POLITECNICO DI TORINO

DIMEAS - Department of Mechanical and Aerospace Engineering

MASTER OF SCIENCE
IN
MECHANICAL ENGINEERING

Master’s Thesis

Modeling, performance assessment and cost estimation of a carbon capture process based on chemisorption with MEA

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Abstract

From the last few couple of decades, it has been seen that the greenhouse gas emissions are risen by more than one quarter and the most contribution in that, is the fast-growing amount of CO_2into the atmosphere. The major part of CO_2 comes from the energy production system. Carbon capture and storage (CCS) technology has been adopted from the last few years to reduce the CO_2 concentration into the atmosphere even though the research is still ongoing to minimize the cost and the energy penalty to facilitate the process. So, the purpose of this thesis is to optimize post-combustion carbon capture system using 30% MEA-solvent as an absorbent. The absorber and stripper columns were modelled with RadFrac and kinetic reactions and the techno-economic analysis has been performed: the purchasing cost of equipment is calculated according to a certain size, and operating conditions.

In this work, a detailed flowsheet model has been built with MEA-solvent by chemical absorption system on Aspen Plus® version 10.0 (Aspentech, USA), with rate-based technique. The final goal was set to get 99% of carbon capture from the flue gas and minimize the heat duty in desorption tower (stripper), in form of re-boiling and condensing. Softening of the flue gas, compression of CO_2 and storage is not the scope of this work.

Acknowledgment

I am grateful for the opportunity to work on CCS post-combustion carbon capture with MEA-solvent by chemical absorption and I feel obliged to sincerely thank Professor Massimo Santarelli for their continuous support and learnings.

I would like to thank Giulio Buffo for his insights, support, and encouragement regarding this work.

Grateful for a loving family and friends.

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Abbreviations

GHG - greenhouse gas
IPCC - intergovernmental panel on climate change,
CCS - carbon capture and storage
ASU - air separation unit
MEA - Monoethanolamine
PSA - Pressure swing adsorption
TSA - Temperature swing adsorption
VPSA - vacuum pressure swing
EOR - enhance oil recovery
FCBM - enhance coalbed methane
GLCD - gas lift advanced dissolution
LMTD - logarithmic mean temperature difference
CAPEX - capital expenditure
BEC - bare erected cost
EPCC - engineering procurement and construction cost
TPC - total plant cost
TOP - total overnight cost
TASC - total overnight cost
SS - Stainless steel
CS - carbon steel
ASME - American society of mechanical engineers
CEPCI - Chemical engineering plant cost index
TOP - total operating cost
FOC - fixed operating cost
VOC - variable operating cost

FCI - fixed capital investment

Chapter 1

Introduction

The biggest problem is the climate change that the world had now the reason behind is man-made things that are contributing to an alarming situation to man itself, the major contribution is the drastically increasing level of CO2 due to burning process, in different ways surrounding, and especially the low-cost fossil fuel and their high availability. Carbon exists in the atmosphere, earth, water in various chemical forms so, it's also essential naturally available products such as biomass, petroleum, and mineral deposit. Over many decades the carbon dioxide has been sub quested by plants and eventually ends up in the form of petroleum and coal behind the face of the earth, highly increasing demand for energy in everyday life can be derived only by producing energy by fossil fuel, in the form of coal, natural gas, and crude oil that have a high concentration of carbon and due to burning these fuel, it emits carbon dioxide to the atmosphere that enhances the greenhouse gas (GHG) effect and cause the climate change, only in the last two decades the rise in GHG is one quarter. According to the intergovernmental panel on climate change (IPCC) issued a report on global warming stated that at the current rate of warming, the global temperature rise should reach the 1.5°C threshold between 2030 and 2052 [2]

Before the start of the industrial revolution, the amount of CO2 content in the atmosphere was 35% lower than the current availability which is about 385 ppmv (Wikipedia, GHG) and this value is exponentially increasing as the man releases the covered CO2 below the surface of the earth, CO2 is known as GHG that traps energy that is being radiated from the earth surface.

Many countries that understand the effect and alarming condition of the CO2 into the atmosphere and understand how difficult it's to overcome the side effect of global warming for the human life, there are already many countries that have started a new concept in the field of energy production such as follows

1. Wind energy
2. Solar energy
3. Tide energy
4. Hydropower energy

These are some clean energy production system that has been started in many countries around the world to overcome the energy demand, despite all these efforts that have been made by many countries the targeted value of the CO2 is so far. Having all these problems that cause global warming the world leader set together and acknowledge that no single solution exists that overcome this situation so a portfolio of CO2 emission control technologies has to be adopted and new legislation has been made to control this global warming and reduce the climate change. In this context, there is a method that's used to capture the carbon dioxide including storage CCS.

1.1 Carbon capture and storage (CCS)

In this technique, waste carbon is captured from industrial stack before extracted to the atmosphere and then transported to the place where it gets stored by injecting it into the ground and permanently trap into the down wells below the surface.CSS utilize different emission sources that include the fossil fuel of power plants, paper, cement, steel, and different other process industries. The basic purpose to use this method is, it allows fossil fuel for the process and helps to reduce the carbon emission in the atmosphere. It's defined as carbon dioxide (CO_2) emissions produced from the use of fossil fuels in electricity generation and industrial process, preventing the carbon dioxide from entering the atmosphere.

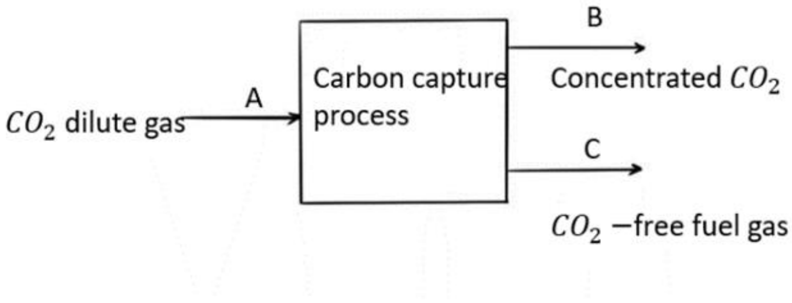


Figure 1 Thermodynamic of CCS process

How much work is required to produce pure CO2 stream "B" from the diluted Stream A is calculated by the following equation

$$W_{sep}{}^{min} = E_{ch,B} + E_{ch,C} - E_{ch,A} [J]$$

$E_{ch,B}$= exergy of pure CO2

$E_{ch,A}$= exergy of gas mixture

$E_{ch,C}$= exergy for gases free from carbon

So, the minimum required work is equal to the exergy difference between the inlet stream and outlet streams as shown above in Figure 1.

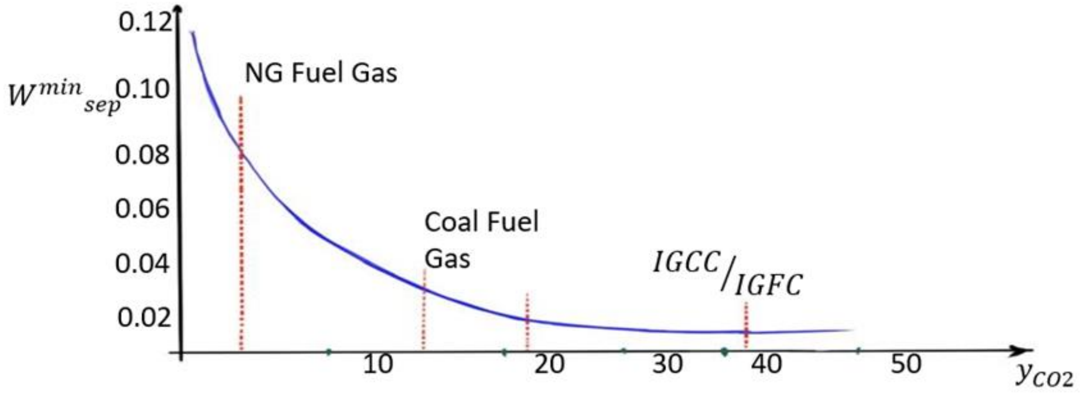


Figure 2 W^{min} changes with respect to the carbon concentration and fuel used from the above graph, the more diluted CO2 is in the exhaust gases it leads to the high separation work to get pure CO2, it also shows that Coal/Biomass gasification process is rich in CO2 content which means that low min work is required to get a higher amount of CO2.

1.2 Carbon Capture and Storage Technologies

There are three different types of carbon capture technologies, which are as follows

1. Post-combustion carbon capture
2. Pre-combustion carbon capture
3. Oxyfuel process

All these processes follow the same CCS methodology that has four different steps

1. Capture the CO2
2. CO2 Compression with liquefaction
3. Pipeline transportation
4. Storage

1.2.1 Pre-combustion:

In the pre-combustion capture the fossil fuels are reformed into a synthesis gas (syngas) comprised mainly of hydrogen and CO_2. The process of decarbonization changes depending on the type of fuel, used in natural gas the syngas is obtained by steam reforming or auto-thermal reforming; if coal is used, it is obtained by gasification [3] The syngas is then further converted to more hydrogen through the water-gas shift (WGS) reaction, resulting in high-pressure CO_2 and H_2. The high partial pressure of

CO_2 leads to the higher driving force, more technologies available for separation and lower costs for compression. However, this technology applies mainly to new plants, as a few gasification plants are currently in operation. The barriers to the commercial application of gasification are the cost of equipment and the extensive requirements of the supporting systems.

1.2.2 Oxyfuel process:

In the oxyfuel capture, also called oxyfiring, pure oxygen is burnt with gaseous fuel resulting in a very high concentration of CO_2 in the flue gases. Since the combustion products consist mainly of water vapor and CO_2, the condensation of the water is the only treatment required and the energy demand for CO_2 separation keeps at relatively low levels. An advantage of this technology is also that it allows the retrofitting and repowering of existing plants. The most common method to produce oxygen for large scale oxyfuel plants is a cryogenic air separation unit (ASU) that has a high cost if a large amount of cryogenic O_2 is required. Since the burning of fossil fuel in an atmosphere of oxygen leads to excessively high temperatures, such as 3500°C, the recycling of cooled CO_2 is required to maintain the temperatures within the limits of the operating temperatures of the combustor materials.

1.2.3 Post-combustion:

Post-combustion capture consists of a downstream process that captures CO_2 from conventional power plants by scrubbing of the flue gases before they are vented. This method doesn’t require fuel treatment and the air is used as the oxidant. The main advantages of post-combustion capture systems are that existing plants can be retrofitted with a capture unit without further rearrangements and that the technology is well established. Barriers to implementation are the very significant amount of thermal and/or electrical energy required and the low CO_2 partial pressure due to the dilution of flue gas and the presence of nitrogen.

