

TECHNICAL AND ECONOMIC CONSIDERATIONS OF POST - COMBUSTION CARBON CAPTURE IN A COAL FIRED POWER PLANT

Mohammed Isah Yakub^{1*}, Mohamed, Samah² and Sule Umar Danladi³

¹Department of Chemical Engineering,

Abubakar Tafawa Balewa University P.M.B 0248 Bauchi Nigeria

²Department of Chemical Engineering, the University of Khartoum, Sudan

³Department of Chemical Engineering, the Federal Polytechnic P.M.B 55 Bida, Nigeria

ABSTRACT

Techno-economic consideration of post combustion carbon capture in a coal fired power plant (500MW) was carried out. Primary (MEA), secondary (DEA) and tertiary (MDEA) amines were analysed regarding their tendency to corrode the equipments and their performance in the absorption and stripping columns using Aspen Hysys® 7.2 software. From the simulation work, DEA showed better performance. The DEA process was optimised to find the optimum parameters that explore the trade-off between efficiency and operating and capital costs at an acceptable operating point. The cost of commissioning and operating the capture plant was also estimated and different cost metrics were calculated for the plant feasibility. The levelized cost of electricity increased by £0.028/kWh compared to the original cost of £0.105/kWh. The cost of CO₂ treatment was found to be £26.8/ton which is higher compared to UK carbon floor price (2012) of £16/ton CO₂. Finally, the decision will be either to operate the power plant without a capture plant and pay the carbon tax or retrofitting the power plant for carbon capture and pay the cost of treating the CO₂. However, if the emphasis is on environmental consideration, the retrofitting becomes the best option.

KEYWORDS: Modelling; simulation; amine; post-combustion; carbon capture, Hysys® 7.2

I. INTRODUCTION

Greenhouse gas emissions are serious dilemma that concerns the scientist and politician together, which accelerate the pace of finding solutions. As such, different regulation are issued a year after a year forcing polluters to cut these emissions. Several methods are currently being investigated towards reducing carbon emissions among which include higher efficiency energy conversion processes, fuel switching and of particular interest, carbon capture and storage, CCS [1]. In the higher efficiency conversion process, carbon reduction is performed prior to the combustion. This operate based upon the increased energy output of power generation plants for every unit of fuel used through more efficient processes which in turn reduces the mass of carbon dioxide released into the atmosphere per unit of energy produced. An example of a higher efficiency conversion process is integrated gasification combined cycles (IGCC) which can be modelled as heat engines [2, 3, 4]. By combining two heat engines cycles, the overall efficiency will increase. However, to implement such processes, it may require installing additional and retrofitting new equipment onto the existing power plant which will result in accrued capital costs. Another solution to increased efficiency with or without combined cycles is to increase the temperature differential between the heat input and the

rejected heat by increasing the temperature and pressure of the cycle but there may not be many materials that can withstand such high levels of the temperature and pressure [2, 3, 4].

Fuel switching is another option for reduction of CO₂ emissions into the atmosphere. The objective of fuel switching is to utilize means that possess lower carbon content in its composition or produces less emission in its energy generation [3, 5]. In doing so, combusting these fuel sources or its generation methods will result in lower CO₂ emissions into the atmosphere. Renewable sources of energy such as solar, wind, wave, etc are examples of fuel switching options as they do not emit CO₂ during the generation. However, despite this simple and straightforward as well as advantageous solution to carbon dioxide emissions, the energy output of these fuels is lower and larger amounts of the fuel sources are required compared to fossil sources. This may result in higher capital and operating costs [3, 5]. Biomass co-firing is one of the options currently used in the industry. It is a process where a proportion of a biomass fuel source is burnt together with a proportion of coal. The energy density of coal is significantly greater than biomass and therefore energy output will be lower. Natural gas is also another alternative fuel as its combustion emits lower levels of harmful substances into the atmosphere. Availability of these alternative fuel sources in the long run may also be an issue [3, 5].

CCS has been considered as an option for limiting CO₂ emissions from the combustion of fossil fuels [1]. This approach includes pre-combustion capture, post-combustion capture and oxy-fuel combustion [2, 3, 4, 5]. In pre-combustion capture, the carbon is removed prior to the combustion process through a series of chemical reactions or processes and it may include the methods of higher energy conversion processes and fuel switching. In post combustion capture, the removal of CO₂ is done after the combustion of the fuel source. In oxy-fuel capture, the fuel source is combusted with O₂/air and the flue gas is cleaned for a carbon free stream [6, 7]. The post-combustion capture technology is more feasible as opposed to the other carbon capture approaches because of its lower cost and retrofitting option. It allows implementation without incurring relatively larger capital and operating costs and allows more plant flexibility as well [7].

Generally, carbon capture plants is still under the development stages, thus great efforts are focused on the most favourable capture method to establish optimum conditions. Hysys® software is used extensively in this direction. Many attempts recently to simulate the carbon capture plant, especially amine based, are carried out for the purposes of feasibility study and the design aiding. Simulation of the amine absorption post-combustion technology is considered as a complex matter, as accurate rigorous models are required [8]. Most modelling focuses mainly on parameter optimization such as the flue gas flow rate and its effect, CO₂ concentration, solvent temperature and pressure, recycle stream, heat and work duties along with other parameters as reported by Singh et al. (2003) [9]; Fisher et al. (2005) [10]. An attempt to analyse a 500MW existing coal power plant with monoethanolamine (MEA) for CO₂ capture in Canada was carried out by Ali (2004) [11] using ASPEN PLUS. The pressure profile for the columns was of utmost priority to analyse the hydrodynamic performance for design purposes. Further development for the same case was done by Alie et al. (2005) [12] to improve the convergence in ASPEN.

Another trial was performed for 600MW bituminous coal fired power plant, using MEA in ASPEN software for column design. The optimized parameter was the thermal energy consumption. 3.0 GJ/ton was found as the minimum achieved by 40 wt% MEA solutions for 13 %v/v CO₂ in flue gas and 90% recovery [13]. A similar attempt was done by Oi, (2007) [14]. The work of Khakdaman et al. (2008) [15] focused on improving efficiency, using a mixture of solvents, diethanolamine (DEA) and N-methyldiethanolamine (MDEA) for a fixed CO₂ recovery and heat duty in columns. The simulation proved that a better capacity could be achieved when solvent blends are used.

In this present study, initial attempts were made at simulating a carbon capture and storage system using a chemical absorption process with chilled ammonia as the solvent. Our investigations revealed that the chilled ammonia process has many advantages over the conventional amine solvents in terms of cost efficiency and technical viability. One example of these advantages is the lower heat of absorption which leaves the process in less need of heating requirements. This is also in accordance with Darde et al (2010) [16]. However, the process met a lot of difficulties due to the absence of suitable property packages in Aspen HYSYS and lack of data (such as interaction parameters to use other conventional fluid packages) and therefore decided to continue with the amine solvents.

The objective of this study was to carry out a technical and economic analysis of carbon capture in 500MW full capacity (420MW generation) coal fired power plant using Hysys® simulation to achieve a above 90% recovery of CO₂ with a purity of 99% using amine solvents.

This article is structured in to different sections: Methodology, results and discussions, conclusion, future work, acknowledgment, references and appendix.

II. METHODOLOGY

2.1 Feed Composition

The plant feed comprises of mainly carbon dioxide, nitrogen, water and oxygen which is obtained from a 500 MW plant. Table 1 below gives a typical composition of the flue gas from a coal fired power plant. The operational window of the power plant is assumed to be 8000 hours annually. The desired removal of carbon dioxide from the inlet stream is to be approximately 90% of the original in-flow in to the absorption column.

2.2 Process Description

The absorber column is operating at certain pressure with a pressure drop across each stage. The absorber column consists of packings which maximise gas/liquid contact and give sufficient liquid hold-up for reaction between CO₂ and amine solvent to take place. The flue gas leaves the top of the absorber column with a largely reduced concentration of CO₂ at a particular temperature and pressure. The gas leaving the top of the absorber is released into the atmosphere with a significantly reduced concentration of CO₂. The solvent, now rich with CO₂ leaves the bottom of the absorber column at a specific temperature and pressure. The bottom stream is then pumped to increase the pressure before entering the 2 way heat exchanger. The 2 way heat exchanger brings into contact the rich-in-CO₂ bottom stream from the absorber column and the lean CO₂ solvent which exits the bottom of the stripper column. The CO₂ rich stream is heated before it enters the stripping unit where CO₂ is stripped off from the amine solvent. The lean solvent is then cooled and reused in the absorption column. A schematic process flow diagram is shown in Figure 1 below.

