equilibrium constants. The ELECTROLYTE-NRTL (or ELECNRTL) property method is used to accommodate interactions with ions in solution. NRTL models non-ideal liquid solutions for ionic interactions in the solution. Aspen Plus™ [72] physical property system contains binary and pair interaction parameters and chemical equilibrium constants for systems containing CO_2, H_2S, MEA, and H_2O with temperatures up to 120°C and amine concentrations up to 50 wt.%.

Aspen Plus™ provides special data packages which are made specifically for a defined process. One of these packages contains components and property definitions specifically defined for amine gas treating processes at specified conditions. The **"emea"** insert package is suitable for the simulation that deals with monoethanolamine (MEA) and acid gases. An **"emea"** package is inserted in the Aspen Plus™ property set prior to the development of the entire flowsheet. The "emea" package provides the thermodynamic properties for the H_2O, MEA, H_2S, and CO_2 system [72]. Henry's law is used for CO_2 and H_2S. N_2 is not included in the "emea" insert package, so it is added as one of the components. N_2 is also defined in the system to obey Henry's law.

The elementary steps for the reaction can be represented by the following equilibrium reactions:

$2H_2O \leftrightarrow H_3O^+ + OH^-$ (R 13)

$CO_2 + 2H_2O \leftrightarrow H_3O^+ + HCO_3^-$ (R 14)

$HCO_3^- + H_2O \leftrightarrow H_3O^+ + CO_3^{-2}$ (R 15)

$MEAH^+ + H_2O \leftrightarrow H_3O^+ + MEA$ (R 16)

$MEACOO^- + H_2O \leftrightarrow HCO_3^- + MEA$ (R 17)

To overcome difficulties in converging the flowsheet, especially due to the recycle stream, the simulation is carried out into three steps [51]:

- Step 1: The stand alone absorber model
- Step 2: The absorber and stripper integrated model, and
- Step 3: The absorber and stripper integrated model with recycle stream.

Since it is extremely difficult to converge the MEA flowsheet with a closed recycle stream, the rational behind this three-step process is to determine very good first initial guesses for each step, the results from the previous step becoming the initial guess for the subsequent step. The details of each step are given below.

Step 1: The Stand Alone Absorber Model

The model of the stand-alone absorber is shown in Figure 3-2.

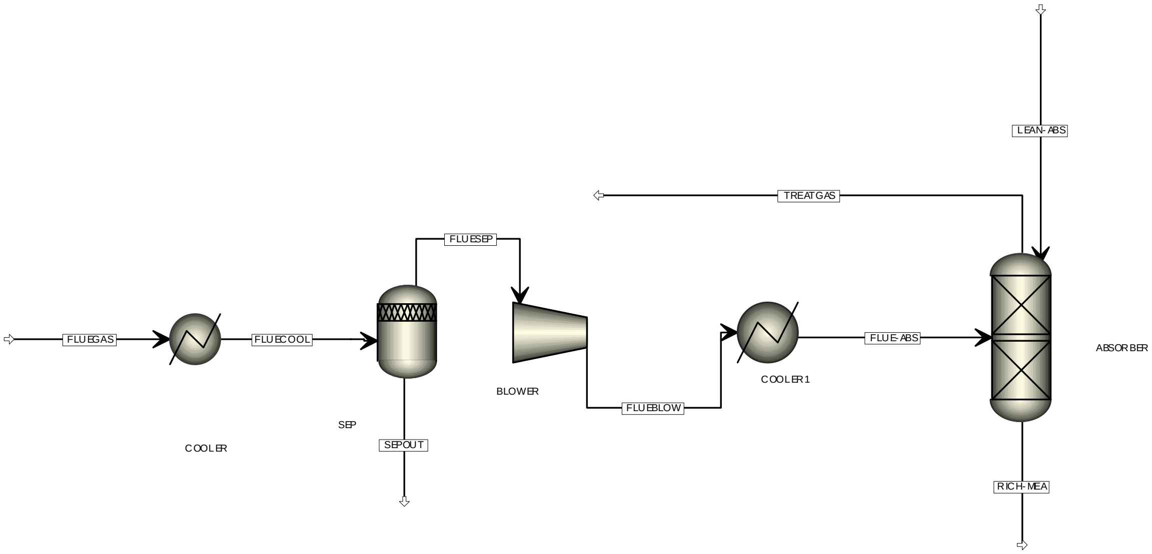

Figure 3-2: Stand-alone absorber model

Table 3-2 and Table 3-3 represent the stream input data and block data, respectively, for Case II (regular plant load data) for the stand-alone absorber model.

FLUEGAS data for the input stream were obtained from the St. Marys Cement plant. A reasonable initial estimate of LEAN-ABS stream is made before the simulation starts. The convergence of the simulation depends on how accurately the initial guess is made. For a desired percentage of carbon dioxide recovery and 30 wt. % MEA concentrations the LEAN-ABS flow rate is calculated. A design specification (DS-1) in Aspen Plus™ is defined for a target recovery of carbon dioxide in the flue gas. The design specification in Aspen Plus™ measures the CO_2 flow rate in TREATGAS and varies the mass flow rate of LEAN-ABS stream to obtain the target carbon dioxide recovery. The simulation for the stand alone absorber is performed for a CO_2 recovery from 55% to 95% and lean loading (mol MEA/ mol CO_2) of LEAN-ABS from 0.05 to 0.3. The mass flow rate of WATER is varied to keep the MEA concentration at 30 wt. % in the LEAN-ABS stream. The total number of trays for the absorber column specification is chosen arbitrarily to be 10. The absorber column pressure is specified at a constant pressure of 1.2 bar.

Table 3-2: Stream input data for stand alone absorber model (Case II)

Stream Name	Stream Properties	Value
FLUE-GAS	Temperature, °C	160
	Pressure, bar	1.013
	Mass Flow, kg/hr	
	H_2O	10631
	CO_2	80335
	N_2	155626
	O_2	6119
LEAN-ABS	Temperature, °C	40
	Pressure, bar	2.1
	Mass Flow, kg/hr	1730020
	Mass Fraction	
	MEA	0.282
	H_2O	0.657
	CO_2	0.061

Table 3-3: Block input data for stand alone absorber model (Case II)

Block Name	Block Properties	Value
COOLER	**Flash specifications:**	
	Temperature, °C	40
	Pressure, bar	1.013
SEP	**Type**	**Flash2**
	Temperature, °C	30
	Pressure, bar	1.013
BLOWER	**Type**	**Isentropic**
	Discharge pressure, bar	1.2
COOLER1	**Flash specifications:**	
	Temperature, °C	40
	Pressure, bar	1.2
ABSORBER	**Type**	**RateFrac™**
	Condenser type	n/a
	Reboiler type	n/a
	Operating Parameters:	
	Temperature, °C	40
	Top segment pressure, bar	1.2
	Number of Segments	10
	Column estimated diameter, m	5.5
	Pressure drop across column, bar	0
	FLUE-ABS feed location	Above segment 1
	LEAN-ABS feed location	Above segment 10

Step 2: Absorber and Stripper Integrated Model

Once the stand alone absorber model has successfully converged; it was integrated with the stripper model along with the following other unit operations: PUMP1, PUMP2, SOL-HEX, SOL-COOL, MIXER, and MCOMPR. The process flowsheet for the absorber and stripper integrated model is shown in Figure 3-3**.** Table 3-4 and Table 3-5 represent the stream input data and block data, respectively, for the additional streams and blocks of absorber and stripper integrated model.

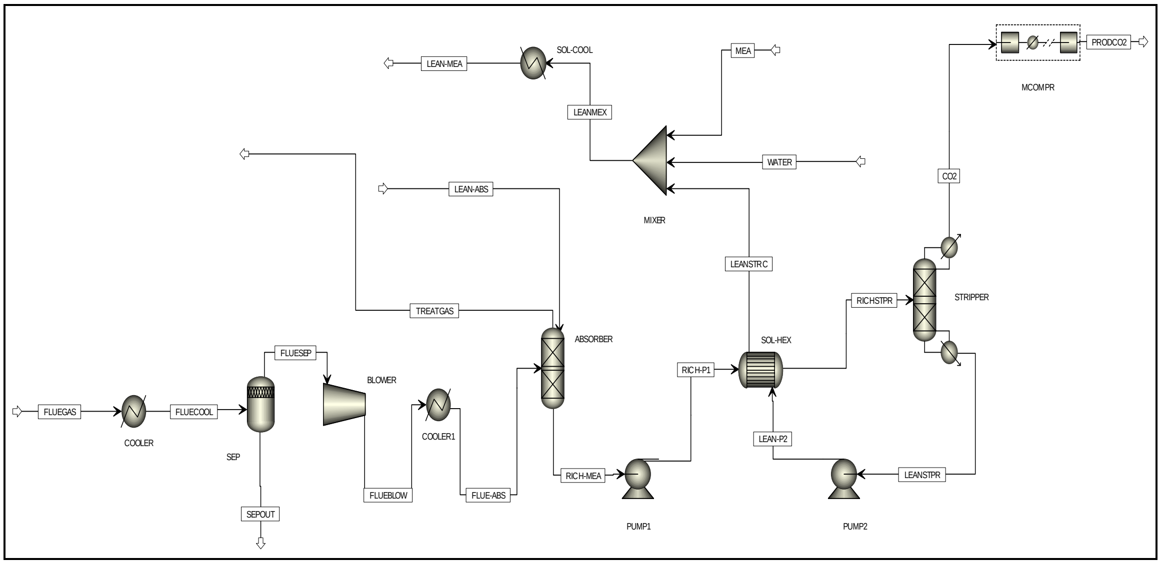

Figure 3-3: Absorber and Stripper integrated model

The purpose of modelling stripper is to minimize the reboiler heat duty while capturing the desired amount of carbon dioxide. The stripper is defined with two design specifications inside the column. The first design specification (DS-1) is to achieve the desired mass flow of CO_2 in the distillate for a set of lean loading by varying the bottoms-to-feed ratio (B: F) at the bottom of the stripper. The other design specification (DS-2) is to attain the desired mole purity of carbon dioxide (typically 98% of CO_2 purity) in the product stream by varying the molar reflux ratio (RR) at the top of the stripper. The stripper operating pressure is specified at 1.9 bar and the total number of tray is arbitrarily chosen to be 12 for the simulation. the rich MEA solution enters above stage 6 of the stripper. There are a partial condenser and a kettle reboiler at the top and bottom of the stripper, respectively in order to control the separation between the carbon dioxide and the MEA. The temperature in the reboiler remains approximately 123^{o}C which readily agrees with the operating conditions found in the literature [31]. Solvent degradation starts if the reboiler temperature exceeds 125^{o}C.

Table 3-4: Stream input data for absorber and stripper integrated model

Stream Name	Stream Properties	Value
WATER	Temperature, °C	40
	Pressure, bar	2.1
	Mass Flow, kg/hr	16,948
	Mass Fraction	
	H_2O	1
MEA	Temperature, °C	40
	Pressure, bar	2.1
	Mass Flow, kg/hr	45
	Mass Fraction	
	MEA	1

Table 3-5: Block input data for absorber and stripper integrated model

Block Name	Block Properties	Value
STRIPPER	**Type**	**RateFrac™**
	Condenser type	Partial vapor
	Reboiler type	Kettle
	Operating Parameters:	
	Number of Segments	12
	Tray type	Valve
	Column estimated diameter, m	6
	Bottom to feed ratio (mole)	0.94
	Reflux ratio (mole)	1.8
	Top segment pressure, bar	1.9
	Pressure drop across column, bar	0
	Feed tray location	7
SOL-HEX	Cold stream outlet Temperature increase, °C	45
SOL-COOL	Temperature, °C	40
	Pressure, bar	2.1
PUMP1/PUMP2	**Type**	**Isentropic**
	Discharge pressure, bar	2.1
MCOMPR	**Type**	**Polytropic**
	No of Stages	4
	Discharge Pressure, bar	150
	Discharge Temperature, °C	30

The hot LEANSTPR stream from the bottom of the stripper is used to preheat the rich MEA stream RICH-P1 to a maximum temperature of 45°C in SOL-HEX, thus reducing the reboiler duty. Make up MEA and WATER streams are mixed with the LEASTRC stream in the MIXER. The combined stream LEANMEX is cooled down to 40°C in the block SOL-COOL. The flow rates of WATER and MEA make up are adjusted in order to control the flow rate of LEAN-MEA to be equal to LEAN-ABS, which will help for converging the flowsheet. Both PUMP1 and PUMP2 are used to change the stream pressure to uphold the desired pressure when any pressure drop is observed. PUMP1 is used to increase the RICH-MEA pressure a little bit higher than the STRIPPER operating pressure to transfer the RICH- MEA to the STRIPPER. On the other hand, PUMP2 is used to compensate for the pressure drops across the unit SOL-HEX, MIXER, SOL-COOL, and ABSORBER.

Step 3: The Absorber and Stripper Integrated Model with Recycle Stream

The absorber and stripper integrated model with recycle stream is shown in Figure 3-4. This model is used once the convergence of the absorber and stripper integrated model is achieved. To ease convergence, the stream LEAN-ABS should match exactly the stream LEAN-MEA. The convergence depends on how accurate the initial guesses are made. When the stream LEAN-ABS exactly matches the stream LEAN-MEA, then the inlet stream LEAN-ABS is removed from the block absorber and the stream LEAN-MEA substitutes LEAN-ABS as recycle stream. The design specifications in the STRIPPER remain the same as those in Step 2 but the design specifications in the ABSORBER are modified. The target carbon dioxide recovery is attained by varying the flow rate of the stream MEA and the make up WATER is varied to retain the MEA concentration at 30 wt % in the LEAN-MEA stream. To ease the convergence of the simulation, the sequence of the block for simulation is defined inside the process convergence options of Aspen Plus$^{\text{TM}}$ as shown in Table 3-6. The stream LEAN-P2 is used as tear stream for convergence C1 and tear stream LEANSTRC is used for convergence C2. The Wegstein method was used for the tear stream convergence.

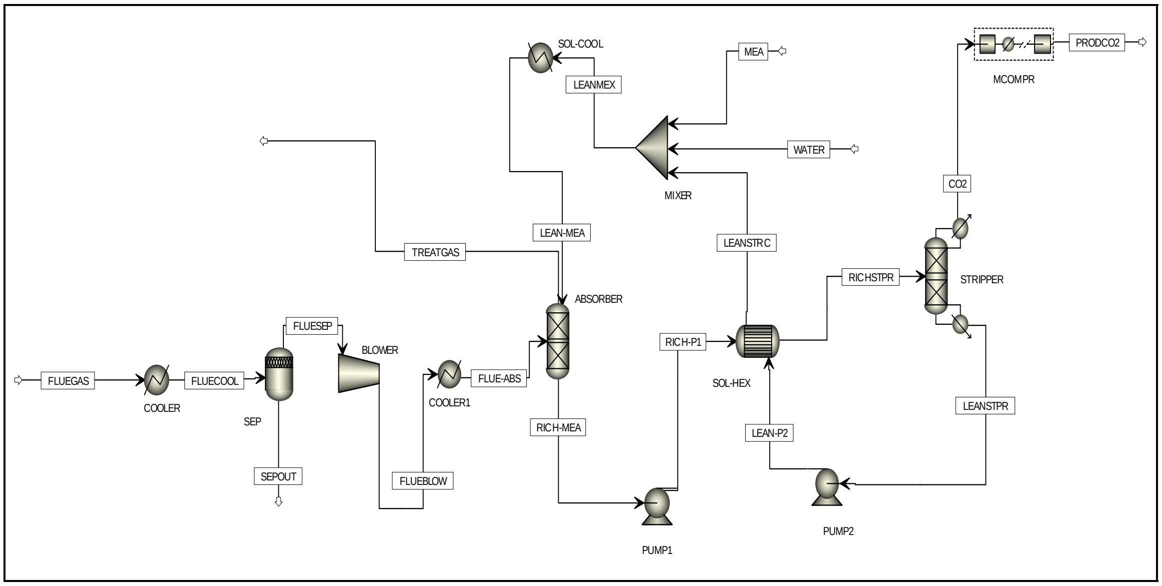

Figure 3-4: Absorber and Stripper integrated model with Recycle stream

Table 3-6: Block Simulation sequence for Absorber and Stripper integrated model with Recycle stream

Loop Return	**Block Type**	**Block**
	Unit Operations	**COOLER**
	Unit Operations	**SEP**
	Unit Operations	**BLOWER**
	Unit Operations	**COOLER1**
Begin	**Convergence**	**C2**
	Unit Operations	**MIXER**
	Unit Operations	**SOL-COOL**
	Unit Operations	**ABSORBER**
	Unit Operations	**PUMP-1**
Begin	**Convergence**	**C1**
	Unit Operations	**SOL-HEX**
	Unit Operations	**STRIPPER**
	Unit Operations	**PUMP-2**
Return to	**Convergence**	**C1**
Return to	**Convergence**	**C2**
	Unit Operations	**MCOMPR**

3.4 Aspen Simulation Parameters for MEA based CO_2 Absorption Model

Aspen Plus™ simulation parameters for MEA based CO_2 capture process for a CO_2 recovery of 85 percent and lean loading (mol CO_2/mol MEA) of 0.30 for Case II is discussed in this section. The simulation results for Case I, Case II and Case III are presented in Appendix B.

Flue gas from the St. Marys cement plant is usually emitted through the stack into the atmosphere at a temperature of 160°C. Before the flue gas is vented to the atmosphere, it passes through the flue gas conditioning units where SO_2 is scrubbed and all the particulate matters are removed. A clean flue gas, free of SO_2 and particulates, is used as the basis for the MEA CO_2 capture process, as shown in Figure 3-5.

The flue gas is cooled down to 40°C before it enters the separator. Some water is removed during the cooling process. The separator is used to condense and separate some of the saturated water of the flue gas. The separator flash temperature is set at 30°C. This reduces the blower power requirement to compress the flue gas before it enters into the absorber. The flue gas from the separator next enters the blower where the flue gas is compressed to a pressure of 1.2 bar. The flue gas from the blower further cools down to 40°C before it enters the absorption column. The CO_2 capture efficiency increases and MEA and water evaporation loss decreases if the temperature in the column is maintained at low temperature. Higher flue gas temperatures in the range of 50°C to 85°C tend to increase MEA and H_2O evaporation losses [31]. The lean solution temperature also increases in this temperature range. The flue gas enters at the bottom of the absorption column and the lean MEA solution flows counter currently at the top of the absorber. The lean MEA solution temperature is maintained at 40°C temperature and at 1.2 bar pressure. It is very important to keep the lean MEA solution temperature as low as possible for two main reasons: (1) to reduce MEA and water make up and (2) to increase CO_2 capture efficiency. A lean MEA solution temperature of 40°C is a good compromise between the above benefits, cooling duty and cooling tower size [31]. The lean MEA solution is composed of 30% wt MEA and has a lean loading of 0.30 mol CO_2/mol of MEA. The total number of trays in the absorber is set at 3 with no condenser and reboiler.

The absorption column operates at a temperature of 40^{o}C and a pressure of 1.20 bar. It is well known that the absorption rate in the absorber increases with partial pressure of CO_2. Therefore, an increase in feed pressure should increase the mass transfer rate in the absorber. However, there is a penalty in the form of blower power associated with this. So, there is no cost benefit in operating the absorber at high pressure [52]. Therefore, the absorption column is set at a pressure of 1.2 bar for this simulation process.

The absorption of CO_2 is an exothermic process where MEA reacts chemically with CO_2 in the column. The principle of CO_2 extraction using amines is based on the reaction of a weak base (MEA) with a weak acid (CO_2) to form a water soluble salt. This reaction can be reversed by varying the temperature [55]. In a commercial unit, a wash section is included at the top of the absorption column to remove H_2O and vapourized the MEA carried by the washed gas. This is done to cool the treated flue gas and to avoid MEA losses [70] [71]. However, this process is not included in the simulation.

The rich MEA stream from the bottom of the absorber then flows to PUMP1 where its pressure is increased to 2.1 bar. Due to the very low pressure of RICH-MEA stream, this pump is needed to transfer the rich solution to the stripper. The RICH-P1 stream from PUMP1 next enters the heat exchanger SOL-HEX, where the temperature of the RICH-P1 stream is increased from 56.9°C to 101.9°C before it enters the STRIPPER. In SOL-HEX, the hot LEANSTPR stream from the bottom of the STRIPPER is used to preheat the RICH-P1 stream. This heat exchanger reduces the reboiler heat duty and reduces the cooling duty of SOL-COOL on the lean solution as well. Make up for the MEA and WATER streams are mixed with LEASTRC stream in the MIXER at a temperature of 40°C and a pressure of 2.1 bar. The flow rates of WATER and MEA make up are set from the calculation in order to control the flow rate of LEAN-MEA to be equal to LEAN-ABS to ease to convergence in the simulation step. The combined stream LEANMEX is cooled down to 40°C in the block SOL-COOL.

The STRIPPER column is modelled with 9 theoretical stages, where the condenser represents the first stage and a kettle reboiler represents the 9th stage. The Stripper operates at a pressure of 1.9 bar. The carbon dioxide is stripped from the MEA solvent by heat when the MEA solution descends through the stripping tower. The heat required by the rich solution regeneration process is provided to the kettle reboiler from an external source. A typical heat source can be a conventional industrial steam generator.

A wash section is also provided in the stripper unit where residual MEA vapour is condensed and returned to the column, which is not shown in Figure 3-5. The stripper column operates best at higher pressures. An increase in the pressure leads to an increase in reboiler temperature and, subsequently the temperature of the whole column. However, solvent degradation occurs at reboiler temperatures greater than 125°C. The pressure limit for the solution in order to prevent temperatures in excess of 125°C is 2 bar, which corresponds to a temperature slightly below 125^{o}C [31]. The lean solution from the bottom of the stripper exits at a temperature of 119.1°C. The vapour exits at the top of the STRIPPER and is sent to a reflux condenser where the steam is condensed and the CO_2 is cooled. A reflux drum is provided to separate the cooled CO_2 and the condensed steam. The condensate is then pumped to the stripper column.

The carbon dioxide exits at the top of the STRIPPER and CO_2 is compressed depending on its utilization. In this study, the product CO_2 is compressed to a pressure of 150 bar and cooled down to a temperature of 30°C for the pipeline transportation. The Aspen simulation results for Case II with 85% CO_2 capture and lean loading of 0.30 is shown in Table 3-7. The simulation results for Case I and Case III is provided in Appendix B.

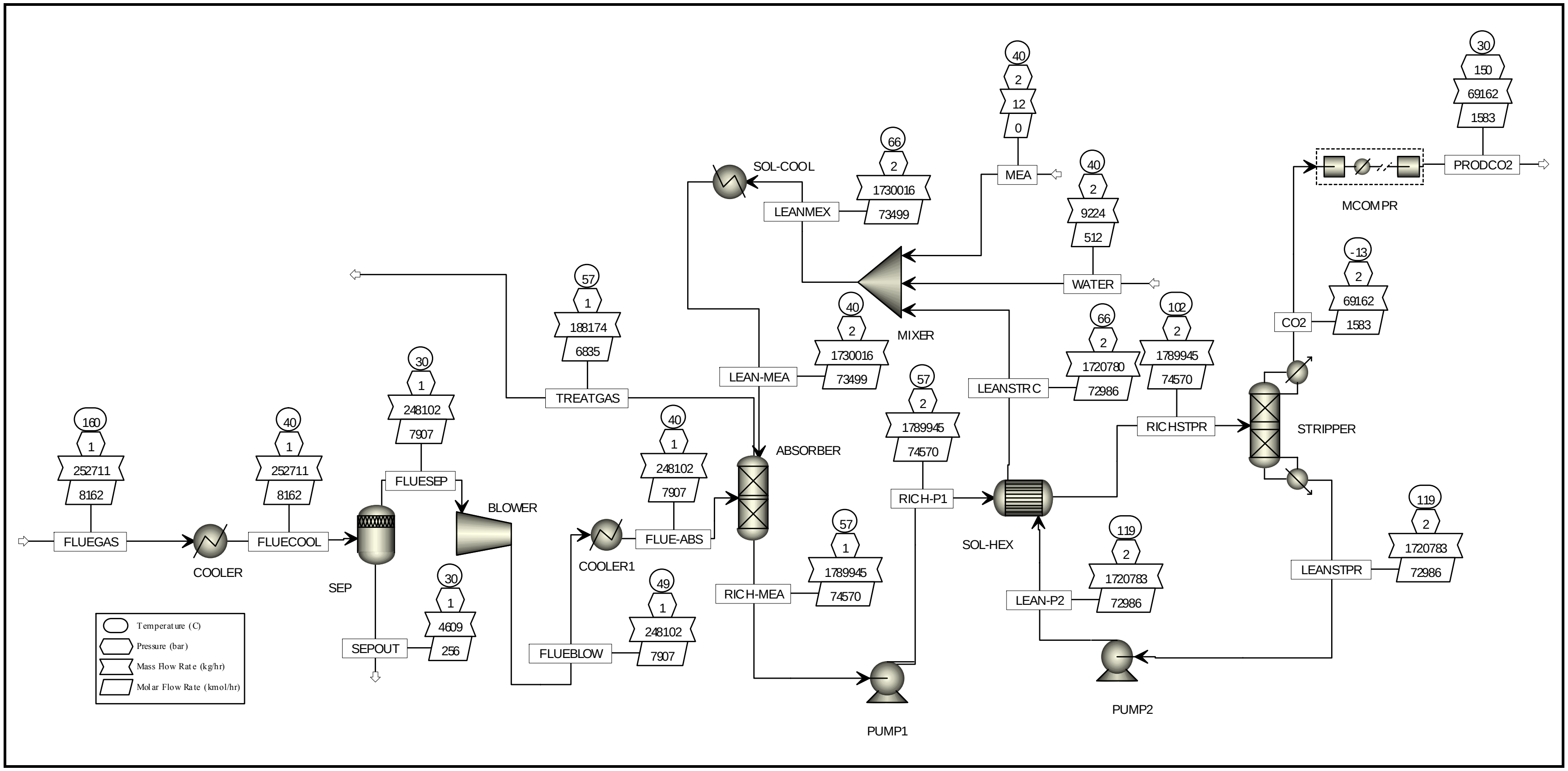

Figure 3-5: Aspen simulation process parameters for MEA based CO_2 capture process

Table 3-7: Aspen simulation results of MEA based CO_2 absorption for Case II (CO_2 recovery: 85% and lean loading: 0.30)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	56.9	119.1	119.1	56.9	101.9	56.7	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	8162	8162	7907	7907	7907	73499	74570	72986	72986	74570	74570	6835	512	0	1583
Mass Flow kg/hr	252711	252711	248102	248102	248102	1730020	1789950	1720780	1720780	1789950	1789950	188174	9224	12	69162
Volume Flow m^3/hr	290000	209428	196399	176408	171293	1733	1898	1818	1818	1898	15231	156035	9	0	221
Enthalpy MMkcal/hr	-197	-205	-191	-190	-190	-5086	-5203	-4939	-4939	-5203	-5128	-73	-35	0	-148
Mass Flow kg/hr															
H_2O	10631	10631	6035	6035	6035	1137190	1128000	1127970	1127970	1128000	1128000	15223	9224	0	35
MEA	0	0	0	0	0	487366	487354	487354	487354	487354	487354	12	0	12	0
CO_2	80335	80335	80334	80334	80334	105462	173749	105464	105464	173749	173749	12047	0	0	68285
N_2	155626	155626	155616	155616	155616	0	767	0	0	767	767	154848	0	0	767
O_2	6119	6119	6118	6118	6118	0	75	0	0	75	75	6043	0	0	75
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.657	0.63	0.655	0.655	0.63	0.63	0.081	1	0	0.001
MEA	0	0	0	0	0	0.282	0.272	0.283	0.283	0.272	0.272	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.061	0.097	0.061	0.061	0.097	0.097	0.064	0	0	0.987
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.823	0	0	0.011
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.032	0	0	0.001
Mole Flow kmol/hr															
H_2O	590	590	335	335	335	63124	62614	62612	62612	62614	62614	845	512	0	2
MEA	0	0	0	0	0	7979	7978	7978	7978	7978	7978	0	0	0	0
CO_2	1825	1825	1825	1825	1825	2396	3948	2396	2396	3948	3948	274	0	0	1552
N_2	5555	5555	5555	5555	5555	0	27	0	0	27	27	5528	0	0	27
O_2	191	191	191	191	191	0	2	0	0	2	2	189	0	0	2
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.859	0.84	0.858	0.858	0.84	0.84	0.124	1	0	0.001
MEA	0	0	0	0	0	0.109	0.107	0.109	0.109	0.107	0.107	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.033	0.053	0.033	0.033	0.053	0.053	0.04	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.809	0	0	0.017
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.028	0	0	0.001

3.5 Simulation Results and Discussions

Initially, the number of trays in the absorber and the stripper was arbitrarily chosen to be 10 and 12, respectively. The reboiler heat duty is strongly dependent on the absorber and stripper number of trays. Simulations were carried out to optimize the number of trays for both the absorber and the stripper columns. Figure 3-6 shows the effect of number of trays in the absorber and stripper columns on the reboiler heat duty.

The effect of number of stripper trays on the reboiler heat duty was studied by varying the stripper tray numbers from 4 to 12 while keeping the absorber trays fixed at 10. Then the absorber number of trays was reduced and fixed at 8 and stripper number of trays was varied from 4 to 12 to study the effect of number of stripper trays on the reboiler duty. Again, the absorber number of trays was reduced and fixed at 6 and the number of trays in the stripper was varied from 4 to 12. It is seen from Figure 3-6 that the reboiler duty can be reduced by increasing the stripper number of trays. As can be seen the reboiler duty can be lowered by increasing the stripper number of trays from 4 to 9. Further increase in the number of trays in the stripper increases the reboiler duty only marginally. An increase in the number of trays in the stripper requires a lower reflux ratio for a given carbon dioxide recovery thereby helping in the reduction of the reboiler heat duty.

Like the stripper trays, an increase in the number of trays in the absorber also reduces the reboiler heat duty as shown in Figure 3-6 but the change is trivial. The reduction in reboiler heat duty is due to the fact that higher number of trays allows lower solution circulation rates in the absorber and thus reduces the heat required by reboiler.

Figure 3-6 shows that a minimum reboiler heat duty is obtained with the absorber number of trays 3 and stripper number of trays 9 for a desired 85% CO_2 capture (Case II).