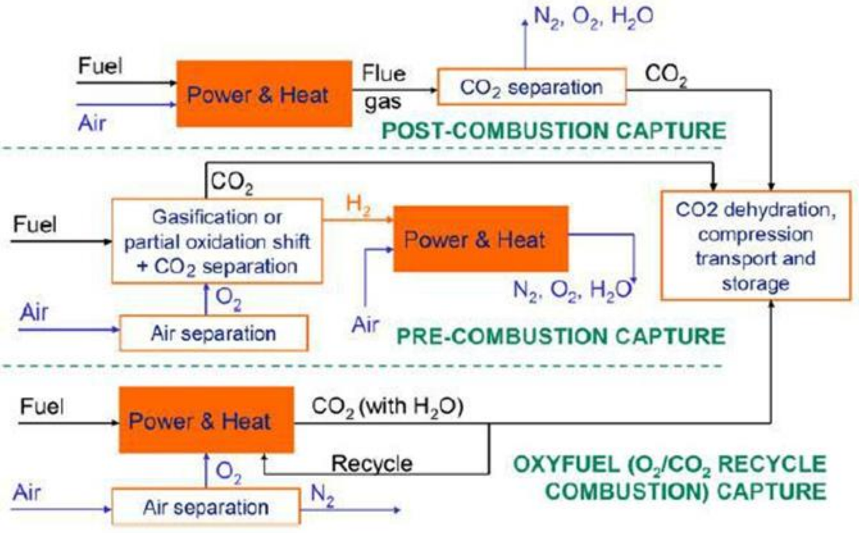


Figure 3 Operating principles of post-combustion capture, pre-combustion capture and oxyfuel capture

All these technologies can be achieved by several ways of carbon dioxide removal processes such as solid adsorption, absorption into a gas-liquid solvent, membranes or other physical or biological separation methods. Since the focus of this thesis is the post-combustion capture of CO_2, several general methods for this technology are briefly discussed, including:

Chemical absorption

Membrane separation

Adsorption.

1.3 Chemical absorption

Absorption processes are the most used technologies for CO_2 capture. Generally, it's divided into two categories. Processes where the solvent chemically reacts with the dissolved gas are referred to as chemical absorption processes. For these applications, alkanolamines are commonly used as reactive absorbents. In the physical absorption processes, the solvent only interacts physically with the dissolved gas. The used solvents have thermodynamic properties as the relative absorption of the gas is favored over the other components of the gas mixture. Some commonly used physical solvents are methanol (Rectisol Process) and glycol ethers (Selexol Process). In many industrial applications, combinations of physical solvents and reactive absorbents may be used.

Many comparative assessment studies have shown that absorption processes based on chemical solvents are currently the preferred option for post-combustion CO_2 capture. Systems with chemical absorption have been in use since the 1930s for the capture of CO_2 from ammonia plants for use in food applications and hence, are a commercially realized technology, though not at the scale required for power plants[3]. The absorption processes in post-combustion capture exploit the reversible nature of the chemical reaction of an aqueous alkaline solvent, usually an amine, with an acid or sour gas, such as CO_2. The gas is separated from the flue gas bypasses the flue gas through a continuous scrubbing system, where the amines react with CO_2 to form water-soluble compounds. Figure 4 shows the process flow diagram for the most commonly used amine, i.e. Monoethanolamine (MEA). The flue gas is bubbled through the solvent in a packed absorber column, where the solvent preferentially removes the CO_2 from the flue gas. Afterward, the solution rich in solvent passes through a regenerator unit called stripper, where the absorbed CO_2 is stripped from the solvent by counterflowing steam at 120-150°C. After cooling, the regenerated solvent is cooled to 40-65°C and is recycled into the absorption column [9]. Water vapor is condensed, leaving at a highly concentrated CO_2 stream, which may be dried and compressed for commercial utilization or transported to safe geological storage.

The MEA process is well-established, but the energy consumptions are substantial. The heat required for the stripping stage depends on the operating conditions, such as the loading, the MEA-mass concentration in water, the temperature and the pressure of the stripper. Many studies have measured the heat of the absorption of the MEA solvent at 80-120 kJ/mol_{CO2} [4]. This heat is provided as steam in the reboiler at the bottom of the stripper. The reboiler duty, the pumping and the compression powers are provided by the power plant resulting in the reduction of its efficiency. Figure 4 also shows some additional equipment needed to maintain the solution quality as a result of the formation of degradation products, corrosion products and the presence of particles. This is generally done using filters, carbon beds and thermally operated reclaimers. Control of degradation and corrosion has been an important aspect in the development of absorption processes over the past few decades.

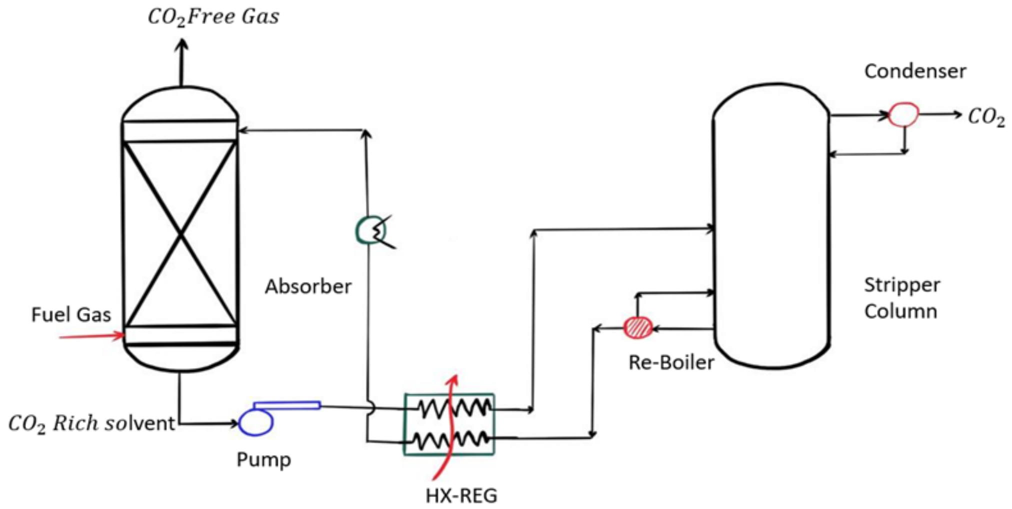


Figure 4Flow diagram of amine-based absorption for post-combustion capture

Some relatively large CCS demonstration projects have been explored that use amine absorption to capture CO_2 from power plant flue gas. Amine absorption is being considered for CO_2 capture because it is a proven technology used successfully to treat industrial gas streams for decades. However, many studies have shown that amine absorption for CO_2 capture is going to be costly and energy intensive. Reports predict that an amine system used to capture 90% of the CO2 in flue gas will have an energy penalty of about 30%. Thus, the main goals of technological improvements should be to reduce the energy and the solvent consumptions and eventually reduce the operation cost and investment. Because amine absorption is a mature technology, there are unlikely to be significant future improvements in this process.

1.4 Membrane separation

Several studies have shown that the gas separation membrane is one of the most promising capture technologies. In a gas separation membrane process, the membranes act as a filter to separate one or more gas components from the feed gas mixture and produce a specific gas-rich permeate stream and a retentate stream poor of the captured gas. The driving force is the pressure difference between the feed side at high pressure and the permeate side at low pressure, as shown below by Fick's law [5]

$$J_i = \frac{P_i^*}{\delta} A_m\left(x_i . P_f - y_i P_p\right) = \frac{P_i^*}{\delta} A_m \Delta P$$

J_i=flux across the membrane

P_i^*=Permeability value for the component i

δ = membrane thickness

A_m=membrane area

x_i, y_i=mole fraction of component "i", in feed side and permeate side

P_f, P_p=pressure at feed and permeate side

Permeability (permeance=permeability/thickness) and selectivity are the two parameters typically used to characterize the membrane performance. Permeance is defined as the flux of a specific gas passing through a membrane at a given pressure and temperature, while selectivity is evaluated by the ratio of the permeability values of different gas species. The separation principle is based on the difference in the transport properties of the gas molecules: for example, CO_2 is a fast diffusing gas molecule compared to N_2 and O_2 in many membrane materials, such as glassy and rubbery polymers and carbon molecular sieve membranes. The parameters that affect the CO_2 purity are the pressure ratio and the membrane CO_2/N_2 selectivity. Using Fick's law, the ratio of the fluxes for the CO_2 to N_2 is

$$\frac{n(CO_2)}{n(N_2)} = \frac{P_{co2}{}^{*}}{P_{N_2}{}^{*}} * \frac{\left(x_{co2} - y_{co2}\frac{P_f}{P_p}\right)}{\left(x_{N_2} - y_{N_2}\frac{P_f}{P_p}\right)}$$

Thus, the ratio of the flux of CO_2 compared to the flux of N_2 (or the purity of CO_2 in the permeate stream) can be increased by increasing the CO_2/N_2 selectivity and/or increasing the pressure ratio (ratio of the feed pressure to the permeate pressure). Furthermore, high CO_2 permeance is also required to reduce the membrane area. Commercially available polymeric gas separation membranes are mostly used with energy demands of 70-75 kWh per ton of recovered CO_2, while typical capture rates are 82-88% [6].

1.5 Adsorption

Solid materials with a high surface area can potentially be used to separate CO_2 from flue gases by physical adsorption. The gas is fed to an adsorbent bed that adsorbs the CO_2, while the rest of the gas passes through. When the bed is fully loaded, the feed is switched to the next bed on the train. The loaded bed is then regenerated by pressure swing adsorption (PSA) or temperature swing adsorption (TSA). In general, PSA is preferred over TSA because TSA systems are larger and involve more heat for the regeneration of the adsorbent[3]. Vacuum pressure swing adsorption (VPSA) is a subset of the larger category of PSA. While PSA vents to atmospheric pressures and uses a pressurized feed gas, VPSA draws the gas through the separation process with a vacuum. For the adsorption systems, the typical recovery of the CO_2 gas can be in the range of 85-90% with energy demands from 160 to 180 kWh/ton CO_2 recovered[6]. A disadvantage in the adsorption systems is that the gases must be treated before treatment in the adsorbent bed since the stability of adsorbents in the presence of impurities is low. The most important design decision when developing an adsorption system is the choice of the adsorbent. Hundreds of materials have been studied as potential CO_2

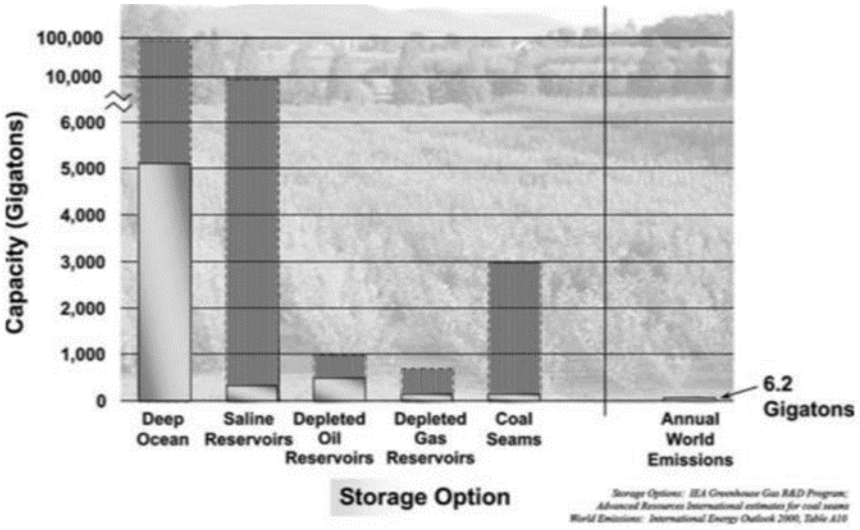
adsorbents but only a few classes of adsorbents are being used to remove CO_2 from gas streams.