Table 1: Coal Fired Power Plant Flue Gas [3]

Capacity (MW)	500
Temperature (°C)	50
Pressure (Kpa)	110
Molar Flow (Kmole/h)	71280
Volume Flow (m ³ /h)	1914000
<i>Composition (mole percent)</i>	
Carbon Dioxide (CO ₂)	15.00
Nitrogen (N ₂)	77.00
Sulphur Dioxide (SO ₂)	0.00001
Nitrogen Oxides (NO _x)	0.00005
Water (H ₂ O)	5.00
Carbon Monoxide (CO)	0.00001
Oxygen (O ₂)	3.00

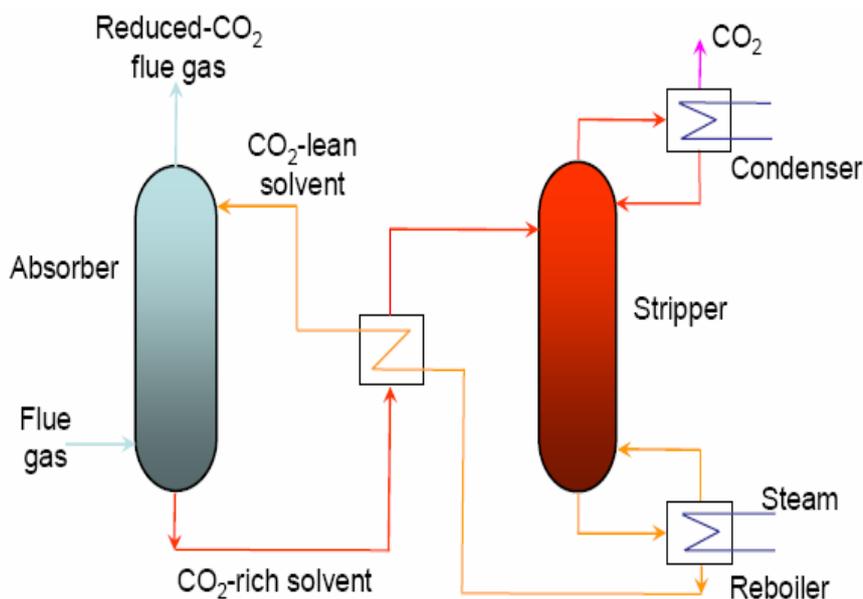


Figure 1: Schematic of amine scrubbing process in CO₂ post-combustion capture

III. RESULTS AND DISCUSSIONS

3.1 Amine Selection

Amine selection for carbon capture is based on three important factors: Corrosiveness, absorption capacity and stripping condition. A study of the absorption and desorption (stripping) capacity of the amines is also necessary to avoid the need for extremely tall towers that can offer the process enough area and residence time for the CO₂ transfer from/towards the solvent. It is well known that corrosion effect of amines varies depending on the class and concentration used. No tertiary amine concentration can have corrosion rates similar to those of primary and secondary amines, so a concentration of 50%w/w as suggested by the literature [17] was used to compare the behaviour of MDEA against MEA and DEA. Similarly, 30%w/w of the secondary amine DEA, and 16.5%w/w MEA (primary amine) were employed. The results of simulations are shown in Figures (2 to 4) for the three different amines.

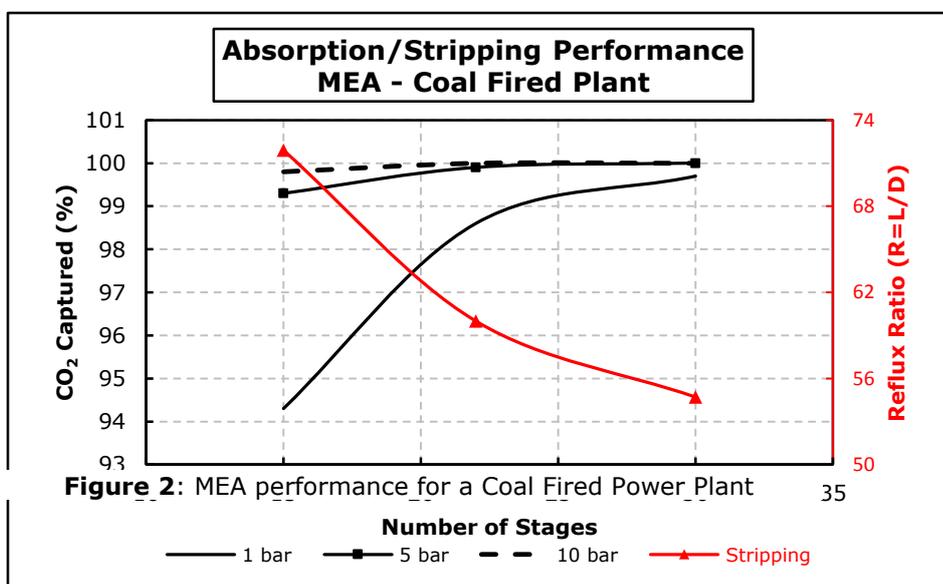


Figure 2: MEA performance for a Coal Fired Power Plant

From the Figure 3, it is seen that a primary amine, MEA is very appropriate for absorb CO₂ in the flue gas with capture over 90% as required (Audit report, 2009). However, from the stripping curve read from the right side of the plot, it is also observed that the system requires huge reflux ratios in separation which suggests that this process is not cost effective.

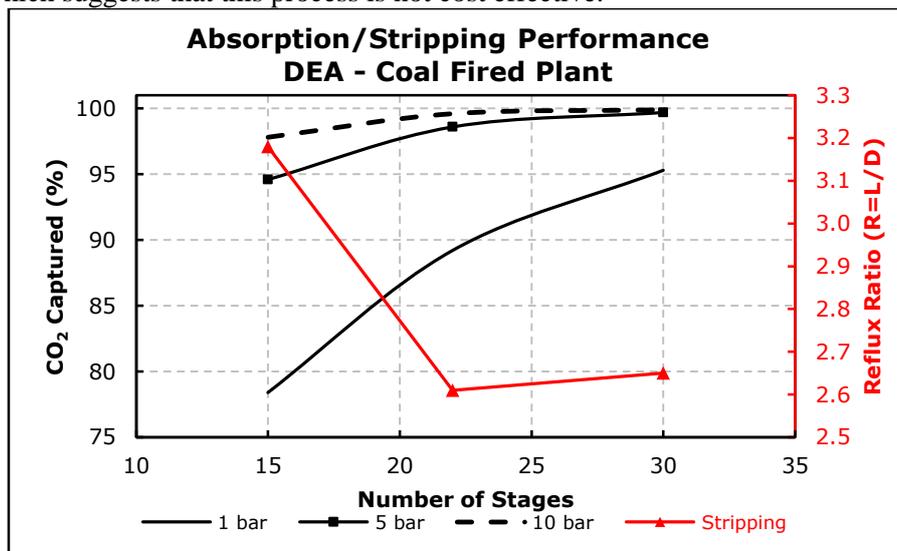


Figure 3: DEA performance for a Coal Fired Power Plant

Performance of absorber and stripper in a carbon capture plants at different pressures, for DEA based solvent showed reduction in affinity with the acid gas, requiring further efforts in absorption. The process requires either high pressures or more than 23 stages in absorption for an atmospheric pressure absorber to captures over 90% CO₂. On the other hand, stripping has been improved to very acceptable levels, with reflux ratios within the range 2–3.2 as shown in the Figure 3.

Tertiary amines as MDEA are very well known to have limited affinity for acid gases in comparison to primary and secondary amines as earlier stated. This fact can easily be observed in Figures 4 above where no operating pressure can achieve the target capture, making of this option the least favourable in terms of absorption rate. Nevertheless, the stripping process is much easier with very low reflux ratio in relatively small columns.

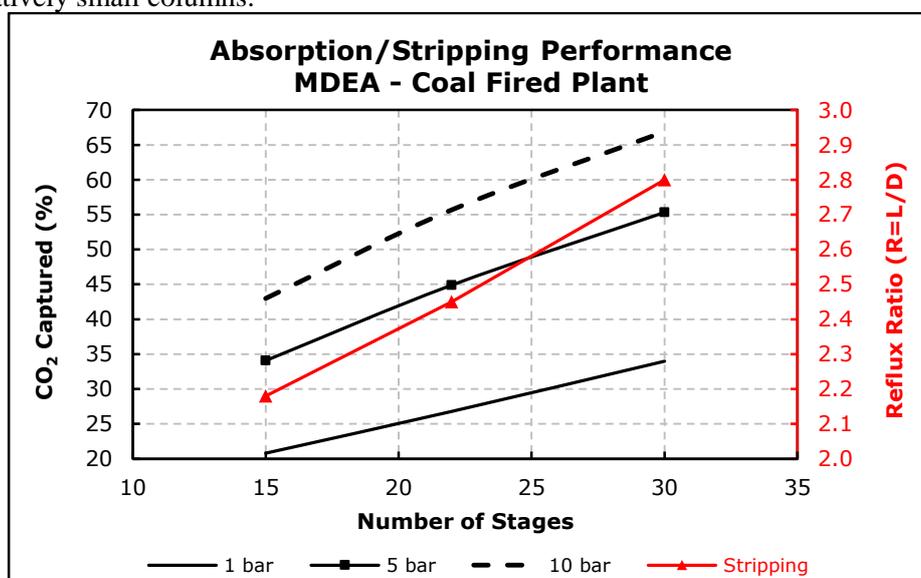


Figure 4: MDEA performance for a Coal Fired Power Plant

It is now clear that using only tertiary amines to capture acid gases is not technically feasible due to the high number of stages needed for absorption at the required level, no matter how easy the

stripping process could be. On the other hand, primary amines showed a good tendency to removing CO₂ from the flue gas with the additional constraint of difficult separation in the recovery process. This would be an advisable option only in case of diluted flue gas stream. The secondary amine DEA showed the best performance in terms of affinity for the acid gas and separation viability with moderate reflux ratio making it suitable for the design considered in this study.

3.2 Process Modelling and Simulation

From the process flow diagram in the Figure 5 below, it is necessary to ensure that the inlet stream can overcome the head loss through the absorption towers. It is also well known that a higher operating pressure in tower will increase the CO₂ absorption into the liquid phase, however, this process is very expensive due to the high energy requirement to increase the pressure of the large flue gas flow rate coming from the power station and therefore lower pressures are preferred despite their lower performance within the absorption equipment, thus, only the necessary head was supplied to the flue gas using blower to travel across the absorption stages. A 2 kPa pressure drop was assumed for each stage in the absorption towers.