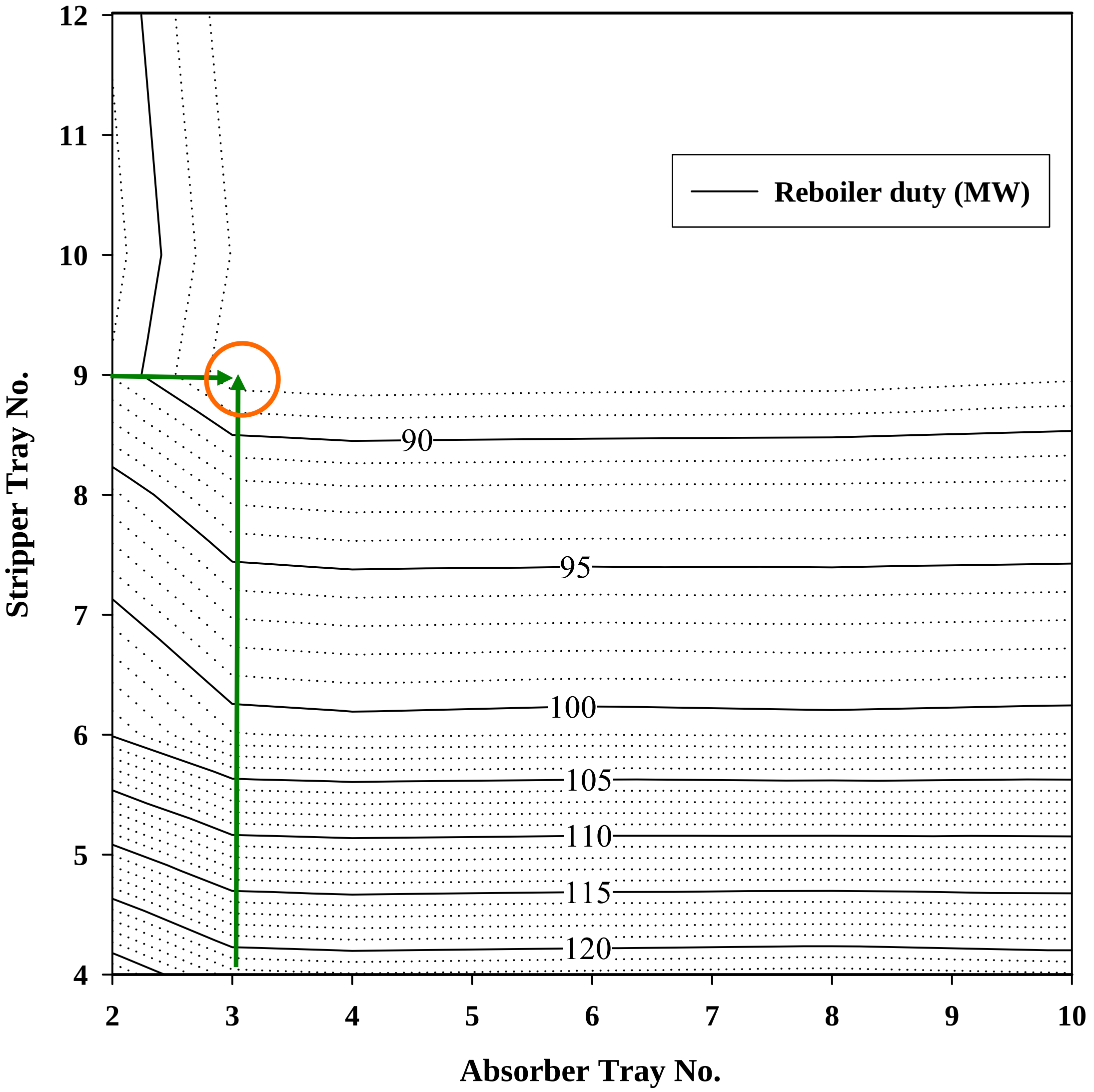

Figure 3-6: The effect of number of trays in the absorber and stripper on the reboiler heat duty

It is seen from Figure 3-7 that the lean solvent flow rate increases with lowering the number of stages required for the separation. When the solvent flow rate increases, the diameter of the column increases subsequently which will increase the capital cost required for the CO_2 capture process.

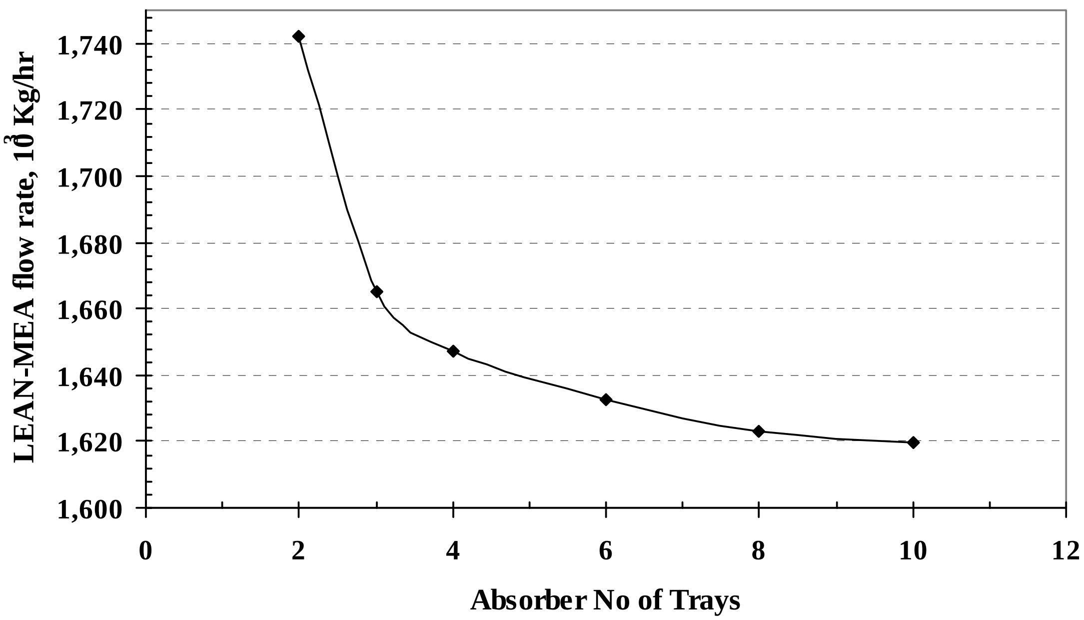

Figure 3-7: LEAN-MEA flow rate for different absorber number of stages

The simulation was conducted with the goal of capturing carbon dioxide from flue gas with minimum reboiler heat duty per unit CO_2 captured. A low CO_2 recovery will result in a large amount of CO_2 recirculated throughout the process and consequently will increase the size of the equipment required. The absorber column itself consumes no energy in terms of a reboiler or a condenser and the only energy consumption takes place in the reboiler of the stripper column. The study was made for a carbon dioxide recovery from 55% to 95% for different lean loadings. Figure 3-8 shows the reboiler duty for CO_2 recovery from 55% to 95% with a lean loading from 0.05 to 0.30 for Case II. As can be seen from Figure 3-8 the minimum reboiler heat duty is required for a CO_2 recovery of 55 percent and lean loading of 0.30. It is observed that at lean loading of 0.30, the reboiler heat duty is not very sensitive to the CO_2 recovery, at least in the range 55% to 95%. It is desirable to capture higher percentage of carbon dioxide. So, 85 percent carbon dioxide recovery with lean loading of 0.30 was chosen for further analysis and cost evaluation.

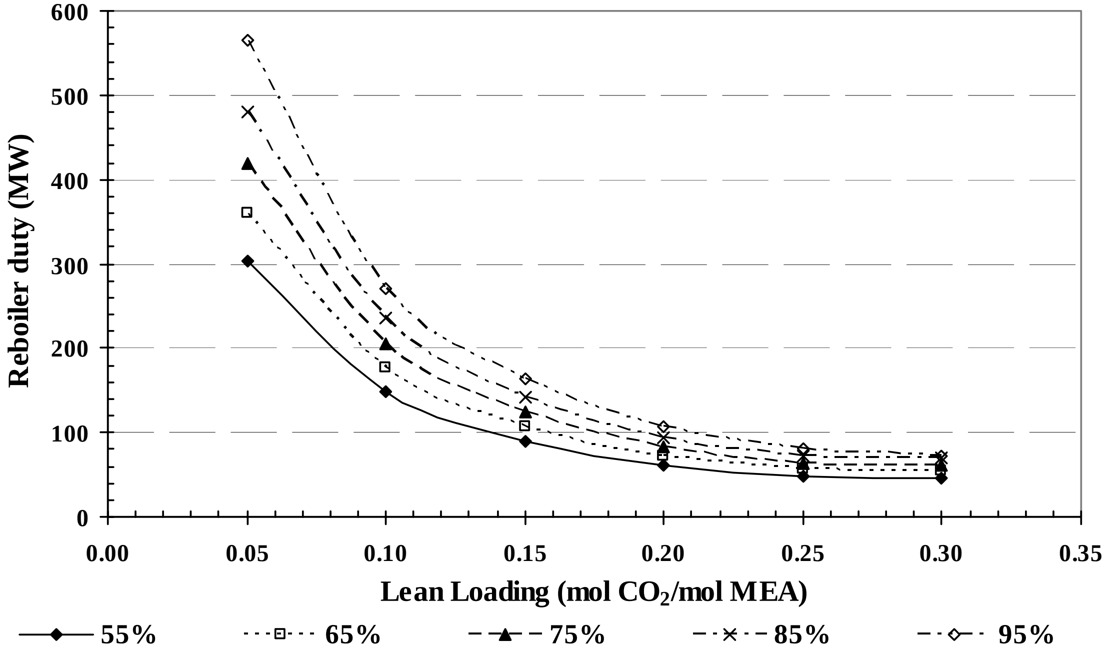

Figure 3-8: Reboiler duty for different CO_2 recovery at various Lean loading

Solvent flow rate is perhaps the most important parameter that affects significantly carbon dioxide capture from flue gas. Figure 3-9 represent the LEAN-MEA or/and RICH-MEA flow rate, respectively, for different lean loadings (mol CO_2/mol MEA). As can be seen from Figure 3-9, the LEAN-MEA flow rate was increased from 732,261 to 1,730,020 kg/hr for Case II when the lean loading was increased from 0.05 to 0.30. Higher CO_2 loadings require greater flow of lean MEA to achieve the same level of CO_2 recovery. The data for Case I and Case III can be seen in Appendix B.

Figure 3-10 shows the relationship between the carbon dioxide loading in rich solvent and lean solvent. It is seen from Figure 3-10 that the rich loading increases with increasing lean loading. But the value of rich loading increase is not much compared to the value of lean loading increase. The rich loading increase is due to significant increase of carbon dioxide recirculation through the columns as the lean loading increases. For Case II, the range of rich loading was increased from 0.4858 to 0.4948 with lean loading increase from 0.05 to 0.30. The rich loading values for Case I and Case III are presented in Appendix B.

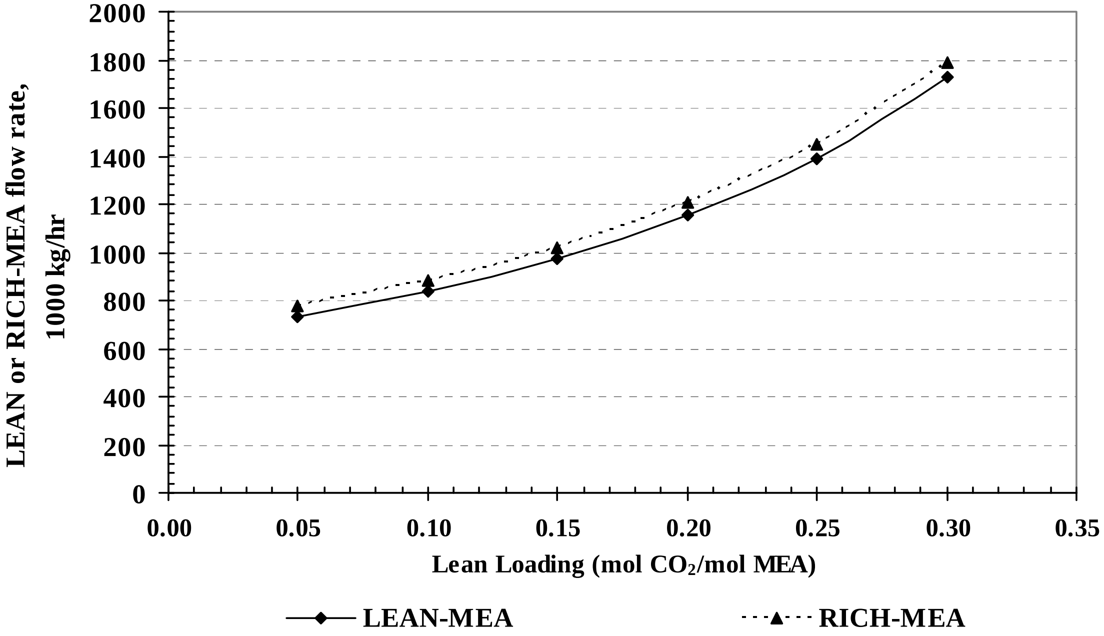

Figure 3-9: LEAN or RICH-MEA flow rate for various lean loading (mol CO_2/mol MEA)

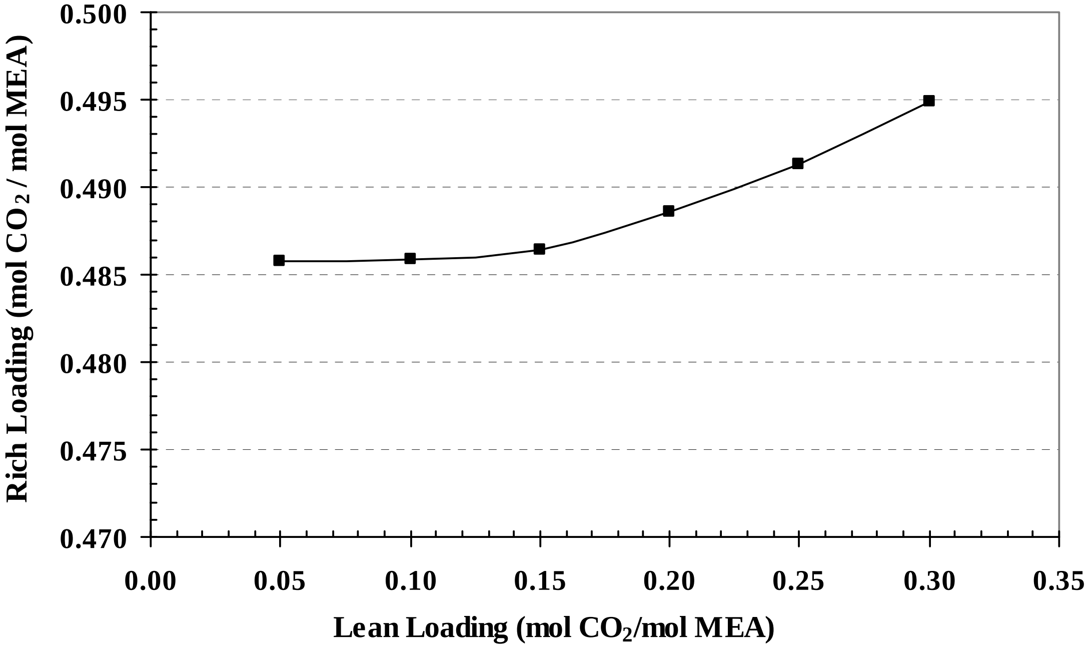

Figure 3-10: Rich loading (mol CO_2/mol MEA) at different lean loading

There is a relationship between the energy requirement or energy consumption per unit CO_2 captured and the CO_2 loading of lean amine solution obtained after the regeneration process. The effect of lean loading (mol CO_2/mol MEA) on reboiler duty for CO_2 capture process was studied by varying the lean loading from 0.05 to 0.30. The carbon dioxide recovery was specified at 85 percent with purity of 98 percent by mole.

Figure 3-11 shows the effect of lean loading on reboiler duty for CO_2 Capture process for Case I, Case II, and Case III. The reboiler duty decreased from 508 to 68.98 MW (For Case II) when the lean loading was varied from 0.05 to 0.30 for a CO_2 recovery of 85% and purity of 98%. As can be seen, the difference in reboiler duty for lean loading of 0.25 and 0.30 is trivial. So the cost evaluation was done for a CO_2 recovery of 85% with lean loading (mol CO_2/mol MEA) of 0.30 for the Case I, Case II and Case III. The simulation did not converge at lean loading of 0.35 for CO_2 recovery of 85% and 98% mole purity.

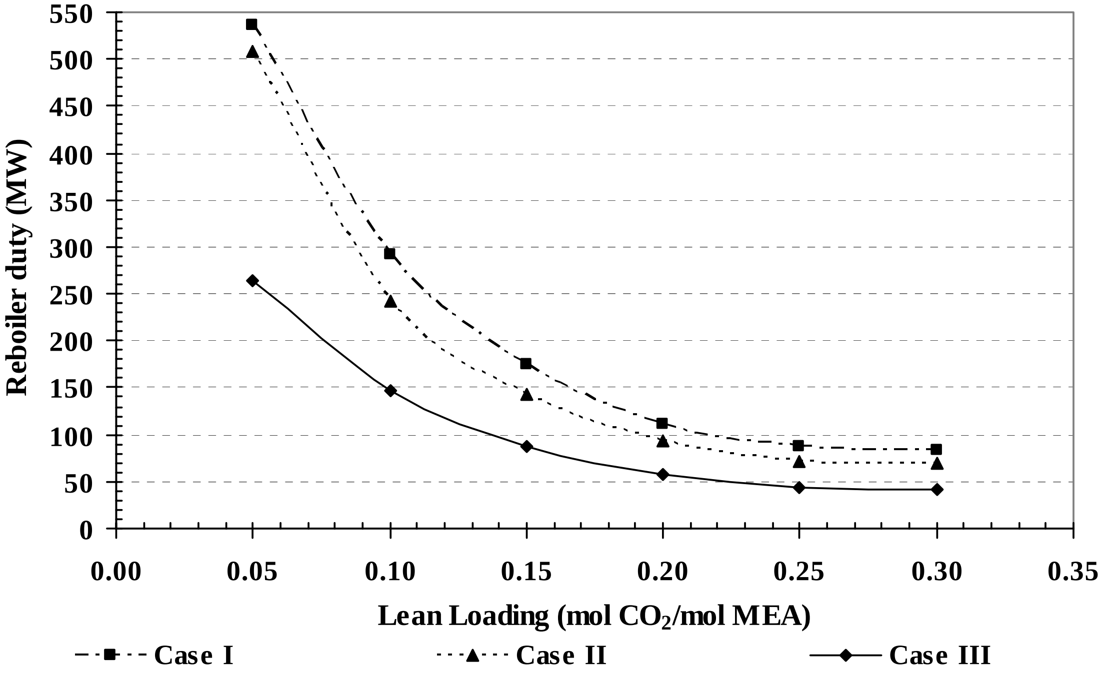

Figure 3-11: Reboiler duty at different lean loading for Case I, Case II and Case III

Figure 3-12 (a)-(c) represents the reboiler duty per tonne of CO_2 captured against different lean loadings for Case I, Case II, and Case III respectively. There are two important factors affecting the value of the reboiler duty. The first one is the circulation rate of the solvent and the second one is the value of the lean loading of the stream leaving the bottom of the stripper and recycled back to the absorber. At a lean loading of 0.05, a huge amount of heat is required at the reboiler to separate the desired carbon dioxide because the specified loading is very low. Even though the solvent circulation rate is low, the reboiler would consume a lot of energy to separate the carbon dioxide until the lean solvent reaches that CO_2 loading. Also at low lean-CO_2 loading, a significant amount of additional energy is required during the regeneration process.

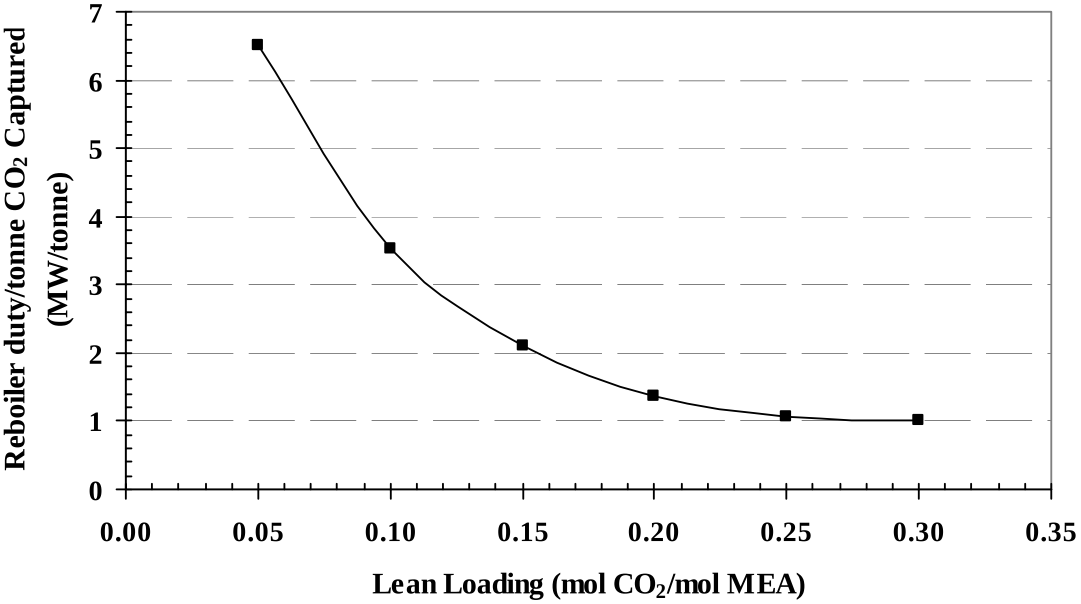

Figure 3-12 (a): Reboiler duty per tonne of CO_2 Captured at various lean loading for Case I

When the lean loading specification is increased, the reboiler duty decreases until it reaches a minimum for a lean loading equal to 0.30. So, the amine solution with a higher CO_2 loading can be regenerated more easily and with lower reboiler duty than with a lower CO_2 loading. The CO_2 loading of lean solution may require to be adjusted to an optimal level where the energy consumption at the regeneration unit is reduced, while maintaining a satisfactory capture performance. This may result in a significant reduction in the overall cost of the capture process.

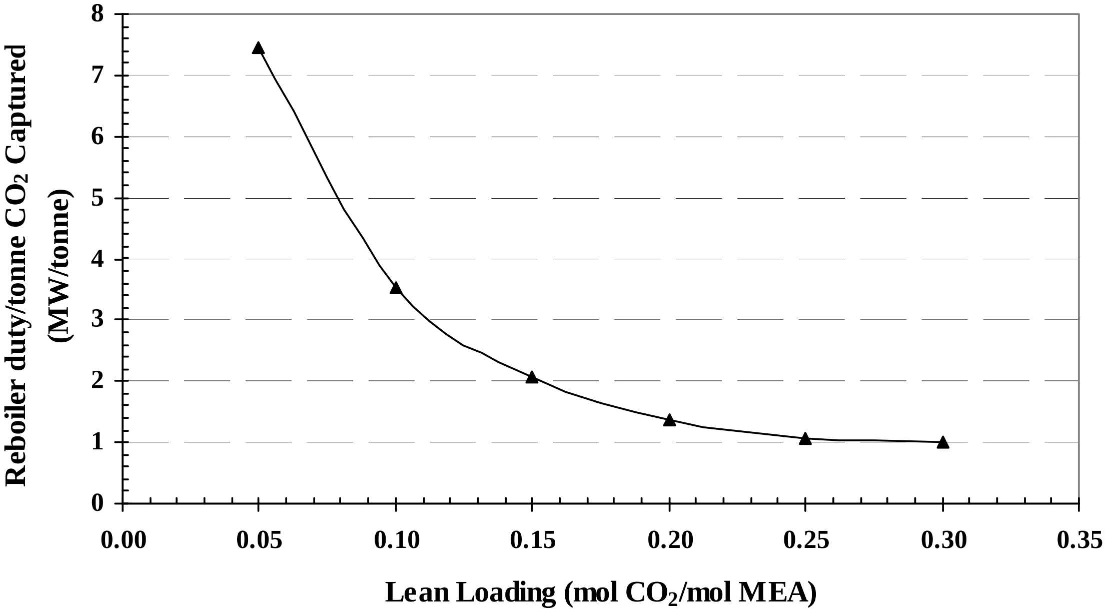

Figure 3-12 (b): Reboiler duty per tonne of CO_2 Captured at different lean loadings for Case II

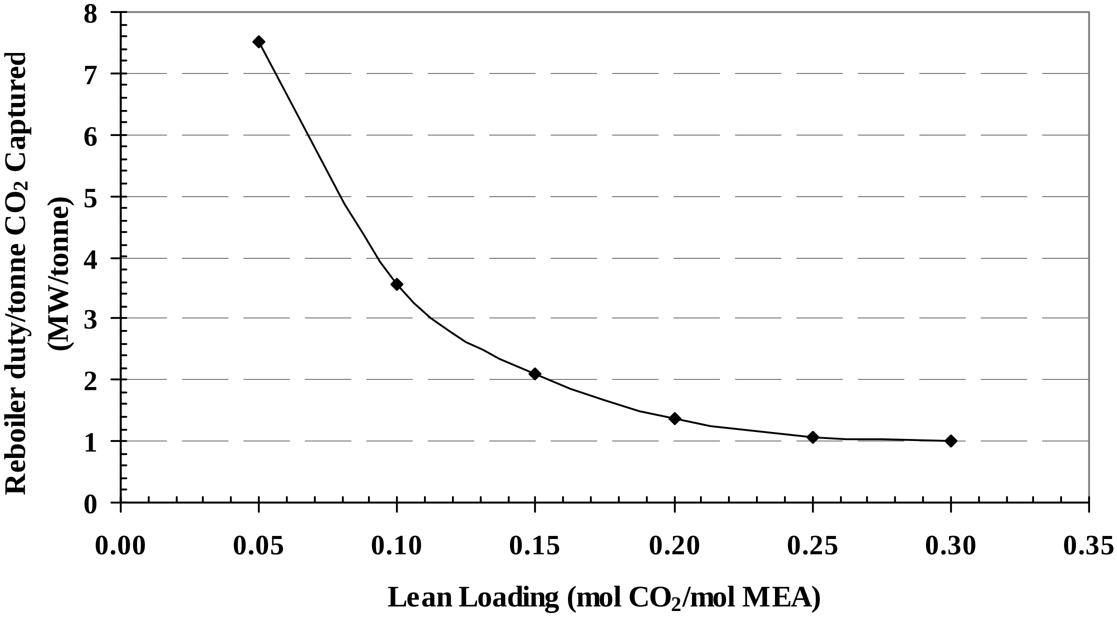

Figure 3-12 (c): Reboiler duty per tonne of CO_2 Captured at different lean loadings for Case III

3.6 Case IV: Switching to a Lower Carbon Content Fuel at Average Plant Load

Cement manufacturing industry emits a large amount of carbon dioxide into the atmosphere because combustion is the heart of this industry and the reaction itself in the kiln also produces huge amount of carbon dioxide. In cement manufacturing operations about 40% of the industry's emissions come from fossil fuel combustion; about 5% from transport of raw materials; and about 5% emissions comes from combustion of fossil fuel required to produce the electricity consumed by the cement manufacturing operations and the remaining cement-related emissions (about 50%) an inherent byproduct originate from the process that converts limestone ($CaCO_3$) to calcium oxide (CaO), the primary precursor to cement [19].

The appropriate selection and use of a fuel has always been and still is a matter of great concern for the cement industry. The current fierce competition in the cement market and the high impact of the item "fuel cost" in the final price of the product is making companies look for the most economic mix to fire in their kilns. Solid pulverized fuels are most frequently used by the cement industry. Coal is the most widely used fuel by the cement industries and among those bituminous coal ranks first in preference, followed by subbituminous coal and lignite. Petcoke's share in the cement production energy matrix grew significantly in the last 10 years. Delayed petcoke, fluidcoke, flexicoke are three different types of petcoke used in cement industry. The use of residual solids such as scrap tires, saw dust, wood chips, cardboard, textiles, and plastic shavings by cement plants has increased noticeably because it provides reduction of cement production costs. Fuel oil is widely used in the cement industry because it is very easy to fire in its liquid phase. In the last few years the share of liquid wastes such as solvents, paint sludge, used lubricant oils etc. has grown considerably in cement plants energetic matrix. The share of natural gas in the cement plant energy matrix has grown slightly in the past few years.

Fuel usage in North American cement plants is approximately 58 percent coal, 13 percent each of petcoke, natural gas, and "other fuels" such as residues and wastes, and 2.5 percent fuel oil [19].

It is unlikely that the cement industry would switch from burning low cost fuel such as coal, coke, petcoke to high cost natural gas. Natural gas costs more than twice or thrice as much per GJ than coal or petcoke, and switching to natural gas would also render the profitability of the plant to be dependent on the likely market fluctuations in the price of natural gas.

However, this section discussed the switching fossil fuel especially coal to less carbon-intensive industrial fuels such as natural gas to see how it can reduce GHG emissions in a cost-effective manner. Natural gas is a naturally occurring gas mixture, consisting mainly of methane. The component, composition, and heating value use for the calculation is shown in Table 3-8 [Union Gas].

Table 3-8: Natural gas composition and properties

Component	Formula	Composition (mol %)
Methane	CH_4	**94.9**
Ethane	C_2H_6	**2.5**
Propane	C_3H_8	**0.20**
iso-Butane	$i\text{-}C_4H_{10}$	**0.03**
neo-Butane	$n\text{-}C_4H_{10}$	**0.03**
iso-Pentane	$i\text{-}C_5H_{12}$	**0.01**
neo-Pentane	$n\text{-}C_5H_{12}$	**0.01**
Hexane	C_6H_{14}	**0.01**
Nitrogen	N_2	**1.60**
Carbon dioxide	CO_2	**0.70**
Oxygen	O_2	**0.02**
Hydrogen	H_2	**trace**
Specific Gravity	**0.585**	
Gross Heating Value MJ/m^3)	**37.8**	
Ignition Point	**593^{o}C**	
Theoretical Flame Temperature	**1960^{o}C**	
Maximum Flame Velocity	**0.30 m/s**	

This section discusses the carbon dioxide emission for the average plant capacity. The average clinker production capacity is 94.2 tonnes of clinker per hour. At average plant load the carbon dioxide production in the kiln from the calcinations reaction is 48.984 tonnes of CO_2 per hour. If coal, petcoke, and coal fines are used as fuel for combustion, the CO_2 generation is 31.351 tonnes of CO_2 per hour. If fuel switching of coal, petcoke, and coal fines to natural gas takes place the carbon dioxide generation from the burning of natural gas is 16.442 tonnes of CO_2 per hour. The carbon dioxide emission reduced 14.909 tonnes per hour which is 18.55 percent reduction of carbon dioxide reduction. Details of the emission reduction calculation are reported in the Appendix D.