1.6 CO_2 transport

Once CO_2 has been captured, it must be transported to a storage facility as gas, liquid or solid. Transportation can be performed via pipeline, ship, rail, truck or a combination of these means. Gas at conditions close to the atmospheric pressure occupies a large volume and needs very large facilities to be transported. If gas is compressed, it occupies less volume and can be transported by pipeline or ship. Volume can be further reduced by liquefaction or solidification. Liquefaction is an established technology for gas transport by ship, while solidification still requires much more energy to be cost-competitive with other options. Transportation of CO_2 by pipeline is the most economical means for moving large quantities of CO_2 for long distances. It is transported at a phase called dense phase which means at physical conditions above the CO_2 critical point. In the dense phase, CO_2 behaves as a compressible liquid with a density of about 900 kg/m^3. The pressure at the injection must be above 70 bar up to about 200 bar to accomplish a dense phase. This means that the pressure drop in the pipeline must be compensated by the CO_2 compressor at the capture plant. To avoid corrosion in the pipeline, the water content in the stream must be below 500 ppm (vol %). This limitation can be achieved using coolers and separators in the capture plant [7].

1.7 CO_2 storage

After recovering the CO_2 gas, it must be stored somewhere to prevent it from appearing in the atmosphere. Many different options for storage CO_2 have been explored, including geological sequestration, terrestrial sequestration, oceanic sequestration, and novel sequestration systems. Geological sequestration is currently the most advanced and the most likely approach to be deployed on a large scale in the future. To be geologically stored, CO_2 needs first to be compressed, usually to a dense fluid state known as ‘supercritical fluid’. CO_2 is injected into underground reservoirs where it is expected to be isolated from the atmosphere for several hundred years. Possible geological storage formations include basins, oil and gas fields, depleted gas fields, saline formations, and coal beds Geological storage is possible both onshore and offshore, with offshore sites accessed through pipelines from shore or offshore platforms



Storage Options: IEA Greenhouse Gas R&D Program; Advanced Resources International estimates for coal seams
World Emissions: International Energy Outlook 2000, Table A10

Figure 5 Large potential worldwide storage capacity[8]

Figure 5 shows that the estimated capacity of geologic formations is so much larger than the annual world emissions, accounting for 6.2 Gtons, to can store CO_2 emissions for many centuries..

Objectives:

In this thesis, only post-combustion carbon capture with MEA chemical absorption, system is considered with MEA used commercially from last few decades despite several disadvantages including the loss of solvent, high volatility, high energy consumption during desorption process, degradation and rusting as well, on the other hand, cost of MEA is very low as compared to the other IL’s. For simulation

Aspenplus® software was used and the final purpose is to achieve 99% carbon capture, while transportation and storage of CO2 is not the scope of this thesis

Chapter 2

Methodology

Process description

In this chapter the process of MEA-solvent chemical absorption,CO_2 separation is described from the exhaust fuel gases of a coal-fired thermal power plant in Italy. A detail description of the gases and the working condition and process is as follows

2.1 Industrial stack

The stack of a thermal power plant that contains the composition of different gases is shown below in table 1 and the composition of exhaust gas is mentation in table 2, the softening process is already done which is not part of this thesis so, ash and other acidic gases are removed before the entering into the carbon capture unit.

Parameter	**Units**	**Values**
Plant size	MW	150
Utilization factor	%	85
Mass-Flow rate of exhaust gases	ton/h	581
Gases temperature	°C	200
Gases pressure	bar	1.01

Table 1Process exhaust fuel gases thermodynamics

Parameter	Mass fraction %
CO_2	0.1155
H_2O	0.0323
N_2	0.8038
O_2	0.0484

Table 2 Exhaust fuel gases composition

2.2 Working principle of MEA absorption system

As in post-combustion carbon system, carbon dioxide is captured after the burning of fuel (e.g. biomass, natural gas, oil or coal) here in this process the most challenging stuff is to separate the CO_2 that is produced during the burning, here the solvent that is used as an absorbent is MEA. A detail description is shown below in the flow diagram, and the goal is extracting the 99% carbon dioxide from the exhaust gases and regeneration of the solvent MEA.

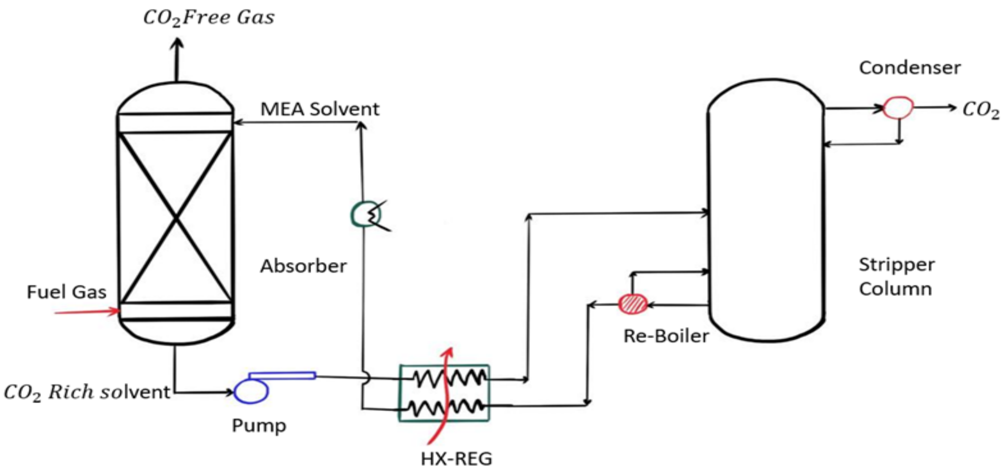


Figure 6post-combustion carbon system with MEA solvent

In process flow sheet fuel gases have temperature about 200 °C and pressure is 1.01 bar but before entering into the absorber column it's 40 °C and 1.1 bar while the pressure drop inside the column is 0.10 so before the exhaust gases enter into the absorber columns it should be compressed and lower in temperature so a compressor and a cooler is installed in-between.

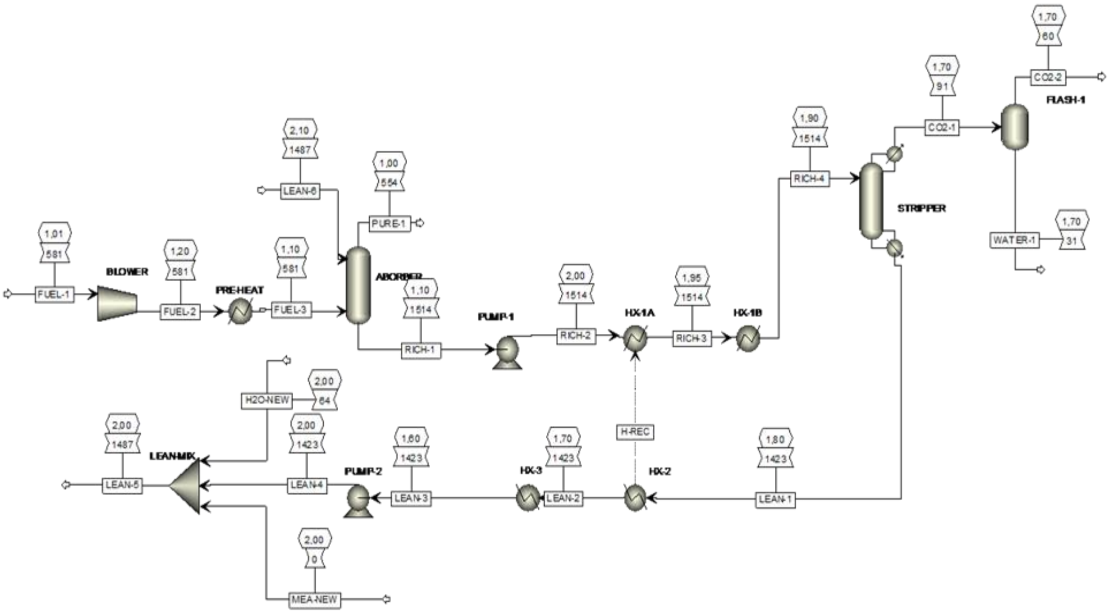


Figure 7 Process flowsheet of MEA absorption system on Aspenplus®

The absorber that is installed is vertically packed column has the specific height and diameter that is described in Table 3, the packing surface area must be enough to absorb the CO_2 from the exhaust fuel gases, these gases enter from the bottom of the absorber column while the solvent will entre from the top of the column in such a way the flow will counter one another while the vent gases leave from the top of the column. MEA, solvent contains 64% water mass concentration in loading with 0.1506 MEA in Lean-6 entre into the column while leaving with a rich concentration of carbon dioxide

Name	Starting stage	Endstage	Mode of operation	Internal Type	Diameter (m)	Packed height(m)	Operating pressure(bar)
Absorber	1	12	Rating	Packed	10.5	20	1

Table 3Absorber column parameter and specification.

Rich solvent leaves from the bottom of the absorber and sent to the pump-1 whose purpose is increasing the pressure up to 2.0 bar the operating pressure of stripper is 1.7 bar and the pressure drop within the column is 0.10 bar here heat recovery heat exchangers are installed between the rich solvent and the stripper, where the heat from the lean solvent that leaves from the bottom of stripper is given to the rich solvent that is coming from the absorber by the help of H-REC so the heat transfer between the rich solvent and lean solvent is about 40÷50°C. before entering the lean solvent into the absorber further cooldown is required to bring it about 40°C and an extra stream of Water+MEA is added with the help of mixer to accommodate a little loss of water and MEA.

The Rich solvent (Rich-3) taken heat by the cross heat exchanger sent to the stripper column which is packed vertical column that contain the reboiler and condenser the solvent entre into the 2nd stage of the stripper column and flows down the column, counter to the direction of the vapor from the reboiler the parameter and specification of stripper is shown in table 4 and table 5

Name	Starting stage	End-stage	Mode of operation	Internal Type	Diameter (m)	Packed height(m)	Material
Stripper	2	11	Rating	Packed	8.0	14	SS

Table 4 Geomaterial parameter of the stripper column

Name	**Operating pressure**	**P.D in Column**	**Mode of Operation**	**Internal Type**	**Reflux Ratio**	**Reboiler Duty (MW)**
Stripper	1.7	0.1	Rating	Packed	0.15	30

Table 5 Operating parameter of the stripper column

the stream from the top of the stripper sent to the flash column to separate the HO_2 and CO_2, a part liquid Reflux(pure water) sent to the stripper, while the pure CO_2 is sent to the compression and then the storage. The cooling water for the heat exchanger is extracting from the river that has a temperature of 15°C and after the heat transfer it is maximum 20÷25°C and the heat exchangers are "Shall & tube".