Amine absorption processes for CO₂ capture are typically carried out with aqueous amine solutions with concentrations between 25-30% by mass, which enters at 45°C on the top of the absorber while the acid gas enters in the tower at the bottom and leaves the equipment at the top as a sweet gas with very low CO₂ concentration (under 1.5% mole fraction). This equipment was designed by modifying the number of stages, until the absorption rose over 90%. It is necessary to clarify that temperatures over 125 °C can cause amine degradation and therefore there is a software limitation associated with the fluid package available for this purpose (Amines Package in Aspen HYSYS), warnings about the accuracy of calculation for the properties of streams above 125 °C are shown during the simulation, this operating condition was established to reach a desired recovery and top composition of CO₂ and the solver was run until convergence was reached. A bottom pressure of 215 kPa and a condenser temperature of 40 °C achieved up to 91% recovery of CO₂ with an initial 3.81 reflux ratio and CO₂ purity on tops of 95.7% by mole. The bottom stream leaving the stripping column contains all the amine used in absorbing the acid gas and was recycled. Enthalpy of the recycled stream was used to preheat incoming feed to the tower as system heat integration.

The amine and water concentrations in the treated gas stream were very low and their accumulation was observed when the recycle unit is connected. As the iterations reach convergence, the amine required to keep the system stable is always under 1 kg/h, typically 0.5 kg/h. On the other hand, the water balance shows that high temperatures in the recycled stream will increase the water loss in absorption pushing for a high water make up, while low temperatures are difficult to reach due to the high cooling demand in this stream. This balance deserves an economic analysis in order to establish the equilibrium point in which the water consumption does not exceed the cooling expenses. Two adjust units were used to vary different parameters in the process, one of them is used to change the inlet gas flow based on the capacity of the power station and evaluate its impact on the capture efficiency. The other was used to vary the inlet flow rate of amine to the absorber to ensure a good CO₂/mol Amine ratio for the absorption units.

The cooling requirements for this process are related to the heats that must be removed from the condenser of the stripping column, where the amine is recovered, and an additional heat that must be taken out in the recycled amine in to the absorber. This heat was taken out from the process by means of heat exchangers. All the heat exchangers were designed to keep a temperature approach of 12.6 °C. The rate at which the makeup components are fed into the process could help reducing the effect of the amine decomposition. In this case, purging some part of the used amine became necessary and a greater make up was required in order to keep the chemical activity of the absorbent in the system.

As the flue gas coming from the power plant consist of low percentage of solids that could not be simulated in Aspen HYSYS, a microfilter casing with special designed cartridges suitable for amines operation was installed in the recycled amine stream coming from the bottom of the stripper, this equipment is expected to retain solids captured from the flue gas during the absorption and degradation products formed during the stripping procedure.

All pressure drops due to friction, tubing accessories and head losses were calculated and imposed on hypothetical valves right after each pump in the process flow diagram. These valves will only account

for the energy required to move the streams through the pipes at the different levels of the plant and do not include the pressure drops within the different pieces of equipment.

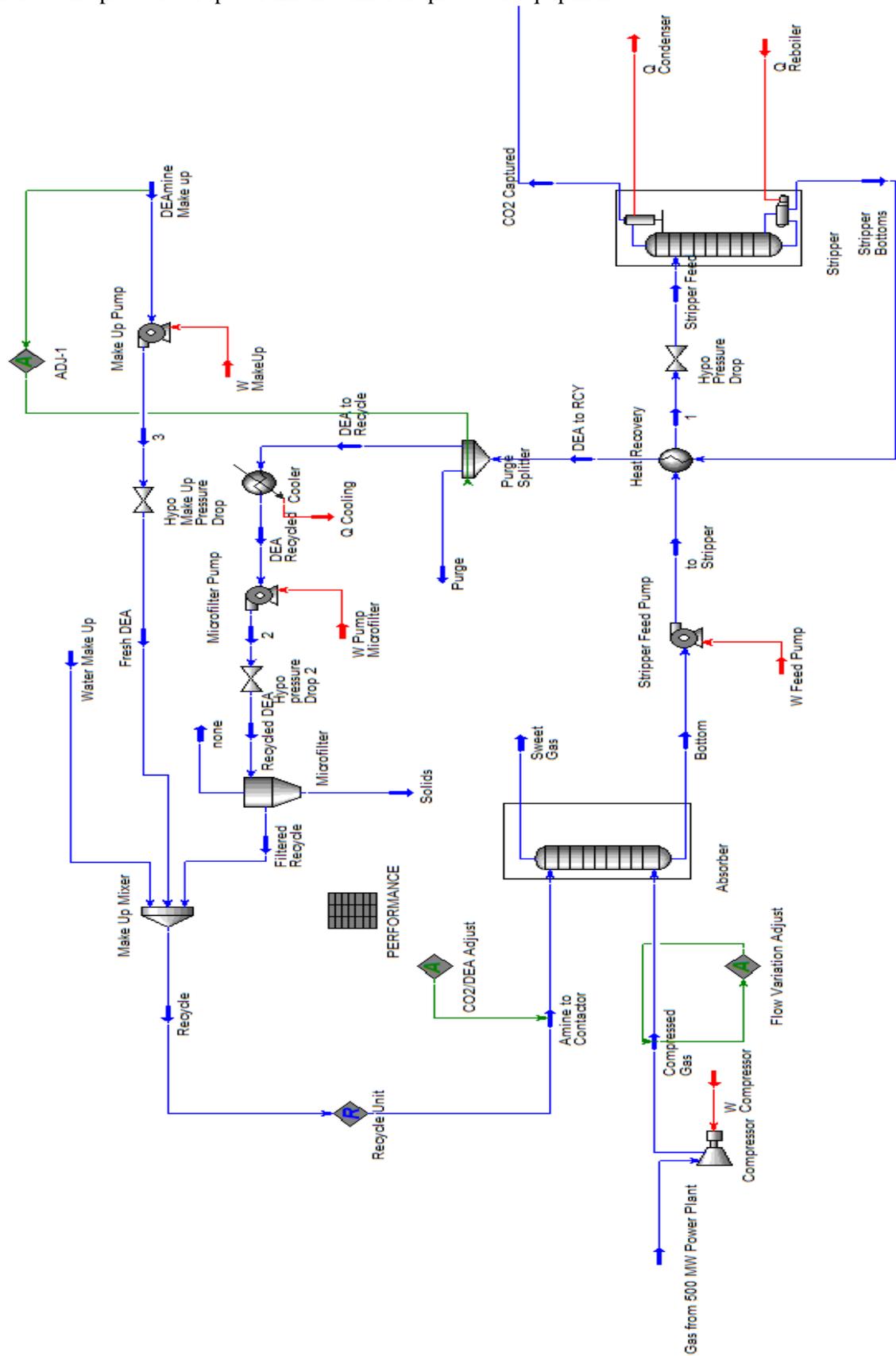


Figure 5: Process Flow Diagram

Table 2: Summary of material stream obtained from Hysys

Material Streams													
	Compressed Gas	Amine to Contactor	DEA to Recycle	Recycled DEA	DEAmine Make up	Recycle	Gas from 500 MW Power Plant	Sweet Gas	Bottom	Fresh DEA			
Vapour Fraction	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	0.0000	0.0000			
Temperature	91.59	47.00	71.15	47.09	25.00	47.00	50.00	46.97	58.96	25.13			
Pressure	160.0	100.0	150.0	435.2	100.0	100.0	110.0	100.0	160.0	100.0			
Molar Flow	7.128e+004	6.965e+005	6.932e+005	6.932e+005	0.8417	6.965e+005	7.128e+004	6.393e+004	7.033e+004	0.8417			
Mass Flow	2.141e+006	1.673e+007	1.667e+007	1.667e+007	88.50	1.673e+007	2.141e+006	1.745e+006	1.713e+007	88.50			
Liquid Volume Flow	2603	1.634e+004	1.628e+004	1.628e+004	8.082e-002	1.634e+004	2603	2113	1.683e+004	8.082e-002			
Heat Flow	7.748e+008	-1.824e+010	-1.660e+010	-1.813e+010	4.593e+004	-1.824e+010	6.837e+008	6.033e+008	-1.807e+010	4.596e+004			
	Water Make Up	Stripper Feed	Stripper Bottoms	CO2 Captured	to Stripper	Purge	DEA to RCY	Filtered Recycle	Solids	none			
Vapour Fraction	0.0000	0.0040	0.0000	0.9999	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000			
Temperature	25.00	110.0	123.7	38.36	59.09	71.15	71.15	47.09	47.09	47.09			
Pressure	100.0	200.0	215.0	150.0	548.8	150.0	150.0	100.0	100.0	100.0			
Molar Flow	3236	7.038e+005	6.933e+005	1.060e+004	7.039e+005	11.91	6.933e+005	6.932e+005	0.0000	0.0000			
Mass Flow	5.830e+004	1.713e+007	1.668e+007	4.542e+005	1.713e+007	286.4	1.668e+007	1.667e+007	0.0000	0.0000			
Liquid Volume Flow	58.41	1.683e+004	1.628e+004	549.5	1.683e+004	0.2796	1.628e+004	1.628e+004	0.0000	0.0000			
Heat Flow	-1.104e+008	-1.457e+010	-1.311e+010	1.049e+008	-1.806e+010	-2.852e+005	-1.660e+010	-1.813e+010	0.0000	0.0000			
	DEA Recycled	1	2	3	4	5	6	7	8	9			
Vapour Fraction	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000			
Temperature	46.86	112.1	47.09	25.13	107.0	107.0	32.00	32.00	32.00	32.00			
Pressure	100.0	528.8	765.9	374.0	336.7	792.2	336.7	336.7	792.2	792.2			
Molar Flow	6.932e+005	7.038e+005	6.932e+005	0.8417	1.060e+004	1.028e+004	1.028e+004	316.7	1.020e+004	83.08			
Mass Flow	1.667e+007	1.713e+007	1.667e+007	88.50	4.542e+005	4.485e+005	4.485e+005	5719	4.470e+005	1505			
Liquid Volume Flow	1.628e+004	1.683e+004	1.628e+004	8.082e-002	549.5	542.8	542.8	5736	541.3	1.511			
Heat Flow	-1.814e+010	-1.457e+010	-1.813e+010	4.596e+004	1.336e+008	1.296e+008	9.912e+007	-1.061e+007	9.828e+007	-2.776e+006			
	10	11	12	13	14	15	16						
Vapour Fraction	1.0000	1.0000	0.0000	1.0000	1.0000	0.0000	1.0000						
Temperature	107.0	32.00	32.00	107.0	32.00	32.00	107.0						
Pressure	1862	1862	1862	4354	4354	4354	9963						
Molar Flow	1.020e+004	1.016e+004	34.37	1.016e+004	1.015e+004	11.96	1.015e+004						
Mass Flow	4.470e+005	4.463e+005	606.7	4.463e+005	4.461e+005	220.7	4.461e+005						
Liquid Volume Flow	541.3	540.7	0.6306	540.7	540.4	0.2230	540.4						
Heat Flow	1.285e+008	9.793e+007	-1.142e+006	1.281e+008	9.781e+007	-3.931e+005	1.280e+008						