3.7 Sources of Waste Heat in The Cement Industry for Steam Generation

Cement manufacturing is a highly energy-intensive process. A schematic diagram for different stages of cement manufacturing process was shown in Figure 2-1. Once the raw materials for cement making have been selected and blended, ground and homogenized into a fine and uniform kiln feed, then the raw materials enter at the upper end of the kiln. The raw materials slides and tumbles into the most progressively hotter zone of the kiln where a flame is produced by precisely controlled burning of powdered coal, oil or gas under forced draft. At the lower end of the kiln, fuels such as powdered coal, fuel oil, and natural gas feed a flame that reaches 1870°C. Here in the hottest part of the kiln, the raw materials reach about 1480°C and become partially molten. This intense heat triggers chemical and physical changes and a series of chemical reactions converts the calcium and silicon oxides into calcium silicates, cement’s primary constituent. At the lower end of the kiln, the raw materials emerge as a new substance: red-hot particles called clinker.

During the clinker-burning process, the combustion products are produced by the burning of the fossil or alternative fuels that leave the kiln as stack gas. The hot flue gas leaves the kiln at a temperature of 800°C and enters into the preheater tower where the flue gas increases the raw material waiting for kiln temperature of 150^oC-250^oC. The flue gas exits the preheater tower at temperature of 445^oC and then enters into the conditioning tower where water is sprayed to cool down the flue gas to a temperature of 263^oC. The flue gas exiting the perheater tower contains finely divided particles of raw mix, fly ash, calcined kiln feed, cement kiln dust (CKD), and volatile organic constituents (VOC) (e.g., potassium sulfate) CO_2, N_2, O_2, SO_x, NO_x. These particles are almost entirely removed from the gas stream before the stack gas is vented to the atmosphere. The conditioning tower serves the purpose of removing the particles from the stack gas. The sludge from the conditioning tower is partially recycled back as cement raw material. A large amount of heat is wasted in the conditioning tower because the flue gas enters into the conditioning tower at 445°C and exits at 263°C. If the plant runs at regular production load, the amount of waste is around 180 KW per tonne of clinker produced. By using improved technology and special type of heat exchanger this waste heat can be recovered. In a few of the European cement plants, exhaust-gas, waste-heat recovery systems have been installed which produce low grade steam for heating nearby residential and commercial buildings. For the present operating condition in St. Marys cement plant, this amount of waste heat can not be recovered. The flue gas after conditioning tower enters to raw mill where it is further cooled down. Then it enters to baghouse to remove the particulates and after that the flue gas emits through the stack at temperature of 160^{o}C. The flue gas from the stack goes to the CO_2 capture process where the flue gas is cooled down to 40^{o}C before it enters into the blower and around 92.35 KW of heat per tonne of clinker produced can be recovered during the flue gas cooling process.

The other source of heat recovery in cement industry is the clinker cooling. At the discharge end of the kiln, the clinker is red hot and contains around 1.168 GJ per tonne of clinker thermal energy [22] and at a temperature of 1400^{o}C. It is equally important that the heat exchange between clinker and air is efficient to ensure proper cooling, and at the same time maximize the recovery of heat to secondary air, tertiary air, and the related process requirement. The clinker is grate-cooled by forced-draft air in the clinker cooler. It is critical that cooling of the clinker is rapid to secure a phase composition that imparts adequate cementitious properties. Different kinds of clinker cooler such as planetary cooler, rotary cooler, shaft cooler, traveling grate cooler, and grate cooler are used for this purpose but the grate cooler is widely used in North America where the clinker and air moves the cross-current direction. This type of cooler can produce clinker discharge temperatures around 80°C. The clinker cooler in St. Marys cement plant can cool down the hot clinker to a temperature of 350^{o}C and then the further clinker cool down in ambient air and a heat of approximately 106 KW per tonne of clinker produced is wasted during this time. This amount of heat can be recovered to produce steam.

If the plant operates in regular load, the amount of heat that can be recovered as waste heat from St. Marys cement plant is about 393 KW per tonne of clinker produced but unfortunately this amount of heat is wasted because of the process constraints. By using improved technology this amount of heat can be recovered and produced steam that can be used in reboiler for CO_2 capture process.

Chapter 4: Economic Evaluation

4.1 Sizing and Costing Using Icarus Process Evaluator (IPE)

Icarus Process Evaluator (IPE) is designed to automate the preparation of detailed designs, estimates, investment analysis and schedules from minimum scope definition, whether from process simulation results or sized equipment lists. IPE uses expert links to affect the automatic transfer of process simulator output results. The simulation results from Aspen Plus™ were sent to IPE as input data for sizing and costing. After loading the process simulator data into IPE, the loaded data is examined using process view to make sure the models and arrangements. The standard model or process component in Aspen Plus is recognized by IPE and would be mapped with the equipments or project components existing in IPE. Mapping (and sizing) can be done either one item at a time or all items at once. Mapping relates each process simulator model to one or more of IPE's list of several hundred types of process equipment. Table 4-1 shows the mapping specifications in IPE for the MEA based carbon dioxide capture process. Size of equipment is a prerequisite to costing and results of size calculations performed during process simulation are loaded automatically by IPE [73]. But the absorber and stripper column were simulated using the RATEFRAC model which were not recognized by IPE. Therefore, the absorber and stripper columns were not sized automatically by IPE. The project components ABSORBER and STRIPPER were then sized manually with DTW TRAYED using IPE's expert sizing programs [73].

The cost of each project component is then estimated in order to evaluate the total cost of the whole capture plant. When the components sizing and costing are completed, the next step is to evaluate the total capital investment and total operating cost. IPE would use all the information from the Aspen simulation results and specifications to evaluate these costs automatically. But the capital cost and operating cost reported by IPE is not exactly the total cost of the whole carbon dioxide capture plant. The costs calculated in IPE are based only on the units used in the Aspen Plus™ simulation. For the whole plant cost evaluation, there are more units that must be taken into account.

Table 4-1: Project Component Map Specifications for CO_2 Capture Process

Process Component	Project Component	Definition
BLOWER	DGC CENTRIF	Heavy duty, low noise blower
MCOMPR	DGC CENTRIF	Centrifugal compressor
COOLER	DHE FLOAT HEAD	Floating head shell and tube exchanger
COOLER1	DHE FLOAT HEAD	Floating head shell and tube exchanger
SOL-HEX	DHE FLOAT HEAD	Floating head shell and tube exchanger
SOL-COOL	DHE FLOAT HEAD	Floating head shell and tube exchanger
PUMP1	DCP CENTRIF	Centrifugal pump
PUMP2	DCP CENTRIF	Centrifugal pump
STRIPPER	RB KETTLE	Kettle type reboiler with floating head
	C	Quoted equipment (Overhead splitter)
	C	Quoted equipment (Bottom splitter)
	CP CENTRIF	Centrifugal pump
	HT HORIZ DRUM	Horizontal drum
	DTW TRAYED	Trayed tower
ABSORBER	DTW TRAYED	Trayed tower
SEP	DVT CYLINDER	Vertical process vessel (knock out drum)

4.2 Cost Analysis

Cost analysis is important to investigate the development of any new process or process modification. This section describes the cost analysis of MEA based CO_2 capture process from cement plant flue gases for Case I, Case II, and Case III. The capture cost was evaluated with the lean loading (mol CO_2/mol MEA) of 0.30 for CO_2 recovery of 85%, and CO_2 product purity of 98 percent by mole. The lean loading of 0.30 is chosen for the cost evaluation because the reboiler heat duty was found minimum at this loading and it was considered the most practical design among other lean loadings. Also a sensitivity analysis was also performed for the lean loading varying from 0.05 to 0.30. Cost evaluation was performed using Icarus Process Evaluator (IPE) with the balance of equipment costing from a number of published sources [33] [50] [70] [71]**.** The total cost (capital + operating) of each cases was translated to $/tonne of CO_2 captured.

4.2.1 Assumptions for the MEA based CO_2 Capture Cost Analysis

The important assumptions and specifications used in carrying out the economic evaluation of MEA based CO_2 capture process are given below:

- Currency Description: U.S. DOLLARS (2005)
- Operating Hours per Year: 8000
- Interest Rate: 7 Percent per Year
- Economic Life of Project: 20 Years
- Salvage Value: 20 Percent of Initial Capital Cost
- Depreciation Method: Straight Line
- Labour Cost: $ 20 / hr / Operator
- Supervisor Cost: $ 35 / hr / Supervisor
- Electricity cost = $ 0.06 / kWh
- Steam Cost = $ 9.18 / ton
- Cooling water cost = $ 0.015 / m^3
- MEA cost = $ 1.44 / kg
- Operating Charges: 25 percent of Operating Labor Costs
- Plant Overhead Cost: 50 Percent of Operating Labor and Maintenance Costs
- General and Administrative Expenses: 8 Percent of Subtotal Operating Costs
- Project Capital Escalation: 5 Percent
- Raw Material Escalation: 3.5 Percent
- Operating and Maintenance Labor Escalation: 3 Percent
- Utilities Escalation: 3 Percent
- Working Capital: 5 Percent of Total Capital Investment
- Construction material for absorber and regenerator are stainless steel in order to prevent the corrosion.

4.2.2 Capital Cost

The summary of capital cost for Case I, Case II, and Case III is shown in Table 4-2.

Table 4-2: Summary of Capital Cost for Case I, Case II, and Case III

Parameter	**Cost ($)**		
	Case I	**Case II**	**Case III**
Direct Costs			
Purchased Equipment	7,548,200	6,667,600	5,645,400
Equipment Installation	72,912	67,329	54,714
Piping	2,476,140	2,427,020	2,600,150
Civil	268,463	239,633	267,347
Steel	54,361	53,150	64,580
Instrumentation	413,260	415,111	414,631
Electrical	450,057	432,724	399,835
Insulation	527,775	501,169	500,808
Paint	43,768	41,526	29,747
Other	3,125,300	2,993,500	2,911,100
Total Direct Cost	**14,980,236**	**13,838,762**	**12,888,312**
Indirect Costs			
Total Design, Eng, Procurement cost	1,342,400	1,320,100	1,339,500
G and A Overheads	409,135	375,560	346,464
Contract Fee	505,825	477,628	454,711
Contingencies	2,702,180	2,497,630	2,327,210
Total Indirect Cost	**4,959,540**	**4,670,918**	**4,467,885**
Fixed Capital Cost			
Working Capital	1,160,370	1,072,530	999,350
Total Capital Cost (IPE)	**21,100,146**	**19,582,210**	**18,355,547**
Reclaimer cost	453,266	375,584	258,940
FGD Equipment cost	6,422,852	5,321,624	3,670,108
CO_2 Drying System cost	5,519,731	4,573,744	3,153,290
Grand Total Capital Cost	**33,495,995**	**29,853,162**	**25,437,885**
Annual Capital cost	**3,161,785**	**2,817,927**	**2,401,156**
Salvage value	6,699,199	5,970,632	5,087,577
Annual Salvage value	**163,413**	**145,641**	**124,101**

The basic plant cost was estimated using Icarus Process Evaluator (IPE) and the balance of equipment cost was estimated from published sources. The baghouse, flue gas desulphurisation unit (FGD), and CO_2 dryer were not included in Aspen Plus™ simulation model to avoid the complexity of the simulation. The solution degradation was not considered during the MEA based CO_2 capture simulation process; so, the reclaimer unit was not modelled and simulated. But in reality the cost for the reclaimer has to be included in the cost evaluation. The CO_2 compressor which compresses the product carbon dioxide to 150 bar was estimated using Icarus Process Evaluator (IPE) but the cost of CO_2 drying system to prevent the corrosion in the compressor was not evaluated by IPE. For the equipments that were not simulated, other sources were used. In particular, to evaluate the carbon dioxide capture cost using amine scrubbing, many information were taken from a report from Fluor Daniel [70] [71]. In order to evaluate the capital cost and the operating cost for these additional units, some assumptions were developed and the reference of these assumptions was from the work of Singh et al. [50]. Because the objective was to estimate the cost of the capture process for different amounts of flue gas from this work, the cost factors were calculated based on the mass or mole of the stream related to those units. Table 4-3 shows capital and operating costing factors for the reclaimer, baghouse, flue gas desulphurization (FGD) unit and drying units. The carbon dioxide pipeline transportation and sequestration costs were not considered in the present cost analysis.

Table 4-3: Capital and Operating Cost factors for Reclaimer, Baghouse and FGD and CO_2 Drying Unit [50] [70] [71]**.**

Unit	Capital Cost	Operating Cost
Reclaimer	\$ 242/(kg mol CO_2 produced/hr)	\$ 2.3/(ton CO_2 produced)
Baghouse & FGD	\$ 652/(kg mol flue gas/hr)	\$ 190/(kg mol flue gas/hr)
Drying	\$ 2,947/(kg mol CO_2 produced/hr)	n/a

4.2.3 The Operating Cost

The summary of capital cost for Case I, Case II, and Case III is shown in Table 4-4.

Table 4-4: Summary of Operating Cost for Case I, Case II, and Case III

Parameters	**Cost ($/yr.)**		
	Case I	**Case II**	**Case III**
Manufacturing Costs			
Direct Production Costs			
Raw material cost	1,042,004	859,454	577,200
Utility cost	17,853,401	15,155,159	11,070,838
Operating Labor cost	640,000	640,000	640,000
Operating Supervision cost	280,000	280,000	280,000
Maintenance and Repair cost	356,000	276,000	284,000
Operating Supplies and Laboratory charge (25% of Operating labor costs)	230,000	230,000	230,000
Total Direct Production Cost	**20,401,405**	**17,440,613**	**12,504,838**
Plant Overhead Cost (50% of Operating Labor and maintenance cost)	638,000	598,000	602,000
General and Administrative cost	1,683,152	1,443,089	1,048,547
Operating cost (at second year)	22,722,557	19,481,702	14,155,385
Annual Operating Cost (over 20 years)	**25,968,634**	**22,264,799**	**16,177,581**
Baghouse, FGD Scrubber chemical cost	1,871,690	1,550,780	1,069,510
MEA make-up cost from degradation	1,516,362	1,256,444	866,511
Total Annual Operating Cost	**29,356,686**	**25,072,023**	**18,113,602**

The raw material cost in Table 4-4includes the MEA cost and the process water cost; the utility cost includes the cooling water cost, electricity cost and the steam cost. As can be seen from Table 4-4that in the operating cost the predominant cost the utility cost. The operating cost for reclaimer as MEA make up cost for the solution degradation, baghouse and FGD has to be included for these units as discussed in section 4.2.2.

4.2.4 CO_2 Capture Cost

The total annual cost consists of three costs:

- Amortised capital cost (which is calculated over 20 years, with 7% interest rate) in which the capital recovery factor is used to determine the amount of each future annuity payment required to dissipate a given present value when the interest rate and number of payments are known [74].
- Annual salvage value which is calculated over 20 years, 7% interest rate and straight line depreciation method is used. Salvage value is assumed as 20 percent of initial capital cost.
- Annual operating cost which includes the raw material cost, utility cost, operating labour and supervision cost, maintenance and repair cost, operating charges, plant overhead cost, general and administrative costs. Operating cost at second year is obtained from IPE and annual operating cost over 20 years is calculated using the annuity factor as 14.2857[73] [74].

The total annual cost for the MEA based CO_2 capture plant is calculated as follows:

Total annual cost = Annual operating cost + Amortised capital cost – Annual salvage value

CO_2 capture cost = Total annual cost/ Total CO_2 Captured

Table 4-5 summarizes the annual capital cost, annual salvage value, annual operating cost, total annual cost and the CO_2 capture cost per tonne of CO_2 captured for Case I, Case II and Case III.

Table 4-5: CO_2 Capture cost for Case I, Case II and Case III

Parameters	Case I	Case II	Case III
Annual Capital Cost	3,161,785	2,817,927	2,401,156
Annual Salvage value	163,413	145,641	124,101
Annual Operating Cost	29,356,686	25,072,023	18,113,602
Total Annual Cost ($)	**32,355,058**	**27,744,309**	**20,390,657**
Total Annual CO_2 Captured (tonne)	659,288	546,280	376,744
$ / tonne of CO_2 Captured	**49**	**51**	**54**

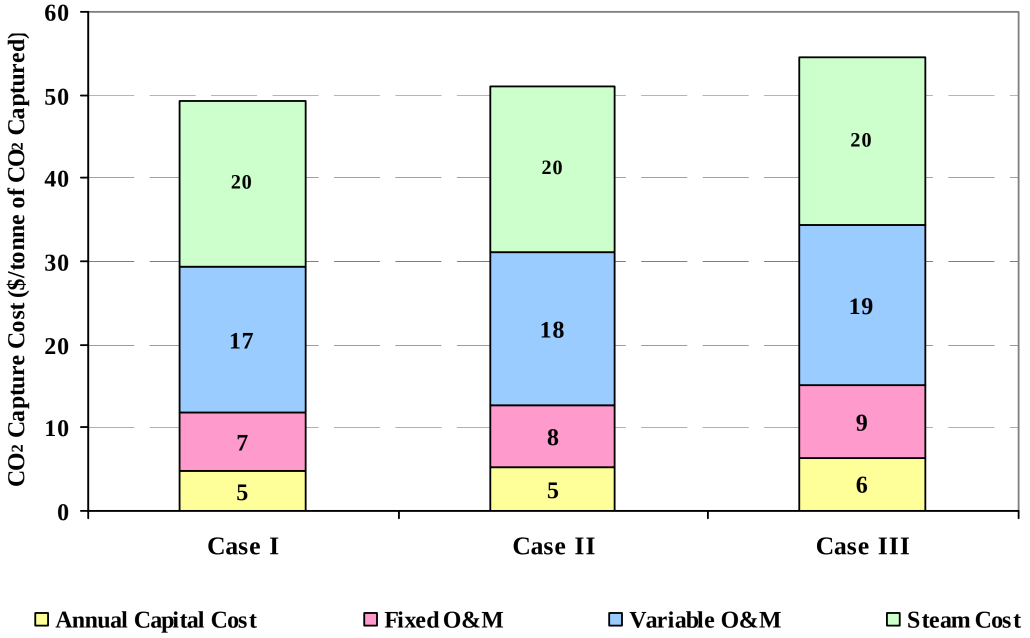

Figure 4-1: Breakdown of CO_2 capture cost per tonne of CO_2 captured for Case I, Case II, and Case III

The CO_2 capture cost are USD 49, 51, and 54 per tonne of CO_2 captured for Case I, Case II, and Case III respectively. Figure 4-1 shows the breakdown of CO_2 capture cost per tonne of CO_2 captured in terms of annual capital cost, fixed operating and maintenance cost, variable operating and maintenance cost and steam cost. As can be seen that for the three cases the steam cost is the highest capture cost in CO_2 capture process and steam cost represents 41%, 39%, and 37% of total CO_2 capture process for Case I, Case II, and Case III, respectively. The variable operating and maintenance cost includes the raw material cost, maintenance and repair cost, plant overhead cost and general and administrative cost is the second highest cost in CO_2 capture process. For Case I, Case II, and Case III the variable operating and maintenance cost represents 35%, 35%, and 35% of the total CO_2 capture cost respectively. The fixed operating and maintenance cost including the operating labour and supervision cost, operating supplies and laboratory charges, baghouse and FGD scrubber chemical cost, and MEA degradation cost, is the third highest cost for the capture process. Overall, the operating cost for CO_2 capture rocess represents 90%, 90%, and 89% of total CO_2 capture cost for Case I, Case II, and Case III, respectively.

4.3 Sensitivity Analysis of CO_2 Capture Cost to Lean Loading

The effect of lean loading on cost of CO_2 capture was studied for case II. The CO_2 recovery was specified at 85 percent and the carbon dioxide product purity at 98 percent by mole. Table 4-6 (a)-(b) shows the summary of capital cost for lean loading (mol CO_2/mol MEA) from 0.05 to 0.30 for Case II. As can be seen, at low lean loading of 0.05 the annual capital cost is the maximum and at lean loading of 0.30 the annual capital cost is minimum. This is due to the fact that at low lean loading of 0.05, huge amount of

Table 4-6 (a): Summary of capital cost for lean loading from 0.05 to 0.15 (Case II)

Parameters	Cost ($)		
	0.05	**0.10**	**0.15**
Direct Costs			
Purchased Equipment	88,432,800	65,496,933	10,324,100
Equipment Installation	2,636,320	1,962,712	112,828
Piping	5,204,580	4,359,448	5,739,840
Civil	11,825,800	9,510,307	763,841
Steel	276,412	211,452	106,606
Instrumentation	467,861	414,947	487,831
Electrical	439,683	486,104	444,477
Insulation	7,901,330	5,894,730	1,012,180
Paint	40,973	30,963	36,547
Other	22,227,800	17,109,509	4,534,900
Total Direct Cost	**139,453,559**	**105,477,104**	**23,563,150**
Indirect Costs			
Total Design, Eng, Procurement Cost	3,114,500	2,314,172	1,502,400
G and A Overheads	4,090,170	3,094,883	661,822
Contract Fee	4,010,920	3,060,552	822,258
Contingencies	25,084,300	18,977,546	4,258,030
Total Indirect Cost	**36,299,890**	**27,447,153**	**7,244,510**
Fixed Capital Cost			
Working Capital	10,258,700	7,761,227	1,741,410
Total Capital Cost (IPE)	**186,012,149**	**140,685,484**	**32,549,070**
Reclaimer cost	375,584	375,584	375,584
FGD Equipment cost	5,321,624	5,321,624	5,321,624
CO_2 Drying System cost	4,573,744	4,573,744	4,573,744
Grand Total Capital Cost	**196,283,101**	**146,986,681**	**42,820,022**
Annual Capital cost	**18,527,736**	**13,874,503**	**4,041,907**
Salvage value	39,256,620	29,397,336	8,564,004
Annual Salvage value	**957,584**	**717,087**	**208,901**

Table 4-6 (b): Summary of capital cost for lean loading from 0.20 to 0.30 (Case II)

Parameters	Cost ($)		
	0.20	**0.25**	**0.30**
Direct Costs			
Purchased Equipment	8,949,000	8,271,600	6,667,600
Equipment Installation	94,800	78,553	67,329
Piping	4,553,770	4,830,450	2,427,020
Civil	601,212	509,724	239,633
Steel	90,306	88,777	53,150
Instrumentation	454,483	476,549	415,111
Electrical	432,626	439,683	432,724
Insulation	887,232	737,286	501,169
Paint	39,259	33,268	41,526
Other	3,945,100	3,945,300	2,993,500
Total Direct Cost	**20,047,788**	**19,411,190**	**13,838,762**
Indirect Costs			
Total Design, Eng, Procurement Cost	1,422,400	1,445,500	1,320,100
G and A Overheads	558,762	538,971	375,560
Contract Fee	714,393	699,891	477,628
Contingencies	3,624,560	3,510,510	2,497,630
Total Indirect Cost	**6,320,115**	**6,194,872**	**4,670,918**
Fixed Capital Cost			
Working Capital	1,556,460	1,507,480	1,072,530
Total Capital Cost (IPE)	**27,924,363**	**27,113,542**	**19,582,210**
Reclaimer cost	375,584	375,584	375,584
FGD Equipment cost	5,321,624	5,321,624	5,321,624
CO_2 Drying System cost	4,573,744	4,573,744	4,573,744
Grand Total Capital Cost	**38,195,315**	**37,384,494**	**29,853,162**
Annual Capital cost	**3,605,368**	**3,528,832**	**2,817,927**
Salvage value	7,639,063	7,476,899	5,970,632
Annual Salvage value	**186,339**	**182,383**	**145,641**

reboiler heat duty is required in order to separate the carbon dioxide until the lean solvent reaches that loading. The reboiler would consume a lot of energy even though the MEA solvent circulation rate is low. If the reboiler heat duty is increased, it would produce high amount of vapour flowing back to the bottom of the stripper column and this leads to an increase in stripper diameter and consequently the direct costs, indirect cost and fixed capital cost increases. As a result the annual capital cost increases and the minimum annual cost is obtained at lean loading of 0.30.

Table 4-7(a): Summary of operating cost for lean loading from 0.05 to 0.15 (Case II)

Parameters	**Cost ($)**		
	0.05	**0.10**	**0.15**
Manufacturing Costs			
Direct Production Costs			
Raw material cost	911,008	922,922	956,043
Utility cost	76,822,945	39,464,477	25,671,256
Operating Labor cost	640,000	640,000	640,000
Operating Supervision cost	280,000	280,000	280,000
Maintenance and Repair cost	1,290,000	454,000	682,000
Operating Supplies and Laboratory charge (25% of Operating labor costs)	230,000	230,000	230,000
Total Direct Production Cost	**80,173,953**	**41,991,399**	**28,459,299**
Plant Overhead Cost (50% of Operating Labor and maintenance cost)	1,105,000	687,000	801,000
General and Administrative cost	6,502,316	3,414,272	2,340,824
Operating cost (at second year)	87,781,269	46,092,671	31,601,123
Annual Operating Cost (over 20 years)	**100,321,438**	**52,677,331**	**36,115,565**
Baghouse, FGD Scrubber chemical cost	1,550,780	1,550,780	1,550,780
MEA make-up cost from degradation	1,256,444	1,256,444	1,256,444
Total Annual Operating Cost	**103,128,662**	**55,484,555**	**38,922,789**

Table 4-7(b): Summary of operating cost for lean loading from 0.20 to 0.30 (Case II)

Parameters	**Cost ($)**		
	0.20	**0.25**	**0.30**
Manufacturing Costs			
Direct Production Costs			
Raw material cost	996,329	757,555	859,454
Utility cost	18,161,251	15,524,925	15,155,159
Operating Labor cost	640,000	640,000	640,000
Operating Supervision cost	280,000	280,000	280,000
Maintenance and Repair cost	980,000	480,000	276,000
Operating Supplies and Laboratory charge (25% of Operating labor costs)	230,000	230,000	230,000
Total Direct Production Cost	**21,287,580**	**17,912,480**	**17,440,612**
Plant Overhead Cost (50% of Operating Labor and maintenance cost)	950,000	700,000	598,000
General and Administrative cost	1,779,006	1,488,998	1,443,089
Operating cost (at second year)	24,016,587	20,101,478	19,481,701
Annual Operating Cost (over 20 years)	**27,447,524**	**22,973,115**	**22,264,799**
Baghouse, FGD Scrubber chemical cost	1,550,780	1,550,780	1,550,780
MEA make-up cost from degradation	1,256,444	1,256,444	1,256,444
Total Annual Operating Cost	**30,254,748**	**25,780,339**	**25,072,023**

Table 4-7 (a)-(b) above shows the summary of operating cost for lean loading (mol CO_2/mol MEA) from 0.05 to 0.30 for Case II. The major operating cost in CO_2 capture process is the steam consumption by the reboiler. The reboiler heat duty decreases from 508 MW to 69 MW when the lean loading is increased from 0.05 to 0.30. Two terms affect the value of the reboiler heat duty: the first one is the circulation rate of the solvent and the second one is the value of the lean loading of the stream leaving the bottom of the stripper and recycled back to the absorber. As huge amount of heat is consumed by the reboiler at low lean loading of 0.05, so the annual operating cost is maximum at lean loading of 0.05 and the operating cost decreases with increasing lean loading and it is minimum at lean loading of 0.30.

Table 4-8 (a)-(b) represents the CO_2 capture cost per tonne of CO_2 captured for lean loading from 0.05 to 0.30 for Case II. The trend of the CO_2 capture cost is the same as the trend of the reboiler heat duty because the steam utility cost at the reboiler represents around 60% to 80% of the total annual capture cost.

Table 4-8 (a): CO_2 capture cost for lean loading from 0.05 to 0.15 (Case II)

Parameters	**0.05**	**0.10**	**0.15**
Annual Capital Cost	18,527,736	13,874,503	4,041,907
Annual Salvage value	957,584	717,087	208,901
Annual Operating Cost	103,128,662	55,484,555	38,922,789
Total Annual Cost ($)	**120,698,814**	**66,134,177**	**42,755,795**
Total Annual CO_2 Captured (tonne)	546,280	546,280	546,280
($) / tonne of CO_2 Captured	**221**	**121**	**78**

Table 4-8 (b): CO_2 capture cost for lean loading from 0.20 to 0.30 (Case II)

Parameters	**0.20**	**0.25**	**0.30**
Annual Capital Cost	3,605,368	3,528,832	2,817,927
Annual Salvage value	186,339	182,383	145,641
Annual Operating Cost	33,673,777	29,126,788	27,744,309
Total Annual Cost ($)	**30,254,748**	**25,780,339**	**25,072,023**
Total Annual CO_2 Captured (tonne)	546,280	546,280	546,280
($) / tonne of CO_2 Captured	**62**	**53**	**51**

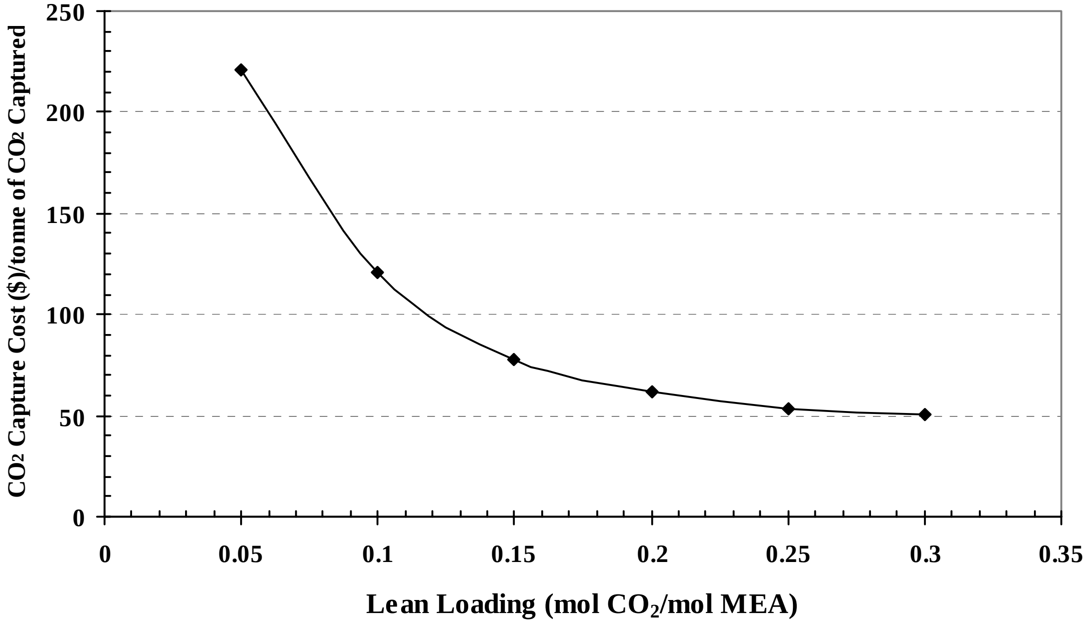

Figure 4-2: CO_2 capture cost per tonne of CO_2 captured at various lean loading (Case II)

Figure 4-2 shows the CO_2 capture cost per tonne of CO_2 captured at lean loading from 0.05 to 0.30. As can be seen from this figure, the lean loading that gives the minimum CO_2 capture cost per tonne of CO_2 captured for a CO_2 recovery of 85 percent is 0.30 which corresponds to a cost of $ 51 per tonne of carbon dioxide captured and the capture cost is maximum at lean loading of 0.05 which corresponds to CO_2 capture cost of $ 221 per tonne of CO_2 captured.