The extracted CO_2 further needs to be dried and compressed for that it's sent to flash tower where water is separated from the mixture of CO_2+water and pure CO_2 is further sent for the storage all this happens in the flash tower in Table 2.6 shows the process condition of the MEA solvent extracting model

Parameter	**Units**	**Values**
Fuel gas inlet temperature	$°C$	40
Fuel gas inlet pressure	bar	1.01
MEA, mass fraction	Wt.%	30
Absorber pressure	bar	1.1
Absorber column pressure drop	bar	0.1

Stripper pressure	bar	1.8
Stripper column pressure drop	bar	0.1
Reboiler duty	MW	50
Carbon capture	%	99
Number of stages	-	12

Table 6 Process conditions of the MEA solvent extracting model

2.3 Reaction chemistry

$$2H_2O \Leftrightarrow H_3O^+ + OH^- \qquad R_1$$

$$H_2O \Leftrightarrow HCO^-{}_3 + CO_3{}^{-2} \qquad R_2$$

$$H_2O \Leftrightarrow H_3O^+ + MEA \qquad R_3$$

While the rate-based reaction are as follows

$$CO_2 + OH^- \Leftrightarrow HCO^-{}_3 \qquad R_4$$

$$HCO^-{}_3 \Leftrightarrow CO_2 + OH^- \qquad R_5$$

$$MEA + CO_2 + H_2O \Leftrightarrow H_3O^+ + MEACOO^- \qquad R_6$$

$$H_3O^+ + MEACOO^- \Leftrightarrow MEA + CO_2 + H_2O \qquad R_7$$

The equilibrium constant for these reactions is calculated by the following equation where T is the temperature in K. The empirical constants A, B, C, and D [9]

$$\ln(K_{eq}) = A + \frac{B}{T} + C.\ln(T) + D.T$$

Reaction	A	B	C	D
R_1	132.889	-13445.9	-22.4773	0
R_2	216.05	-12431.7	-354819	0
R_3	-3.03833	-7008.36	0	-0.0031349

Reaction	K	E
R_4	4.32E+13	5.55E+07
R_5	2.38E+17	1.23E+08
R_6	9.77E+10	4.13E+07
R_7	3.23E+19	6.55E+07

2.4 Process simulation system

MEA, absorption Process was carried out Aspenplus®version10.1.0 using the ELECNRTL & thermodynamic model vapor & liquid and the process flow sheet is simulated as an open-loop system because it's easier for convergence and allows multiple runs quickly. The calculation in aspen plus has been carried out as apparent components which means that the results are reported as neutral components (MEA and CO_2), not any other form. The apparent component has been defined in the property set with names CO2WXAPP,$CO2XAPP$, $H2OWXAPP$, $H2OXAPP$, MEAWXAPP, MEAXAPP which calculates the molar flow rate of the CO_2, water and MEA, while the weighted concentration of MEA in water can be calculated by MEAXAPP and H2OXAPP.

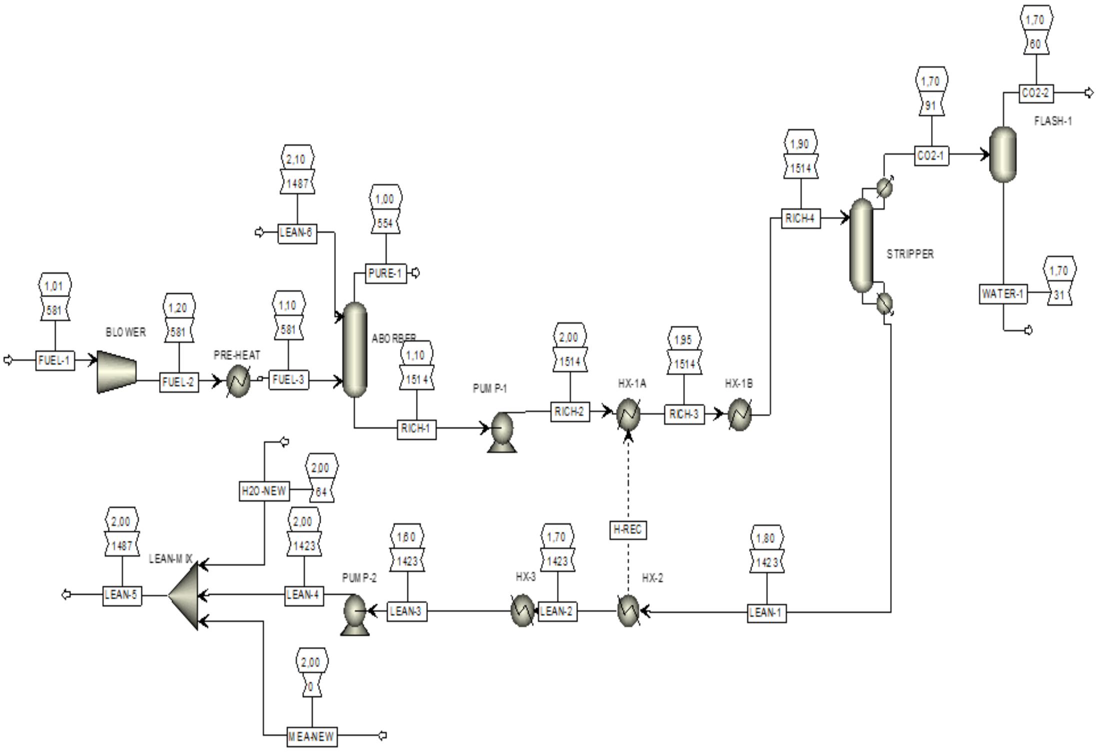


Figure 8Model MEA adsorption flow sheet build by using the aspen plus software

The fuel gases entre into the blower/compressor whose anisotropic and mechanical efficiency 0.80 and 0.90 respectively, both columns absorber and stripper are RADFRAC but both are different because the absorber column does not contain the reboiler and condenser while the stripper modeled as partial-vapor condenser and kettle reboiler for calculation rate-based methodology was adopted and the number of stages is equal in each column. Fuel gases cooler is (PRE-HEATER) and solvent cooler HX-3 both are modeled as a heater without losses while the rest of other heaters are shell &tube, heat exchanger. The dehydrator is modeled as a flash column and pumps have mechanical efficiency is 90.

The required design specification is described below that is imposed in the process flow sheet:

- The require CO_2 capture rate is achieved by design specification which is imposed to regulate the amount of CO2 that is extracted from the top of the absorber column by just changing the flow of MEA into the absorber by the help of stream LEAN-6.

- In the stripper column, reboiler duty depends upon the composition of the stream LEAN-1 which is the output of the stripper column and it is imposed to get the desired value of the amine and then sent into the absorber column.
- The temperature difference between the input and output in the heat exchanger depends upon the working fluids and the water that is used for this purpose will be extracted to the river so the maximum rise of the temp of water should not be larger than $25°C$.
- In LEAN-MIX block a stream of MEA-NEW is also added to minimize the losses of the solvent by just adding a proper portion of the water and it’s accrued by varying the amount of water in MAKE-UP.
- The final task is to get the 99% pure CO2 and regeneration of the MEA.

2.5 Equilibrium-based versus rate-based

An equilibrium model was developed using the RadFrac equilibrium model in aspen plus with a theoretical number of stages, contrary to the rate-based model and in these stages liquid and water was assumed to be at equilibrium due to infinitely fast mass transportation while on the other hand in rate based the actual mass transportation is considered in defined number of stage. In rate-based reactive columns gives an accurate process condition as compared to the equilibrium model so that’s why the rate-based approach is more suitable as compared to the equilibrium-based model.

Reaction in Equilibrium model:

$$2\ H2O \Leftrightarrow OH- + H3O+$$

$$H2O + HCO3- \Leftrightarrow CO3-- + H3O+$$

$$2\ H2O + CO2 \Leftrightarrow HCO3- + H3O+$$

$$\mathrm{H2O + MEACOO\text{-}} \Leftrightarrow \mathrm{MEA + HCO3\text{-}}$$

$$\mathrm{H2O + MEA+} \Leftrightarrow \mathrm{MEA + H3O+}$$

Reaction in Rate-based model:

$$MEA^{+} + \mathrm{H2O} \Leftrightarrow \mathrm{MEA + H3O+}$$

$$\mathrm{MEACOO\text{-} + H2O} \Leftrightarrow \mathrm{MEA + HCO3\text{-}}$$

$$\mathrm{2\ H2O} \Leftrightarrow \mathrm{OH\text{-} + H3O+}$$

$$\mathrm{CO2 + 2\ H2O} \Leftrightarrow \mathrm{HCO3\text{-} + H3O+}$$

$$\mathrm{HCO3\text{-} + H2O} \Leftrightarrow \mathrm{CO3\text{--} + H3O+}$$

$$\mathrm{MEA + CO2 + H2O} \Leftrightarrow \mathrm{MEACOO\text{-} + H3O+}$$

$$\mathrm{MEACOO\text{-} + H3O+} \Leftrightarrow \mathrm{MEA + CO2 + H2O}$$

$$\mathrm{CO2 + OH\text{-}} \Leftrightarrow \mathrm{HCO3\text{-}}$$

$$\mathrm{HCO3\text{-}} \Leftrightarrow \mathrm{CO2 + OH\text{-}}$$

In rate based chemical reaction that is taking part to get the actual heat and mass transfer[9]

2.6 Packing of columns

In the rate-based model, the geometry of the column is defined that is appropriate according to the number of stages, these packing parameters[9] are as below in the table2.7 and according to column sized using generalized pressure drop correlation. For the absorber, column diameter was 10.5m and the height was 20m and the stripper diameter was 9.5m and the height was 17m, and the number of stages is 12.

Type	Vender	Material	Dimension
Absorber Packing IMPT	KOCH	Metal	0.625-IN (16-MW)
Stripper Packing FLEXI PACK	KOCH	Metal	1Y

Table 7packing parameters characteristic both absorber and stripper column

2.7 Stripper

The stripper process flow diagram is shown below figure 9 where the rich amine (RICH-4) enters into the tower and go down through the packing inside the tower, while due to re-boiling in the bottom of the tower a percentage of lean water evaporate into steam, this steam provided the heat to enhance the desorption process.

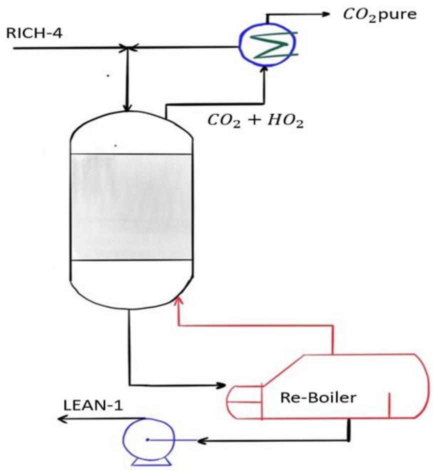


Figure 9 Flow within the column containing re-boiler and condenser

to the rich amine and increases the temperature reverse these chemical reactions and decrease the solubility of CO2 in liquid and increase in the steam this is what we called as a de-absorption process so, as liquid-rich amine and steam counter interact due to flow between the tower the concentration of CO2 increase in the stream.

A condenser that is installed on the top of the column which condensed this mixture of steam and MEA into the liquid the above de-absorption model is just like the absorption by just changing the temperature the purpose of the reboiler and condenser is to provide the initial inlet boundary condition to the stripper as the diameter of the stripper tower is lower than the absorber tower that leads volume flow rate is lower than the absorber as gases passes up through the tower due to this lower rate the condensed Water+MEA get diluted and increase the liquid flow rate into the tower. This de-absorption process is operating at 1.9 bar and when the temperature increases the rate of reaction get faster than the absorption, here the equilibrium constant act as a controlling factor which detects the amount of liquid CO2 from the bounded MEACOO^{-} and $HCO3^{-}$.

Reboiler:

The consumption of heat is important in this column because of the major portion of the heat is within the reboiler, this heat energy is divided into two parts.