Table 3: Summary of energy streams

Energy Streams									
		Q Cooling	W Compressor	W MakeUp	Q Reboiler	Q Condenser	W Feed Pump	W Pump Microfilter	Q-100
Heat Flow	kJ/h	1.540e+009	9.106e+007	28.94	2.870e+009	1.308e+009	8.497e+006	1.441e+007	2.872e+007
		Q-101	Q-102	Q-103	Q-104	Q-105	Q-106	Q-107	Q-108
Heat Flow	kJ/h	-4.509e+007	3.047e+007	-3.408e+007	3.026e+007	-3.175e+007	3.017e+007	-3.069e+007	3.014e+007

3.3 Equipment Specification

Detail description of the procedure on how to determine all the parameters used in this section are shown in the Appendix

3.3.1 Absorption column

Assumptions and design considerations:

- CO₂ absorption is carried out in a continuous counter current packed column, to achieve high contact area between the flue gas and the solvent.
- Both structured and random packing could be used, but random is chosen because it is cheaper. Plastic Intalox[®] saddles are used to overcome the corrosive nature of Amines, beside its high efficiency compared to ceramic [18].
- DEA absorption has no tendency to produce foams in the absorption process [19].

Table 4: Absorption column specification

<u>Column Details</u>		
Number of columns	-	8
Number of stages	-	30 ¹
Internals	-	Packed ²
Internals Type	-	Intalox saddles (3_inche) ²
HETP	-	0.8468
Foaming Factor	-	1 ²
Maximum flooding	%	84 ³
Material of construction	-	Stainless steel
<u>Column sizing</u>		
Section diameter	m	6 ¹
Section height	m	25.5 ¹

¹ Obtained from the simulation

² Assumed

³ Calculated in appendix

3.3.2 Stripper column

Assumptions and design considerations:

- Solvent recovery is carried out in a stage wise distillation column.
- Sieve plate is used among the different plate types, as it is cheap, has less tendency for fouling and suitable for most applications [18].
- DEA absorption has no tendency to produce foams in the absorption process [19].

Table 5: Stripper column specification

<u>Column Details</u>		
Number of columns	-	8
Number of stages	-	30 ¹
Internals	-	plate ²
Internals Type	-	Sieve plate ²
Tray spacing	m	0.6096 ²
Tray thickness	mm	3.175 ²
Foaming Factor	-	1

Maximum flooding	%	65 ³
Column sizing		
Section diameter	m	6 ¹
Section height	m	18 ¹

¹ Obtained from the simulation² Assumed³ Calculated in appendix

3.3.3 Compressors

Table 6: Compressor specification

Specification		Flue Gas Compression	Compression to Storage
<i>Operating Conditions</i>			
Q _{in}	m ³ /h	1'739,485	175,712
Q _{out}	m ³ /h	1'350,182	2,463
P _{out}	bar	1.6	100
ΔP	bar	0.5	98.45
Pressure Ratio	-	3:2	4:1
Inlet Temperature	°C	50	38.4
Polytropic Efficiency	%	82.57	81.7
<i>Compressor Details</i>			
Single/Multi stage	-	Single Stage	Multi Stage
Intercooling?	y/n	No	Yes
Material	-	Carbon Steel	Carbon Steel
Type	-	Centrifugal fan / Axial blower	Centrifugal Compressor
<i>Compressor Disposition</i>			
Number of compressors	-	5	9
Series	-	0	3
Parallel	-	5	3

3.3.4 Pumps and pipelines

Table 7: Pump specifications data

Specification		Stripper Feed	Make Up	Recycle
<i>Piping and Head Losses</i>				
Flow rate	m ³ /h	16,390	0.079	16,223
Piping Inside Diameter	in	21.6	0.1	32.3
ΔP _{equipment}	kPa	20.0	0.0	450.2
ΔP _{static}	kPa	174.3	274.0	252.1
ΔP _{friction}	kPa	86.2	0.0	97.0
<i>Pump Details</i>				
Pumps required	-	9	2	4
Material	-	Stainless Steel 316	Stainless Steel 316	Stainless Steel 316
Type of pump	-	Single Stage (3500 rpm)	Single Stage (3500 rpm)	Single Stage (1750 rpm)
Specific Speed	rpm	10824	76	7274
Type of Impeller	-	Axial	Radial	Axial
Efficiency	%	82	60	82

3.3.5 Microfilter

Table 8: Microfiltration system specification

Specification	Units	Value
Volumetric flow rate	m ³ /h	16,449
Number of Systems Installed	-	4
Cartridges per System	-	19
Casing Pressure Drop	kPa	30.4
Cartridges Pressure Drop	kPa	304.8
Pressure Drop	kPa	335.2

3.3.6 Heat Exchangers

The four heat exchangers are: the recovery heat exchanger, cooler, stripper condenser and stripper reboiler.

Table 9: Heat recovery exchanger specification

Shell side conditions			
Parameters		Inlet	Outlet
Shell side flow	kmol/hr	703849.23	703849.23
Temperature	°C	59.09	110.00
Pressure	kPa	548.80	200.00
Tube side conditions			
Parameters		Inlet	Outlet
Tube side flow	kmol/hr	693251.16	693239.26
Temperature	°C	123.72	71.15
Pressure	kPa	215.00	150.00

Table 10: Cooler specification

Shell side conditions (CO ₂ lean solvent)			
Parameters		Inlet	Outlet
Shell side flow	kmol/hr	893694.71	893694.71
Temperature	°C	25	25
Pressure	kPa	101.3	101.3
Tube side conditions (Cooling water)			
Parameters		Inlet	Outlet
Tube side flow	kmol/hr	693239.26	693239.26
Temperature	°C	71.153	46.86
Pressure	kPa	150	100

Table 11: Stripper condenser specification

Parameter	Unit	Value
Flow into condenser	kg/h	960514.75
Flow out of condenser	kg/h	454175.65
Reflux	kg/h	506339.10
Condenser Inlet temperature	°C	103.43
Condenser Outlet temperature	°C	38.28

Table 12: Stripper reboiler specification

Parameter	Unit	Value
Flow into reboiler	kg/h	17972649.79
Flow out of stripper bottoms	kg/h	16675110.21
Total flow out	kg/h	34647759.99
Temperature into boiler	°C	123.07
Temperature in boiler	°C	123.72

Table 13: Heat exchanger area specifications

	Area
Heat Recovery Exchanger	60.32 m ²
Cooler Design	60.32 m ²
The stripper condenser	117.6 m ²
Stripper reboiler	52,640m ²

3.4 Energy Requirements

3.4.1 Heating - steam

The heat supplied to the reboiler in the stripping columns will be given to the process according to the specifications below as calculated in the Appendix

Table 14: Steam specifications for the CCS process

Specification	Value
<i><u>Operating Conditions</u></i>	
Reboiler Temperature	°C 123.7
Min Temperature of Steam	°C 130.7
Saturation Pressure @ 130.7°C	kPa 279.6
Steam Pressure Used	kPa 300
Steam Flow Rate	ton/h 1310

3.4.2 Cooling – water from cooling towers

The most common and efficient refrigerant used for plants when the temperatures are above ambient 25 °C and below 50 °C is water, as mentioned in Sinnott (2005) [18] pp. 769, where cases with temperatures above 65 °C are suggested to be cooled in a cheaper way by means of air (using fans). However, the only stream suitable to be cooled down with air (DEA recycled) can be refrigerated by using heat exchange nets and a complete calculation of the cooling water requirements and the possibilities for reuse in the process is presented in Appendix, the results are summarized below in Table 14.

Table 15: Cooling requirements for the CCS process simulated in Aspan-HYSYS

Specification	Condenser	Cooler	Compression
<i><u>Operating Conditions</u></i>			
Heat Flow Rate	MJ/h 1,307,881	1,540,199	141,604
Water Flow Rate	ton/h 49,873	43,916	< 5957
Water Inlet Temperature	°C 25.0	31.4	31.4
Source* : Fresh/Reuse	- Fresh	<i>Reuse</i>	<i>Reuse</i>
Water Outlet Temperature	°C 31.4	39.9	37.2

*Notes

Fresh : Water coming directly from cooling towers

Reuse: Water coming from a previous heat exchanger

3.5 Cost Analysis

Assumptions and economic parameter used

Discount rate:

The discount rate of a project is linked to its market risk. The market risk is usually a systematic risk that includes technology maturity, policy risk and others. Therefore, the final discount rate is calculated according to Oxera (2011) [20] as follows:

$$\text{Discount rate} = \text{Risk-free rate} + \text{compensation for risk} \text{ ----- (1)}$$

Coal CCS is considered as a high risk, immature technology with a great uncertainty in the estimation of its feasibility. It is found that in 2011 the rate for a coal CCS plant ranged from 12 to 17% [20]. An average discount rate of 15% was used in this study.