Figure 4-3 is a breakdown of CO_2 capture cost per tonne of CO_2 captured at different lean loading for Case II. It is seen from Figure 4-3 that at lean loading of 0.05 the steam cost corresponds to 78 percent of the total operating cost and 67 percent of the total capture cost. At a lean loading of 0.30 the steam cost carries about 43.5 percent of total operating cost and 39.2 percent of total CO_2 capture cost respectively which represents the minimum cost for this case.

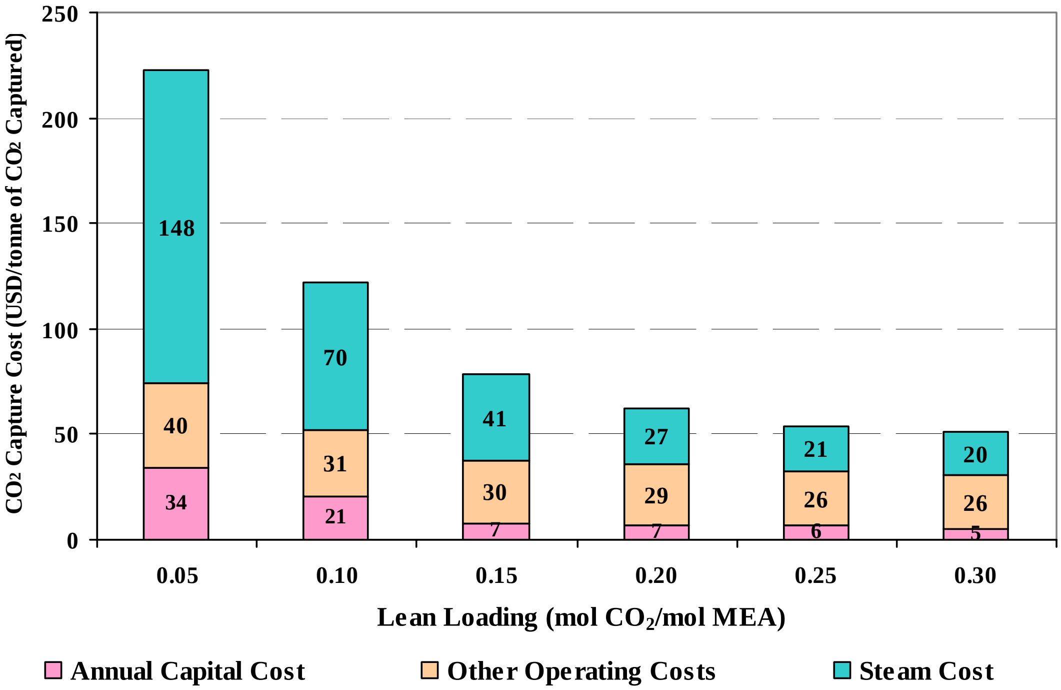

Figure 4-3: Break down of CO_2 capture cost per tonne of CO_2 captured at various lean loading (Case II)

4.4 Comparison of CO_2 Capture Cost with Other Studies

Many researchers dealt with the design and cost of CO_2 capture for fossil fuel- based power plants. The results of these studies (i.e. flue gas from burning of coal) are used for comparison with the data obtained in the present study. CO_2 capture costs are compared to the studies from the University of Waterloo (Singh et al., 2003 [50]), Mariz (1998) [71], and Alstom (Nsakala et al., 2001 [32]). Figure 4-3 compares the different results in terms of USD per tonne of CO_2 avoided for the amine case and it ranges from USD 49 to USD 55 per tonne of CO_2 avoided.

CO_2 Capture from an existing coal fired power plant was studied by Singh et al (2003) [50] with the conventional amine scrubbing approach as well as an emerging alternative commonly known as O_2/CO_2 recycle combustion. The cost of CO_2 capture however, for the amine case was reported to USD 53 per tonne of CO_2 avoided. The CO_2 composition in the flue gas was 14.59% on molar basis.

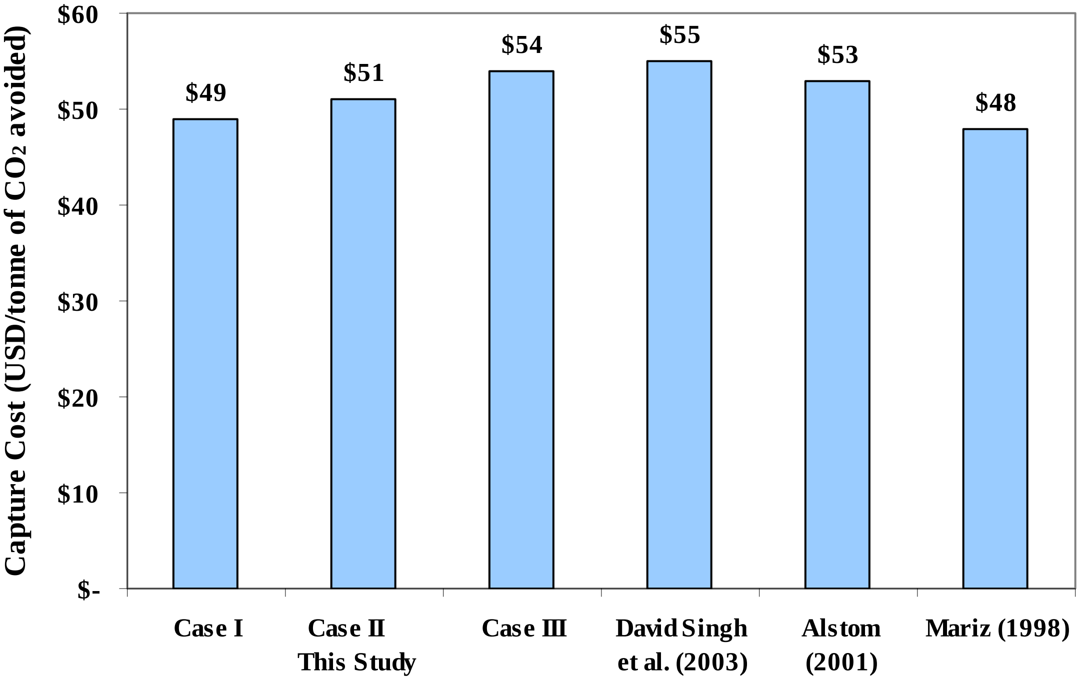

Figure 4-4: Comparison of CO_2 capture cost

Mariz (1998) [71] illustrates the variation in total plant cost with volume percent of CO_2 in the flue gas for a fixed plant size of 290 ton/day. There is an increase of about 14% in the total plant cost when the CO_2 composition of the flue gas varies from 13% to 8%. A summary of the operating costs for a 13 vol% CO_2 in the flue gas producing 907 ton/day CO_2 is also provided by the author. Mariz (1998) also suggested economic scaling in approximating the total plant cost relative to a fixed plant size. The CO_2 capture cost with this condition based on the data given by Mariz (1998) is approximately $48/ton CO_2 captured.

In the Alstom case (Nsakala et al., 2001 [32]), which had a higher capital cost and lower operating cost, reported a CO_2 capture cost of USD 53 per tonne of CO_2 captured.

The value obtained in this study is $ 49, 51 and 54 per tonne CO_2 captured for Case I, Case II and Case III, respectively which lies in the range estimated by other researchers. This points out that, although the concentration of CO_2 in the flue gas from the cement plant was higher than in the case of power plant, it did not help reducing the capture cost.

Chapter 5: Conclusions

The design and costing of a MEA based CO_2 capture process for cement plant was studied using Aspen Plus™ and Icarus Process Evaluator (IPE). Four cases were considered all to reach a CO_2 purity of 98% i) the plant operates at the highest capacity ii) the plant operates at average load iii) the plant operates at minimum operating capacity and iv) fuel switching to a lower carbon content fuel at average plant load. A number of important conclusions can be drawn with respect to the validity and performance of MEA based CO_2 capture process for cement plant.

1. Design and costing of CO_2 capture from cement plant flue gas is similar to design and costing of capturing carbon dioxide from power plant flue gases by using MEA based CO_2 absorption process.
2. The CO_2 capture cost per tonne of CO_2 captured was found \$ 49, 51, and 54 for Case I, Case II, and Case III respectively. The operating cost for capturing CO_2 represented approximately 90% of total CO_2 capture cost for all cases. The stem cost was the highest cost in CO_2 capture process and represented 41%, 39%, and 37% of total CO_2 capture process for Case I, Case II, and Case III respectively. The variable operating and maintenance cost was the second highest cost in CO_2 capture process and represented around 35% of the total CO_2 capture cost for three cases.
3. Lots of waste heat is available in cement making process but it is too difficult and costly to recover waste heat to generate steam to be used in the reboiler for CO_2 capture process. If the plant operates at regular load, around 37 MW of waste heat per hour can be recovered from St. Marys cement but unfortunately, because of the process constraints, this amount of heat is wasted. By using improved technology this amount of waste heat can be recovered to produce steam.
4. Switching high carbon content fuel like coal, petcoke, coal fines to natural gas, the carbon dioxide emissions can be reduced by 18.55 percent at the average plant load operation.

5. The capture cost obtained in this study was within the range obtained by other researchers. Singh, et al (2003) with the conventional amine scrubbing reported the CO_2 capture cost USD 53 per tonne of CO_2 avoided. Mariz (1998) reported the CO_2 capture cost approximately \$48/ton CO_2 captured. The value obtained in this study lied in the range estimated by other researchers.

CO_2 capture is dependent on economic recovery of heat to satisfy parasitic heat requirements of MEA process. To decide what is the best operating condition for carbon dioxide capture process for cement plant, not only the minimum capture cost should be considered, but also the maximum steam that can be supplied and the maximum annual cost per year that can be paid.

Chapter 6: Recommendations

Based on the results of this study, the following recommendations are made.

1. MEA based CO_2 capture process is an energy intensive process and large amount of heat is required by the reboiler to separate CO_2. In cement plants huge amounts of heat are wasted but due to process constraints cannot be recovered. A detailed design needs to be investigated to recover this waste heat to generate steam by using improved technology.
2. For further study, the whole capture plant which consists of Baghouse, FGD unit, CO_2 dryer should be incorporated in the simulation process. The cost should be evaluated together to augment the accuracy of the estimation.
3. In this study only the MEA based CO_2 capture process was studied but other available CO_2 separation techniques like membrane separation, cryogenic separation, hybrid process, and physical absorption using Selexol should be studied to find out the best capture process for capturing CO_2 from flue gas.
4. In this study, monoethanolamine (MEA) was selected as viable solvent for capturing carbon dioxide from flue gases. The effect of other solvents such as Diethanolamine (DEA), Diisopropanolamine (DIPA), Diglycolamine (DGA), Methyldiethanolamine (MDEA), and Triethanolamine (TEA) on the CO_2 capture cost should to be studied.
5. In this study the absorber and stripper pressure was specified at 1.2 bar and 1.9 bar, respectively. The effect of absorber and stripper pressure on CO_2 capture cost needs to be studied to find out the optimum absorber and stripper pressures.
6. Evaluate the design and cost by using plate-type absorber and a stripper with other types of trays (i.e. valve tray, sieve tray, tray spacing, weir height etc.) as well as structured packing which has been shown studied to have a greater mass transfer coefficient and higher surface area.