1. Heat consumed to raise the temperature of the liquid from outlet temperature of stripper up to reboiler temperature.
2. Heat consumed to vaporizing the vapor flow

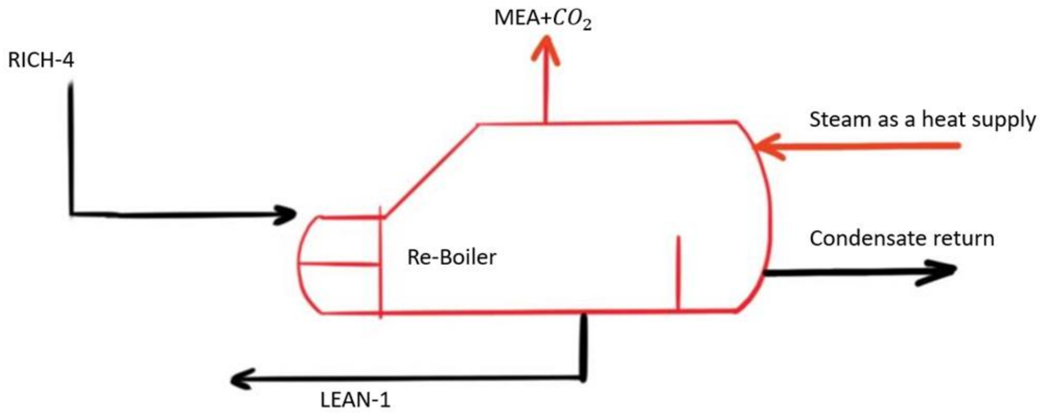


Figure 10 Re-boiler working inside the stripper column

Condenser:

The top of the stripper column and the condenser condense a percentage of MEA and water, as the output of the absorber is an inlet of the stripper and this condensed liquid mix with the rich MEA, this not actually the system is constructed to avoid this the inlet of the stripper is not exactly at the top of the tower but a few meters down the assumption that the flow gets to mix and provided the inlet stream from the top of the stripper is only because to make it easier to design.

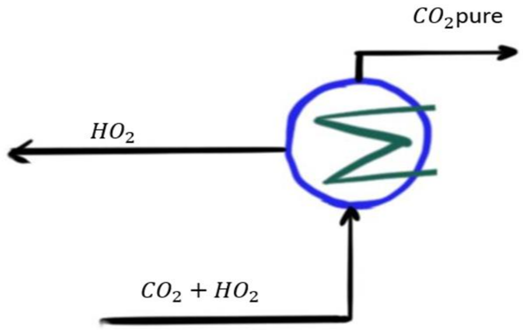


Figure 11 Fluid flow within the condenser inside the stripper column

2.8 Heat exchanger

There are four heat exchangers in the process with the names as

1. Pre-heater
2. HX-1B
3. HX-1A/HX-2
4. HX-3

The first heat exchanger transfer heat from hot stream FUEL-2 coming from the compressor with cold water before it gets entre into the absorption tower, to absorb the maximum CO2 from the exhaust at such a lower temperature, so how much temperature is differ can be calculated by logarithmic mean temperature difference(LMTD)method [10] and the flow between the liquid is counter-current

$$\Delta T_M = \frac{(T_{Hin} - t_{COUT}) - (T_{HOUT} - t_{Cin})}{ln\frac{(T_{Hin} - t_{COUT})}{(T_{HOUT} - t_{Cin})}}$$

T_{Hin}=FUEL-2

T_{HOUT}=FUEL-3

t_{Cin}=WATER

t_{COUT}=WATER

and how much heat is transfer[10] is also calculated by the following

$$Q = U * A * \Delta T_M$$

A=Area of heat transfer (m^2)

Q=Mean heat transfer rate(W)

U=Heat transfer coefficient ($^{W}/_{m^2\text{K}}$)

ΔT_M= Logarithmic mean temperature difference (LMTD)

HX-1B:

In HX-1B the hot fluid is superheated steam and the cold at the inlet is RICH-3 and at the outlet is RICH-4, because the stream that is going inside the stripper column needs to be at a higher temperature that can attain, to minimize the heat penalty.

T_{Hin}=Steam at 150°C

T_{HOUT}=Steam at 130°C

t_{Cin}=RICH-3

t_{COUT}=RICH-4

HX-2/HX-1B:

These heat exchangers are working as a heat recovering from the outlet stream of the stripper and sent it to the HX-1A supplies to the inlet of the stripper to minimize the thermal penalty.

T_{Hin}=LEAN-1

T_{HOUT}=LEAN-2

t_{Cin}=WATER

t_{COUT}=WATER

HX-3:

In this heat exchanger, the hot fluid is LEAN-2 AND LEAN-3 while the clod fluid is the water.

T_{Hin}=LEAN-2

T_{HOUT}=LEAN-3

t_{Cin}=WATER

t_{COUT}=WATER

Chapter 3

Techno-economic analysis

This chapter explains Cost analysis considers all types of costs that will participate in capital cost/investment that is required for the construction and operation of the plant generally, these are subdivided into two major types.

1.CAPEX

2.OPEX

CAPEX

CAPEX is the capital cost of the plant which is further sub-divided into different types of cost as shown in figure 3.1

BEC : Bare Erected Cost

EPCC : Engineering, Procurement, and construction cost

TPC : Total Plant Cost

TOP : Total Overnight Capital

TASC: Total As-spent capital

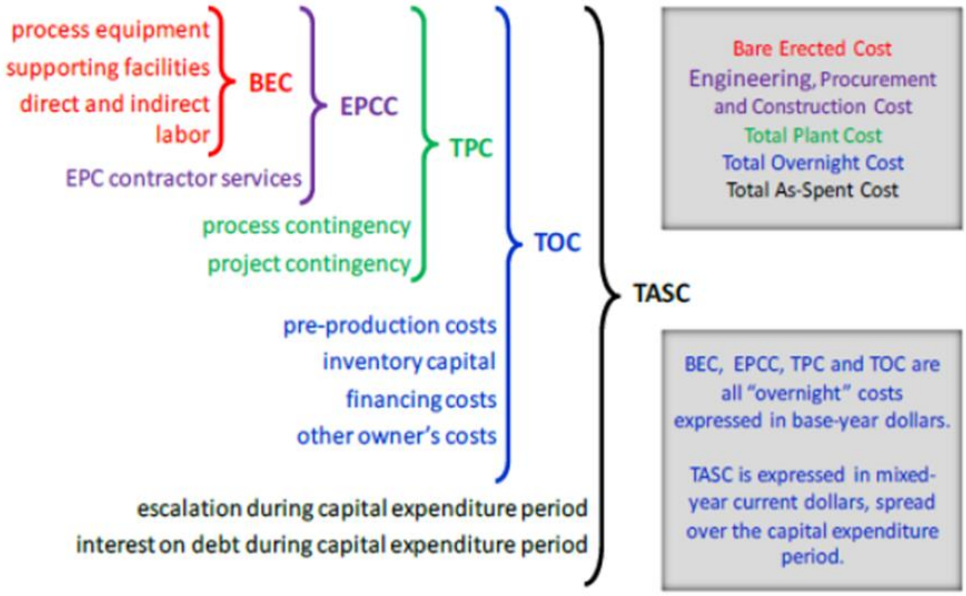


Figure 12Different types of cost that take part in CAPEX [11]

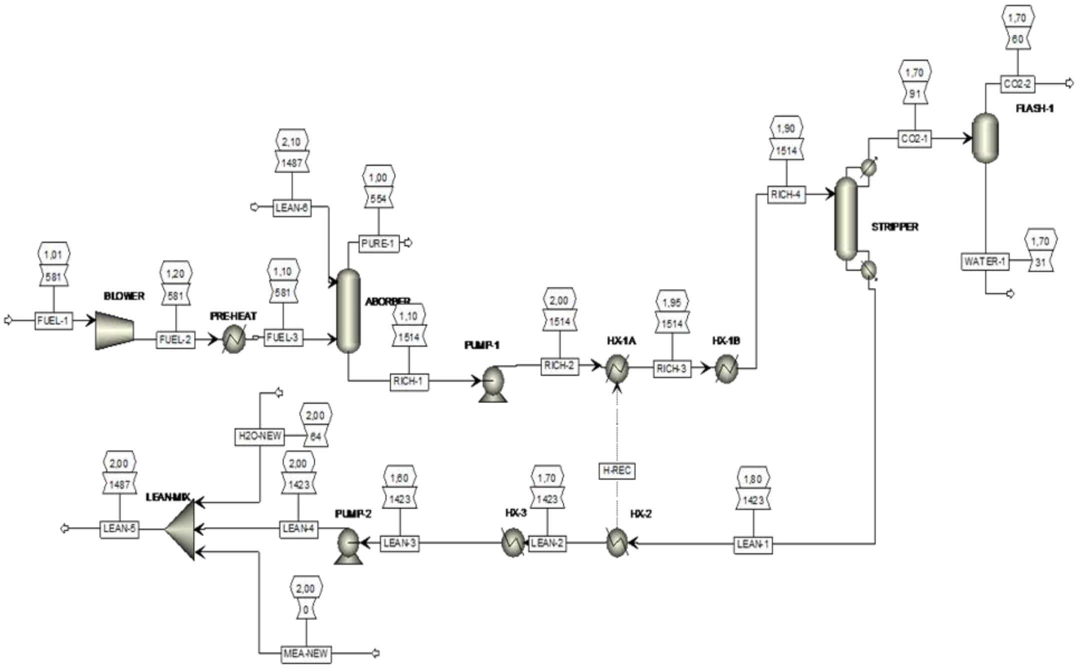


Figure 13Process flowsheet of the CCS plant

The table below contains all types of components that involve in the process and the cost must calculate to find the capital cost of the plant

Q.NO	**Component Detail**	**Quantity**
1	Blower/Compressor	01
2	Heat Exchanger	04
3	Pumps	02
4	Columns	03

Figure 14 Different types of components from the process flow sheet

The cost of each component varies concerning the thermodynamic conditions i.e. temperature, pressure, flow rate, etc. and material of component even the same type of component may contain the different cost due to different conditions. [1]

$$\boldsymbol{C_{BEC=}C_P^0 . F_{BEC}} \qquad \textbf{3.1}$$

$$\boldsymbol{log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3 . [log_{10}A]^2} \qquad \textbf{3.2}$$

Equation, 3.1 and 3.2 are used to calculate the cost of the component that involves different other parameters. Where "A" is an unknown parameter which changes concerning the specific component while K_1, $K_2$and$K_3$are the constants that refer to a specific type of component. "C_P^0" is the base cost of the component that is normally provided by the manufacturer and F_{BEC} is the factor that includes the pressure and material factor

3.1 Blower/Compressor

To calculate the BEC of the blower by using standard equations 3.1 and 3.2.

$$C_{BEC}=C_P^0 . F_{BEC}$$

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3 . [log_{10}A]^2$$

$$F_{BEC} = F_M . F_P$$

Specification:

Component: Compressor.

Type: Centrifugal axial and reciprocating.

Base Parameter(A): Power

Range of validity: (450-3000) KW.