Useful life time:

As the capital cost for this plant may be very high, we try to make the plant life time extend as long as the equipment can withstand the corrosion, degradation and other effects that shorten the life time of the plant. Usually many studies expected a life time for a CCS from 30 to 40 [21, 22]. In this study a life time of 25 years is suggested, after which an equipment replacement may be implemented to extend the life time of the plant which will add another term to the plant cost.

Table 16 Assumed values for cost estimation

parameter	Value
Discount rate	15%
Plant lifetime	25 years
Year of plant start-up	2015
Plant Capacity	500 MW
Capacity factor	90% ¹
Net energy output	420MW ²

¹Assumed

²Calculated in appendix

Cost measures:

To characterize CO₂ capture, the following are some metrics of cost.

Capital cost: It is known as the first cost or the investment cost. It could be expressed in cost per year or per kW [23].

Cost of CO₂ avoided: This cost shows the average cost of reducing unit mass of CO₂ from the atmosphere, providing the same kW of electricity from the plant without carbon capture (reference plant). It is calculated by the following formula [23]:

$$\text{Cost of CO}_2 \text{ avoided (}\text{£/ton CO}_2\text{)} = \frac{\text{COC}_{\text{capture}} - \text{COC}_{\text{ref}}}{\text{Co}_2/\text{kWh}_{\text{ref}} - \text{Co}_2/\text{kWh}_{\text{capture}}} \text{ ----- (2)}$$

Where, CO₂/kWh is CO₂ mass emission rate per kW generated.

Cost of CO₂ captured: This based on mass of CO₂ captured or removed instead of CO₂ avoided, and it could be found by:

$$\text{Cost of CO}_2 \text{ captured (£/ton CO}_2\text{)} = \frac{\text{COE}_{\text{capture}} - \text{COE}_{\text{ref}}}{\text{CO}_2/\text{kWh}_{\text{capture}}} \quad (3)$$

Where COE is the levelized cost of electricity (cost/kWh)

Table 17 below shows a summary of the obtained cost and a detailed calculation is showed in the appendix

Table 17: Results of the cost estimation

Cost	value
Total capital investment (£)	417,974,468.1053
Total product cost (£/year)	104,115,032.2031
Electricity cost before CCS (£/kW.h)	0.1051
Additional Electricity cost after CCS (£/kW.h)	0.028
Carbon floor price (£/ton of CO ₂)	16.00
Cost of treating a ton of CO ₂ (£/ton of CO ₂)	26.80

IV. CONCLUSION

Simulation of amine based post-combustion carbon capture plant in a 500MW coal fired power plant was carried using Aspen Hysys^R. Preliminary analysis of different amine solvent was conducted and diethanolamine (DEA) showed better performance. The plant was found to require eight absorber and eight stripping units. Equipment specification and economic analysis were carried out. The decision in this case (higher cost for treating CO₂ than carbon floor price) will be at the discretion of the plant stakeholders either by operating without a capture plant and paying the carbon tax or by retrofitting the power plant for carbon capture and paying the cost of treating the CO₂. The costing of the capture plant may hinder the feasibility of the retrofitting option. However, if the emphasis is on environmental consideration, the retrofitting option is one they should consider.

V. FUTURE WORK

Techno-economic analysis of carbon capture in a gas fired power plant of similar capacity will be an interesting study. This will provide basis for comparison between the two sources of fuel.

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AUTHORS BIOGRAPHY

Isah Yakub Mohammed was born in Bida, Niger State, Nigeria on the 22nd March, 1982. He obtained BEng and MSc degrees in Chemical Engineering from Federal University of Technology Minna, Nigeria and the University of Nottingham United Kingdom in 2006 and 2012 respectively. He is currently a PhD student in Chemical Engineering at University of Nottingham and a lecturer at Abubakar Tafawa Balewa University Bauchi, Nigeria. Isah Yakub Mohammed is a registered Engineer with Council for the regulation of Engineering in Nigeria (COREN), a Coporate member of Nigerian Society of Engineers



(NSE) and Nigerian society of Chemical Engineers (NSChE). Member American Institute of Chemical Engineers (AIChE), International Water Association (IWA) and Associate Member Institution of Chemical Engineers (ICHEM).

Samah Mohamed was born in Sudan. She obtained BSc and MSc degrees in Chemical Engineering from University of Khartoum, Sudan and the University of Nottingham United Kingdom in 2009 and 2012 respectively. She is currently at King Abdullah University of Science and Technology (KAUST) Saudi Arabia as a graduate student for her PhD degree. Samah is a lecture at university of Khartoum.



Umar Sule Danladi was born in Bida, Niger State Niger, Nigeria on the 6th June, 1965. He obtained MEng degree in Chemical Engineering from Federal University of Technology Minna, Nigeria in 2010. He is currently a Senior Lecturer at the Federal Polytechnic Bida, Nigeria. He is also a Coporate member of Nigerian Society of Engineers (NSE) and Nigerian society of Chemical Engineers (NSChE).

APPENDIX

DETAILED DESIGN CALCULATION AND COSTING

1. Absorption Column

Design based on the following data from a stage below the feed position:

	Mass flow rate (kg/h)	Volumetric flow rate (m ³ /h)	Calculated density (kg/m ³)
Liquid	1.975e6	1905	1027.3
gas	2.209e5	1.673e5	1.32

To size the absorber using HYSYS®, the following are the parameters that need to be calculated.

Pressure drop in mmH₂O/plate:

$$\begin{aligned} \Delta P / \text{plate} &= 2 \text{ kPa} \\ \Delta P \text{ in mmH}_2\text{O} &= \rho g h = 1000 \times 9.81 \times h \\ h &= 203.9 \text{ mmH}_2\text{O} \\ \Delta P \text{ in mmH}_2\text{O/meter of packed height} &= 203.9 \text{ mmH}_2\text{O} / 0.8468 \text{ m} \\ \Delta P / \text{plate} &= 240.8 \text{ mmH}_2\text{O/m} \end{aligned}$$

Flooding percentage:

$$\% \text{ flooding} = \left[\frac{K_4 \text{ at design pressure drop}}{K_4 \text{ at flooding}} \right]^{0.5}$$

The constant K₄ is estimated from the following figure [18]:

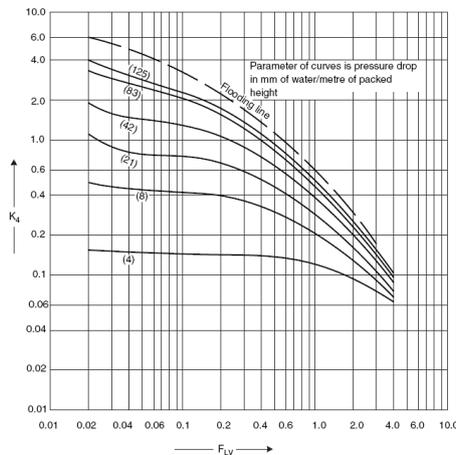


Figure 1: K₄ constant for flooding calculations

Where $F_{LV} = \frac{L_w^*}{V_w^*} \sqrt{\frac{\rho_v}{\rho_L}}$ (L_w and V_w are liquid and vapour mass flow rate, kg/s).

$$F_{LV} = 0.32$$

From the graph at $F_{LV} = 0.32$ and $240.8 \text{ mmH}_2\text{O}/\text{m}$ curve

	F_{LV}	Curve	K_4
At design pressure drop	0.32	240.8 mmH ₂ O/m	1.3
At flooding	0.32	(Flooding curve)	1.8

$$\% \text{ flooding} = \left[\frac{K_4 \text{ at design pressure drop}}{K_4 \text{ at flooding}} \right]^{0.5} = 84.98\%$$

Cost estimation:

Wall thickness is calculated (Sinnott, 2005): $t = \frac{P_i D_i}{2S - P_i}$

Where, P_i is the internal pressure, and

S is the maximum allowable stress (For stainless steel = 20 ksi = 1.397.5 kPa)

$$t = 0.0054\text{m (adding 2mm for corrosion allowance)}$$

The cost formula used [18]:

	Cost = a + b.S ⁿ					
	a	b	N	S	S*	Cost
Vessel	15000	68.00	0.85	Shell mass (kg)	19.8e3 kg	3.205e5
Packing	0.0	1800	1.00	Volume (m ³)	2675	4.186e6
Total Cost £/year						£5.136e6

*S calculations:

$$\text{Shell mass} = \pi D L t_w \rho = 19.8\text{e}3 \text{ kg } [\rho = \text{metal density (8000kg/m}^3)]$$

$$\text{Packing volume} = \pi r^2 L = 2675 \text{ m}^3$$

$$\text{Cost for 8 columns} = \text{£}5.136\text{e}6 \times 8 = \text{£}41.08 \text{ million}$$

$$\text{Cost at (2011) using Table II-3} = \text{£}50.94 \text{ million}$$

2. Stripper Column

Design based on a stage below the feed position:

	Mass flow rate (kg/h)	Volumetric flow rate (m ³ /h)	Calculated density (kg/m ³)
Liquid	2.077e6	2605	1005.8
gas	1.333e5	1.11e5	1.2

To size the absorber using HYSYS®, the following are the parameters that need to be calculated.