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Appendix A: Aspen Simulation Report

```
ASPEN PLUS   PLAT: WIN32     VER: 12.1                              02/28/2005

                          CO2 CAPTURE PROCESS FOR CASE II
                               RUN CONTROL SECTION

 RUN CONTROL INFORMATION
 -----------------------------------------

 THIS COPY OF ASPEN PLUS LICENSED TO UNIV OF WATERLOO

 TYPE OF RUN: NEW

 INPUT FILE NAME: _4444mfu.inm

 OUTPUT PROBLEM DATA FILE NAME: _4444mfu
 LOCATED IN:

 PDF SIZE USED FOR INPUT  TRANSLATION:

   NUMBER OF FILE RECORDS (PSIZE) =     0
   NUMBER OF IN-CORE RECORDS       =   256
   PSIZE NEEDED FOR SIMULATION    =   256

 CALLING PROGRAM NAME:         apmain

 LOCATED IN:                   C:\PROGRA~1\ASPENT~1\ASPENP~2.1\Engine\xeq

 SIMULATION REQUESTED FOR ENTIRE FLOWSHEET

 DESCRIPTION
 -------------------
      H2O-MEA-H2S-CO2 PROPERTY METHOD: ELECNRTL TEMPERATURE: UP TO 120
      C MEA CONCENTRATION UP TO 50 WT.%

 FLOWSHEET CONNECTIVITY BY STREAMS
 ---------------------------------------------------------

   STREAM      SOURCE       DEST         STREAM       SOURCE       DEST
   FLUEGAS       ----       COOLER       WATER          ----       MIXER
   MEA           ----       MIXER        FLUECOOL     COOLER       SEP
   FLUESEP       SEP        BLOWER        SEPOUT       SEP          ----
   FLUEBLOW    BLOWER       COOLER1       FLUE-ABS    COOLER1      ABSORBER
   TREATGAS    ABSORBER       ----       RICH-MEA     ABSORBER     PUMP1
   RICH-P1     PUMP1        SOL-HEX      LEANSTRC      SOL-HEX     MIXER
   RICHSTPR    SOL-HEX      STRIPPER     LEAN-P2      PUMP2        SOL-HEX
   CO2         STRIPPER     MCOMPR       LEANSTPR     STRIPPER     PUMP2
   LEANMEX     MIXER        SOL-COOL     LEAN-MEA     SOL-COOL     ABSORBER
    PRODCO2    MCOMPR         ----
```

```
FLOWSHEET CONNECTIVITY BY BLOCKS
------------------------------------------------------

 BLOCK           INLETS                        OUTLETS
 COOLER          FLUEGAS                       FLUECOOL
 SEP             FLUECOOL                      FLUESEP SEPOUT
 BLOWER          FLUESEP                       FLUEBLOW
 COOLER1         FLUEBLOW                      FLUE-ABS
 ABSORBER        FLUE-ABS     LEAN-MEA         TREATGAS RICH-MEA
 PUMP1           RICH-MEA                      RICH-P1
 SOL-HEX         LEAN-P2      RICH-P1          LEANSTRC RICHSTPR
 PUMP2           LEANSTPR                      LEAN-P2
 STRIPPER        RICHSTPR                      CO2 LEANSTPR
 MIXER           LEANSTRC     MEA WATER        LEANMEX
 SOL-COOL        LEANMEX                       LEAN-MEA
 MCOMPR          CO2                           PRODCO2

CONVERGENCE STATUS SUMMARY
-----------------------------------------------

 DESIGN-SPEC SUMMARY
 ====================

 DESIGN                                                                CONV
 SPEC          ERROR          TOLERANCE      ERR/TOL       VARIABLE    STAT   BLOCK
 --------      ----------     -----------    -----------   ----------  -----  ----------

MEACON        -0.21544E-08    0.10000E-03   -0.21544E-04   9223.6       #     $OLVER01
 RECOVERY      0.41429E-04    0.10000E-03    0.41429       12.047       #     $OLVER02

 TEAR STREAM SUMMARY
 ====================

 STREAM        MAXIMUM                      MAXIMUM      VARIABLE                  CONV
 ID            ERROR          TOLERANCE     ERR/TOL      ID                  STAT  BLOCK
 ------        -------        ---------     --------     --------            ----  -----

LEAN-P2       0.13322E-04    0.66565E-04   0.20014      CO2 MOLEFLOW          #    C-1
 LEANSTRC     0.16758E-05    0.12201E-04   0.13735      MASS ENTHALPY         #    C-2

 # = CONVERGED
 * = NOT CONVERGED

 LB = AT LOWER BOUNDS
 UB = AT UPPER BOUNDS

DESIGN-SPEC:  MEACON
--------------------------------

SAMPLED VARIABLES:

   MEACON   : MEA MASSFLOW IN STREAM LEANMEX SUBSTREAM MIXED
   WATER    : H2O MASSFLOW IN STREAM LEANMEX SUBSTREAM MIXED
```

```
SPECIFICATION:

   MAKE MEACON / (MEACON+WATER)          APPROACH       0.30000
    WITHIN                  0.00010000

  MANIPULATED VARIABLES:

    VARY     : TOTAL MASSFLOW IN STREAM WATER SUBSTREAM MIXED
    LOWER LIMIT =          1.00000                                   KG/HR
    UPPER LIMIT =         50,000.0                                   KG/HR
    FINAL VALUE =         9,223.58                                   KG/HR

  VALUES OF ACCESSED FORTRAN VARIABLES:

  VARIABLE                VALUE AT START       FINAL VALUE            UNITS
                             OF LOOP
   --------------         -------------------------    -------------------          ---------
   MEACON                     487366               487366               KG/HR
   WATER                    0.113719E+07         0.113719E+07           KG/HR


DESIGN-SPEC:  RECOVERY
 -------------------------------------

  SAMPLED VARIABLES:

    CO2IN         : CO2 MOLEFLOW  IN  STREAM  FLUE-ABS   SUBSTREAM  MIXED
   CO2OUT        : CO2 MOLEFLOW  IN  STREAM  TREATGAS  SUBSTREAM  MIXED

  SPECIFICATION:

    MAKE (CO2IN-CO2OUT)/CO2IN        APPROACH       0.85000
     WITHIN                  0.00010000

  MANIPULATED VARIABLES:

    VARY     : TOTAL MASSFLOW IN STREAM MEA SUBSTREAM MIXED
   LOWER LIMIT =                  1.00000                            KG/HR
   UPPER LIMIT =                  100.000                            KG/HR
   FINAL VALUE =                  12.0470                            KG/HR

  VALUES OF ACCESSED FORTRAN VARIABLES:

   VARIABLE               VALUE AT START       FINAL VALUE            UNITS
                             OF LOOP
   ---------------        -------------------------    -------------------           --------
   CO2IN                     1825.36              1825.36             KMOL/HR
   CO2OUT                    273.728              273.728             KMOL/HR
```

```
CONVERGENCE BLOCK:  C-1
 ----------------------------------------
    Tear Stream:            LEAN-P2
    Tolerance used:         0.100D-03
    Trace molefrac:         0.100D-05

    MAXIT=   50 WAIT    1 ITERATIONS BEFORE ACCELERATING
    QMAX = 0.00E+00 QMIN = -5.0
    METHOD: WEGSTEIN      STATUS: CONVERGED
    TOTAL NUMBER OF ITERATIONS:     1

                  *** FINAL VALUES ***

   VARIABLE                          VALUE        PREV VALUE             ERR/TOL

 TOTAL MOLEFLOW     KMOL/HR         7.2986+04        7.2986+04              9.0801-03
 H2O MOLEFLOW       KMOL/HR         6.2612+04        6.2612+04              2.9301-03
 MEA MOLEFLOW       KMOL/HR         7978.4633        7978.4633              -4.3192-05
 H2S MOLEFLOW       KMOL/HR         0.0              0.0                    0.0
 CO2 MOLEFLOW       KMOL/HR         2396.3739        2396.3259              0.2001
 HCO3- MOLEFLOW     KMOL/HR         0.0              0.0                    0.0
 MEACOO- MOLEFLOW KMOL/HR           0.0              0.0                    0.0
 MEA+ MOLEFLOW      KMOL/HR         0.0               0.0                   0.0
 CO3-2 MOLEFLOW     KMOL/HR         0.0              0.0                    0.0
 HS- MOLEFLOW        KMOL/HR        0.0              0.0                    0.0
 S-2 MOLEFLOW       KMOL/HR         0.0               0.0                   0.0
 H3O+ MOLEFLOW      KMOL/HR         0.0               0.0                   0.0
 OH- MOLEFLOW       KMOL/HR         0.0              0.0                    0.0
 N2 MOLEFLOW        KMOL/HR         1.5665-06        1.4944-06              482.2252     T
 O2 MOLEFLOW        KMOL/HR         7.6531-07        7.3343-07              434.6528     T
 PRESSURE           ATM             2.0725           2.0725                 0.0
 MASS ENTHALPY      CAL/GM          -2870.2284       -2870.1972             -0.1086

    T - SIGNIFIES COMPONENT IS A TRACE COMPONENT

                  *** ITERATION HISTORY ***

    TEAR STREAMS:

    ITERATION           MAX-ERR/TOL           STREAM ID    VARIABLE
    ----------------    --------------------    ---------------    ----------------
         1                  0.2001               LEAN-P2    CO2 MOLEFLOW

 CONVERGENCE BLOCK:  C-2
 ----------------------------------------
    Tear Stream:            LEANSTRC
    Tolerance used:         0.100D-03
    Trace molefrac:         0.100D-05

   MAXIT=   50 WAIT     1 ITERATIONS BEFORE ACCELERATING
    QMAX = 0.00E+00 QMIN = -5.0
    METHOD: WEGSTEIN      STATUS: CONVERGED
    TOTAL NUMBER OF ITERATIONS:     1
```

```
                    *** FINAL VALUES ***

  VARIABLE                           VALUE        PREV VALUE     ERR/TOL

TOTAL MOLEFLOW     KMOL/HR       7.2986+04       7.2986+04       -7.8311-05
H2O MOLEFLOW       KMOL/HR       6.2612+04       6.2612+04       -8.8655-05
MEA MOLEFLOW       KMOL/HR       7978.4633       7978.4633       -2.0519-05
H2S MOLEFLOW       KMOL/HR       0.0             0.0             0.0
CO2 MOLEFLOW       KMOL/HR       2396.3259       2396.3259       0.0
HCO3- MOLEFLOW      KMOL/HR      0.0             0.0             0.0
MEACOO- MOLEFLOW KMOL/HR         0.0             0.0             0.0
MEA+ MOLEFLOW      KMOL/HR       0.0             0.0             0.0
CO3-2 MOLEFLOW     KMOL/HR       0.0             0.0             0.0
HS- MOLEFLOW       KMOL/HR       0.0             0.0             0.0
S-2 MOLEFLOW       KMOL/HR       0.0             0.0             0.0
H3O+ MOLEFLOW      KMOL/HR       0.0             0.0             0.0
OH- MOLEFLOW       KMOL/HR       0.0             0.0             0.0
N2 MOLEFLOW        KMOL/HR       1.4944-06       1.5667-06       -461.6191    T
O2 MOLEFLOW        KMOL/HR       7.3343-07       7.6540-07       -417.6874    T
PRESSURE           ATM           2.0725          2.0725          0.0
MASS ENTHALPY      CAL/GM        -2914.0672      -2914.1072      0.1374

   T - SIGNIFIES COMPONENT IS A TRACE COMPONENT

                    *** ITERATION HISTORY ***

   TEAR STREAMS:

   ITERATION         MAX-ERR/TOL         STREAM ID    VARIABLE
   ----------------  ------------------  -----------  ------------
        1                 0.1374         LEANSTRC    MASS ENTHALPY

CONVERGENCE BLOCK:  $OLVER01
-----------------------------------------------
   SPECS: MEACON
   MAXIT=   30          STEP-SIZE=            1.0000            % OF RANGE
                        MAX-STEP=             100.              % OF RANGE
                        XTOL=                 1.000000E-08
   THE NEW ALGORITHM WAS USED WITH BRACKETING=NO
   METHOD: SECANT             STATUS: CONVERGED
   TOTAL NUMBER OF ITERATIONS:                1

                    *** FINAL VALUES ***

  VARIABLE                            VALUE         PREV VALUE     ERR/TOL

TOTAL MASSFL     KG/HR               9223.5780       9223.5780       -2.1544-05
```

```
                    *** ITERATION HISTORY ***

DESIGN-SPEC ID: MEACON

   ITERATION         VARIABLE              ERROR         ERR/TOL
   ---------------   ---------------       ----------    -------------
        1               9224               -0.2154E-08   -0.2154E-04

 CONVERGENCE BLOCK:  $OLVER02
 ------------------------------------------------
   SPECS: RECOVERY

   MAXIT=   30       STEP-SIZE=            1.0000        % OF RANGE
                     MAX-STEP=             100.          % OF RANGE
                     XTOL=                 1.000000E-08

   THE NEW ALGORITHM WAS USED WITH BRACKETING=NO
   METHOD: SECANT            STATUS: CONVERGED
   TOTAL NUMBER OF ITERATIONS:             1

                    *** FINAL VALUES ***

  VARIABLE                          VALUE        PREV VALUE          ERR/TOL

 TOTAL MASSFL        KG/HR          12.0470      12.0470             0.4143

                    *** ITERATION HISTORY ***

   DESIGN-SPEC ID: RECOVERY

   ITERATION         VARIABLE              ERROR         ERR/TOL
   ----------------  ---------------       ----------    ------------

      1                 12.05              0.4143E-04    0.4143

 COMPUTATIONAL SEQUENCE
 ----------------------------------------

 SEQUENCE USED WAS:
   COOLER SEP BLOWER COOLER1
   C-2
   |       $OLVER02
   |       |                 $OLVER01 MIXER
   |       |                 (RETURN $OLVER01)
   |       |                 SOL-COOL ABSORBER
   |       (RETURN $OLVER02)
   |       PUMP1
   |       C-1 SOL-HEX STRIPPER PUMP2
   |       (RETURN C-1)
   (RETURN C-2)
   MCOMPR
```

```
OVERALL FLOWSHEET BALANCE
-----------------------------------------------

            ***  MASS AND ENERGY BALANCE  ***
                                            IN              OUT              RELATIVE DIFF

  CONVENTIONAL COMPONENTS (KMOL/HR)
    H2O                                 1102.12         1102.10          0.161425E-04
    MEA                                 0.197221        0.197272          -0.257672E-03
    H2S                                 0.000000E+00    0.000000E+00     0.000000E+00
    CO2                                 1825.39         1825.34          0.262742E-04
    HCO3-                               0.000000E+00    0.000000E+00     0.000000E+00
    MEACOO-                             0.000000E+00    0.000000E+00     0.000000E+00
    MEA+                                0.000000E+00    0.000000E+00     0.000000E+00
    CO3-2                               0.000000E+00    0.000000E+00     0.000000E+00
    HS-                                 0.000000E+00    0.000000E+00     0.000000E+00
    S-2                                 0.000000E+00    0.000000E+00     0.000000E+00
    H3O+                                0.000000E+00    0.000000E+00     0.000000E+00
    OH-                                 0.000000E+00    0.000000E+00     0.000000E+00
    N2                                  5555.38         5555.38          -0.121337E-10
    O2                                  191.217         191.217          -0.125239E-10

  TOTAL BALANCE
    MOLE (KMOL/HR)                      8674.30         8674.24          0.757417E-05
    MASS (KG/HR   )                     261946.         261944.          0.926961E-05
    ENTHALPY (CAL/SEC)                  -0.644611E+08  -0.661614E+08     0.256992E-01

COMPONENTS
--------------------

  ID          TYPE  FORMULA               NAME OR ALIAS          REPORT NAME
 H2O          C     H2O                         H2O                   H2O
 MEA          C     C2H7NO                      C2H7NO                MEA
 H2S          C     H2S                         H2S                   H2S
 CO2          C     CO2                         CO2                   CO2
 HCO3-        C     HCO3-                       HCO3-                 HCO3-
 MEACOO-      C     MISSING                     MISSING               MEACOO-
 MEA+         C     MISSING                     MISSING               MEA+
 CO3-2        C     CO3-2                       CO3-2                 CO3-2
 HS-          C     HS-                         HS-                   HS-
 S-2          C     S-2                         S-2                   S-2
 H3O+         C     H3O+                        H3O+                  H3O+
 OH-          C     OH-                         OH-                   OH-
 N2           C     N2                          N2                    N2
 O2           C     O2                          O2                    O2

 LISTID       SUPERCRITICAL COMPONENT LIST
  EMEA        CO2 H2S

BLOCK:  ABSORBER MODEL: RATEFRAC
-------------------------------------------------------------

 INLETS               - FLUE-ABS COLUMN         1        SEGMENT      4
                        LEAN-MEA COLUMN         1        SEGMENT      1
```

```
OUTLETS                 - TREATGAS COLUMN          1        SEGMENT      1
                          RICH-MEA COLUMN          1        SEGMENT      3

PROPERTY OPTION SET:       ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
HENRY-COMPS ID      :      EMEA
CHEMISTRY ID:              EMEA    - APPARENT COMPONENTS

                    *** MASS AND ENERGY BALANCE ***

                                 IN             OUT            RELATIVE DIFF
 TOTAL BALANCE
   MOLE (KMOL/HR)              81405.1        81405.1         0.000000E+00
   MASS (KG/HR   )             0.197812E+07   0.197812E+07   -0.235406E-15
   ENTHALPY (CAL/SEC)         -0.146565E+10  -0.146565E+10   -0.114416E-07

                      ****   INPUT PARAMETERS   ****

NUMBER OF COLUMNS                                             1
TOTAL NUMBER OF SEGMENTS                                      3
NUMBER OF INTERCONNECTING STREAMS                             0
MAXIMUM NO OF RATEFRAC BLOCK ITERATIONS                       50
MAXIMUM NO OF FLASH ITERATIONS                                30
RATEFRAC BLOCK TOLERANCE                                      0.00010000
FLASH TOLERANCE                                               0.00010000

                **** INPUT PARAMETERS FOR COLUMN 1  ****

CONDENSER DUTY                         CAL/SEC                0.0
REBOILER DUTY                          CAL/SEC                0.0
DISTILLATE VAPOR MOLE FRACTION                                1.0000

TOTAL NUMBER OF EQUILIBRIUM SEGMENTS                          0
TOTAL NUMBER OF NON-EQUILIBRIUM SEGMENTS                      3
TOTAL NUMBER OF TRAY SEGMENTS                                 3
TOTAL NUMBER OF TRAYS                                         3

                     **** TRAY PARAMETERS ****
SEGMENT NUMBER(S)                                    1-  3
TYPE OF TRAYS                                        BUBBLE CAP
NUMBER OF TRAYS PER SEGMENT                          1.0000
COLUMN DIAMETER                        METER         7.1628
COLUMN DIAMETER ESTIMATE               METER         6.0000
PERCENT FLOODING                                     80.000
EXIT WEIR HEIGHT                       METER         0.50800E-01
SYSTEM FACTOR                                        1.0000
NUMBER OF PASSES                                     1
MIXING OPTION USED FOR VAPOR                         COMPLETE
MIXING OPTION USED FOR LIQUID                        COMPLETE
MASS-TRANSFER SUBROUTINE                             BUILT-IN
      CORRELATION                                    AICHE
HEAT-TRANSFER SUBROUTINE                             BUILT-IN
      CORRELATION                                    COLBURN
INTERFACIAL-AREA SUBROUTINE                          BUILT-IN
      CORRELATION                                    AICHE
```

```
                    **** PRESSURE SPECIFICATIONS ****

 COLUMN                  SEGMENT

    1                       1             1.18431        ATM

                       ****************************
                       ********* RESULTS *********
                       ****************************

**** COMPONENT SPLIT FRACTIONS ****

                               OUTLET STREAMS
                               --------------------------
                    TREATGAS                  RICH-MEA

 COMPONENT:
      H2O           .13316E-01                .98668
      MEA           .24725E-04                .99998
      CO2           .64839E-01                .93516
      N2            .99507                    .49301E-02
      O2            .98781                    .12189E-01

                     **** COLUMN 1 RESULTS ****

 TOP SEGMENT       LIQUID TEMP               K              329.317
                   VAPOR TEMP                K              329.843
 BOTTOM SEGMENT    LIQUID TEMP               K              330.060
                   VAPOR TEMP                K              328.834
 TOP SEGMENT       LIQUID FLOW           KMOL/HR            74,869.0
                   VAPOR FLOW            KMOL/HR            6,835.43
 BOTTOM SEGMENT    LIQUID FLOW           KMOL/HR            74,569.6
                   VAPOR FLOW            KMOL/HR            8,427.95
 DISTILLATE VAPOR MOLE FRACTION                             1.00000

SEGM   TEMPERATURE                      ENTHALPY              HEAT DUTY
     LIQUID     VAPOR    PRESSURE     LIQUID         VAPOR      LIQUID     VAPOR
       K          K        ATM        CAL/MOL        CAL/MOL    CAL/SEC   CAL/SEC

 1     329.3      329.8    1.1843    -6.9436+04     -1.0681+04

3      330.1      328.8    1.1843    -6.9778+04     -2.5368+04

SEGM         FLOW  RATE            FEED    RATE                  PRODUCT RATE
             LIQUID     VAPOR      LIQUID    VAPOR    MIXED      LIQUID    VAPOR
             KMOL/HR   KMOL/HR    KMOL/HR  KMOL/HR  KMOL/HR   KMOL/HR  KMOL/HR

 1           0.7487E+05    6835   7.3499+04                                  6835.4341

 3           0.7457E+05    8428              7906.5644              7.4570+04
```

```
    **** BULK X-PROFILE ****

 SEGMENT        H2O        MEA        CO2          N2           O2

   1            0.8478     0.1066     0.4521E-01   0.4381E-03   0.3731E-04
   3            0.8397     0.1070     0.5294E-01   0.3673E-03   0.3125E-04

    **** BULK Y-PROFILE ****

 SEGMENT        H2O        MEA        CO2          N2           O2

   1            0.1236     0.2886E-04 0.4005E-01   0.8087       0.2763E-01
   3            0.1188     0.1212E-04 0.1993       0.6592       0.2269E-01

BLOCK:  BLOWER   MODEL: COMPR
-----------------------------

 INLET STREAM:            FLUESEP
 OUTLET STREAM:           FLUEBLOW
 PROPERTY OPTION SET:     ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:          EMEA
 CHEMISTRY ID:            EMEA     - APPARENT COMPONENTS


                        *** MASS AND ENERGY BALANCE ***

                          IN              OUT             RELATIVE DIFF

  TOTAL BALANCE
    MOLE (KMOL/HR)        7906.56         7906.56         0.000000E+00
    MASS (KG/HR   )       248102.         248102.         0.000000E+00
    ENTHALPY (CAL/SEC)    -0.529556E+08   -0.526383E+08   -0.599132E-02


*** INPUT DATA ***

 ISENTROPIC CENTRIFUGAL COMPRESSOR
  OUTLET PRESSURE  ATM                     1.18431
  ISENTROPIC EFFICIENCY                    0.72000
  MECHANICAL EFFICIENCY                    1.00000

                 *** RESULTS ***

  INDICATED HORSEPOWER REQUIREMENT  KW                1,328.36
  BRAKE      HORSEPOWER REQUIREMENT        KW         1,328.36
  NET WORK REQUIRED                        KW         1,328.36
  POWER LOSSES                             KW         0.0
  ISENTROPIC HORSEPOWER REQUIREMENT        KW         956.421
  CALCULATED OUTLET TEMP       K                      322.451
  ISENTROPIC TEMPERATURE       K                      317.082
  EFFICIENCY (POLYTR/ISENTR) USED                     0.72000
  OUTLET VAPOR FRACTION                               1.00000
  IDEAL HEAD DEVELOPED, M-KGF/KG                      1,415.14
```

```
  ACTUAL HEAD DEVELOPED, M-KGF/KG                        1,965.48
  MECHANICAL EFFICIENCY USED                             1.00000
  INLET HEAT CAPACITY RATIO                              1.36117
  INLET VOLUMETRIC FLOW RATE , L/MIN                     3,273,320
  OUTLET VOLUMETRIC FLOW RATE, L/MIN                     2,940,140
  INLET COMPRESSIBILITY FACTOR                           0.99858
  OUTLET COMPRESSIBILITY FACTOR                          0.99867
  AV. ISENT. VOL. EXPONENT                               1.36168
  AV. ISENT. TEMP EXPONENT                               1.36170
  AV. ACTUAL VOL. EXPONENT                               1.57581
  AV. ACTUAL TEMP EXPONENT                               1.57453

 BLOCK:  COOLER   MODEL: HEATER
 ------------------------------------------------

  INLET STREAM:                  FLUEGAS
  OUTLET STREAM:                 FLUECOOL
  PROPERTY OPTION SET:           ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
  HENRY-COMPS ID:                EMEA
  CHEMISTRY ID:                  EMEA     - APPARENT COMPONENTS


             *** MASS AND ENERGY BALANCE ***

                            IN             OUT            RELATIVE DIFF
  TOTAL BALANCE
    MOLE (KMOL/HR)          8162.12        8162.12        0.000000E+00
    MASS (KG/HR   )         252711.        252711.        0.000000E+00
    ENTHALPY (CAL/SEC)     -0.547870E+08  -0.568774E+08   0.367524E-01

              *** INPUT DATA ***

  TWO    PHASE TP FLASH
  SPECIFIED TEMPERATURE                    K              313.150
  SPECIFIED PRESSURE                       ATM            1.00000
  MAXIMUM NO. ITERATIONS                                  30
  CONVERGENCE TOLERANCE                                   0.00010000


               *** RESULTS ***

 OUTLET TEMPERATURE        K                              313.15
 OUTLET PRESSURE           ATM                            1.0000
 HEAT DUTY                 CAL/SEC                        -0.20904E+07
 OUTLET VAPOR FRACTION                                    1.0000
 PRESSURE-DROP CORRELATION PARAMETER                      0.00000E+00
```

```
  V-L PHASE EQUILIBRIUM:

COMP          F(I)              X(I)            Y(I)            K(I)

 H2O       0.72301E-01       0.99867       0.72301E-01     0.73675E-01
 CO2       0.22364           0.99292E-04   0.22364         2292.3
  N2       0.68063           0.11287E-02   0.68063         613.69
  O2       0.23427E-01       0.98936E-04   0.23427E-01     240.97

 BLOCK:  COOLER1 MODEL: HEATER
 -------------------------------------------------

  INLET STREAM:              FLUEBLOW
  OUTLET STREAM:             FLUE-ABS
  PROPERTY OPTION SET:       ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
  HENRY-COMPS ID:            EMEA
  CHEMISTRY ID:              EMEA     - APPARENT COMPONENTS


             *** MASS AND ENERGY BALANCE ***

                             IN              OUT           RELATIVE DIFF
   TOTAL BALANCE
     MOLE (KMOL/HR)          7906.56         7906.56         0.000000E+00
     MASS (KG/HR   )         248102.         248102.         0.000000E+00
     ENTHALPY (CAL/SEC)    -0.526383E+08   -0.527921E+08   0.291301E-02


               *** INPUT DATA ***

TWO    PHASE TP FLASH
SPECIFIED TEMPERATURE              K                       313.150
SPECIFIED PRESSURE                 ATM                     1.18431
MAXIMUM NO ITERATIONS                                      30
CONVERGENCE TOLERANCE                                      0.00010000

                *** RESULTS ***

OUTLET TEMPERATURE                 K                       313.15
OUTLET PRESSURE                    ATM                     1.1843
HEAT DUTY                          CAL/SEC                  -0.15378E+06
OUTLET VAPOR FRACTION                                      1.0000
PRESSURE-DROP CORRELATION PARAMETER                        0.00000E+00


  V-L PHASE EQUILIBRIUM:

COMP             F(I)            X(I)            Y(I)            K(I)

 H2O          0.42369E-01     0.99766         0.42369E-01     0.62288E-01
 CO2          0.23087         0.17425E-03     0.23087         1943.4
  N2          0.70258         0.19871E-02     0.70258         518.58
  O2          0.24181E-01     0.17418E-03     0.24181E-01     203.62
```

```
BLOCK:  MCOMPR   MODEL: MCOMPR
----------------------------------------------------

 INLET STREAMS:       CO2             TO STAGE              1
 OUTLET STREAMS:      PRODCO2         FROM STAGE            4
 PROPERTY OPTION SET:        ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:             EMEA
 CHEMISTRY ID:               EMEA     - APPARENT COMPONENTS

              *** MASS AND ENERGY BALANCE ***

                              IN             UT           ELATIVE DIFF

  TOTAL BALANCE
    MOLE (KMOL/HR)           583.25          583.25          .000000E+00
    MASS (KG/HR   )          9161.7          9161.7          .000000E+00
    ENTHALPY (CAL/SEC)       0.406925E+08    0.410489E+08    .868273E-02


                 *** INPUT DATA ***

 ISENTROPIC COMPRESSOR USING ASME METHOD
 NUMBER OF STAGES                                    4
 FINAL PRESSURE, ATM                                 148.038


              COMPRESSOR SPECIFICATIONS PER STAGE

NUMBER                   EFFICIENCY            EFFICIENCY

  1                        1.000                 0.8000
  2                        1.000                 0.8000
  3                        1.000                 0.8000
  4                        1.000                 0.8000


              COOLER SPECIFICATIONS PER STAGE

 STAGE              PRESSURE DROP                  TEMPERATURE
 NUMBER                  ATM                            K

  1                 0.0000E+00                     313.2
  2                 0.0000E+00                     313.2
  3                 0.0000E+00                     313.2
  4                 0.0000E+00                     303.2


                 *** RESULTS ***

 FINAL PRESSURE, ATM                                 148.038
 TOTAL WORK REQUIRED, KW                             6,124.54
 TOTAL COOLING DUTY, CAL/SEC                         -1,819,240
```

```
               *** PROFILE ***

               COMPRESSOR PROFILE

 STAGE          OUTLET          PRESSURE           OUTLET
 NUMBER        PRESSURE          RATIO           TEMPERATURE
                 ATM                                  K

  1            5.589            2.981                350.3
  2            16.66            2.981                414.5
  3            49.66            2.981                416.8
  4            148.0            2.981                419.3

 STAGE                 INDICATED              BRAKE
 NUMBER                HORSEPOWER           HORSEPOWER
                          KW                   KW

  1                      1453                 1453
  2                      1714                 1714
  3                      1621                 1621
  4                      1336                 1336

               COOLER PROFILE

 STAGE        OUTLET         OUTLET      COOLING      VAPOR
 NUMBER    TEMPERATURE      PRESSURE      LOAD      FRACTION
                K              ATM       CAL/SEC

  1            313.2          5.589     -.1544E+06    1.000
  2            313.2          16.66     -.4495E+06    1.000
  3            313.2          49.66     -.5350E+06    1.000
  4            303.2          148.0     -.6803E+06    1.000


BLOCK:  MIXER    MODEL: MIXER
----------------------------------------------

 INLET STREAMS:            LEANSTRC    MEA        WATER
 OUTLET STREAM:            LEANMEX
 PROPERTY OPTION SET:      ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:           EMEA
 CHEMISTRY ID:             EMEA     - APPARENT COMPONENTS


            *** MASS AND ENERGY BALANCE ***

                              IN              OUT                RELATIVE DIFF


  TOTAL BALANCE
    MOLE (KMOL/HR)         73498.5         73498.5              -0.777648E-08
    MASS(KG/HR   )         0.173002E+07    0.173002E+07         -0.636008E-08
    ENTHALPY (CAL/SEC)     -0.140258E+10   -0.140260E+10        0.137109E-04
```

```
*** INPUT DATA ***

  TWO    PHASE     FLASH
  MAXIMUM NO ITERATIONS                         30
  CONVERGENCE TOLERANCE                         0.00010000
  OUTLET PRESSURE:         MINIMUM OF INLET STREAM PRESSURES

 BLOCK:  PUMP1    MODEL: PUMP
 -------------------------------------------

  INLET STREAM:            RICH-MEA
  OUTLET STREAM:           RICH-P1
  PROPERTY OPTION SET:     ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
  HENRY-COMPS ID:          EMEA
  CHEMISTRY ID:            EMEA     - APPARENT COMPONENTS


             *** MASS AND ENERGY BALANCE ***

                           IN             OUT                  RELATIVE DIFF

   TOTAL BALANCE
     MOLE (KMOL/HR)        74569.6        74569.6                 0.000000E+00
     MASS (KG/HR   )       0.178994E+07   0.178994E+07            0.000000E+00
     ENTHALPY (CAL/SEC)    -0.144537E+10  -0.144535E+10           -0.810900E-05


               *** INPUT DATA ***

   OUTLET PRESSURE   ATM                        1.97385
   DRIVER EFFICIENCY                            1.00000
   FLASH SPECIFICATIONS:
   LIQUID PHASE CALCULATION
   NO FLASH PERFORMED
   MAXIMUM NUMBER OF ITERATIONS                 30
   TOLERANCE                                    0.00010000

               *** RESULTS ***

   VOLUMETRIC FLOW RATE   L/MIN                          31,630.9
   PRESSURE CHANGE        ATM                            0.78954
   NPSH AVAILABLE         M-KGF/KG                       0.19790
   FLUID POWER            KW                             42.1745
   BRAKE POWER            KW                             49.0757
   ELECTRICITY            KW                             49.0757
   PUMP EFFICIENCY USED                                  0.85938
   NET WORK REQUIRED      KW                             49.0757
   HEAD DEVELOPED         M-KGF/KG                       8.64952
```

```
BLOCK:  PUMP2    MODEL: PUMP
-------------------------------------------
 INLET STREAM:              LEANSTPR
 OUTLET STREAM:             LEAN-P2
 PROPERTY OPTION SET:       ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:            EMEA
 CHEMISTRY ID:              EMEA     - APPARENT COMPONENTS

             *** MASS AND ENERGY BALANCE ***

                              IN              OUT              RELATIVE DIFF

  TOTAL BALANCE
    MOLE (KMOL/HR)          72986.4         72986.4             0.000000E+00
    MASS (KG/HR   )         0.172078E+07    0.172078E+07        0.000000E+00
    ENTHALPY (CAL/SEC)      -0.137196E+10   -0.137196E+10       -0.205341E-05


               *** INPUT DATA ***

  OUTLET PRESSURE   ATM                                 2.07254
  DRIVER EFFICIENCY                                     1.00000
  FLASH SPECIFICATIONS:
  LIQUID PHASE CALCULATION
  NO FLASH PERFORMED
  MAXIMUM NUMBER OF ITERATIONS                          30
  TOLERANCE                                             0.00010000

               *** RESULTS ***

  VOLUMETRIC FLOW           RATE  L/MIN                 30,296.0
  PRESSURE CHANGE           ATM                         0.19738
  NPSH AVAILABLE            M-KGF/KG                    0.0
  FLUID POWER               KW                          10.0987
  BRAKE POWER               KW                          11.7511
  ELECTRICITY               KW                          11.7511
  PUMP EFFICIENCY USED                                  0.85938
  NET WORK REQUIRED         KW                          11.7511
  HEAD DEVELOPED            M-KGF/KG                    2.15436


BLOCK:  SEP      MODEL: FLASH2
-------------------------------------------
 INLET STREAM:                   FLUECOOL
 OUTLET VAPOR STREAM:            FLUESEP
 OUTLET LIQUID STREAM:           SEPOUT
 PROPERTY OPTION SET:            ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:                 EMEA
 CHEMISTRY ID:                   EMEA     - APPARENT COMPONENTS
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```
   *** MASS AND ENERGY BALANCE ***

                               IN            OUT               RELATIVE DIFF
 TOTAL BALANCE
   MOLE (KMOL/HR)              8162.12       8162.12             0.000000E+00
   MASS (KG/HR   )             252711.       252711.             -0.505719E-11
   ENTHALPY (CAL/SEC)          -0.568774E+08 -0.577884E+08       0.157640E-01


               *** INPUT DATA ***

 TWO    PHASE TP FLASH
 SPECIFIED TEMPERATURE    K                          303.150
 SPECIFIED PRESSURE       ATM                        1.00000
 MAXIMUM NO ITERATIONS                               30
 CONVERGENCE TOLERANCE                               0.00010000


               *** RESULTS ***

 OUTLET TEMPERATURE       K                          303.15
 OUTLET PRESSURE          ATM                        1.0000
 HEAT DUTY                CAL/SEC                    -0.91098E+06
 VAPOR FRACTION                                      0.96869


 V-L PHASE EQUILIBRIUM:

 COMP             F(I)          X(I)          Y(I)          K(I)

  H2O             0.72301E-01   0.99837       0.42369E-01   0.42438E-01
  CO2             0.22364       0.12740E-03   0.23087       1812.3
  N2              0.68063       0.13791E-02   0.70258       509.47
  O2              0.23427E-01   0.12292E-03   0.24181E-01   196.72

BLOCK:  SOL-COOL MODEL: HEATER
----------------------------------------------------

 INLET STREAM:            LEANMEX
 OUTLET STREAM:           LEAN-MEA
 PROPERTY OPTION SET:     ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:          EMEA
 CHEMISTRY ID:            EMEA     - APPARENT COMPONENTS

             *** MASS AND ENERGY BALANCE ***

                                IN            OUT              RELATIVE DIFF
 TOTAL BALANCE
   MOLE (KMOL/HR)              73498.5       73498.5             0.000000E+00
   MASS (KG/HR   )             0.173002E+07  0.173002E+07        0.000000E+00
   ENTHALPY (CAL/SEC)          -0.140260E+10 -0.141285E+10       0.725641E-02
```

```
                              *** INPUT DATA ***

 TWO    PHASE TP FLASH
 SPECIFIED TEMPERATURE                    K                   313.150
 SPECIFIED PRESSURE                       ATM                 2.07254
 MAXIMUM NO ITERATIONS                                        30
 CONVERGENCE TOLERANCE                                        0.00010000


                   *** RESULTS ***

 OUTLET TEMPERATURE        K                                  313.15
 OUTLET PRESSURE           ATM                                2.0725
 HEAT DUTY                 CAL/SEC                            -0.10252E+08
 OUTLET VAPOR FRACTION                                        0.00000E+00
 PRESSURE-DROP CORRELATION PARAMETER                          0.00000E+00


V-L PHASE EQUILIBRIUM:

  COMP            F(I)          X(I)          Y(I)          K(I)

  H2O             0.85884       0.85884       0.99238       0.36528E-01
  MEA             0.10856       0.10856       0.47144E-03   0.13729E-03
  CO2             0.32604E-01   0.32604E-01   0.71433E-02   0.69260E-02
  N2              0.21316E-10   0.21316E-10   0.38825E-06   575.79
  O2              0.10414E-10   0.10414E-10   0.74504E-07   226.16


BLOCK:  SOL-HEX MODEL: HEATX
 -----------------------------------------------

 HOT SIDE:
---------------

  INLET STREAM:              LEAN-P2
  OUTLET STREAM:             LEANSTRC
  PROPERTY OPTION SET:       ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
  HENRY-COMPS ID:            EMEA
  CHEMISTRY ID:              EMEA     - APPARENT COMPONENTS

COLD SIDE:
---------------

  INLET STREAM:              RICH-P1
  OUTLET STREAM:             RICHSTPR
  PROPERTY OPTION SET:       ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
  HENRY-COMPS ID:            EMEA
  CHEMISTRY ID:              EMEA     - APPARENT COMPONENTS
```

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                         *** MASS AND ENERGY BALANCE ***

                               IN              OUT                 RELATIVE DIFF
    TOTAL BALANCE
      MOLE (KMOL/HR)         147556          147556                  0.449133E-06
      MASS (KG/HR   )        0.351073E+07    0.351073E+07            0.694769E-06
      ENTHALPY (CAL/SEC)     -0.281731E+10   -0.281729E+10           -0.595376E-05

                    *** INPUT DATA ***

    FLASH SPECS FOR HOT SIDE:

  TWO    PHASE     FLASH
    MAXIMUM NO ITERATIONS                                   30
    CONVERGENCE TOLERANCE                                   0.00010000

    FLASH SPECS FOR COLD SIDE:

  TWO    PHASE     FLASH
    MAXIMUM NO ITERATIONS                                   30
    CONVERGENCE TOLERANCE                                   0.00010000

    FLOW DIRECTION AND SPECIFICATION:

      COUNTERCURRENT   HEAT EXCHANGER
      SPECIFIED COLD TEMP CHANGE
      SPECIFIED VALUE          K                            45.0000
      LMTD CORRECTION FACTOR                                1.00000

    PRESSURE SPECIFICATION:

      HOT  SIDE PRESSURE DROP        ATM                    0.0000
      COLD SIDE PRESSURE DROP        ATM                    0.0000


    HEAT TRANSFER COEFFICIENT SPECIFICATION:

      HOT LIQUID        COLD LIQUID        CAL/SEC-SQCM-K        0.0203
      HOT 2-PHASE       COLD LIQUID        CAL/SEC-SQCM-K        0.0203
      HOT VAPOR         COLD LIQUID        CAL/SEC-SQCM-K        0.0203
      HOT LIQUID        COLD 2-PHASE       CAL/SEC-SQCM-K        0.0203
      HOT 2-PHASE       COLD 2-PHASE       CAL/SEC-SQCM-K        0.0203
      HOT VAPOR         COLD 2-PHASE       CAL/SEC-SQCM-K        0.0203
      HOT LIQUID        COLD VAPOR         CAL/SEC-SQCM-K        0.0203
      HOT 2-PHASE       COLD VAPOR         CAL/SEC-SQCM-K        0.0203
      HOT VAPOR         COLD VAPOR         CAL/SEC-SQCM-K        0.0203
```

```
                    *** OVERALL RESULTS ***

STREAMS:
                   --------------------------------------------
                   |                                          |
LEAN-P2 ---------->|                 HOT                      |---------> LEANSTRC
T= 3.9221D+02      |                                          |          T= 3.3960D+02
P= 2.0725D+00      |                                          |          P= 2.0725D+00
V= 0.0000D+00      |                                          |          V= 0.0000D+00
                   |                                          |
RICHSTPR <---------|                 COLD                     |<--------- RICH-P1
T= 3.7508D+02      |                                          |          T= 3.3008D+02
P= 1.9738D+00      |                                          |          P= 1.9738D+00
V= 1.1562D-02      |                                          |          V= 0.0000D+00
                   --------------------------------------------

DUTY AND AREA:

 CALCULATED HEAT DUTY                     CAL/SEC         20969519.0945
 CALCULATED (REQUIRED) AREA               SQM             8964.2223
 ACTUAL EXCHANGER AREA                    SQM             8964.2223
 PER CENT OVER-DESIGN                                     0.0000

HEAT TRANSFER COEFFICIENT:

 AVERAGE COEFFICIENT (DIRTY)   CAL/SEC-SQCM-K             0.0203
 UA (DIRTY)                    CAL/SEC-K                  1819907.5572

LOG-MEAN TEMPERATURE DIFFERENCE:

 LMTD CORRECTION FACTOR                                   1.0000
 LMTD (CORRECTED)              K                          11.5223
 NUMBER OF SHELLS IN SERIES                               1

PRESSURE DROP:

 HOTSIDE, TOTAL           ATM                             0.0000
 COLDSIDE, TOTAL          ATM                             0.0000

PRESSURE DROP PARAMETER:

 HOT SIDE:                     0.00000E+00
 COLD SIDE:                    0.00000E+00
```

```
                 *** ZONE RESULTS ***

 TEMPERATURE LEAVING EACH ZONE:

                                          HOT
                 ------------------------------------------------------------------
                 |                         |                          |
LEAN-P2          |      LIQ                |       LIQ                |  LEANSTRC
------>          |                         |                          |------>
 392.2           |                   354.6 |                          |   339.6
                 |                         |                          |
RICHSTPR         |      BOIL               |       LIQ                |  RICH-P1
<------          |                         |                          |<------
 375.1           |                   345.5 |                          |   330.1
                 |                         |                          |
                 ------------------------------------------------------------------
                                          COLD


 ZONE HEAT TRANSFER AND AREA:

 ZONE        HEAT DUTY           AREA          DTLM       AVERAGE U              UA
             CAL/SEC             SQM           K       CAL/SEC-SQCM-K   CAL/SEC-K

  1          15099019.574        5855.2967     12.7017        0.0203      1188736.5499
  2          5870499.521         3108.9256     9.3010         0.0203       631171.0074


BLOCK:  STRIPPER MODEL: RATEFRAC
----------------------------------------------------

 INLETS      - RICHSTPR    COLUMN       1       SEGMENT     4
 OUTLETS     - CO2         COLUMN       1       SEGMENT     1
               LEANSTPR    COLUMN       1       SEGMENT     9
 PROPERTY OPTION SET:      ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:           EMEA
 CHEMISTRY ID:             EMEA     - APPARENT COMPONENTS


                *** MASS AND ENERGY BALANCE ***

                           IN              OUT                 RELATIVE DIFF

  TOTAL BALANCE
    MOLE (KMOL/HR)         74569.6         74569.6                 0.000000E+00
    MASS(KG/HR   )         0.178994E+07    0.178994E+07            0.260154E-15
    ENTHALPY (CAL/SEC)     -0.142438E+10   -0.141265E+10           -0.823803E-02
```

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    ****   INPUT PARAMETERS   ****

NUMBER OF COLUMNS                                      1
TOTAL NUMBER OF SEGMENTS                               9
NUMBER OF INTERCONNECTING STREAMS                      0
MAXIMUM NO OF RATEFRAC BLOCK ITERATIONS                80
MAXIMUM NO OF FLASH ITERATIONS                         30
RATEFRAC BLOCK TOLERANCE                               0.00010000
FLASH TOLERANCE                                        0.00010000

    **** INPUT PARAMETERS FOR COLUMN 1 ****

MOLAR BOTTOMS / FEED RATIO                             0.9500
MOLAR REFLUX RATIO                                     0.6500
DISTILLATE VAPOR MOLE FRACTION                         1.0000
TOTAL NUMBER OF EQUILIBRIUM SEGMENTS                   2
EQUILIBRIUM SEGMENT NUMBERS                            1       9
TOTAL NUMBER OF NON-EQUILIBRIUM SEGMENTS               7
TOTAL NUMBER OF TRAY SEGMENTS                          7
TOTAL NUMBER OF TRAYS                                  7


    **** TRAY PARAMETERS ****

SEGMENT NUMBER(S)                                      2-  8
TYPE OF TRAYS                                          BUBBLE CAP
NUMBER OF TRAYS PER SEGMENT                            1.0000
COLUMN DIAMETER                         METER          4.4196
COLUMN DIAMETER ESTIMATE                METER          5.5000
PERCENT FLOODING                                       80.000
EXIT WEIR HEIGHT                        METER          0.50800E-01
SYSTEM FACTOR                                          1.0000
NUMBER OF PASSES                                       1
MIXING OPTION USED FOR VAPOR                           COMPLETE
MIXING OPTION USED FOR LIQUID                          COMPLETE
MASS-TRANSFER SUBROUTINE                               BUILT-IN
      CORRELATION                                      AICHE
HEAT-TRANSFER SUBROUTINE                               BUILT-IN
      CORRELATION                                      COLBURN
INTERFACIAL-AREA SUBROUTINE                            BUILT-IN
      CORRELATION                                      AICHE

    **** MANIPULATED VARIABLES ****

  VARY      1      MOLE-RR OF COLUMN=1

  VARY      2      MOLE-B:F OF COLUMN=1
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    **** DESIGN SPECIFICATIONS ****

   SPEC        1        MOLE-FRAC (0.9800) OF COMP CO2 IN PRODUCT STREAM=CO2

   SPEC        2        MASS-FLOWS (6.8285+04 KG/HR) OF COMP CO2 IN PRODUCT
                        STREAM=CO2


    **** PRESSURE SPECIFICATIONS ****

  COLUMN              SEGMENT

     1                   1                  1.87515    ATM


                    ****************************
                    ********* RESULTS *********
                    ****************************

  ****   COMPONENT SPLIT FRACTIONS   ****

                    OUTLET STREAMS
                    -------------------------

                          CO2                LEANSTPR
  COMPONENT:
  H2O                     .31112E-04         .99997
  MEA                     .59720E-15         1.0000
  CO2                     .39301             .60699
  N2                      1.0000             .57198E-07
  O2                      1.0000             .32841E-06



    **** MANIPULATED VARIABLES ****

  1            MOLE-RR            0.83214
  2            MOLE-B:F           0.97877


    **** DESIGN SPECIFICATIONS ****

       SPEC-TYPE   UNITS   SPECIFIED VALUE   CALCULATED VALUE   ABSOLUTEERROR
 1     MOLE-FRAC              0.98000            0.98000            -0.11102E-15
 2     MASS-FLOW   KG/HR      68285.             68285.              0.00000E+00
```

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                        **** COLUMN 1 RESULTS ****

 TOP SEGMENT          LIQUID TEMP           K                260.133
                      VAPOR TEMP            K                260.133
BOTTOM SEGMENT        LIQUID TEMP           K                392.212
                      VAPOR TEMP            K                392.212
TOP SEGMENT           LIQUID FLOW        KMOL/HR             1,317.48
                      VAPOR FLOW          KMOL/HR            1,583.25
BOTTOM SEGMENT        LIQUID FLOW        KMOL/HR             72,986.4
                      VAPOR FLOW          KMOL/HR            4,773.38
CONDENSER DUTY (W/O SUBCOOL)             CAL/SEC             -4,743,590
REBOILER DUTY                            CAL/SEC             0.164773+08
MOLAR REFLUX RATIO                                           0.83214
DISTILLATE VAPOR MOLE FRACTION                               1.00000
MOLAR BOILUP RATIO                                           0.065401