The following table is the values of the constant that is taken from Turotan, and by choosing the appropriate type of the compressor

Type	**Description**	k_1	k_2	k_3	**Capacity/Units**	**Min**	**Max**
Compressor	centrifugal	2.2897	1.3604	0.1027	KW	450	3000

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3 . [log_{10}A]^2$$

$$C_P^0 = 10^{K_1 + K_2 log_{10}(A) + K_3 . [log_{10}A]^2}$$

$$C_P^0 = 881985.5\$$$

This cost is against maximum power which is 3MW but here, it is required 5.386MW almost double so to get the exact value, apply the six-tenths rule [12]

$$\frac{C_1}{C_0} = \left(\frac{S_1}{S_0}\right)^n$$

Equipment Type	**Range of correlation**	**Units of Capacity**	**Cost exponent "n"**
Reciprocating Compressor with motor drive	0.75 to1490	kW	0.84
Heat exchanger shell and tube carbon steel	1.9 to1860	m^2	0.59
Vertical tank carbon steel	0.4 to76	m^3	0.30
Centrifugal blower	0.24-71	$m^3/_S$	0.6
Jacket kettle glass lined	0.2 to3.8	m^3	0.48

Table 8 Different values for the coefficient of the exponent for component

$$C_P^0 = C_1 = 1252992.5\$$$

The working pressure of the compressor is 1bar and the temperature is about $200°C$, So that the material that is selected for this compressor is Carbon steel.

Material	$\boldsymbol{F_M}$	**Operating Temperature°*C***	**Note**
CS	3.8	<350	-
SS	8	<1100	-
Ni Alloy	15.8	<1300	Special applications

So, the material factor is 3.8.

$$F_M = 3.8$$

$$F_p = (10)^{C_1 + C_2 logP + C_3 (log_{10} P)^2}$$

From Turton, C_1, C_2 and C_3 are equal to zero so the pressure factor is 1.

$$C_{BEC=} C_P^0 . F_M . F_P$$

$$C_{BEC=} 1252992.5 * 3.8 * 1$$

$$\boldsymbol{C_{BEC,compressor}} = \mathbf{476k\$}$$

3.2 Pumps

The BEC of the pump will calculate by the same methodology as in the case of the compressor but, variable "A" will be different than the compressor.

Specification:

Component: Pumps

Type: Reciprocating

Base Parameter(A): Shaft Power $[KW]$

Range of Validity: 0.1-200$[KW]$

So, in this case, the standard equation for BEC of the component will be modified as

$$C_{BEC=}C_P^0(B_1+B_2 * F_m * F_p)$$

Here $B_1 and B_2$ are the "bare module factor" of the component which also depends upon the type of component so in our case it is as follows while the reset of other factors is the same as it was in the case of the compressor.

Equipment	**Type**	$\boldsymbol{B_1}$	$\boldsymbol{B_2}$
Pumps	Reciprocating	1.89	1.35
	Positive displacement	1.89	1.35
	Centrifugal	1.89	1.35

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3.[log_{10}A]^2$$

$$F_p = (10)^{C_1 + C_2 logP + C_3(log_{10}P)^2}$$

In the case of the pump, the base cost for Pump is calculated by using the same standard equation 3.1 and 3.2 while the values of constants can be found from the following table. 9 and 10 both for the base cost and the pressure factor.

Pump1:

Now to calculate the BEC of "pump1" first to know about the required power which is, in this case, is 54 kW and C_P^0 is calculated by using standard equation 3.1

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3.[log_{10}A]^2$$

Component	Type	K_1	K_2	K_3	Parameter	Units	Mini	Max
Pumps	Reciprocating	3,8696	0,3161	0.122	Shaft Power	KW	0.1	200
	Positive displacement	3..4771	0.135	0.1438	Shaft Power	KW	1	100
	Centrifugal	3.3892	0.0536	0.1538	Shaft Power	KW	1	300

Table 9Types of different pumps that are available with standard capacity

$$C_P^0 = 60715.5\$$$

Pressure Factor:

For the calculation of the pressure factor, we will use the following table according to our model pumps and operating parameters.

$$F_p = (10)^{C_1 + C_2 logP + C_3 (log_{10}P)^2}$$

Component	**Type**	C_1	C_2	C_3	**Pressure bar**
Pumps	Reciprocating	0	0	0	P<10
		-0.245382	0.259016	-0.01363	P<10 P<100
	Positive displacement	0	0	0	P<10
		-0.245382	0.29016	-0.01363	P<10 P<100
	Centrifugal	0	0	0	P<10
		-0.3935	0.3957	-0.00226	P<10 P<100

Table 10 Pressure factor against standard type and capacity

As it is mention in the above table according to the type and operating pressure the value of constants C_1, C_2 and C_3 is equal to zero because the operating pressure is lower than P<10

So,

$$F_p = 1$$

Material Factor:

The material factor also depends upon the equipment, its type, and the operating conditions. Here, it is CS and its value of F_M=3.8.

$$C_{BEC} = C_P^0 * (B_1 + B_2 F_m . F_p)$$

Equipment	**Type**	B_1	B_2
	Reciprocating	1.89	1.35

Pumps	Positive displacement	1.89	1.35
	Centrifugal	1.89	1.35

$$C_{BM} = 60715.5 * (1.89 + 1.35 * 3.8 * 1)$$

$$\boldsymbol{C_{BECP1} = 426K\$}$$

Pump2:

In this case of pump2 according to model, the required capacity is 21KW so against this value the base cost of the pump2 will be calculated by using the standard equation 3.1.

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3.[log_{10}A]^2$$

Component	**Type**	$\boldsymbol{K_1}$	$\boldsymbol{K_2}$	$\boldsymbol{K_3}$	**Parameter**	**Units**	**Mini**	**Max**
Pumps	Reciprocating	3,8696	0,3161	0.1220	Shaft Power	KW	0.1	200
	Positive displacement	3..4771	0.135	0.1438	Shaft Power	KW	1	100
	Centrifugal	3.3892	0.0536	0.1538	Shaft Power	KW	1	300

$$C_P^0 = 31673.7\$$$

And the pressure factor, in this case, will be calculated again as it's done before in case of pump1 which is same because the operating pressure is P<10 while the value of a material factor is F_M=3.8 because the material is also the same for both pumps.

$$F_p = 1, F_M = 3.8$$

$$C_{BEC2} = C_P^0 . (B_1 + B_2 F_m . F_p)$$

$$C_{BEC2} = 31673.7 * (1.89 + 1.35 * 3.8 * 1)$$

$$\boldsymbol{C_{BECP2} = 222K\$}$$

3.3 Heat Exchangers

In this model, four different heat exchange is used that have the following names.

- Pre-heat heat exchanger
- HX-1B heat exchanger
- HX-2 heat exchanger
- HX-3 heat exchanger

Specification:

Component: Heat Exchanger

Type: Shall & Tube

Base Parameter(A): Area (m^2)

Range of Validity: 10-1000[m^2]

Pre-Heater/Cooler:

To calculate the base cost of the heat exchanger same standard equations 3.1and 3.2 are utilized but the unknown parameter "A" is different in each heat exchanger.

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3.[log_{10}A]^2$$

$$C_P^0 = 10^{K_1 + K_2 log_{10}(A) + K_3.[log_{10}A]^2}$$

To calculate the unknown parameter area "A" of heat transfer, the following relationship is used

$$Q = U * A * \Delta T_M$$

A=Area of heat transfer (m^2)

Q=Mean heat transfer rate(W)

U=Heat transfer coefficient ($W/_{m^2 \mathrm{K}}$)

ΔT_M=Logrithmic mean temperature difference (LMTD)

ΔT_M requires the temperature difference between the hot fluid and cold fluid then applies the LMTD for Counter-current flow [10]

$$\Delta T_M = \frac{(T_{Hin} - t_{COUT}) - (T_{HOUT} - t_{Cin})}{ln\frac{(T_{Hin} - t_{COUT})}{(T_{HOUT} - t_{Cin})}}$$

T_{Hin}=Hot fluid gas inlet temperature

T_{HOUT}=Hot fluid gas outlet temperature

t_{Cin}=Cold fluid inlet temperature

t_{COUT}= Cold fluid outlet temperature

Fluid	Inlet Temp. $°C$	Outlet Temp. $°C$
Hot Gas	228	40
Water	15	20

$$\Delta T_M = 86.38k$$

"U" is the heat transfer coefficient which depends upon the type of fluids that are interacting normally it is, Liquid-to-liquid, Gas-to-liquid, and Gas-to-gas heat transfer while in this heat exchanger it is Gas-to-liquid and its range is (10-250) $W/_{m^2K}$,

U=130$W/_{m^2}$K

$$\Delta T_M = 86.38k$$

$$Q = 32.116MW$$

$$Q = U * A * \Delta T_M$$

So, the unknown, A=2859.9m^2, now applies the equation 3.1

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3.[log_{10}A]^2$$

Type	K_1	K_2	K_3	**Area**	**Mini**	**Maxi**
U-Tube	4.1884	-0.2503	0.1974	m^2	10	1000

$$C_P^0 = 47302.7\$$$

This cost is against an area of $1000m^2$ so to get the exact value use following expression

$$C_1 = C_0 * \left(\frac{S_1}{S_0}\right)^{\boldsymbol{n}}$$

The value of “n” is chosen from the table below related to the type of equipment

Equipment Type	**Range of correlation**	**Units of Capacity**	**Cost exponent “n”**
Reciprocating Compressor with motor drive	0.75 to1490	kW	0.84
Heat exchanger shell and tube carbon steel	1.9 to1860	m^2	0.59
Vertical tank carbon steel	0.4 to76	m^3	0.30
Centrifugal blower	0.24-71	$m^3/_s$	0.6
Jacket kettle glass lined	0.2 to3.8	m^3	0.48

Table 11Different values for the coefficient of exponent w.r.t. component

$$C_P^0 = C_1 = 87929.16\$$$

$$C_{BEC} = C_P^0.(B_1 + B_2 * F_M * F_P)$$

Equipment	**Type**	$\boldsymbol{B_1}$	$\boldsymbol{B_2}$
Heat exchangers	Shall & Tube	1.63	1.66

Pressure and Material Factor:

From Turton, if the pressure is below 5bar the constants C1, C2, and C3 are equal to zero in the case of Heat Exchanger

$$F_p = (10)^{C_1 + C_2 logP + C_3 * (log_{10}P)^2}$$

$$F_P = 1$$

$$F_M = 3.8$$

$$B_1 = 1.63 \quad B_2 = 1.66$$

$$C_{BEC} = 87929.16 * (1.63 + 1.66 * 3.8 * 1)$$

$$\boldsymbol{C_{BEC(pre-heater)} = 697K\$}$$

HX-1B:

Following is the table that contains information about the temperature of the fluids and the heat duty is equal to Q=27.440Mw and for steam to liquid overall heat transfer coefficient is U=2200-35000 $W/_{m^2K}$.

Fluid	**Inlet Temp. °C**	**Outlet Temp. °C**
MEA+CO_2	97	107
Steam	150	130

$$\Delta T_M = 37.73\text{k}$$

And the heat transfer area is

$$Q = U * A * \Delta T_M$$

$$A = 330.57m^2$$

Type	$\boldsymbol{K_1}$	$\boldsymbol{K_2}$	$\boldsymbol{K_3}$	**Area**	**Mini**	**Maxi**
U-Tube	4.1884	-0.2503	0.1974	m^2	10	1000

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3.[log_{10}A]^2$$

$$C_P^0 = 64253.9\$$$

$$F_P=1$$

$$F_M = 3.8$$

$$C_{BEC} = C_P^0.(B_1+B_2 * F_M * F_P)$$

$$C_{BEC} = 64253.9(1.63 + 1.66 * 3.8 * 1)$$

$$\boldsymbol{C_{BEC(HX-1B)}=510K\$}$$

HX-2:

Following is the table that contains information about the temperature of both fluids and the heat duty is equal to Q=79.192Mw and the overall heat transfer coefficient is U=1000-2500 $W/_{m^2K}$ in the case of ammonia to liquid fluid heat transfer.