Flooding percentage:

Assume 85% flooding initially

Thus actual real flooding [18]:

$$\text{Flooding \%} = \frac{U_n^*}{U_f}$$

$$\text{Flooding velocity} = U_f = K_1^* \sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$$

K_1 obtained from the graph [18]:

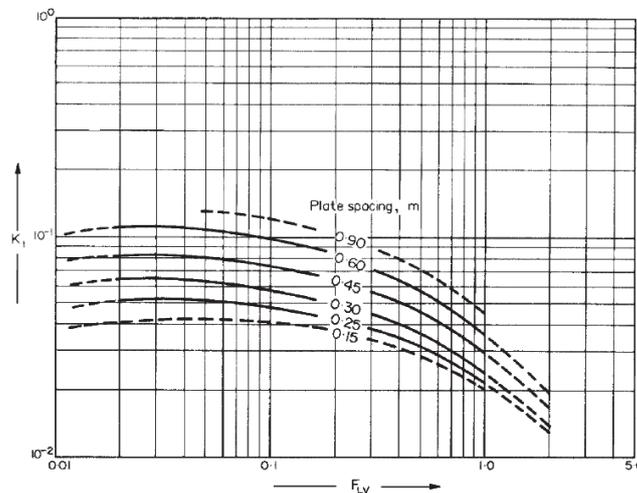


Figure 2: K4 constant for flooding calculations

Where $F_{LV} = \frac{L_w}{V_w} \sqrt{\frac{\rho_v}{\rho_L}}$ (L_w and V_w are liquid and vapour mass flow rate, kg/s).

$$F_{LV} = 0.54$$

Plate spacing is assumed = 0.6096 m

From the graph:

$$K_1 = 0.058$$

Corrected $K_1^* = K_1 \left[\frac{\sigma}{0.02} \right]^{0.2}$ [σ is the liquid surface tension]

$$K_1^* = 0.069$$

$$\text{Flooding velocity} = U_f = K_1^* \sqrt{\frac{\rho_L - \rho_v}{\rho_v}} = 1.996 \text{ m/s}$$

Thus, the velocity $U_n = \text{flooding\%} \times U_f = 1.697 \text{ m/s}$

The velocity $U_n = 1.697 \text{ m/s}$

$$\text{After sizing the velocity} = U_n^* = \frac{\text{flow rate}}{\text{cross sectional area}} = \frac{(1.11e5/3600)}{\pi 3^2} = 1.1 \text{ m/s}$$

$$\text{Thus, actual flooding \%} = \frac{U_n^*}{U_f} = 65\%$$

Cost estimation:

Wall thickness is calculated [18]: $t = \frac{P_i D_i}{2S - P_i}$

Where, P_i is the internal pressure, and

S is the maximum allowable stress (For stainless steel = 15 ksi = 1.034e5 kPa)

$$t = 0.0085 \text{ m (adding 2mm for corrosion allowance)}$$

The cost formula used [18]:

$$\text{Cost} = a + b.S^n$$

	a	b	N	S	S*	Cost
Vessel	15000	68.00	0.85	Shell mass (kg)	2.359e4 kg	3.693e5
Trays	110.0	380.0	1.80	Diameter (m)	6	1.039e4×30
Total Cost £/year						£6.811e5

*S calculations:

$$\text{Shell mass} = \pi D L t_w \rho = 2.359e5 \text{ kg } [\rho = \text{metal density (8000kg/m}^3\text{)}]$$

$$\text{Cost for 8 columns} = 6.811e5 \times 8 = \text{£}5.449 \text{ million}$$

$$\text{Cost at (2011) using Table II-3} = \text{£}6.76 \text{ million}$$

3. Compressors

Compressors are needed in the process flow sheet for pumping the gas streams through the absorbers and finally to increase the pressure of the carbon captured in order to store it underground. As these two processes are well distinguished for the pressure range allowed, it is possible to use Figure 3 and Table 1 in order to define each type of compressor.

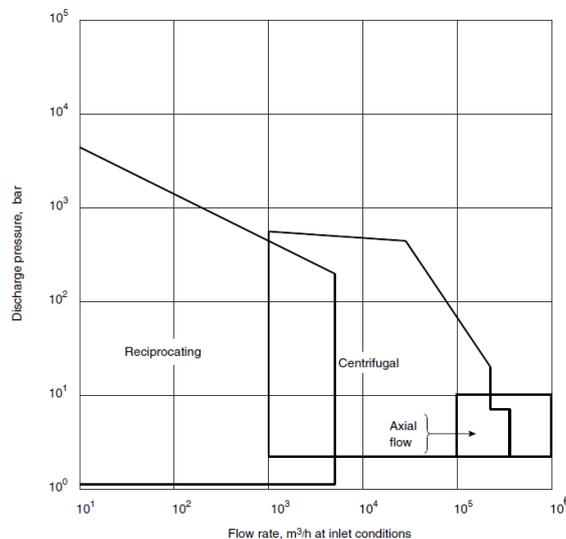


Figure 3: Compressor Operating ranges for selection [18, pp. 477] and [24 pp. 1]

Table 1: Operating ranges for compressors and blowers from [18]

Type of Compressor	Normal maximum speed (rpm)	Normal maximum capacity (m ³ /h)	Normal maximum pressure (differential) (bar)	
			Single stage	Multiple stage
<i>Displacement</i>				
1. Reciprocating	300	85,000	3.50	5000
2. Sliding Vane	300	3,400	3.50	8
3. Liquid Ring	200	2,550	0.70	1.7
4. Rootes	250	4,250	0.35	1.7
5. Screw	10,000	12,750	3.50	17
<i>Dynamic</i>				
6. Centrifugal Fan	1,000	170,000		0.2
7. Turbo Blower	3,000	8,500	0.35	1.7
8. Turbo compressor	10,000	136,000	3.50	100
9. Axial flow fan	1,000	170,000	0.35	2
10. Axial flow blower	3,000	170,000	3.50	10

Taking the volumetric flow rate at the inlet of the absorbers, we have:

$$Q_{Act} = 1,350.182 \text{ m}^3/h$$

To establish the pressure of this stream, it is necessary to analyze the absorbers and their outlets and recalculate back the necessary pressure to overcome the head losses caused by the packing material. A pressure drop of 2 kPa/stage and an atmospheric outlet for the sweet gas were assumed for the gas through the absorbers.

Accounting for 30 stages in the absorption operation,

$$\begin{aligned}
 P_{inlet\ Abs} &= P_{outlet} + \Delta P_{Absorption} \\
 \Delta P_{Absorption} &= N_{Stages} * \Delta P \text{ [kPa/stage]} \\
 &= (30 * 2) \text{ kPa} \\
 &= 60 \text{ kPa} \\
 P_{outlet} &= 100 \text{ kPa (atmospheric)} \\
 P_{inlet\ Abs} &= 100 \text{ kPa} + 60 \text{ kPa} \\
 P_{inlet\ Abs} &= 160 \text{ kPa}
 \end{aligned}$$

From Table 4-1, it is possible to get the pressure of the flue gas coming from the power plant and therefore a pressure increase can be calculated for this compressor,

$$\begin{aligned}
 P_{flue\ Gas} &= 110 \text{ kPa} \\
 \Delta P_{Compressor} &= P_{inlet\ Abs} - P_{flue\ Gas} = (160 - 110) \text{ kPa} \\
 \Delta P_{Compressor} &= 50 \text{ kPa} = 0.5 \text{ bar}
 \end{aligned}$$

By using these values in Figure 3 and Table 1, it can be seen that only dynamic compressors can fit the process conditions because of their high volumetric flow rate and low differential pressure. It is also possible to realize that several blowers in parallel are needed in order to achieve the total volumetric flow rate and therefore data for the maximum actual capacity of blowers is needed. From Figure 3 and considering the limitations for these data [25]:

Centrifugal Compressors

Axial (inline) centrifugal gas compressor with motor driver. Excludes intercoolers and knock-out drums.

	<i>Design Basis</i>	<i>Simulation Conditions</i>
Year	1st Quarter 1998 Dollars	Nov 2011 (last CEPCI available)
Material	Carbon Steel	Carbon Steel
Inlet Temperature	68 F	122 F
Inlet Pressures	14.7/ 14.7/ 190 psia	15.95 psia
Pressure Ratios	3:1/ 10:1/ 10:1	1.5:1
Molecular Weight	29	30.03
Specific Heat Ratio	1.4	1.38

Cost estimation:

With exception to the higher inlet temperature and the low pressure ratio suggesting that no multiple stages are required, this equipment seems to fit all the process conditions to calculate accurately the cost. In the case of this simulation, the installed cost was taken from Table 2, source for Figure 4 and finally these costs will be brought to present according to the last Chemical Engineering Plant Cost Indexes available, shown in Table 3.

**Centrifugal Compressor
Purchased Equipment Cost**

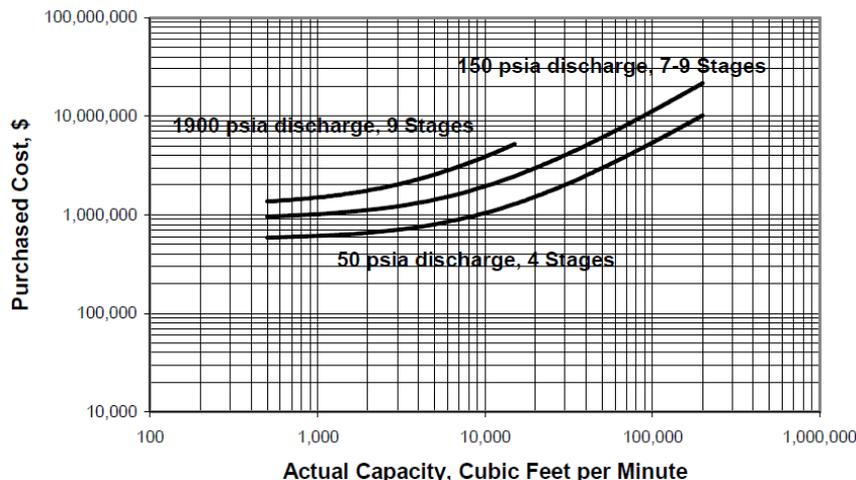


Figure 4: Cost plot for centrifugal compressors [25]

Table 2: Data for constructing Figure 4. [25]