SEGM   TEMPERATURE                              ENTHALPY          HEAT DUTY
       LIQUID      VAPOR       PRESSURE       LIQUID      VAPOR    LIQUID     VAPOR
        K           K           ATM          CAL/MOL    CAL/MOL   CAL/SEC CAL/SEC

1      260.1       260.1       1.8752   -6.9079+04     -9.2527+04   -4.7436+06
2      362.0       371.2       1.8752   -6.7152+04     -7.5990+04
3      370.4       373.9       1.8752   -6.6998+04     -7.4523+04
4      375.3       375.4       1.8752   -6.8663+04     -7.5060+04
5      376.4       376.6       1.8752   -6.8618+04     -7.4273+04
8      387.4       388.0       1.8752   -6.8033+04     -6.5103+04
9      392.2       392.2       1.8752   -6.7671+04     -6.1152+04   1.6477+07

SEGM    FLOW RATE                  FEED     RATE                       PRODUCT RATE
        LIQUID      VAPOR         LIQUID     VAPOR      MIXED        LIQUID      VAPOR
       KMOL/HR    KMOL/HR      KMOL/HR    KMOL/HR    KMOL/HR     KMOL/HR     KMOL/HR

1      1317            1583                                                      1583.2483
2      1550            2901
3      1570            3134                          921.7492
4      0.7533E+05      2232                          7.3648+04
5      0.7552E+05      2339
8      0.7776E+05      3735
9      0.7299E+05      4773                                          7.2986+04


   **** BULK X-PROFILE ****

SEGMENT    H2O            MEA            CO2            N2             O2

  1        0.9950         0.4290E-03     0.4382E-02     0.1573E-03     0.3698E-04
  2        0.9985         0.5194E-03     0.9701E-03     0.2396E-04     0.5080E-05
  3        0.9985         0.6941E-03     0.8282E-03     0.1209E-04     0.2393E-05
  4        0.8475         0.1059         0.4655E-01     0.2871E-05     0.5564E-06
  5        0.8483         0.1057         0.4602E-01     0.5666E-06     0.1145E-06
  8        0.8594         0.1027         0.3787E-01     0.2688E-08     0.5988E-09
  9        0.8579         0.1093         0.3283E-01     0.2146E-10     0.1049E-10
```

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     **** BULK Y-PROFILE ****

   SEGMENT      H2O           MEA          CO2           N2            O2

     1        0.1230E-02    0.3009E-14    0.9800        0.1730E-01    0.1472E-02
     2        0.4526        0.1948E-03    0.5369        0.9513E-02    0.8202E-03
     3        0.4946        0.2570E-03    0.4956        0.8751E-02    0.7462E-03
     4        0.5025        0.3474E-03    0.4966        0.4760E-03    0.8831E-04
     5        0.5251        0.3901E-03    0.4744        0.9245E-04    0.1791E-04
     8        0.7756        0.1155E-02    0.2232        0.3706E-06    0.8032E-07
     9        0.8831        0.1956E-02    0.1150        0.4346E-07    0.9594E-08


  CO2 FLUE-ABS FLUEBLOW FLUECOOL FLUEGAS
  -------------------------------------------------------------------

  STREAM ID          CO2           FLUE-ABS      FLUEBLOW      FLUECOOL   FLUEGAS

  FROM :             STRIPPER      COOLER1       BLOWER        COOLER       ----
  TO   :             MCOMPR        ABSORBER      COOLER1       SEP        COOLER

  SUBSTREAM: MIXED
  PHASE:             VAPOR         VAPOR         VAPOR         VAPOR        VAPOR
  COMPONENTS: KMOL/HR

    H2O              1.9480        334.9921      334.9921      590.1325     590.1325
    MEA              4.7647-12     0.0           0.0           0.0          0.0
    H2S              0.0           0.0           0.0           0.0          0.0
    CO2              1551.5833     1825.3596     1825.3596     1825.3921   1825.3921
    HCO3-            0.0           0.0           0.0           0.0          0.0
    MEACOO-          0.0           0.0           0.0           0.0          0.0
    MEA+             0.0           0.0           0.0           0.0          0.0
    CO3-2            0.0           0.0           0.0           0.0          0.0
    HS-              0.0           0.0           0.0           0.0          0.0
    S-2              0.0           0.0           0.0           0.0          0.0
    H3O+             0.0           0.0           0.0           0.0          0.0
    OH-              0.0           0.0           0.0           0.0          0.0
    N2               27.3866       5555.0273     5555.0273     5555.3797   5555.3797
    O2               2.3304        191.1855      191.1855      191.2169    191.2169

 TOTAL FLOW:

    KMOL/HR          1583.2483     7906.5644     7906.5644      8162.1213   8162.1213
    KG/HR            6.9162+04     2.4810+05     2.4810+05     2.5271+05   2.5271+05
    L/MIN            2.9632+05     2.8549+06     2.9401+06     3.4905+06   4.8333+06
  STATE VARIABLES:
    TEMP   K         260.1327      313.1500      322.4513      313.1500    433.1500
    PRES   ATM       1.8752        1.1843        1.1843        1.0000       1.0000
    VFRAC            1.0000        1.0000        1.0000        1.0000       1.0000
    LFRAC            0.0           0.0           0.0           0.0          0.0
    SFRAC
              0.0           0.0           0.0           0.0           0.0
  ENTHALPY:
    CAL/MOL          -9.2527+04     -2.4037+04   -2.3967+04    -2.5086+04  -2.4164+04
```

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  CAL/GM               -2118.1206      -766.0216       -763.7902       -810.2492      -780.4706
  CAL/SEC              -4.0692+07      -5.2792+07      -5.2638+07      -5.6877+07     -5.4787+07
 ENTROPY:
  CAL/MOL-K            -1.5957         1.3422          1.5626          1.4853            3.9752
  CAL/GM-K             -3.6528-02      4.2774-02       4.9796-02       4.7973-02         0.1284
 DENSITY:
  MOL/CC               8.9051-05       4.6158-05       4.4820-05        3.8973-05      2.8145-05
  GM/CC                3.8901-03       1.4484-03       1.4064-03       1.2067-03       8.7142-04
 AVG MW                43.6834         31.3792         31.3792         30.9614           30.9614

 MIXED    SUBSTREAM PROPERTIES:

 *** VAPOR PHASE   ***

 VOLFLMX  L/MIN        2.9632+05       2.8549+06       2.9401+06       3.4905+06      4.8333+06
 MASSFLMX KG/HR        6.9162+04       2.4810+05       2.4810+05       2.5271+05      2.5271+05
 KMX  KCAL-M/HR-SQ 1.1773-02           2.0437-02       2.1029-02       2.0346-02      2.7915-02
 MUMX     CP           1.3160-02       1.7663-02       1.8089-02       1.7599-02      2.2812-02
 CPMX     KJ/KG-K      0.8146          1.0027          1.0061          1.0183         1.0586
 MWMX                  43.6834         31.3792         31.3792         30.9614        30.9614

STREAM ID              CO2             FLUE-ABS        FLUEBLOW        FLUECOOL   FLUEGAS

 *** LIQUID PHASE ***

VOLFLMX  L/MIN         MISSING         MISSING         MISSING         MISSING    MISSING
 MASSFLMX KG/HR       MISSING          MISSING         MISSING         MISSING    MISSING
 KMX  KCAL-M/HR-SQ  MISSING            MISSING         MISSING         MISSING    MISSING
 SIGMAMX  DYNE/CM   MISSING            MISSING         MISSING         MISSING    MISSING
 MUMX     CP           MISSING         MISSING         MISSING         MISSING    MISSING
 CPMX     KJ/KG-K      MISSING         MISSING         MISSING         MISSING    MISSING
 MWMX                  MISSING         MISSING         MISSING         MISSING    MISSING

 TOTAL STREAM PROPERTIES:

 ***  ALL PHASES   ***

 MASSFLOW KG/HR
     H2O               35.0946         6034.9763       6034.9763       1.0631+04      1.0631+04
    MEA                2.9105-10       0.0             0.0             0.0               0.0
    H2S                0.0             0.0             0.0             0.0               0.0
    CO2                6.8285+04       8.0334+04       8.0334+04       8.0335+04       8.0335+04
    HCO3-              0.0             0.0             0.0             0.0               0.0
    MEACOO-            0.0             0.0             0.0             0.0               0.0
    MEA+               0.0             0.0             0.0             0.0               0.0
    CO3-2              0.0             0.0             0.0             0.0               0.0
    HS-                0.0             0.0             0.0             0.0               0.0
    S-2                0.0             0.0             0.0             0.0               0.0
    H3O+               0.0             0.0             0.0             0.0               0.0
    OH-                0.0             0.0             0.0             0.0               0.0
    N2                 767.1930        1.5562+05       1.5562+05       1.5563+05      1.5563+05
    O2                 74.5685         6117.7068       6117.7068       6118.7120       6118.7120
 MASSVFRA              1.0000          1.0000          1.0000          1.0000            1.0000
 MASSSFRA              0.0             0.0             0.0             0.0               0.0
 MASSFLMX KG/HR        6.9162+04       2.4810+05       2.4810+05       2.5271+05      2.5271+05
```

```
VOLFLMX  L/MIN        2.9632+05       2.8549+06       2.9401+06       3.4905+06      4.8333+06
TEMP     K            260.1327        313.1500        322.4513        313.1500        433.1500
PRES     ATM          1.8752          1.1843          1.1843          1.0000          1.0000
MWMX                  43.6834         31.3792         31.3792         30.9614         30.9614


FLUESEP LEAN-MEA LEAN-P2 LEANMEX LEANSTPR
------------------------------------------------------------------------

STREAM ID              FLUESEP        LEAN-MEA        LEAN-P2    LEANMEX    LEANSTPR

FROM:                 SEP             SOL-COOL        PUMP2       MIXER        STRIPPER
TO   :                BLOWER          ABSORBER        SOL-HEX    SOL-COOL    PUMP2


MAX CONV. ERROR:       0.0            0.0             2.0014-05          0.0            0.0
SUBSTREAM: MIXED
PHASE:                VAPOR           LIQUID           LIQUID          LIQUID       LIQUID
COMPONENTS: KMOL/HR
  H2O                 334.9921        6.3124+04       6.2612+04       6.3124+04   6.2612+04
  MEA                 0.0             7978.6605       7978.4633       7978.6605    7978.4633
  H2S                 0.0             0.0             0.0             0.0           0.0
  CO2                 1825.3596       2396.3259       2396.3739       2396.3259    2396.3739
  HCO3-               0.0             0.0             0.0             0.0           0.0
  MEACOO-             0.0             0.0             0.0             0.0            0.0
  MEA+                0.0             0.0             0.0             0.0            0.0
  CO3-2               0.0             0.0             0.0             0.0            0.0
  HS-                 0.0             0.0             0.0             0.0            0.0
  S-2                 0.0             0.0             0.0             0.0            0.0
  H3O+                0.0             0.0             0.0             0.0            0.0
  OH-                 0.0             0.0             0.0             0.0            0.0
  N2                  5555.0273       1.5667-06       1.5665-06       1.5667-06    1.5665-06
  O2                  191.1855        7.6540-07       7.6531-07       7.6540-07    7.6531-07
TOTAL FLOW:
  KMOL/HR             7906.5644       7.3499+04       7.2986+04       7.3499+04   7.2986+04
  KG/HR               2.4810+05       1.7300+06       1.7208+06       1.7300+06   1.7208+06
  L/MIN               3.2733+06       2.8887+04       3.0296+04       2.9294+04    3.0296+04
STATE VARIABLES:
  TEMP   K            303.1500        313.1500        392.2133        339.3859     392.2118
  PRES   ATM          1.0000          2.0725          2.0725          2.0725        1.8752
  VFRAC               1.0000          0.0             0.0             0.0           0.0
  LFRAC               0.0             1.0000          1.0000          1.0000        1.0000
  SFRAC               0.0              0.0            0.0             0.0           0.0
ENTHALPY:
  CAL/MOL             -2.4112+04      -6.9202+04      -6.7671+04      -6.8700+04  -6.7671+04
  CAL/GM              -768.3939       -2940.0152      -2870.2284      -2918.6812  -2870.2343
  CAL/SEC             -5.2956+07      -1.4129+09      -1.3720+09      -1.4026+09  -1.3720+09
ENTROPY:
  CAL/MOL-K           1.4361          -63.8544        -59.3665        -62.2792       -59.3898
  CAL/GM-K            4.5765-02       -2.7128         -2.5180         -2.6459         -2.5190
DENSITY:
  MOL/CC              4.0258-05       4.2406-02       4.0152-02       4.1816-02      4.0152-02
  GM/CC               1.2633-03       0.9982          0.9467          0.9843          0.9467
AVG MW                31.3792         23.5381         23.5768         23.5381         23.5768
```

```
MIXED    SUBSTREAM PROPERTIES:

*** VAPOR PHASE   ***

VOLFLMX  L/MIN        3.2733+06          MISSING          MISSING          MISSING    MISSING
MASSFLMX KG/HR       2.4810+05           MISSING          MISSING          MISSING    MISSING
KMX  KCAL-M/HR-SQ  1.9800-02             MISSING          MISSING          MISSING    MISSING
MUMX     CP             1.7200-02        MISSING          MISSING          MISSING    MISSING
CPMX     KJ/KG-K          0.9986         MISSING          MISSING          MISSING    MISSING

STREAM ID                 FLUESEP        LEAN-MEA         LEAN-P2          LEANMEX   LEANSTPR

MWMX                      31.3792        MISSING          MISSING          MISSING      MISSING

*** LIQUID PHASE ***

VOLFLMX  L/MIN            MISSING        2.8887+04        3.0296+04        2.9294+04   3.0296+04
MASSFLMX KG/HR            MISSING        1.7300+06        1.7208+06        1.7300+06   1.7208+06
KMX  KCAL-M/HR-SQ         MISSING        0.3939           0.4015           0.4031       0.4015
SIGMAMX  DYNE/CM          MISSING        71.1954          56.6635          66.3802     56.6638
MUMX     CP               MISSING        1.2185           0.3479           0.7324       0.3479
CPMX     KJ/KG-K          MISSING        3.3601           3.5457           3.4115       3.5480
MWMX                      MISSING        23.5381          23.5768          23.5381     23.5768

TOTAL STREAM PROPERTIES:

*** ALL PHASES   ***

MASSFLOW KG/HR
    H2O                   6034.9763      1.1372+06        1.1280+06        1.1372+06   1.1280+06
    MEA                   0.0            4.8737+05        4.8735+05        4.8737+05   4.8735+05
    H2S                   0.0            0.0              0.0              0.0               0.0
    CO2                   8.0334+04      1.0546+05        1.0546+05        1.0546+05   1.0546+05
    HCO3-                 0.0            0.0              0.0              0.0               0.0
    MEACOO-               0.0            0.0              0.0              0.0               0.0
    MEA+                  0.0            0.0              0.0              0.0               0.0
    CO3-2                 0.0            0.0              0.0              0.0               0.0
    HS-                   0.0            0.0              0.0              0.0               0.0
    S-2                   0.0            0.0              0.0              0.0               0.0
    H3O+                  0.0            0.0              0.0              0.0               0.0
    OH-                   0.0            0.0              0.0              0.0               0.0
    N2                    1.5562+05      4.3889-05        4.3882-05        4.3889-05   4.3882-05
    O2                    6117.7068      2.4492-05        2.4489-05        2.4492-05   2.4489-05
MASSVFRA                  1.0000         0.0              0.0              0.0               0.0
MASSSFRA                  0.0            0.0              0.0              0.0               0.0
MASSFLMX KG/HR            2.4810+05      1.7300+06        1.7208+06        1.7300+06   1.7208+06
VOLFLMX  L/MIN            3.2733+06      2.8887+04        3.0296+04        2.9294+04   3.0296+04
TEMP     K                303.1500       313.1500         392.2133         339.3859    392.2118
PRES     ATM              1.0000         2.0725           2.0725           2.0725      1.8752
MWMX                      31.3792        23.5381          23.5768          23.5381     23.5768
```

```
LEANSTRC MEA PRODCO2 RICH-MEA RICH-P1
 -----------------------------------------------------------

 STREAM ID           LEANSTRC      MEA          PRODCO2        RICH-MEA    RICH-P1

 FROM:               SOL-HEX       ----         MCOMPR         ABSORBER    PUMP1
 TO  :               MIXER         MIXER        ----           PUMP1       SOL-HEX


 MAX CONV. ERROR:    1.3735-05     0.0          0.0            0.0            0.0
 SUBSTREAM: MIXED
 PHASE:              LIQUID        LIQUID       VAPOR          LIQUID         LIQUID
 COMPONENTS: KMOL/HR
   H2O               6.2612+04     0.0          1.9480         6.2614+04    6.2614+04
   MEA               7978.4633     0.1972       4.7647-12      7978.4633    7978.4633
   H2S               0.0           0.0          0.0            0.0            0.0
   CO2               2396.3259     0.0          1551.5833      3947.9572    3947.9572
   HCO3-             0.0           0.0          0.0            0.0            0.0
   MEACOO-           0.0           0.0          0.0            0.0            0.0
   MEA+              0.0           0.0          0.0            0.0            0.0
   CO3-2             0.0           0.0          0.0            0.0            0.0
   HS-               0.0           0.0          0.0            0.0            0.0
   S-2               0.0           0.0          0.0            0.0            0.0
   H3O+              0.0           0.0          0.0            0.0            0.0
   OH-               0.0           0.0          0.0            0.0            0.0
   N2                1.4944-06     0.0          27.3866        27.3866        27.3866
   O2                7.3343-07     0.0          2.3304         2.3304         2.3304
 TOTAL FLOW:
   KMOL/HR           7.2986+04     0.1972       1583.2483      7.4570+04   7.4570+04
   KG/HR             1.7208+06     12.0470      6.9162+04      1.7899+06   1.7899+06
   L/MIN             2.9142+04     0.1333       3678.4549      3.1631+04   3.1630+04
 STATE VARIABLES:
   TEMP   K          339.5992      313.1500     303.1500       330.0598    330.0778
   PRES   ATM        2.0725        2.0725       148.0385       1.1843      1.9738
   VFRAC              0.0          0.0          1.0000         0.0            0.0
   LFRAC             1.0000        1.0000       0.0            1.0000      1.0000
   SFRAC             0.0           0.0          0.0            0.0            0.0
 ENTHALPY:
   CAL/MOL           -6.8704+04    -6.4774+04   -9.3337+04     -6.9778+04  -6.9777+04
   CAL/GM            -2914.0670    -1060.4093   -2136.6728     -2906.9715  -2906.9479
   CAL/SEC           -1.3929+09    -3548.5421   -4.1049+07      -1.4454+09  -1.4454+09
 ENTROPY:
   CAL/MOL-K         -62.4506      -119.3998    -10.6792       -72.2535  -72.2526
   CAL/GM-K          -2.6488       -1.9547      -0.2445        -3.0101   -3.0101
 DENSITY:
   MOL/CC            4.1742-02      2.4662-02   7.1735-03      3.9292-02  3.9292-02
   GM/CC             0.9841        1.5065       0.3134         0.9431      0.9432
 AVG MW              23.5768       61.0837      43.6834        24.0037     24.0037
```

```
MIXED    SUBSTREAM PROPERTIES:

*** VAPOR PHASE   ***

VOLFLMX  L/MIN        MISSING        MISSING        3678.4549      MISSING    MISSING
MASSFLMX KG/HR        MISSING        MISSING        6.9162+04      MISSING    MISSING
KMX KCAL-M/HR-SQ      MISSING        MISSING        1.4608-02      MISSING    MISSING
MUMX     CP           MISSING        MISSING        1.5255-02      MISSING    MISSING
CPMX     KJ/KG-K      MISSING        MISSING        11.5900        MISSING    MISSING

STREAM ID             LEANSTRC       MEA            PRODCO2        RICH-MEA   RICH-P1

MWMX                  MISSING        MISSING        43.6834        MISSING     MISSING


*** LIQUID PHASE ***

VOLFLMX  L/MIN        2.9142+04      0.1333         MISSING        3.1631+04  3.1630+04
MASSFLMX KG/HR        1.7208+06      12.0470        MISSING        1.7899+06  1.7899+06
KMX  KCAL-M/HR-SQ     0.4024         0.2057         MISSING        0.4883     0.4883
SIGMAMX  DYNE/CM      66.3478        46.6759        MISSING        70.2787    70.2754
MUMX     CP           0.7338         11.3428        MISSING        1.1774     1.1770
CPMX     KJ/KG-K      3.4079         2.8736         MISSING        3.1492     3.1491
MWMX                  23.5768        61.0837        MISSING        24.0037    24.0037

TOTAL STREAM PROPERTIES:

*** ALL PHASES   ***

MASSFLOW KG/HR
   H2O                1.1280+06      0.0            35.0946        1.1280+06  1.1280+06
   MEA                4.8735+05      12.0470        2.9105-10      4.8735+05  4.8735+05
   H2S                0.0            0.0            0.0            0.0              0.0
   CO2                1.0546+05      0.0            6.8285+04      1.7375+05  1.7375+05
   HCO3-              0.0            0.0            0.0            0.0              0.0
   MEACOO-            0.0            0.0            0.0            0.0              0.0
   MEA+               0.0            0.0            0.0            0.0              0.0
   CO3-2              0.0            0.0            0.0            0.0              0.0
   HS-                0.0            0.0            0.0            0.0              0.0
   S-2                0.0            0.0            0.0            0.0              0.0
   H3O+               0.0            0.0            0.0            0.0              0.0
   OH-                0.0            0.0            0.0            0.0              0.0
   N2                 4.1863-05      0.0            767.1930       767.1931   767.1931
   O2                 2.3469-05      0.0            74.5685        74.5685    74.5685
MASSVFRA              0.0            0.0            1.0000         0.0              0.0
MASSSFRA              0.0            0.0            0.0            0.0              0.0
MASSFLMX KG/HR        1.7208+06      12.0470        6.9162+04      1.7899+06   1.7899+06
VOLFLMX  L/MIN        2.9142+04      0.1333         3678.4549      3.1631+04   3.1630+04
TEMP     K            339.5992       313.1500       303.1500       330.0598    330.0778
PRES     ATM          2.0725         2.0725         148.0385       1.1843      1.9738
MWMX                  23.5768        61.0837        43.6834        24.0037     24.0037
```

```
RICHSTPR SEPOUT TREATGAS WATER
-----------------------------

STREAM ID              RICHSTPR        SEPOUT          TREATGAS        WATER
FROM:                  SOL-HEX         SEP             ABSORBER        ----
TO   :                 STRIPPER        ----            ----            MIXER

SUBSTREAM: MIXED

PHASE:                 MIXED           LIQUID          VAPOR           LIQUID

COMPONENTS: KMOL/HR

  H2O                  6.2614+04       255.1405        845.0126        511.9864
  MEA                  7978.4633       0.0             0.1973          0.0
  H2S                  0.0             0.0              0.0            0.0
  CO2                  3947.9572       3.2558-02       273.7283        0.0
  HCO3-                0.0             0.0             0.0             0.0
  MEACOO-              0.0             0.0             0.0             0.0
  MEA+                 0.0             0.0             0.0             0.0
  CO3-2                0.0             0.0             0.0             0.0
  HS-                  0.0             0.0             0.0             0.0
  S-2                  0.0             0.0             0.0             0.0
  H3O+                 0.0             0.0             0.0             0.0
  OH-                  0.0             0.0             0.0             0.0
  N2                   27.3866         0.3524          5527.6407       0.0
  O2                   2.3304          3.1412-02       188.8552        0.0

TOTAL FLOW:

  KMOL/HR              7.4570+04       255.5569        6835.4341       511.9864
  KG/HR                1.7899+06       4608.7374       1.8817+05       9223.5780
  L/MIN                2.5385+05       77.5476         2.6006+06       154.9120

STATE VARIABLES:

  TEMP   K             375.0776        303.1500        329.8429        313.1500
  PRES   ATM           1.9738          1.0000          1.1843          2.0725
  VFRAC                1.1562-02       0.0             1.0000          0.0
  LFRAC                0.9884          1.0000          0.0             1.0000
  SFRAC                0.0             0.0             0.0             0.0
ENTHALPY:
  CAL/MOL              -6.8765+04      -6.8079+04      -1.0681+04      -6.7998+04
  CAL/GM               -2864.7734      -3775.0283      -387.9773       -3774.4550
  CAL/SEC              -1.4244+09      -4.8328+06      -2.0280+07      -9.6706+06
ENTROPY:
  CAL/MOL-K            -66.8879        -38.6418        0.4057          -38.0849
  CAL/GM-K             -2.7866         -2.1427         1.4737-02       -2.1140
DENSITY:
  MOL/CC               4.8960-03       5.4925-02       4.3807-05       5.5084-02
  GM/CC                0.1175          0.9905          1.2060-03       0.9923
AVG MW                 24.0037         18.0341         27.5291         18.0153
```

```
MIXED    SUBSTREAM PROPERTIES:

*** VAPOR PHASE   ***

VOLFLMX  L/MIN        2.2215+05     MISSING       2.6006+06     MISSING
MASSFLMX KG/HR        2.6958+04     MISSING       1.8817+05     MISSING
KMX  KCAL-M/HR-SQ     2.0799-02     MISSING       2.2755-02     MISSING
MUMX     CP           1.7287-02     MISSING       1.8809-02     MISSING
CPMX     KJ/KG-K      1.2007        MISSING       1.0965        MISSING
MWMX                  31.2684       MISSING       27.5291       MISSING

STREAM ID             RICHSTPR      SEPOUT        TREATGAS      WATER


*** LIQUID PHASE ***

VOLFLMX  L/MIN        3.1698+04     77.5476       MISSING       154.9120
MASSFLMX KG/HR        1.7630+06     4608.7374     MISSING       9223.5780
KMX  KCAL-M/HR-SQ     0.4646        0.4587        MISSING       0.5382
SIGMAMX  DYNE/CM      61.4512       71.7032       MISSING       69.8444
MUMX     CP           0.5255        0.8129        MISSING       0.6711
CPMX     KJ/KG-K      3.3256        4.1648        MISSING       4.1721
MWMX                  23.9187       18.0341       MISSING       18.0153


TOTAL STREAM PROPERTIES:

*** ALL PHASES   ***

MASSFLOW KG/HR
   H2O                1.1280+06     4596.4267     1.5223+04     9223.5780
   MEA                4.8735+05     0.0           12.0501       0.0
   H2S                 0.0          0.0           0.0           0.0
   CO2                1.7375+05     1.4329        1.2047+04     0.0
   HCO3-              0.0           0.0           0.0           0.0
   MEACOO-            0.0           0.0           0.0           0.0
   MEA+               0.0           0.0           0.0           0.0
   CO3-2              0.0           0.0           0.0           0.0
   HS-                0.0           0.0           0.0           0.0
   S-2                0.0           0.0           0.0           0.0
   H3O+               0.0           0.0           0.0           0.0
   OH-                0.0           0.0           0.0           0.0
   N2                 767.1930      9.8727        1.5485+05     0.0
   O2                 74.5685       1.0052        6043.1384     0.0
MASSVFRA              1.5061-02     0.0           1.0000        0.0
MASSSFRA              0.0           0.0           0.0           0.0
MASSFLMX KG/HR        1.7899+06     4608.7374     1.8817+05     9223.5780
VOLFLMX  L/MIN        2.5385+05     77.5476       2.6006+06     154.9120
TEMP     K            375.0776      303.1500      329.8429      313.1500
PRES     ATM          1.9738        1.0000        1.1843        2.0725
MWMX                  24.0037       18.0341       27.5291       18.0153
```

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BLOCK STATUS
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 **************************************************************************
 *                                         *
 * Calculations were completed normally                    *
 *                                             *
 * All Unit Operation blocks were completed normally              *
 *                                             *
 * All streams were flashed normally                      *
 *                                             *
 * All Convergence blocks were completed normally               *
 *                                         *
 **************************************************************************
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Appendix B: Aspen Simulation Results For Case I, Case II, and Case III

Aspen simulation results of MEA based CO_2 absorption for Case I (CO_2 recovery: 85% and lean loading: 0.05)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	59.5	123.1	123.1	59.5	104.5	70.7	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	9851	9851	9542	9542	9542	38362	38425	36514	36514	38425	38425	9479	1845	1	1911
Mass Flow kg/hr	304996	304996	299433	299433	299433	880953	930857	847614	847614	930857	930857	249530	33232	85	83243
Volume Flow m^3/hr	350000	252757	237033	212907	206733	811	989	828	828	989	9211	225450	33	0	277
Enthalpy MMkcal/hr	-238	-247	-230	-229	-229	-2608	-2681	-2419	-2419	-2681	-2641	-157	-125	0	-180
Mass Flow kg/hr															
H_2O	12831	12831	7284	7284	7284	610178	577363	576925	576925	577363	577363	40099	33232	0	437
MEA	0	0	0	0	0	261513	261427	261427	261427	261427	261427	86	0	85	0
CO_2	96956	96956	96954	96954	96954	9262	91672	9261	9261	91672	91672	14544	0	0	82411
N_2	187824	187824	187812	187812	187812	0	360	0	0	360	360	187452	0	0	360
O_2	7385	7385	7383	7383	7383	0	35	0	0	35	35	7349	0	0	35
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.693	0.62	0.681	0.681	0.62	0.62	0.161	1	0	0.005
MEA	0	0	0	0	0	0.297	0.281	0.308	0.308	0.281	0.281	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.011	0.098	0.011	0.011	0.098	0.098	0.058	0	0	0.99
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.751	0	0	0.004
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.029	0	0	0
Mole Flow kmol/hr															
H_2O	712	712	404	404	404	33870	32049	32024	32024	32049	32049	2226	1845	0	24
MEA	0	0	0	0	0	4281	4280	4280	4280	4280	4280	1	0	1	0
CO_2	2203	2203	2203	2203	2203	210	2083	210	210	2083	2083	330	0	0	1873
N_2	6705	6705	6704	6704	6704	0	13	0	0	13	13	6691	0	0	13
O_2	231	231	231	231	231	0	1	0	0	1	1	230	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.883	0.834	0.877	0.877	0.834	0.834	0.235	1	0	0.013
MEA	0	0	0	0	0	0.112	0.111	0.117	0.117	0.111	0.111	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.005	0.054	0.006	0.006	0.054	0.054	0.035	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.706	0	0	0.007
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.024	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case I (CO_2 recovery: 85% and lean loading: 0.10)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	59.9	122.9	122.9	59.9	104.9	69.1	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	9851	9851	9542	9542	9542	43677	43943	42032	42032	43943	43943	9276	1643	1	1911
Mass Flow kg/hr	304996	304996	299433	299433	299433	1008200	1061780	978519	978519	1061780	1061780	245848	29601	67	83264
Volume Flow m^3/hr	350000	252757	237033	212907	206733	946	1128	974	974	1128	10785	219604	30	0	276
Enthalpy MMkcal/hr	-238	-247	-230	-229	-229	-2981	-3065	-2797	-2797	-3065	-3019	-145	-112	0	-179
Mass Flow kg/hr															
H_2O	12831	12831	7284	7284	7284	690795	661586	661185	661185	661586	661586	36492	29601	0	401
MEA	0	0	0	0	0	296055	295986	295986	295986	295986	295986	69	0	67	0
CO_2	96956	96956	96954	96954	96954	21348	103759	21348	21348	103759	103759	14543	0	0	82411
N_2	187824	187824	187812	187812	187812	0	412	0	0	412	412	187400	0	0	412
O_2	7385	7385	7383	7383	7383	0	40	0	0	40	40	7344	0	0	40
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.685	0.623	0.676	0.676	0.623	0.623	0.148	1	0	0.005
MEA	0	0	0	0	0	0.294	0.279	0.302	0.302	0.279	0.279	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.021	0.098	0.022	0.022	0.098	0.098	0.059	0	0	0.99
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.762	0	0	0.005
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.03	0	0	0
Mole Flow kmol/hr															
H_2O	712	712	404	404	404	38345	36724	36701	36701	36724	36724	2026	1643	0	22
MEA	0	0	0	0	0	4847	4846	4846	4846	4846	4846	1	0	1	0
CO_2	2203	2203	2203	2203	2203	485	2358	485	485	2358	2358	330	0	0	1873
N_2	6705	6705	6704	6704	6704	0	15	0	0	15	15	6690	0	0	15
O_2	231	231	231	231	231	0	1	0	0	1	1	229	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.878	0.836	0.873	0.873	0.836	0.836	0.218	1	0	0.012
MEA	0	0	0	0	0	0.111	0.11	0.115	0.115	0.11	0.11	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.011	0.054	0.012	0.012	0.054	0.054	0.036	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.721	0	0	0.008
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.025	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case I (CO_2 recovery: 85% and lean loading: 0.15)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	60	122.5	122.5	60	104	67	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	9851	9851	9542	9542	9542	50319	50814	48903	48903	50814	50814	9048	1412	1	1911
Mass Flow kg/hr	304996	304996	299433	299433	299433	1167300	1225050	1141760	1141760	1225050	1225050	241685	25433	84	83293
Volume Flow m^3/hr	350000	252757	237033	212907	206733	1114	1301	1156	1156	1301	11388	212905	26	0	275
Enthalpy MMkcal/hr	-238	-247	-230	-229	-229	-3446	-3543	-3268	-3268	-3543	-3492	-132	-96	0	-179
Mass Flow kg/hr															
H_2O	12831	12831	7284	7284	7284	791472	766333	765979	765979	766333	766333	32423	25433	0	354
MEA	0	0	0	0	0	339227	339174	339174	339174	339174	339174	53	0	84	0
CO_2	96956	96956	96954	96954	96954	36605	119020	36606	36606	119020	119020	14540	0	0	82414
N_2	187824	187824	187812	187812	187812	0	480	0	0	480	480	187332	0	0	480
O_2	7385	7385	7383	7383	7383	0	46	0	0	46	46	7337	0	0	46
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.678	0.626	0.671	0.671	0.626	0.626	0.134	1	0	0.004
MEA	0	0	0	0	0	0.291	0.277	0.297	0.297	0.277	0.277	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.031	0.097	0.032	0.032	0.097	0.097	0.06	0	0	0.989
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.775	0	0	0.006
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.03	0	0	0.001
Mole Flow kmol/hr															
H_2O	712	712	404	404	404	43933	42538	42518	42518	42538	42538	1800	1412	0	20
MEA	0	0	0	0	0	5553	5553	5553	5553	5553	5553	1	0	1	0
CO_2	2203	2203	2203	2203	2203	832	2704	832	832	2704	2704	330	0	0	1873
N_2	6705	6705	6704	6704	6704	0	17	0	0	17	17	6687	0	0	17
O_2	231	231	231	231	231	0	1	0	0	1	1	229	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.873	0.837	0.869	0.869	0.837	0.837	0.199	1	0	0.01
MEA	0	0	0	0	0	0.11	0.109	0.114	0.114	0.109	0.109	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.017	0.053	0.017	0.017	0.053	0.053	0.037	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.739	0	0	0.009
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.025	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case I (CO_2 recovery: 85% and lean loading: 0.20)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	59.7	121.9	121.9	59.8	104.8	64.2	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	9851	9851	9542	9542	9542	59131	59894	57983	57983	59894	59894	8780	1149	1	1911
Mass Flow kg/hr	304996	304996	299433	299433	299433	1378540	1441160	1357830	1357830	1441160	1441160	236814	20701	38	83329
Volume Flow m^3/hr	350000	252757	237033	212907	206733	1338	1530	1396	1396	1530	14726	204903	21	0	273
Enthalpy MMkcal/hr	-238	-247	-230	-229	-229	-4064	-4176	-3891	-3891	-4176	-4113	-117	-78	0	-179
Mass Flow kg/hr															
H_2O	12831	12831	7284	7284	7284	924948	904561	904272	904272	904561	904561	27671	20701	0	289
MEA	0	0	0	0	0	396406	396368	396368	396368	396368	396368	39	0	38	0
CO_2	96956	96956	96954	96954	96954	57184	139600	57189	57189	139600	139600	14538	0	0	82411
N_2	187824	187824	187812	187812	187812	0	573	0	0	573	573	187239	0	0	573
O_2	7385	7385	7383	7383	7383	0	55	0	0	55	55	7328	0	0	55
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.671	0.628	0.666	0.666	0.628	0.628	0.117	1	0	0.003
MEA	0	0	0	0	0	0.288	0.275	0.292	0.292	0.275	0.275	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.041	0.097	0.042	0.042	0.097	0.097	0.061	0	0	0.989
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.791	0	0	0.007
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.031	0	0	0.001
Mole Flow kmol/hr															
H_2O	712	712	404	404	404	51342	50211	50195	50195	50211	50211	1536	1149	0	16
MEA	0	0	0	0	0	6490	6489	6489	6489	6489	6489	1	0	1	0
CO_2	2203	2203	2203	2203	2203	1299	3172	1299	1299	3172	3172	330	0	0	1873
N_2	6705	6705	6704	6704	6704	0	20	0	0	20	20	6684	0	0	20
O_2	231	231	231	231	231	0	2	0	0	2	2	229	0	0	2
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.868	0.838	0.866	0.866	0.838	0.838	0.175	1	0	0.008
MEA	0	0	0	0	0	0.11	0.108	0.112	0.112	0.108	0.108	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.022	0.053	0.022	0.022	0.053	0.053	0.038	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.761	0	0	0.011
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.026	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case I (CO_2 recovery: 85% and lean loading: 0.25)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	58.6	120.9	120.9	58.6	103.6	60.9	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	9851	9851	9542	9542	9542	71388	72411	70500	70500	72411	72411	8520	888	0	1911
Mass Flow kg/hr	304996	304996	299433	299433	299433	1672350	1739720	1656330	1656330	1739720	1739720	232064	16005	1	83385
Volume Flow m^3/hr	350000	252757	237033	212907	206733	1650	1846	1728	1728	1846	16362	196915	16	0	271
Enthalpy MMkcal/hr	-238	-247	-230	-229	-229	-4924	-5050	-4750	-4750	-5050	-4975	-103	-60	0	-179
Mass Flow kg/hr															
H_2O	12831	12831	7284	7284	7284	1110550	1094760	1094570	1094570	1094760	1094760	23082	16005	0	191
MEA	0	0	0	0	0	475883	475859	475859	475859	475859	475859	24	0	1	0
CO_2	96956	96956	96954	96954	96954	85911	168321	85909	85909	168321	168321	14545	0	0	82411
N_2	187824	187824	187812	187812	187812	0	713	0	0	713	713	187099	0	0	713
O_2	7385	7385	7383	7383	7383	0	69	0	0	69	69	7314	0	0	69
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.664	0.629	0.661	0.661	0.629	0.629	0.099	1	0	0.002
MEA	0	0	0	0	0	0.285	0.274	0.287	0.287	0.274	0.274	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.051	0.097	0.052	0.052	0.097	0.097	0.063	0	0	0.988
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.806	0	0	0.009
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.032	0	0	0.001
Mole Flow kmol/hr															
H_2O	712	712	404	404	404	61645	60768	60758	60758	60768	60768	1281	888	0	11
MEA	0	0	0	0	0	7791	7790	7790	7790	7790	7790	0	0	0	0
CO_2	2203	2203	2203	2203	2203	1952	3825	1952	1952	3825	3825	331	0	0	1873
N_2	6705	6705	6704	6704	6704	0	25	0	0	25	25	6679	0	0	25
O_2	231	231	231	231	231	0	2	0	0	2	2	229	0	0	2
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.864	0.839	0.862	0.862	0.839	0.839	0.15	1	0	0.006
MEA	0	0	0	0	0	0.109	0.108	0.111	0.111	0.108	0.108	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.027	0.053	0.028	0.028	0.053	0.053	0.039	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.784	0	0	0.013
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.027	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case I (CO_2 recovery: 85% and lean loading: 0.30)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	56.9	119.1	119.1	56.9	101.9	56.9	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	9851	9851	9542	9542	9542	88536	89819	87908	87908	89819	89819	8260	628	0	1911
Mass Flow kg/hr	304996	304996	299433	299433	299433	2083980	2156120	2072650	2072650	2156120	2156120	227294	11318	15	83469
Volume Flow m^3/hr	350000	252757	237033	212907	206733	2088	2286	2189	2189	2286	18386	188656	11	0	266
Enthalpy MMkcal/hr	-238	-247	-230	-229	-229	-6127	-6268	-5949	-5949	-6268	-6177	-89	-43	0	-178
Mass Flow kg/hr															
H_2O	12831	12831	7284	7284	7284	1369860	1358580	1358540	1358540	1358580	1358580	18557	11318	0	44
MEA	0	0	0	0	0	587082	587067	587067	587067	587067	587067	15	0	15	0
CO_2	96956	96956	96954	96954	96954	127040	209455	127044	127044	209455	209455	14540	0	0	82411
N_2	187824	187824	187812	187812	187812	0	923	0	0	923	923	186889	0	0	923
O_2	7385	7385	7383	7383	7383	0	90	0	0	90	90	7294	0	0	90
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.657	0.63	0.655	0.655	0.63	0.63	0.082	1	0	0.001
MEA	0	0	0	0	0	0.282	0.272	0.283	0.283	0.272	0.272	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.061	0.097	0.061	0.061	0.097	0.097	0.064	0	0	0.987
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.822	0	0	0.011
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.032	0	0	0.001
Mole Flow kmol/hr															
H_2O	712	712	404	404	404	76039	75413	75410	75410	75413	75413	1030	628	0	2
MEA	0	0	0	0	0	9611	9611	9611	9611	9611	9611	0	0	0	0
CO_2	2203	2203	2203	2203	2203	2887	4759	2887	2887	4759	4759	330	0	0	1873
N_2	6705	6705	6704	6704	6704	0	33	0	0	33	33	6671	0	0	33
O_2	231	231	231	231	231	0	3	0	0	3	3	228	0	0	3
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.859	0.84	0.858	0.858	0.84	0.84	0.125	1	0	0.001
MEA	0	0	0	0	0	0.109	0.107	0.109	0.109	0.107	0.107	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.033	0.053	0.033	0.033	0.053	0.053	0.04	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.808	0	0	0.017
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.028	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case II (CO_2 recovery: 85% and lean loading: 0.05)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	59.7	123.1	123.1	59.7	104.7	70.6	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	8162	8162	7907	7907	7907	31886	31953	30370	30370	31953	31953	7839	1512	1	1583
Mass Flow kg/hr	252711	252711	248102	248102	248102	732261	773883	704909	704909	773883	773883	206480	27247	47	68974
Volume Flow m^3/hr	290000	209428	196399	176408	171293	674	822	689	689	822	7717	186363	27	0	229
EnthalpyMMkcal/hr	-197	-205	-191	-190	-190	-2168	-2229	-2012	-2012	-2229	-2196	-129	-103	0	-149
Mass Flow kg/hr															
H_2O	10631	10631	6035	6035	6035	507145	480226	479864	479864	480226	480226	32954	27247	0	362
MEA	0	0	0	0	0	217348	217277	217277	217277	217277	217277	71	0	47	0
CO_2	80335	80335	80334	80334	80334	7769	76053	7768	7768	76053	76053	12050	0	0	68285
N_2	155626	155626	155616	155616	155616	0	298	0	0	298	298	155317	0	0	298
O_2	6119	6119	6118	6118	6118	0	29	0	0	29	29	6089	0	0	29
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.693	0.621	0.681	0.681	0.621	0.621	0.16	1	0	0.005
MEA	0	0	0	0	0	0.297	0.281	0.308	0.308	0.281	0.281	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.011	0.098	0.011	0.011	0.098	0.098	0.058	0	0	0.99
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.752	0	0	0.004
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.029	0	0	0
Mole Flow kmol/hr															
H_2O	590	590	335	335	335	28151	26657	26636	26636	26657	26657	1829	1512	0	20
MEA	0	0	0	0	0	3558	3557	3557	3557	3557	3557	1	0	1	0
CO_2	1825	1825	1825	1825	1825	177	1728	177	177	1728	1728	274	0	0	1552
N_2	5555	5555	5555	5555	5555	0	11	0	0	11	11	5544	0	0	11
O_2	191	191	191	191	191	0	1	0	0	1	1	190	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.883	0.834	0.877	0.877	0.834	0.834	0.233	1	0	0.013
MEA	0	0	0	0	0	0.112	0.111	0.117	0.117	0.111	0.111	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.006	0.054	0.006	0.006	0.054	0.054	0.035	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.707	0	0	0.007
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.024	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case II (CO_2 recovery: 85% and lean loading: 0.10)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	60	122.9	122.9	60.1	105.1	68.9	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	8162	8162	7907	7907	7907	36274	36508	34925	34925	36508	36508	7672	1348	1	1583
Mass Flow kg/hr	252711	252711	248102	248102	248102	837323	881969	812978	812978	881969	881969	203456	24284	56	68992
Volume Flow m^3/hr	290000	209428	196399	176408	171293	785	937	809	809	937	9006	181556	24	0	228
EnthalpyMMkcal/h	-197	-205	-191	-190	-190	-2476	-2546	-2324	-2324	-2546	-2508	-120	-92	0	-149
Mass Flow kg/hr															
H_2O	10631	10631	6035	6035	6035	573704	549749	549417	549417	549749	549749	29990	24284	0	332
MEA	0	0	0	0	0	245873	245816	245816	245816	245816	245816	57	0	56	0
CO_2	80335	80335	80334	80334	80334	17746	86030	17745	17745	86030	86030	12050	0	0	68285
N_2	155626	155626	155616	155616	155616	0	341	0	0	341	341	155274	0	0	341
O_2	6119	6119	6118	6118	6118	0	33	0	0	33	33	6085	0	0	33
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.685	0.623	0.676	0.676	0.623	0.623	0.147	1	0	0.005
MEA	0	0	0	0	0	0.294	0.279	0.302	0.302	0.279	0.279	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.021	0.098	0.022	0.022	0.098	0.098	0.059	0	0	0.99
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.763	0	0	0.005
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.03	0	0	0
Mole Flow kmol/hr															
H_2O	590	590	335	335	335	31845	30516	30497	30497	30516	30516	1665	1348	0	18
MEA	0	0	0	0	0	4025	4024	4024	4024	4024	4024	1	0	1	0
CO_2	1825	1825	1825	1825	1825	403	1955	403	403	1955	1955	274	0	0	1552
N_2	5555	5555	5555	5555	5555	0	12	0	0	12	12	5543	0	0	12
O_2	191	191	191	191	191	0	1	0	0	1	1	190	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.878	0.836	0.873	0.873	0.836	0.836	0.217	1	0	0.012
MEA	0	0	0	0	0	0.111	0.11	0.115	0.115	0.11	0.11	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.011	0.054	0.012	0.012	0.054	0.054	0.036	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.722	0	0	0.008
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.025	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case II (CO_2 recovery: 85% and lean loading: 0.15)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	60.1	122.5	122.5	60.1	105.1	66.8	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	8162	8162	7907	7907	7907	41822	42247	40664	40664	42247	42247	7482	1155	1	1583
Mass Flow kg/hr	252711	252711	248102	248102	248102	970144	1018260	949248	949248	1018260	1018260	199984	20815	44	69014
Volume Flow m^3/hr	290000	209428	196399	176408	171293	926	1081	961	961	1081	10481	175961	21	0	228
EnthalpyMMkcal/h	-197	-205	-191	-190	-190	-2864	-2946	-2717	-2717	-2946	-2901	-109	-79	0	-149
Mass Flow kg/hr															
H_2O	10631	10631	6035	6035	6035	657841	637280	636988	636988	637280	637280	26595	20815	0	293
MEA	0	0	0	0	0	281850	281806	281806	281806	281806	281806	43	0	44	0
CO_2	80335	80335	80334	80334	80334	30453	98739	30454	30454	98739	98739	12048	0	0	68285
N_2	155626	155626	155616	155616	155616	0	398	0	0	398	398	155218	0	0	398
O_2	6119	6119	6118	6118	6118	0	38	0	0	38	38	6079	0	0	38
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.678	0.626	0.671	0.671	0.626	0.626	0.133	1	0	0.004
MEA	0	0	0	0	0	0.291	0.277	0.297	0.297	0.277	0.277	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.031	0.097	0.032	0.032	0.097	0.097	0.06	0	0	0.989
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.776	0	0	0.006
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.03	0	0	0.001
Mole Flow kmol/hr															
H_2O	590	590	335	335	335	36516	35374	35358	35358	35374	35374	1476	1155	0	16
MEA	0	0	0	0	0	4614	4613	4613	4613	4613	4613	1	0	1	0
CO_2	1825	1825	1825	1825	1825	692	2244	692	692	2244	2244	274	0	0	1552
N_2	5555	5555	5555	5555	5555	0	14	0	0	14	14	5541	0	0	14
O_2	191	191	191	191	191	0	1	0	0	1	1	190	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.873	0.837	0.87	0.87	0.837	0.837	0.197	1	0	0.01
MEA	0	0	0	0	0	0.11	0.109	0.113	0.113	0.109	0.109	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.017	0.053	0.017	0.017	0.053	0.053	0.037	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.741	0	0	0.009
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.025	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case II (CO_2 recovery: 85% and lean loading: 0.20)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	59.9	121.9	121.9	59.9	104.9	63.7	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	8162	8162	7907	7907	7907	49543	50209	48625	48625	50209	50209	7241	919	1	1583
Mass Flow kg/hr	252711	252711	248102	248102	248102	1155270	1207760	1138720	1138720	1207760	1207760	195611	16548	42	69047
Volume Flow m^3/hr	290000	209428	196399	176408	171293	1122	1282	1172	1172	1282	12530	168742	17	0	226
Enthalpy MMkcal/hr	-197	-205	-191	-190	-190	-3406	-3501	-3264	-3264	-3500	-3448	-95	-62	0	-148
Mass Flow kg/hr															
H_2O	10631	10631	6035	6035	6035	774746	758462	758226	758226	758462	758462	22319	16548	0	236
MEA	0	0	0	0	0	331945	331914	331914	331914	331914	331914	32	0	42	0
CO_2	80335	80335	80334	80334	80334	48580	116860	48575	48575	116860	116860	12053	0	0	68285
N_2	155626	155626	155616	155616	155616	0	480	0	0	480	480	155136	0	0	480
O_2	6119	6119	6118	6118	6118	0	46	0	0	46	46	6071	0	0	46
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.671	0.628	0.666	0.666	0.628	0.628	0.114	1	0	0.003
MEA	0	0	0	0	0	0.287	0.275	0.291	0.291	0.275	0.275	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.042	0.097	0.043	0.043	0.097	0.097	0.062	0	0	0.989
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.793	0	0	0.007
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.031	0	0	0.001
Mole Flow kmol/hr															
H_2O	590	590	335	335	335	43005	42101	42088	42088	42101	42101	1239	919	0	13
MEA	0	0	0	0	0	5434	5434	5434	5434	5434	5434	1	0	1	0
CO_2	1825	1825	1825	1825	1825	1104	2655	1104	1104	2655	2655	274	0	0	1552
N_2	5555	5555	5555	5555	5555	0	17	0	0	17	17	5538	0	0	17
O_2	191	191	191	191	191	0	1	0	0	1	1	190	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.868	0.839	0.866	0.866	0.839	0.839	0.171	1	0	0.008
MEA	0	0	0	0	0	0.11	0.108	0.112	0.112	0.108	0.108	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.022	0.053	0.023	0.023	0.053	0.053	0.038	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.765	0	0	0.011
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.026	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case II (CO_2 recovery: 85% and lean loading: 0.25)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	58.7	120.9	120.9	58.8	103.8	60.6	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	8162	8162	7907	7907	7907	59321	60187	58604	58604	60187	60187	7041	714	0	1583
Mass Flow kg/hr	252711	252711	248102	248102	248102	1389830	1445970	1376880	1376880	1445970	1445970	191962	12855	14	69092
Volume Flow m^3/hr	290000	209428	196399	176408	171293	1371	1534	1437	1437	1534	13837	162598	13	0	224
Enthalpy MMkcal/hr	-197	-205	-191	-190	-190	-4092	-4197	-3949	-3949	-4197	-4135	-84	-49	0	-148
Mass Flow kg/hr															
H_2O	10631	10631	6035	6035	6035	922728	909960	909802	909802	909960	909960	18803	12855	0	157
MEA	0	0	0	0	0	395417	395396	395396	395396	395396	395396	21	0	14	0
CO_2	80335	80335	80334	80334	80334	71689	139968	71683	71683	139968	139968	12055	0	0	68285
N_2	155626	155626	155616	155616	155616	0	592	0	0	592	592	155023	0	0	592
O_2	6119	6119	6118	6118	6118	0	57	0	0	57	57	6060	0	0	57
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.664	0.629	0.661	0.661	0.629	0.629	0.098	1	0	0.002
MEA	0	0	0	0	0	0.285	0.273	0.287	0.287	0.273	0.273	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.052	0.097	0.052	0.052	0.097	0.097	0.063	0	0	0.988
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.808	0	0	0.009
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.032	0	0	0.001
Mole Flow kmol/hr															
H_2O	590	590	335	335	335	51219	50510	50502	50502	50510	50510	1044	714	0	9
MEA	0	0	0	0	0	6473	6473	6473	6473	6473	6473	0	0	0	0
CO_2	1825	1825	1825	1825	1825	1629	3180	1629	1629	3180	3180	274	0	0	1552
N_2	5555	5555	5555	5555	5555	0	21	0	0	21	21	5534	0	0	21
O_2	191	191	191	191	191	0	2	0	0	2	2	189	0	0	2
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.863	0.839	0.862	0.862	0.839	0.839	0.148	1	0	0.006
MEA	0	0	0	0	0	0.109	0.108	0.11	0.11	0.108	0.108	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.027	0.053	0.028	0.028	0.053	0.053	0.039	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.786	0	0	0.013
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.027	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case II (CO_2 recovery: 85% and lean loading: 0.30)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	56.9	119.1	119.1	56.9	101.9	56.7	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	8162	8162	7907	7907	7907	73499	74570	72986	72986	74570	74570	6835	512	0	1583
Mass Flow kg/hr	252711	252711	248102	248102	248102	1730020	1789950	1720780	1720780	1789950	1789950	188174	9224	12	69162
Volume Flow m^3/hr	290000	209428	196399	176408	171293	1733	1898	1818	1818	1898	15231	156035	9	0	221
Enthalpy MMkcal/hr	-197	-205	-191	-190	-190	-5086	-5203	-4939	-4939	-5203	-5128	-73	-35	0	-148
Mass Flow kg/hr															
H_2O	10631	10631	6035	6035	6035	1137190	1128000	1127970	1127970	1128000	1128000	15223	9224	0	35
MEA	0	0	0	0	0	487366	487354	487354	487354	487354	487354	12	0	12	0
CO_2	80335	80335	80334	80334	80334	105462	173749	105464	105464	173749	173749	12047	0	0	68285
N_2	155626	155626	155616	155616	155616	0	767	0	0	767	767	154848	0	0	767