Fluid	**Inlet Temp.** $\boldsymbol{°C}$	**Outlet Temp.** $\boldsymbol{°C}$
MEA+CO_2	118	59
RICH2-RICH3	44	97

Apply the LMTD of a Counter-current flow, to get ΔT_M the cooling fluid is water and the overall heat transfer coefficient U=1000-2500.

$$\Delta T_M = \frac{(T_{Hin} - t_{COUT}) - (T_{HOUT} - t_{Cin})}{ln\frac{(T_{Hin} - t_{COUT})}{(T_{HOUT} - t_{Cin})}}$$

$$\Delta T_M = 17.83\text{K}$$

And the area of heat transfer is,

$$A = 4441.5m^2$$

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3.[log_{10}A]^2$$

$$C_P^0 = 796526.0\$$$

This cost is against an area of $1000m^2$ so to get exact value according to the required area by using the following expression

$$C_1 = C_0 * \left(\frac{S_1}{S_0}\right)^{\boldsymbol{n}}$$

$$C_P^0 = C_1 = 3537770.2\$$$

$$F_P\text{=}1$$

$$F_M = 1.8$$

$$C_{BEC} = C_P^0.(B_1 + B_2 * F_M * F_P)$$

$$C_{BEC} = 3537770.2 * (1.63 + 1.66 * 3.8 * 1)$$

$$\boldsymbol{C_{BEC(HX-2)} = 28,082K}\$$$

HX-3:

Following is the table that contains information about the temperature of both fluid at the inlet and the outlet while the heat duty is equal to Q=24.712Mw and U =1000-2500 $W/_{m^2K}$.

Fluid	**Inlet Temp. °C**	**Outlet Temp. °C**
MEA+CO_2	59	40
Water	15	20

just apply the LMTD of a Counter-current flow, to get ΔT_M and in above MEA+CO_2 is the hot fluid while water is the clod fluid.

$$\Delta T_M = \frac{(T_{Hin} - t_{COUT}) - (T_{HOUT} - t_{Cin})}{ln\frac{(T_{Hin} - t_{COUT})}{(T_{HOUT} - t_{Cin})}}$$

$$\Delta T_M = 31.48\text{K}$$

And the area of heat transfer is

$$A = 785m^2$$

$$C_P^0 = 129987.0\$$$

$$F_P\text{=1}$$

$$F_M = 1.8$$

$$C_{BEC} = C_P^0.(B_1{+}B_2)F_M * F_P$$

$$C_{BEC} = 129987.0(1.63 + 1.66 * 3.8 * 1)$$

$$\boldsymbol{C_{BEC(HX-3)} = 1031.8K\$}$$

The total BEC of the heat exchanger is the sum of the individual heat exchanger as per plant requirement.

$$BEC_{total} = C_{BEC(PRE-HEATER)} + C_{BEC(HX-1B)} + C_{BEC(HX-2)} + C_{BEC(HX-3)}$$

$$\boldsymbol{BEC_{T,HX} = 30,322K\$}$$

3.4 Columns

In flowsheet, the three most important columns are absorber, stripper/regenerator, and separator columns to clean the gas and regenerate absorbent fluid MEA.

There are different types of the absorber available in the standard form

- Absorber
- Refluxed absorber
- Reboiler absorber

While for the Stripper we have a distillation column, and for separator use simple flash block, the cost of each column will be different due to height, diameter, thickness and the material for all the columns is SS.

3.5 Absorber

BEC of the column is directly related to the geometry and the operating condition for every single column.

Name	Starting stage	End stage	Mode of operation	Internal Type	Diameter (m)	Packed height(m)	Operating pressure (bar)
Absorber	1	12	Rating	Packed	10.5	20	1

To calculate C_P^0 the same standard equation, 3.1 is used.

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + K_3.[log_{10}A]^2$$

The parameter "A" is calculated by choosing the appropriate type of component in the list of general component datasheet so in this model, the absorber is chosen as a process vessel and the value of constants is defined according to its type, while "A" is volume parameter.

Inlet Stream= Fuel-3 and lean-6

Outlet Stream=Pure-1 and rich-1

Take an assumption the residence time which will be T=0.5Sec because it directly related to the size of the column and also the reaction within the column the volume flow rate changes due to the change of fluid and thermodynamically properties of the fluid are also changes, so the value of "A" for the C_P^0 will be the mean value of the volume flow rate.

$$G_{Vi} = G_{V,Fuel-3} + G_{V,Lean-6}$$

$$G_{Vi} = 132.547871m^3$$

$$G_{VO} = G_{V,Pure-1} + G_{V,Rich-1}$$

$$G_{VO} = 156.9458m^3$$

$$G_{V,m} = \frac{G_{Vi}+G_{VO}}{2}=144.75m^3$$

So, A=144.75 is used to calculate the base cost for the absorber column and the BEC will be calculated by using the same standard equation 3.1

$$C_{BEC} = C_P^0 * F_M * F_P$$

3.6 Pressure Factor in Vessel

Here the F_P is the pressure factor and it directly affects the cost, the cost of equipment will increase if the pressure increases so the pressure factor is calculated by using the following formula this is the standard formula according to ASME[1] for the calculation of pressure factor in the process vessel

$$F_P = \frac{\frac{(P+1)D}{2[850-0.6(P+1)]}+0.00315}{0.0063}$$

In the above relation, P=1which is operating pressure and D=10.5 is the diameter of the pressure Vessel and

The values of these parameters are found from the above table as mention according to our plant.

$$F_P = 1.996 \approx 2$$

Material factor:

The material factor depends upon the type of equipment and the operating pressure so according to these parameters the value is as below

Material	F_M	**Operating Temperature°C**	**Note**
CS	1	<350	-

SS	3.1	<1100	-
Ni Alloy	7.1	<1300	Special applications

The value of F_M=3.1 because of the chosen material is SS even though the operating temperature is lower but only because of the corrosive property of MEA, while SS has more resistance against it.

$$C_{BEC} = C_P^0 * (B_1 + B_2 * F_M * F_P)$$

$$C_P^0 = 10^{3.4974+0.4485\log(144.75)+0.1074[\log(144.75)]^2}$$

$$C_P^0 = 92747.0\$$$

and, the BEC of the absorber column is calculated as

$$C_{BEC} = C_P^0 * (B_1 + B_2 * F_M * F_P)$$

$$C_{BEC} = 92747.02 * (2.25 + 1.82 * 3.1 * 2.0)$$

$$\boldsymbol{C_{BEC,absorber} = 1255K\$}$$

3.7 Stripper

The purpose of stripper column is to regeneration of the absorbent fluid which is, “MEA” because it’s directly related to the operating cost, while on the other hand, the size of stripper column is also important because here the most important thing is to reduce the operating cost as lower as possible by getting the higher recovery of the MEA.

So as in the case of the absorber column, the BEC of Stripper is calculated first by knowing the base cost of the Stripper by using the standard equation **3.1**and **3.2.**

$$C_{BEC=}C_P^0.F_{BM}$$

$$log_{10}C_P^0 = K_1 + K_2log_{10}(A) + K_3.[log_{10}A]^2$$

Geometrical parameter and thermodynamic properties are as follows

Name	Starting stage	End-stage	Mode of operation	Internal Type	Diameter (m)	Packed height(m)	Material
Stripper	2	11	Rating	Packed	8.0	14	SS

Name	Operating pressure	P.D in Column	Mode of Operation	Internal Type	Reflux Ratio	Reboiler Duty (MW)
Stripper	1.7	0.1	Rating	Packed	0.15	30

We can calculate the “A” by the same way as it has been calculated in the case of an absorber.

Inlet stream= Rich-4

Outlet stream=$CO_2 - 1$ and Lean-1

$$G_{Vi} = G_{V,Rich-4}$$

$$G_{Vi} = 6{,}54 m^3$$

$$G_{VO} = G_{V,CO_2-1} + G_{V,Lean-1}$$

$$G_{VO} = 15.78 m^3$$

$$G_{V,m} = \frac{G_{Vi}+G_{VO}}{2} = 11.16 m^3$$

So, by using the value A=$11.16 m^3$ in base cost equation, we can calculate the value of C_P^0 which is

$$log_{10} C_P^0 = K_1 + K_2 log_{10}(A) + K_3 . [log_{10} A]^2$$

$$C_P^0 = 12161.8$$

$$C_{BEC} = C_P^0 * (B_1 + B_2 * F_M * F_P)$$

Pressure Factor:

Here the F_P is the pressure factor and it is directly related to the cost as the cost of equipment is increasing as the pressure increases so the pressure factor is calculated by using the following formula

$$F_P = \frac{\frac{(P+1)D}{2[850 - 0.6(P+1)]} + 0.00315}{0.0063}$$

D=8.0 is the diameter of the vessel and P is operating pressure, so the pressure factor is

$$F_P = 2.52$$

Material Factor:

The value of a material factor is the same as it was in absorber because all the parameters are the same, so it is $F_M = 3.1$

$$C_{BEC} = 12163.82 * (2.25 + 1.82 * 3.1 * 2.52)$$

$$\boldsymbol{C_{BEC,stripper} = 200k\$}$$

3.8 Flash Column

The BEC of the flash column is calculated by using the standard equation, 3.1 before this it's necessary to calculate the base cost of the component by the following equation

$$log_{10}C_P^0 = K_1 + K_2 log_{10}(A) + [K_3 log_{10}A]^2$$

$$C_P^0 = K_1 + K_2 log_{10}(A) + K_3[log_{10}A]^2$$

Here "A" is the mean volume flow rate of the fluid which we know from the process flow diagram

Inlet stream= $CO_2 - 1$

Outlet stream=$CO_2 - 2$ and Water-1

$$G_{V,m} = \frac{G_{Vi}+G_{VO}}{2}=10.01m^3$$

So, the value “A=10.01” for the case of the flash column by using this value we can calculate the value of base cost.

$$C_P^0 = 11308.3\$$$

Pressure Factor:

In flash, take some assumption that the diameter of the column on the bases of the volume flow rate that is 6m and the relative pressure P=0 so the pressure factor is found as follow

$$F_P = \frac{\frac{(P+1)D}{2[850-0.6(P+1)]} + 0.00315}{0.0063}$$

$$F_P = 1.06$$

Material Factor:

. The value of a material factor is the same as it was in absorber and stripper

$$C_{BEC} = 11308.3 * (2.25 + 1.82 * 3.1 * 1.06)$$

$$\boldsymbol{C_{BEC_{Flash}} = 95k\$}$$

3.9 Total BEC of all Equipment

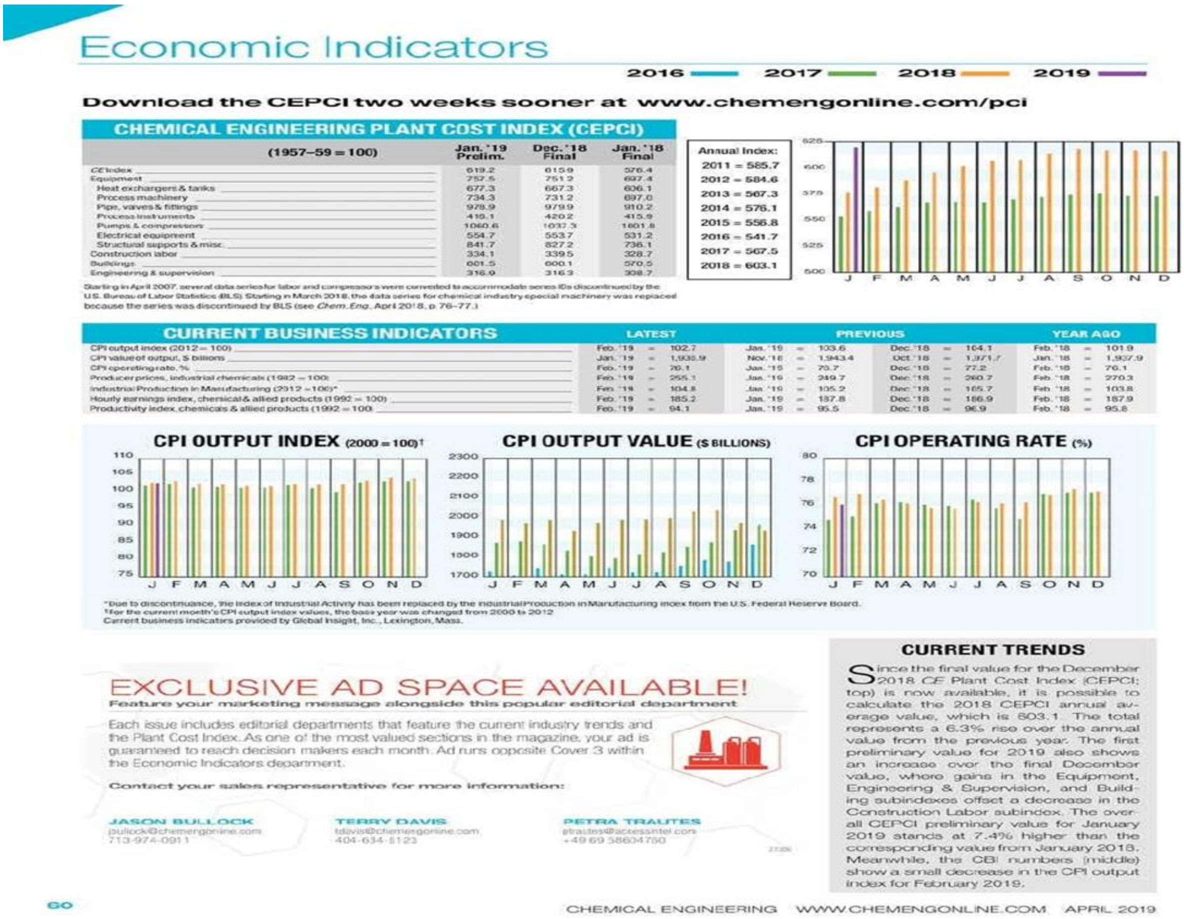
For the total BEC sum up all individual types of all components according to their size that is already define and calculated in previous passages.

Q.NO	Equipment	Type	Quantity	BEC ($)
1	Blower	Compressor	01	476111.5
2	Pump1	Pump	01	226222.8
3	Pump2	Pump	01	222349.3
4	Pre-heater	Heat Exchanger	01	697981.6
5	HX-1B	Heat Exchanger	01	510047.4
6	HX-2	Heat Exchanger	01	28082819.8
7	HX-3	Heat Exchanger	01	1031836.8
8	Absorber	Column/Vessel	01	1255238.1
9	Stripper	Column/Vessel	01	200311.5
10	Flash	Column/Vessel	01	95674.3

This total **BEC=31,869$\boldsymbol{k}$$** is in 2001 so to calculate the value in 2018 by CEPCI, currently, the available cost index from the CEPCI.

$$C_{1,2018} = C_0 * \left[\frac{I_1}{I_0}\right]$$

Here the subscript "1" shows the required year which is 2018 while "0" shows the base year which is 2001, in this case, the base year values "C" denotes the cost and "I" cost index for the respective year so the final value in the year 2018.



Economic Indicators

Download the CEPCI two weeks sooner at www.chemengonline.com/pci

CHEMICAL ENGINEERING PLANT COST INDEX (CEPCI)

(1957–59 = 100)	Jan. '19 Prelim.	Dec. '18 Final	Jan. '18 Final
CE Index	619.2	615.9	576.4
Equipment	757.5	751.2	697.4
Heat exchangers & tanks	677.3	667.3	606.1
Process machinery	734.3	731.2	697.0
Pipe, valves & fittings	978.9	979.9	910.2
Process instruments	419.1	420.2	415.9
Pumps & compressors	1060.6	1037.3	1001.8
Electrical equipment	554.7	553.7	531.2
Structural supports & misc.	841.7	827.2	736.1
Construction labor	334.1	339.5	328.7
Buildings	601.5	600.1	570.5
Engineering & supervision	316.9	316.3	308.7

Annual Index:
2011 = 585.7
2012 = 584.6
2013 = 567.3
2014 = 576.1
2015 = 556.8
2016 = 541.7
2017 = 567.5
2018 = 603.1

Starting in April 2007, several data series for labor and compressors were converted to accommodate series IDs discontinued by the U.S. Bureau of Labor Statistics (BLS). Starting in March 2018, the data series for chemical industry special machinery was replaced because the series was discontinued by BLS (see *Chem. Eng.*, April 2018, p. 76–77.)

CURRENT BUSINESS INDICATORS

	LATEST			PREVIOUS						YEAR AGO		
CPI output index (2012 = 100)	Feb. '19	=	102.7	Jan. '19	=	103.6	Dec. '18	=	104.1	Feb. '18	=	101.9
CPI value of output, $ billions	Jan. '19	=	1,935.9	Nov. '18	=	1,943.4	Oct. '18	=	1,971.7	Jan. '18	=	1,937.9
CPI operating rate, %	Feb. '19	=	70.1	Jan. '19	=	70.7	Dec. '18	=	77.2	Feb. '18	=	76.1
Producer prices, industrial chemicals (1982 = 100)	Feb. '19	=	255.1	Jan. '19	=	249.7	Dec. '18	=	260.7	Feb. '18	=	270.3
Industrial Production in Manufacturing (2012 = 100)*	Feb. '19	=	104.8	Jan. '19	=	105.2	Dec. '18	=	165.7	Feb. '18	=	103.8
Hourly earnings index, chemical & allied products (1992 = 100)	Feb. '19	=	185.2	Jan. '19	=	187.8	Dec. '18	=	186.9	Feb. '18	=	187.9
Productivity index, chemicals & allied products (1992 = 100)	Feb. '19	=	94.1	Jan. '19	=	95.5	Dec. '18	=	96.9	Feb. '18	=	95.8

*Due to discontinuance, the Index of Industrial Activity has been replaced by the Industrial Production in Manufacturing index from the U.S. Federal Reserve Board.
†For the current month's CPI output index values, the base year was changed from 2000 to 2012
Current business indicators provided by Global Insight, Inc., Lexington, Mass.

CURRENT TRENDS

Since the final value for the December 2018 *CE* Plant Cost Index (CEPCI; top) is now available, it is possible to calculate the 2018 CEPCI annual average value, which is 603.1. The total represents a 6.3% rise over the annual value from the previous year. The first preliminary value for 2019 also shows an increase over the final December value, where gains in the Equipment, Engineering & Supervision, and Building subindexes offset a decrease in the Construction Labor subindex. The overall CEPCI preliminary value for January 2019 stands at 7.4% higher than the corresponding value from January 2018. Meanwhile, the CBI numbers (middle) show a small decrease in the CPI output index for February 2019.

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Figure 15 cost index for the year 2018 by CEPCI [12]

$$I_1 = 603.1$$

$$I_0 = 394.3$$

$$\boldsymbol{C_{1,2018} = 48,746k\$}$$

Capital cost of plant:

The above mention cost is not the capital cost for this plant, by including some other costs the NETL report[13] to get the actual capital cost so, to calculate the capital cost of the plant it should require the following cost with specific percentages. EPCC is the sum of BEC+EPC which is further utilized to calculate process and project contingencies. Process contingencies are the sum of BEC, ECP, Process, contingencies.

TPC total plant cost comprises the EPCC plus process contingencies and project contingencies costs.

Parameter	Range from NETL report	Description	%age
Engineering, Procurement, construction cost (EPC)	8%	% of BEC	3899717.4
Process contingencies	5%	% of EPCC	2632309.2
Project contingencies	15%	% of EPCC	1151064.5
Owner cost	20.2%	% of TPC	1782615.2

The capital cost of this plant is **CAPEX= 58,212*k*$** for the year 2018.

OPEX

The operating cost(OPEX) is further divided into two categories, fixed operating cost and the variable operating cost [14]

TOP=FOC+VOC

Previous section the capital investment/total capital cost of the plant has been calculated, so the annual capital cost is calculated by the equation

$$ACC = \frac{TCC}{(1+r)^n - 1 \big/ r(1+0.1)^n}$$

In the above equation, the annual capital cost depends upon the two important parameters r and n where r is the interest rate and n are the life of the plant. The value of r and n is 10% and 30% respectively [15]

Variable operating costs (VOC)	
MEA, make-up US$/ton CO_2	1250
Steam cost US$/ton CO_2	14.5
Cooling water cost US$/ton CO_2	0.329
H_2O make-up US$/ton CO_2	0.5
Electrical cost US$/kWh CO_2	0.07
Fixed operating cost (FOC)	
Operator labor cost	3% of FCI
Maintenance	34.65 US$/h

*FCI is the fixed capital investment

While the total cost of the plant is calculated by the added CAPEX and OPEX, the main parameter in operating cost is the cost of the absorbent solvent which in this case MEA and operator wages.

Conclusion

For the removal of anthropogenic CO_2 selection of absorbent is a critical step in retrofitting current power plants with CO_2 capture process, it's not straight forward to choose the most promising one because of the multiple parameters that affect, the overall performance and the cost. Indeed, there is and will be no single ideal absorbent, so a trade-off between the cost, thermal plenty, and regeneration of solvent is taken into consideration.

The optimization of the plant is done here by changing the number of stages in columns (absorber and stripper) which leads the changing of the height and the diameter of these columns which leads to the minimize the energy penalty in stripper column for the regeneration of the absorbent solvent MEA which is directly related to the operating cost, then the techno-economic analysis is done for the construction of the plant in 2019 regarding that the cost of each component is calculated according to thermodynamic of that component and size, for this the cost is calculated by applying Turton [1] for 2001 then we calculated for the year 2019 by applying the six-tenths rule [12].

Items	Standard model		Optimized model	
	Absorber	Stripper	Absorber	Stripper
Number of stages	14	14	12	12
Pressure (bar)	1.1	1.7	1.1	1.7
Temperature changes with the column(°C)	12.42	32.31	11.44	20.31

Heat duty of condenser	0	0	0	0
Reboiler duty (Mw)	0	70	0	50
Dimension of column Absorber (m)	Height	Diameter	Height	Diameter
	20	10.5	20	9.5
Dimension of column Stripper (m)	Height	Diameter	Height	Diameter
	17	9.5	14	8.0
ΔQ_c	0	-13.83	0	-5.52
ΔQ_R (Mw)	0	55.53	0	50.17
Energy Consumption (Mw)	-1.53E-07/0.0088		-1.62E07/0.0090	
Reflux Ratio	2.93	0.5	2.945	0.15

The softening, storage of CO_2 and transportation to the site is not considered in this work. The cost of CO_2 extraction to the atmosphere is large in a developed country so the capital cost accommodates by not paying the cost of extraction of CO_2 to the atmosphere.

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