Actual Capacity (cfm)	Purchased Equipment Cost (USD)	Installed Cost (USD)
500	595,400	702,700
1,000	626,400	749,300
5,000	719,700	907,100
10,000	1,114,800	1,339,000
50,000	2,699,800	3,247,700
100,000	5,275,800	6,142,000
150,000	8,722,600	9,735,100
200,000	9,627,600	10,980,400

Table 3: Chemical Engineering Plant Cost Indexes CEPCI [26]

Year	CE Index
1998	389.5
2001	394.3
2004	444.2
2005	468.2
Nov 2011	550.8

From Table 2, the maximum volumetric flow rate available for costing is:

$$Q_{\max} = 200,000 \text{ ft}^3/\text{min} = 339,802 \text{ m}^3/\text{h}$$

Therefore, the number of compressors in parallel required is given by:

$$\begin{aligned} \text{Number of compressors needed} &= Q_{\text{Act}} / Q_{\text{max}} = 1,350,182 / 339,802 \\ &= 3.973 \sim 4 \text{ Compressors} \end{aligned}$$

$$\begin{aligned} \text{Actual Number of compressors} &= \text{Compressors needed} + 1 \text{ (spare equipment)} \\ \text{Actual number of compressors} &= 5 \text{ Compressors} \end{aligned}$$

$$\begin{aligned} \text{Flow rate per compressor} &= Q_{\text{Act}} / \# \text{Compressors} \\ &= 1,350,182 / 4 \\ \text{Flow rate per compressor} &= 337,545 \text{ m}^3/\text{h} = 198,672 \text{ cfm} \end{aligned}$$

$$\text{Cost per installed compressor} = 10,947,320 \text{ USD (interpolated from Table 2)}$$

$$\text{Total cost of compressors} = 10.95 \text{ million USD} * 5 = 54.74 \text{ million USD (1998)}$$

$$\text{Total cost to present} = 54.74 \text{ m USD} * (550.8/389.5) * (1\text{£}/1.571\text{USD})$$

$$\text{Total Capital Cost} = 52.85 \text{ million £}$$

It is also possible to estimate and input into the simulation the polytropic efficiency for a compressor with a determined inlet flow rate by reading the y-axis in Figure 5 as follows,

$$\text{Suction flow rate} = 1,739,485 \text{ m}^3/\text{h} \text{ (from Aspen HYSYS simulation)}$$

$$\text{Suction flow rate} = 255,956 \text{ acfm}$$

$$\text{Actual Polytropic Efficiency: } 0.8257 \text{ (read from Fig. 5)}$$

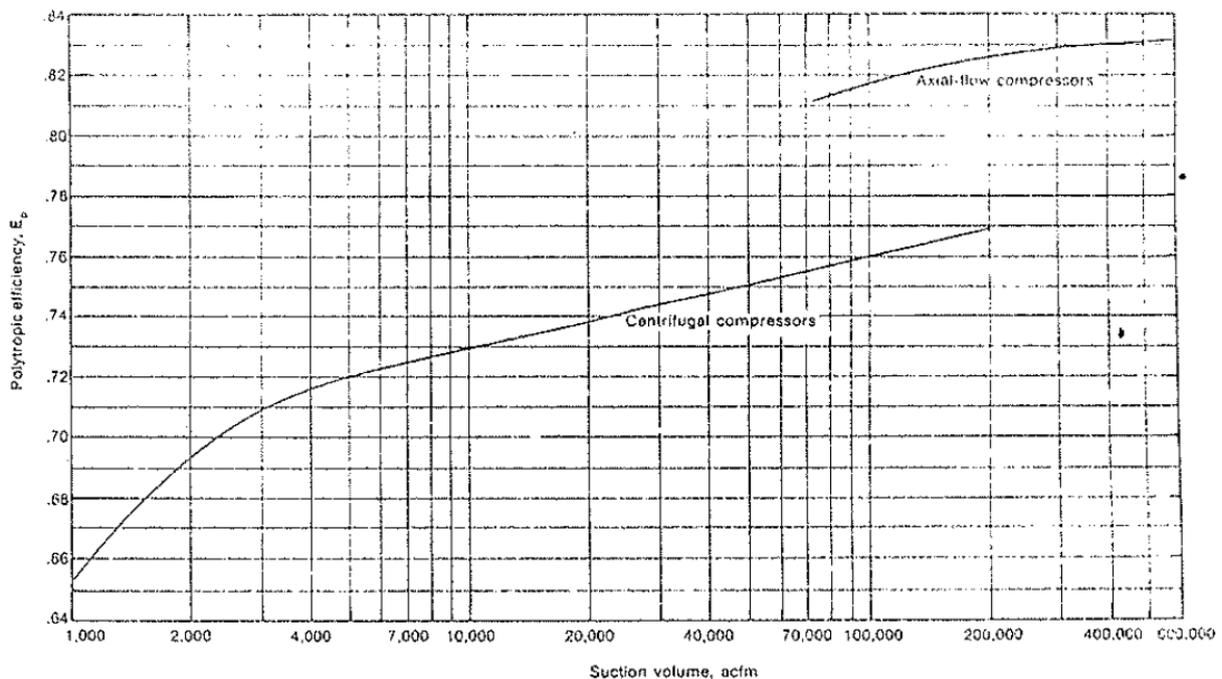


Figure 5: Approximate polytropic efficiencies for centrifugal and axial-flow compressors [24]

Taking the power requirements for the compressor from the simulation in Aspen HYSYS and applying the data from literature:

$$\text{Power price: } 0.05 \text{ \$/kW.h [27]}$$

$$\text{Power Consumption: } 25,296 \text{ kW (from Aspen HYSYS simulation)}$$

$$\text{CO}_2 \text{ Captured} = 445.56 \text{ tons/h (from Aspen HYSYS simulation)}$$

$$\text{Power Cost} = 0.05 \text{ USD/kW.h} * 25,296 \text{ kW} * (1\text{£}/1.571\text{USD}) * (550.8/468.2)$$

$$\text{Power Cost} = 947.13 \text{ £/h}$$

$$\text{Power Cost} = 2.13 \text{ £/ton CO}_2 \text{ Captured}$$

4. Pumps

These are necessary in the process flow sheet to move the absorbent solution through the pipelines and other major pieces of equipment where high pressure drops are present, such as heat exchangers and filters. Tall

towers do not offer losses due to friction but for static head that has to be overcome in order to distribute the liquid at the top of the equipment.

The most common used pumps are centrifugal and these can be selected according to the flow rate and the total head according to Figure 6

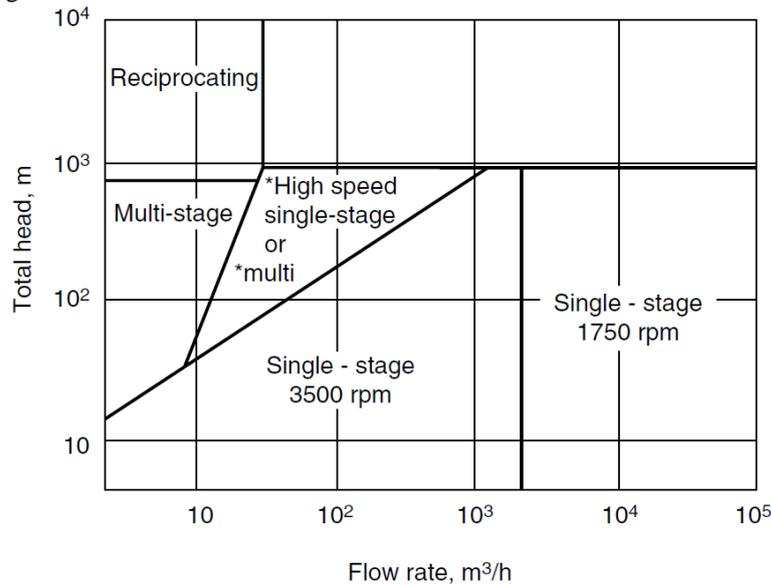


Figure 6: Guideline for Centrifugal pump selection [18] pp. 200.

The materials for pumps construction and other major pieces of equipment have to be resistant to corrosiveness and some authors [28] have suggested that stainless steel is suitable for rich amine streams (high CO₂ content, velocities in pipes between 5 – 8 ft/s) and carbon steel can be used for lean amine streams (low CO₂ content, maximum velocity in pipes 3 ft/s), in addition, the velocity of the fluid must be considered for amine solutions flowing through pipes to avoid further damage for erosion and corrosion effects. Thus, the pump design is intrinsically related to the pipelines design not only for accounting for friction losses but for materials of construction, velocities and heights of interconnected pieces of equipment as well.

Recycled Amine pump:

Taking the flow rate from the simulation,

$$Q = 16,223 \text{ m}^3/\text{h}$$

Despite the CO₂ load (and implicitly the corrosiveness) of this steam is the lowest in all the flow sheet, this point of the process is the one with the greatest erosion effects due to the stream solids content because it includes the microfiltration equipment inlet (where solids are collected and removed from the stream) so stainless steel piping will be used in this and every other point of flow of amine in the system. In order to keep the velocity in the range suggested by the literature as discussed before, the area for the flow and therefore the pipe diameter should be:

$$v = 7 \text{ ft/s (2.134 m/s) Velocity chosen}$$

$$A_{\text{flow}} = Q / v = (16.223 \text{ m}^3/\text{h}) / (2.134 \text{ m/s}) = \pi \cdot D^2/4$$

$$D = 0.82 \text{ m} \sim 32.3 \text{ in}$$

Before a centrifugal pump type can be chosen from Figure 6, it is necessary to calculate the total head required and therefore an estimation of pressure drops from pipelines and heights of the different equipment must be done. For this purpose, it is assumed to have a plant as shown in Figure 7 below where the distance between connected equipments was set to 1.5 times the biggest dimension of both in the attachment direction, allowing for maintenance/replacement if necessary.