O_2	6119	6119	6118	6118	6118	0	75	0	0	75	75	6043	0	0	75
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.657	0.63	0.655	0.655	0.63	0.63	0.081	1	0	0.001
MEA	0	0	0	0	0	0.282	0.272	0.283	0.283	0.272	0.272	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.061	0.097	0.061	0.061	0.097	0.097	0.064	0	0	0.987
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.823	0	0	0.011
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.032	0	0	0.001
Mole Flow kmol/hr															
H_2O	590	590	335	335	335	63124	62614	62612	62612	62614	62614	845	512	0	2
MEA	0	0	0	0	0	7979	7978	7978	7978	7978	7978	0	0	0	0
CO_2	1825	1825	1825	1825	1825	2396	3948	2396	2396	3948	3948	274	0	0	1552
N_2	5555	5555	5555	5555	5555	0	27	0	0	27	27	5528	0	0	27
O_2	191	191	191	191	191	0	2	0	0	2	2	189	0	0	2
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.859	0.84	0.858	0.858	0.84	0.84	0.124	1	0	0.001
MEA	0	0	0	0	0	0.109	0.107	0.109	0.109	0.107	0.107	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.033	0.053	0.033	0.033	0.053	0.053	0.04	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.809	0	0	0.017
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.028	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case III (CO_2 recovery: 85% and lean loading: 0.05)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	60.2	123.1	123.1	60.2	105.2	70.3	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	5629	5629	5453	5453	5453	22097	22166	21074	21074	22166	22166	5384	1024	1	1092
Mass Flow kg/hr	174283	174283	171105	171105	171105	507457	536560	488992	488992	536560	536560	142001	18442	41	47568
Volume Flow m^3/hr	200000	144433	135448	121661	118133	467	570	478	478	570	5447	127881	19	0	158
Enthalpy MMkcal/hr	-136	-141	-131	-131	-131	-1503	-1546	-1396	-1396	-1546	-1523	-88	-70	0	-103
Mass Flow kg/hr															
H_2O	7332	7332	4162	4162	4162	351466	333300	333050	333050	333300	333300	22328	18442	0	250
MEA	0	0	0	0	0	150605	150556	150556	150556	150556	150556	49	0	41	0
CO_2	55404	55404	55403	55403	55403	5386	52480	5387	5387	52480	52480	8309	0	0	47093
N_2	107328	107328	107321	107321	107321	0	205	0	0	205	205	107116	0	0	205
O_2	4220	4220	4219	4219	4219	0	20	0	0	20	20	4199	0	0	20
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.693	0.621	0.681	0.681	0.621	0.621	0.157	1	0	0.005
MEA	0	0	0	0	0	0.297	0.281	0.308	0.308	0.281	0.281	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.011	0.098	0.011	0.011	0.098	0.098	0.059	0	0	0.99
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.754	0	0	0.004
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.03	0	0	0
Mole Flow kmol/hr															
H_2O	407	407	231	231	231	19509	18501	18487	18487	18501	18501	1239	1024	0	14
MEA	0	0	0	0	0	2466	2465	2465	2465	2465	2465	1	0	1	0
CO_2	1259	1259	1259	1259	1259	122	1192	122	122	1192	1192	189	0	0	1070
N_2	3831	3831	3831	3831	3831	0	7	0	0	7	7	3824	0	0	7
O_2	132	132	132	132	132	0	1	0	0	1	1	131	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.883	0.835	0.877	0.877	0.835	0.835	0.23	1	0	0.013
MEA	0	0	0	0	0	0.112	0.111	0.117	0.117	0.111	0.111	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.006	0.054	0.006	0.006	0.054	0.054	0.035	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.71	0	0	0.007
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.024	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case III (CO_2 recovery: 85% and lean loading: 0.10)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	60.4	122.9	122.9	60.4	105.4	68.6	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	5629	5629	5453	5453	5453	25177	25361	24269	24269	25361	25361	5269	909	1	1092
Mass Flow kg/hr	174283	174283	171105	171105	171105	581192	612388	564808	564808	612388	612388	139908	16372	32	47581
Volume Flow m^3/hr	200000	144433	135448	121661	118133	545	650	562	562	650	6335	124549	16	0	158
Enthalpy MMkcal/hr	-136	-141	-131	-131	-131	-1718	-1768	-1615	-1615	-1768	-1742	-81	-62	0	-103
Mass Flow kg/hr															
H_2O	7332	7332	4162	4162	4162	398172	382057	381828	381828	382057	382057	20277	16372	0	229
MEA	0	0	0	0	0	170627	170589	170589	170589	170589	170589	39	0	32	0
CO_2	55404	55404	55403	55403	55403	12392	59484	12391	12391	59484	59484	8311	0	0	47093
N_2	107328	107328	107321	107321	107321	0	236	0	0	236	236	107085	0	0	236
O_2	4220	4220	4219	4219	4219	0	23	0	0	23	23	4196	0	0	23
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.685	0.624	0.676	0.676	0.624	0.624	0.145	1	0	0.005
MEA	0	0	0	0	0	0.294	0.279	0.302	0.302	0.279	0.279	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.021	0.097	0.022	0.022	0.097	0.097	0.059	0	0	0.99
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.765	0	0	0.005
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.03	0	0	0
Mole Flow kmol/hr															
H_2O	407	407	231	231	231	22102	21207	21195	21195	21207	21207	1126	909	0	13
MEA	0	0	0	0	0	2793	2793	2793	2793	2793	2793	1	0	1	0
CO_2	1259	1259	1259	1259	1259	282	1352	282	282	1352	1352	189	0	0	1070
N_2	3831	3831	3831	3831	3831	0	8	0	0	8	8	3823	0	0	8
O_2	132	132	132	132	132	0	1	0	0	1	1	131	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.878	0.836	0.873	0.873	0.836	0.836	0.214	1	0	0.012
MEA	0	0	0	0	0	0.111	0.11	0.115	0.115	0.11	0.11	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.011	0.053	0.012	0.012	0.053	0.053	0.036	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.726	0	0	0.008
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.025	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case III (CO_2 recovery: 85% and lean loading: 0.15)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	60.4	122.5	122.5	60.4	105.4	66.5	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	5629	5629	5453	5453	5453	29004	29318	28226	28226	29318	29318	5139	778	0	1092
Mass Flow kg/hr	174283	174283	171105	171105	171105	672841	706392	658796	658796	706392	706392	137554	14019	29	47596
Volume Flow m^3/hr	200000	144433	135448	121661	118133	642	750	667	667	750	7323	120747	14	0	157
Enthalpy MMkcal/hr	-136	-141	-131	-131	-131	-1987	-2044	-1886	-1886	-2044	-2013	-74	-53	0	-102
Mass Flow kg/hr															
H_2O	7332	7332	4162	4162	4162	456211	442396	442194	442194	442396	442396	17977	14019	0	202
MEA	0	0	0	0	0	195465	195436	195436	195436	195436	195436	29	0	29	0
CO_2	55404	55404	55403	55403	55403	21165	68258	21165	21165	68258	68258	8309	0	0	47093
N_2	107328	107328	107321	107321	107321	0	275	0	0	275	275	107046	0	0	275
O_2	4220	4220	4219	4219	4219	0	27	0	0	27	27	4193	0	0	27
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.678	0.626	0.671	0.671	0.626	0.626	0.131	1	0	0.004
MEA	0	0	0	0	0	0.291	0.277	0.297	0.297	0.277	0.277	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.031	0.097	0.032	0.032	0.097	0.097	0.06	0	0	0.989
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.778	0	0	0.006
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.03	0	0	0.001
Mole Flow kmol/hr															
H_2O	407	407	231	231	231	25324	24557	24546	24546	24557	24557	998	778	0	11
MEA	0	0	0	0	0	3200	3199	3199	3199	3199	3199	0	0	0	0
CO_2	1259	1259	1259	1259	1259	481	1551	481	481	1551	1551	189	0	0	1070
N_2	3831	3831	3831	3831	3831	0	10	0	0	10	10	3821	0	0	10
O_2	132	132	132	132	132	0	1	0	0	1	1	131	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.873	0.838	0.87	0.87	0.838	0.838	0.194	1	0	0.01
MEA	0	0	0	0	0	0.11	0.109	0.113	0.113	0.109	0.109	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.017	0.053	0.017	0.017	0.053	0.053	0.037	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.744	0	0	0.009
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.025	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case III (CO_2 recovery: 85% and lean loading: 0.20)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	59.9	121.9	121.9	60	105	63.7	40	40	40
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	5629	5629	5453	5453	5453	34056	34513	33421	33421	34513	34513	4995	634	0	1092
Mass Flow kg/hr	174283	174283	171105	171105	171105	793948	830120	782503	782503	830120	830120	134932	11424	21	47618
Volume Flow m^3/hr	200000	144433	135448	121661	118133	770	881	805	805	881	8388	116431	12	0	140
Enthalpy MMkcal/hr	-136	-141	-131	-131	-131	-2341	-2406	-2243	-2243	-2406	-2370	-66	-43	0	-102
Mass Flow kg/hr															
H_2O	7332	7332	4162	4162	4162	532706	521444	521280	521280	521444	521444	15425	11424	0	164
MEA	0	0	0	0	0	228303	228282	228282	228282	228282	228282	21	0	21	0
CO_2	55404	55404	55403	55403	55403	32939	80033	32940	32940	80033	80033	8308	0	0	47093
N_2	107328	107328	107321	107321	107321	0	329	0	0	329	329	106992	0	0	329
O_2	4220	4220	4219	4219	4219	0	32	0	0	32	32	4187	0	0	32
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.671	0.628	0.666	0.666	0.628	0.628	0.114	1	0	0.003
MEA	0	0	0	0	0	0.288	0.275	0.292	0.292	0.275	0.275	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.041	0.096	0.042	0.042	0.096	0.096	0.062	0	0	0.989
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.793	0	0	0.007
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.031	0	0	0.001
Mole Flow kmol/hr															
H_2O	407	407	231	231	231	29570	28945	28935	28935	28945	28945	856	634	0	9
MEA	0	0	0	0	0	3738	3737	3737	3737	3737	3737	0	0	0	0
CO_2	1259	1259	1259	1259	1259	748	1819	748	748	1819	1819	189	0	0	1070
N_2	3831	3831	3831	3831	3831	0	12	0	0	12	12	3819	0	0	12
O_2	132	132	132	132	132	0	1	0	0	1	1	131	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.868	0.839	0.866	0.866	0.839	0.839	0.171	1	0	0.008
MEA	0	0	0	0	0	0.11	0.108	0.112	0.112	0.108	0.108	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.022	0.053	0.022	0.022	0.053	0.053	0.038	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.765	0	0	0.011
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.026	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case III (CO_2 recovery: 85% and lean loading: 0.25)

	FLUE GAS	FLUE COOL	FLUE SEP	FLUE BLOW	FLUE ABS	LEAN MEA	RICH MEA	LEAN STPR	LEAN P2	RICH P1	RICH STPR	TREAT GAS	WATER	MEA	PROD CO_2
Temperature C	160	40	30	49.3	40	40	59.1	120.9	120.9	59.1	104.1	60	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	5629	5629	5453	5453	5453	40867	41487	40395	40395	41487	41487	4833	472	0	1092
Mass Flow kg/hr	174283	174283	171105	171105	171105	957455	996591	948942	948942	996591	996591	131969	8495	1	47649
Volume Flow m^3/hr	200000	144433	135448	121661	118133	945	1057	990	990	1057	9805	111403	9	0	155
Enthalpy MMkcal/hr	-136	-141	-131	-131	-131	-2819	-2893	-2722	-2722	-2893	-2850	-57	-32	0	-102
Mass Flow kg/hr															
H_2O	7332	7332	4162	4162	4162	635693	627302	627192	627192	627302	627302	12553	8495	0	110
MEA	0	0	0	0	0	272483	272468	272468	272468	272468	272468	15	0	1	0
CO_2	55404	55404	55403	55403	55403	49279	96375	49282	49282	96375	96375	8306	0	0	47093
N_2	107328	107328	107321	107321	107321	0	406	0	0	406	406	106915	0	0	406
O_2	4220	4220	4219	4219	4219	0	39	0	0	39	39	4180	0	0	39
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.664	0.629	0.661	0.661	0.629	0.629	0.095	1	0	0.002
MEA	0	0	0	0	0	0.285	0.273	0.287	0.287	0.273	0.273	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.051	0.097	0.052	0.052	0.097	0.097	0.063	0	0	0.988
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.81	0	0	0.009
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.032	0	0	0.001
Mole Flow kmol/hr															
H_2O	407	407	231	231	231	35286	34821	34814	34814	34821	34821	697	472	0	6
MEA	0	0	0	0	0	4461	4461	4461	4461	4461	4461	0	0	0	0
CO_2	1259	1259	1259	1259	1259	1120	2190	1120	1120	2190	2190	189	0	0	1070
N_2	3831	3831	3831	3831	3831	0	15	0	0	15	15	3817	0	0	15
O_2	132	132	132	132	132	0	1	0	0	1	1	131	0	0	1
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.863	0.839	0.862	0.862	0.839	0.839	0.144	1	0	0.006
MEA	0	0	0	0	0	0.109	0.108	0.11	0.11	0.108	0.108	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.027	0.053	0.028	0.028	0.053	0.053	0.039	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.79	0	0	0.013
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.027	0	0	0.001

Aspen simulation results of MEA based CO_2 absorption for Case III (CO_2 recovery: 85% and lean loading: 0.30)

	FLUE GAS	**FLUE COOL**	**FLUE SEP**	**FLUE BLOW**	**FLUE ABS**	**LEAN MEA**	**RICH MEA**	**LEAN STPR**	**LEAN P2**	**RICH P1**	**RICH STPR**	**TREAT GAS**	**WATER**	**MEA**	**PROD CO_2**
Temperature C	160	40	30	49.3	40	40	57	119.1	119.1	57	102	56.4	40	40	30
Pressure bar	1.013	1.013	1.013	1.2	1.2	2.1	1.2	1.9	2.1	2	2	1.2	2.1	2.1	150
Mole Flow kmol/hr	5629	5629	5453	5453	5453	50634	51381	50289	50289	51381	51381	4706	345	0	1092
Mass Flow kg/hr	174283	174283	171105	171105	171105	1191820	1233300	1185610	1185610	1233300	1233300	129625	6215	7	47697
Volume Flow m^3/hr	200000	144433	135448	121661	118133	1194	1308	1252	1252	1308	10635	107335	6	0	152
Enthalpy MMkcal/hr	-136	-141	-131	-131	-131	-3504	-3585	-3403	-3403	-3585	-3533	-50	-23	0	-102
Mass Flow kg/hr															
H_2O	7332	7332	4162	4162	4162	783421	777236	777211	777211	777236	777236	10346	6215	0	25
MEA	0	0	0	0	0	335750	335741	335741	335741	335741	335741	9	0	7	0
CO_2	55404	55404	55403	55403	55403	72652	119745	72652	72652	119745	119745	8310	0	0	47093
N_2	107328	107328	107321	107321	107321	0	528	0	0	528	528	106793	0	0	528
O_2	4220	4220	4219	4219	4219	0	51	0	0	51	51	4168	0	0	51
Mass Frac															
H_2O	0.042	0.042	0.024	0.024	0.024	0.657	0.63	0.656	0.656	0.63	0.63	0.08	1	0	0.001
MEA	0	0	0	0	0	0.282	0.272	0.283	0.283	0.272	0.272	0	0	1	0
CO_2	0.318	0.318	0.324	0.324	0.324	0.061	0.097	0.061	0.061	0.097	0.097	0.064	0	0	0.987
N_2	0.616	0.616	0.627	0.627	0.627	0	0	0	0	0	0	0.824	0	0	0.011
O_2	0.024	0.024	0.025	0.025	0.025	0	0	0	0	0	0	0.032	0	0	0.001
Mole Flow kmol/hr															
H_2O	407	407	231	231	231	43486	43143	43142	43142	43143	43143	574	345	0	1
MEA	0	0	0	0	0	5497	5496	5496	5496	5496	5496	0	0	0	0
CO_2	1259	1259	1259	1259	1259	1651	2721	1651	1651	2721	2721	189	0	0	1070
N_2	3831	3831	3831	3831	3831	0	19	0	0	19	19	3812	0	0	19
O_2	132	132	132	132	132	0	2	0	0	2	2	130	0	0	2
Mole Frac															
H_2O	0.072	0.072	0.042	0.042	0.042	0.859	0.84	0.858	0.858	0.84	0.84	0.122	1	0	0.001
MEA	0	0	0	0	0	0.109	0.107	0.109	0.109	0.107	0.107	0	0	1	0
CO_2	0.224	0.224	0.231	0.231	0.231	0.033	0.053	0.033	0.033	0.053	0.053	0.04	0	0	0.98
N_2	0.681	0.681	0.703	0.703	0.703	0	0	0	0	0	0	0.81	0	0	0.017
O_2	0.023	0.023	0.024	0.024	0.024	0	0	0	0	0	0	0.028	0	0	0.001

Appendix C: Sensitivity Analysis of CO_2 Capture Cost to Lean Loading (Case III)

Summary of capital cost for lean loading from 0.05 to 0.15 (Case III)

Parameter	Cost ($)		
	0.05	**0.10**	**0.15**
Direct Costs			
Purchased Equipment	14,821,000	11,450,000	6,589,000
Equipment Installation	213,089	152,818	68,058
Piping	6,063,610	4,696,510	4,338,500
Civil	1,550,830	1,056,490	401,909
Steel	152,930	114,626	74,074
Instrumentation	496,930	459,375	454,726
Electrical	400,125	414,697	401,703
Insulation	1,833,480	1,233,360	668,727
Paint	36,790	34,228	35,595
Other	5,796,200	4,475,400	3,480,000
Total Direct Cost	**31,364,984**	**24,087,504**	**16,512,292**
Indirect Costs			
Total Design, Eng, Procurement Cost	1,625,200	1,484,400	1,351,000
G and A Overheads	892,194	678,093	454,839
Contract Fee	1,067,880	818,707	617,647
Contingencies	5,665,260	4,349,330	2,989,410
Total Indirect Cost	**9,250,534**	**7,330,530**	**5,412,896**
Fixed Capital Cost			
Working Capital	2,316,920	1,867,690	1,283,710
Total Capital Cost (IPE)	**42,932,438**	**33,285,724**	**23,208,898**
Reclaimer cost	258,940	258,940	258,940
FGD Equipment cost	3,670,108	3,670,108	3,670,108
CO_2 Drying System cost	3,153,290	3,153,290	3,153,290
Grand Total Capital Cost	**50,014,776**	**40,368,062**	**30,291,236**
Annual Capital cost	**4,721,041**	**3,810,459**	**2,859,278**
Salvage value	10,002,955	8,073,612	6,058,247
Annual Salvage value	**244,001**	**196,939**	**147,778**

Summary of capital cost for lean loading from 0.20 to 0.30 (Case III)

Parameter	Cost ($)		
	0.20	**0.25**	**0.30**
Direct Costs			
Purchased Equipment	6,781,600	5,733,500	5,645,400
Equipment Installation	68,851	53,158	54,714
Piping	3,386,990	2,877,390	2,600,150
Civil	434,665	282,527	267,347
Steel	78,996	66,856	64,580
Instrumentation	444,684	435,307	414,631
Electrical	414,632	399,835	399,835
Insulation	637,718	503,343	500,808
Paint	28,822	28,615	29,747
Other	3,273,100	3,016,100	2,911,100
Total Direct Cost	**15,550,057**	**13,396,631**	**12,888,312**
Indirect Costs			
Total Design, Eng, Procurement Cost	1,357,500	1,347,200	1,339,500
G and A Overheads	425,777	361,483	346,464
Contract Fee	558,489	493,385	454,711
Contingencies	2,810,840	2,422,760	2,327,210
Total Indirect Cost	**5,152,606**	**4,624,828**	**4,467,885**
Fixed Capital Cost			
Working Capital	1,207,030	1,040,380	999,350
Total Capital Cost (IPE)	**21,909,693**	**19,061,839**	**18,355,547**
Reclaimer cost	258,940	258,940	258,940
FGD Equipment cost	3,670,108	3,670,108	3,670,108
CO_2 Drying System cost	3,153,290	3,153,290	3,153,290
Grand Total Capital Cost	**28,992,031**	**26,144,177**	**25,437,885**
Annual Capital cost	**2,736,643**	**2,467,825**	**2,401,156**
Salvage value	5,798,406	5,228,835	5,087,577
Annual Salvage value	**141,440**	**127,547**	**124,101**

Summary of operating cost for lean loading from 0.05 to 0.15 (Case III)

Parameters	**Cost ($/yr.)**		
	0.05	**0.10**	**0.15**
Manufacturing Costs			
Direct Production Costs			
Raw material cost	727,494	648,369	644,715
Utility cost	54,065,897	27,728,480	18,229,002
Operating Labor cost	640,000	640,000	640,000
Operating Supervision cost	280,000	280,000	280,000
Maintenance and Repair cost	1,040,000	759,000	387,000
Operating Supplies and Laboratory charge (25% of Operating labor costs)	230,000	230,000	230,000
Total Direct Production Cost	56,255,897	29,637,480	19,766,002
Plant Overhead Cost (50% of Operating Labor and maintenance cost)	980,000	839,500	653,500
General and Administrative cost	4,578,872	2,438,158	1,633,560
Operating cost (at second year)	61,814,769	32,915,139	22,053,062
Annual Operating Cost (over 20 years)	70,645,441	37,617,297	25,203,496
Baghouse, FGD Scrubber chemical cost	1,069,510	1,069,510	1,069,510
MEA make-up cost from degradation	866,511	866,511	866,511
Total Annual Operating Cost	72,581,463	39,553,318	27,139,517

CO_2 capture cost for lean loading from 0.05 to 0.15 (Case III)

Parameters	**0.05**	**0.10**	**0.15**
Annual Capital Cost	4,721,041	3,810,459	2,859,278
Annual Salvage value	244,001	196,939	147,778
Annual Operating Cost	72,581,463	39,553,318	27,139,517
Total Annual Cost ($)	**77,058,503**	**43,166,838**	**29,851,017**
Total Annual CO_2 Captured (tonne)	376,744	376,744	376,744
$ / tonne of CO_2 Captured	**205**	**115**	**79**

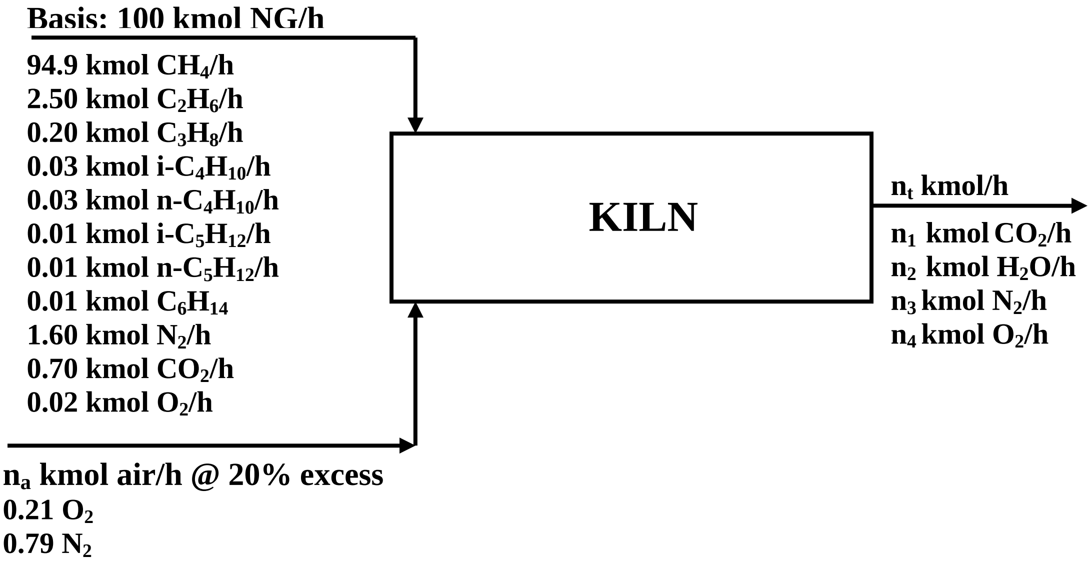
Summary of operating cost for lean loading from 0.20 to 0.30 (Case III)

Parameters	Cost ($/yr.)		
	0.20	0.25	0.30
Manufacturing Costs			
Direct Production Costs			
Raw material cost	594,485	582,942	577,200
Utility cost	13,807,562	11,356,829	11,070,838
Operating Labor cost	640,000	640,000	640,000
Operating Supervision cost	280,000	280,000	280,000
Maintenance and Repair cost	338,000	297,000	284,000
Operating Supplies and Laboratory charge (25% of Operating labor costs)	230,000	230,000	230,000
Total Direct Production Cost	15,295,562	12,803,829	12,504,838
Plant Overhead Cost (50% of Operating Labor and maintenance cost)	629,000	608,500	602,000
General and Administrative cost	1,273,965	1,072,986	1,048,547
Operating cost (at second year)	17,198,527	14,485,315	14,155,385
Annual Operating Cost (over 20 years)	19,655,457	16,554,644	16,177,581
Baghouse, FGD Scrubber chemical cost	1,069,510	1,069,510	1,069,510
MEA make-up cost from degradation	866,511	866,511	866,511
Total Annual Operating Cost	21,591,478	18,490,665	18,113,602

CO_2 capture cost for lean loading from 0.20 to 0.30 (Case III)

Parameters	0.20	0.25	0.30
Annual Capital Cost	2,736,643	2,467,825	2,401,156
Annual Salvage value	141,440	127,547	124,101
Annual Operating Cost	21,591,478	18,490,665	18,113,602
Total Annual Cost ($)	**24,186,681**	**20,830,943**	**20,390,657**
Total Annual CO_2 Captured (tonne)	376,744	376,744	376,744
$ / tonne of CO_2 Captured	**64**	**55**	**54**

Appendix D: Results of Switching to a Lower Carbon Content Fuel at Average Plant Load (Case II)



Combustion Reactions

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$$

$$C_2H_6 + \frac{7}{2}O_2 \rightarrow 2CO_2 + 3H_2O$$

$$C_3H_8 + 5O_2 \rightarrow 3CO_2 + 4H_2O$$

$$C_4H_{10} + \frac{13}{2}O_2 \rightarrow 4CO_2 + 5H_2O$$

$$C_5H_{12} + 8O_2 \rightarrow 5CO_2 + 6H_2O$$

$$C_6H_{14} + \frac{19}{2}O_2 \rightarrow 6CO_2 + 7H_2O$$

Calcinations Reaction

$$CaCO_3 \rightarrow CaO + CO_2$$

Basis: 100 kmol Natural Gas per hour

Theoretical O_2 *required* :

$$\frac{(0.949)\,(100)\ \text{kmol}\ CH_4}{\text{h}}\left|\frac{2\ \text{kmol}\ O_2}{1\ \text{kmol}\ CH_4}\right.+\frac{(0.025)\,(100)\ \text{kmol}\ C_2H_6}{\text{h}}\left|\frac{3.5\ \text{kmol}\ O_2}{1\ \text{kmol}\ C_2H_6}\right.+$$

$$\frac{(0.002)\,(100)\ \text{kmol}\ C_3H_8}{\text{h}}\left|\frac{5\ \text{kmol}\ O_2}{1\ \text{kmol}\ C_3H_8}\right.+\frac{(0.0006)\,(100)\ \text{kmol}\ C_4H_{10}}{\text{h}}\left|\frac{6.5\ \text{kmol}O_2}{1\ \text{kmol}\ C_4H_{10}}\right.+$$

$$\frac{(0.0002)\,(100)\ \text{kmol}\ C_5H_{12}}{\text{h}}\left|\frac{8\ \text{kmol}\ O_2}{1\ \text{kmol}\ C_5H_{12}}\right.+\frac{(0.0001)\,(100)\ \text{kmol}\ C_6H_{14}}{\text{h}}\left|\frac{9.5\ \text{kmol}\ O_2}{1\ \text{kmol}\ C_6H_{14}}\right.$$

$$= 200.195\ \frac{kmol\ O_2}{h}$$

$$\text{Air feed rate, } n_a = \frac{200.195\ \text{kmol}\ O_2}{\text{h}}\left|\frac{1\ \text{kmol air}}{0.21\ \text{kmol}\ O_2}\right|\frac{1.20\ \text{kmol air}}{1\ \text{kmol air}}$$

$$= 1143.97\,\frac{kmol\ air}{h}$$

CO_2 Balance:

$$n_1\ \frac{\text{kmol}\ CO_2}{\text{h}} =$$

$$\frac{(0.949)\,(100)\ \text{kmol}\ CH_4}{\text{h}}\left|\frac{1\ \text{kmol}\ CO_2}{1\ \text{kmol}\ CH_4}\right.+\frac{(0.025)\,(100)\ \text{kmol}\ C_2H_6}{\text{h}}\left|\frac{2\ \text{kmol}\ CO_2}{1\ \text{kmol}\ C_2H_6}\right.+$$

$$\frac{(0.002)\,(100)\ \text{kmol}\ C_3H_8}{\text{h}}\left|\frac{3\ \text{kmol}\ CO_2}{1\ \text{kmol}\ C_3H_8}\right.+\frac{(0.0006)\,(100)\ \text{kmol}\ C_4H_{10}}{\text{h}}\left|\frac{4\ \text{kmol}\ CO_2}{1\ \text{kmol}\ C_4H_{10}}\right.+$$

$$\frac{(0.0002)\,(100)\ \text{kmol}\ C_5H_{12}}{\text{h}}\left|\frac{5\ \text{kmol}\ CO_2}{1\ \text{kmol}\ C_5H_{12}}\right.+\frac{(0.0001)\,(100)\ \text{kmol}\ C_6H_{14}}{\text{h}}\left|\frac{6\ \text{kmol}\ CO_2}{1\ \text{kmol}\ C_6H_{14}}\right.+$$

$$\left[\frac{(0.007)\,(100)\ \text{kmol}\ CO_2}{\text{h}}\right]_{in}$$

$$n_1 = 101.6\ \frac{kmol\ CO_2}{h}$$

H_2O Balance:

$$n_2 \frac{\text{kmol H}_2\text{O}}{\text{h}} = \frac{(0.949)\,(100)\,\text{kmol CH}_4}{\text{h}}\left|\frac{2\,\text{kmol H}_2O}{1\,\text{kmol CH}_4}\right. + \frac{(0.025)\,(100)\,\text{kmol C}_2\text{H}_6}{\text{h}}\left|\frac{3\,\text{kmol H}_2O}{1\,\text{kmol C}_2\text{H}_6}\right. +$$

$$\frac{(0.002)\,(100)\,\text{kmol C}_3\text{H}_8}{\text{h}}\left|\frac{4\,\text{kmol H}_2O}{1\,\text{kmol C}_3\text{H}_8}\right. + \frac{(0.0006)\,(100)\,\text{kmol C}_4\text{H}_{10}}{\text{h}}\left|\frac{5\,\text{kmol H}_2O}{1\,\text{kmol C}_4\text{H}_{10}}\right. +$$

$$\frac{(0.0002)\,(100)\,\text{kmol C}_5\text{H}_{12}}{\text{h}}\left|\frac{6\,\text{kmol H}_2O}{1\,\text{kmol C}_5\text{H}_{12}}\right. + \frac{(0.0001)\,(100)\,\text{kmol C}_6\text{H}_{14}}{\text{h}}\left|\frac{7\,\text{kmol H}_2O}{1\,\text{kmol C}_6\text{H}_{14}}\right.$$

$$\text{n}_2 = 198.59\,\frac{\text{kmol H}_2\text{O}}{\text{h}}$$

N_2 Balance:

$$n_3 \frac{\text{kmol N}_2}{\text{h}} = \left[\frac{(1143.97)\,\text{kmol air}}{\text{h}}\left|\frac{0.79\,\text{kmol N}_2}{1\,\text{kmol air}}\right.\right] + \left[\frac{(100)\,\text{kmol fuel}}{\text{h}}\left|\frac{0.016\,\text{kmol N}_2}{1\,\text{kmol fuel}}\right.\right]$$

$$\text{n}_3 = 905.337\,\frac{\text{kmol N}_2}{\text{h}}$$

O_2 Balance:

$$n_4 \frac{\text{kmol O}_2}{\text{h}} = \left[\frac{(1143.97)\,\text{kmol air}}{\text{h}}\left|\frac{0.21\,\text{kmol O}_2}{1\,\text{kmol air}}\right.\right] + \left[\frac{(100)\,\text{kmol fuel}}{\text{h}}\left|\frac{0.0002\,\text{kmol O}_2}{1\,\text{kmol fuel}}\right.\right]$$

$$n_4 = 240.254\,\frac{\text{kmol O}_2}{\text{h}}$$

$$\text{Total, n}_t = \text{n}_1 + \text{n}_2 + \text{n}_3 + \text{n}_4$$

$$= 101.6 + 198.59 + 905.337 + 240.254$$

$$= 1445.781\,\frac{\text{kmol}}{\text{h}}$$

$$\text{Kiln Thermal Efficiency} \;=\; 790.00\,\frac{\text{kcal}}{\text{kg clinker produced}}$$

$$\text{Natural Gas heating value} = 37.8\,\frac{\text{MJ}}{\text{m}^3\,(\text{STP})}$$

$$= \frac{37.8 \text{ x } 10^6 \;\; \text{J}}{\text{m}^3}\left|\frac{0.23901 \;\; \text{cal}}{1 \;\; \text{J}}\right|\frac{1 \;\; \text{kcal}}{10^3 \;\; \text{cal}}$$

$$= 9034.578\,\frac{\text{kcal}}{\text{m}^3\,(\text{STP})}$$

For Case II (average plant load) : Clinker Production Rate = 94.2 tonne per hour

$$\text{Natural Gas Consumption} \;=\; \frac{94.2 \text{ tonne clinker}}{\text{h}}\left|\frac{10^3 \text{ kg clinker}}{1 \text{ tonne clinker}}\right|\frac{790.00 \text{ kcal}}{\text{kg clinker}}\left|\frac{\text{m}^3\,(\text{STP})}{9034.578 \text{ kcal}}\right.$$

$$\approx 8237 \; SCMH$$

$$\approx \frac{8237 \; m^3\,(STP)}{h}\left|\frac{1 \; kmol}{22.4 \; x \, 10^3 \;\; liters\,(STP)}\right|\frac{10^3 \;\; liters}{1 \, m^3}$$

$$\approx 367.724\,\frac{kmol}{h}$$

Scale Factor (SF) = 3.6777

$$\text{Natural Gas Required} = 367.724\,\frac{\text{kmol}}{\text{h}}$$

$$\text{Air Flow Required} = 4207.21\,\frac{\text{kmol}}{\text{h}}$$

$$\text{CO}_2 : \; (101.6)\,(3.6777)\,\frac{\text{kmol CO}_2}{\text{h}}$$

$$= 373.607\,\frac{\text{kmol CO}_2}{\text{h}}$$

$$\approx 16442.444\,\frac{\text{kg CO}_2}{\text{h}}$$

$$\text{H}_2\text{O} : \; (198.59)\,(3.6777)\,\frac{\text{kmol H}_2 O}{\text{h}}$$

$$= 730.263\,\frac{\text{kmol H}_2 O}{\text{h}}$$

$$\approx 13156.418\,\frac{\text{kg H}_2 O}{\text{h}}$$

$$\text{N}_2 : \; (905.337)\,(3.6777)\,\frac{\text{kmol N}_2}{\text{h}}$$

$$= 3329.142\,\frac{\text{kmol N}_2}{\text{h}}$$

$$\approx 93282.558\,\frac{\text{kg N}_2}{\text{h}}$$

$$\text{O}_2 : \; (240.254)\,(3.6777)\,\frac{\text{kmol O}_2}{\text{h}}$$

$$= 883.472\,\frac{\text{kmol O}_2}{\text{h}}$$

$$\approx 28271.104\,\frac{\text{kg O}_2}{\text{h}}$$

$$\text{CO}_2\ \text{Generation from Calcination} = \frac{94.2\ \text{tonne clinker}}{\text{h}}\left|\frac{10^3\,\text{kg clinker}}{1\ \text{tonne clinker}}\right|\frac{0.52\,\text{kg CO}_2}{1\,\text{kg clinker}}$$

$$\approx 48984.0\,\frac{\text{kg CO}_2}{\text{h}}$$

For Natural Gas Fuel:

$$\text{Total CO}_2\ \text{generation} = \left[48984.0 + 16442.444\right]\frac{\text{kg CO}_2}{\text{h}}$$

$$\approx 65426.444\ \frac{kg\ CO_2}{h}$$

Case II: For Petcoke, coke, and coal fines Fuel:

$$\text{Total CO}_2\ \text{generation} = 80335\ \frac{\text{kg CO}_2}{\text{h}}$$

$$\text{CO}_2\ \text{emissions reduced} = (80335.0 - 65426.444)\frac{\text{kg CO}_2}{\text{h}}$$

$$\approx 14909.0\ \frac{kg\ CO_2}{h}$$

$$Percent\ CO_2\ emissions\ reduced = \frac{(80335.0 - 65426.444)}{80335.0}\,x\,100$$

$$\approx 18.55$$