Table 4: Accessories in Figure 7 for friction loss determination in Lean Amine pump

ACCESSORIES	Equivalent Diameters	BEFORE PUMP		AFTER PUMP	
		#	Eq. Length (m)	#	Eq. Length (m)
Pipelines Length (m)	-	213.5	213.5	109.8	109.8
90° standard long elbow	23	2.0	37.7	2.0	37.7
Tee-entry from leg	60			4.0	196.8
Tee-entry into leg	90	9.0	664.2	12.0	885.5

Union and coupling	2	5.0	8.2	6.0	9.8
Sharp reduction (tank outlet)	25	8.0	164.0		
Sudden expansion (tank inlet)	50			8.0	328.0
TOTAL EQUIVALENT LENGTH			1087.6		1567.6

$$\text{Total length (L)} = 1087.6 \text{ m} + 1567.6 \text{ m} = 2655 \text{ m}$$

To calculate the friction losses it is necessary to apply [18]:

$$\Delta P_f = 8.f.(L/D)(\rho v^2/2)$$

Where,

$$\rho = 1027.9 \text{ kg/m}^3 \text{ (taken from the simulation)}$$

$$\mu = 1.55 \text{ cP (taken from the simulation)}$$

$$\text{Absolute roughness} = 0.046 \text{ mm (taken from Sinnott, 2005 [18], pp 202)}$$

$$\text{Pipe roughness} = \text{Absolute roughness/Diameter} = 4.6 \times 10^{-5} / 0.82 = 5.61 \times 10^{-5}$$

$$\text{Pipe roughness} \sim 0.0001$$

$$Re = \rho v D / \mu = 1,157,597 \sim 1.16 \times 10^6$$

$$f = 0.0016 \text{ (read from Figure 7)}$$

Then,

$$\Delta P_f = 96.97 \text{ kPa}$$

Taking from the simulation the pressure drops for the flow of the solution through the heat exchangers and microfilters,

$$\Delta P_{eq} = \Delta P_{H.E} + \Delta P_{COOLER} + \Delta P_{MICROFILTER}$$

$$\Delta P_{eq} = 65 \text{ kPa} + 50 \text{ kPa} + 335.2 \text{ kPa}$$

$$\Delta P_{eq} = 450.2 \text{ kPa}$$

To calculate the static head necessary to take the fluid from the bottom of the stripper to the top of the absorbers for recycling, it is necessary to check Figure 6 to get:

$$\Delta P_h = \rho g \cdot \Delta h = 1027.9 \text{ kg/m}^3 * 9.81 \text{ kg/m.s}^2 * (26-1) \text{ m}$$

$$\Delta P_h = 252.1 \text{ kPa}$$

The total head to define the type of pump from Fig.6 is defined by,

$$h \text{ (m)} = (\Sigma \Delta P_i) / \rho \cdot g = (96.97 + 450.2 + 252.1) / (1027.9 * 9.81) \text{ m}$$

$$h = 79.27 \text{ m}$$

And the number of pumps required from the process flow sheet is 3, so the flow per pump is:

$$Q_p = Q / 3 = 4056 \text{ m}^3/\text{h}$$

$$\text{Type of Pump: Single Stage (1750 rpm)}$$

It must be noticed that the number of pumps do not affect the calculations made for the pressure drop based on the initial volumetric flow rate because the flow is only divided to feed the microfilters and it is converged into one stream after the microfiltration systems. To calculate the specific speed commonly used by pump manufacturers, and define the type of impeller:

$$N'_s = 1.73 \times 10^4 \frac{NQ^{1/2}}{(gh)^{3/4}}$$

Where,

$$N = \text{revolutions per second} \quad (1750 \text{ rpm})$$

$$Q = \text{flow, m}^3/\text{s} \quad (4056 \text{ m}^3/\text{h})$$

$$h = \text{head, m} \quad (79.3 \text{ m})$$

$$g = \text{gravitational acceleration m/s}^2 \quad (9.81 \text{ m/s}^2)$$

And using the appropriate units,

$$N'_s = 7274 \text{ rpm}$$

The pump impeller type can be defined from Table 5 as:

Type of Impeller	Specific Speed
Radial	400 - 1000
Mixed Flow	1000 - 7000
Axial	> 7000

Type of Impeller: Axial

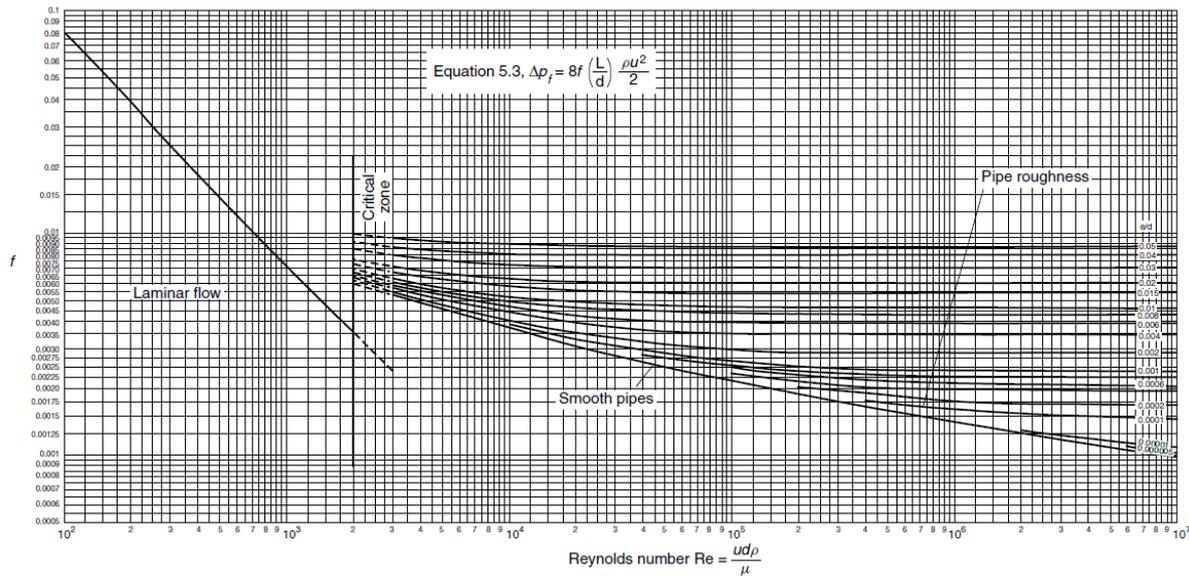


Figure 7: Fanning factor for friction pressure drop calculations [18]

Cost estimation

Single and multistage centrifugal pumps for process or general service when flow/head conditions exceed general service. Split casing not a cartridge or barrel. Includes standard motor driver.

	<i>Design Basis</i>	<i>Simulation Conditions</i>
Year	1st Quarter 1998 Dollars	Nov 2011 (last CEPCI available)
Material	Carbon Steel	Stainless Steel
Design Temperature	120 F	117 F
Design Pressure	150 psig	111 psia
Liquid Specific Gravity	1	1.03
Efficiency	<50 GPM = 60% 50 – 199 GPM = 65% 100 – 500 GPM = 75% > 500 GPM = 82%	17,859 GPM = 82%
Driver Type	Standard motor	
Seal Type	Single mechanical seal	

With exception to the material of construction which can be corrected with an appropriate factor, this description fits all the process conditions to calculate accurately the cost. The installed cost was taken from Table 6. with extrapolation and using the 0.6 rule, and finally these costs will be brought to present according to the last Chemical Engineering Plant Cost Indexes available, shown previously in Table 3.

Table 6: Costing data for centrifugal pumps [25]

Actual Capacity (gpm)	Purchased Equipment Cost (USD)	Installed Cost (USD)
3,000	15,200	58,100
4,000	19,500	72,300
5,000	23,800	77,100
6,000	28,400	93,400
7,000	37,800	103,000
8,000	41,300	119,700
9,000	47,300	126,200
10,000	51,200	144,800

From Table 6, the maximum volumetric flow rate available for costing is:

$$Q_{\max} = 10,000 \text{ gpm} = 2271 \text{ m}^3/\text{h}$$

Which is less than the actual flow rate required (4056 m³/h).

Therefore, by using the 0.6 rule for the actual volumetric flow rate per pump:

$$\begin{aligned} \text{Extrapolated Cost} &= \text{Max Cost} * (4056/Q_{\max})^{0.6} \\ \text{Extrapolated Cost} &= 144,800 * (4056/2271)^{0.6} \\ \text{Cost per installed Pump} &= 0.2051 \text{ million USD} \end{aligned}$$

$$\begin{aligned} \text{Actual Number of pumps} &= \text{Pumps needed} + 1 \text{ (spare equipment)} \\ \text{Actual number of pumps} &= 4 \text{ Pumps} \end{aligned}$$

$$\begin{aligned} \text{Total cost of pumps} &= 0.2051 \text{ million USD} * 4 = 0.8204 \text{ million USD (1998)} \\ \text{Total cost to present} &= 0.8204 \text{ m USD} * (550.8/389.5) * (1\text{£}/1.571\text{USD}) \\ \text{Partial Total Capital Cost} &= 0.8202 \text{ million £} \end{aligned}$$

$$\begin{aligned} \text{Total cost corrected by Material} &= \text{Total cost} * \text{Correction factor (S.S. 316)} \\ \text{Total cost corrected by Material} &= 0.8202 \text{ m£} * 1.8 \text{ [25] pp. 45} \\ \text{Total Capital Cost} &= 1.426 \text{ million £} \end{aligned}$$

Taking the power requirements for the pumping process from the simulation in Aspen HYSYS and applying the data from literature:

$$\begin{aligned} \text{Power price} &= 0.05 \text{ \$/kW.h [27]} \\ \text{Power Consumption} &= 4,002 \text{ kW (from Aspen HYSYS simulation)} \\ \text{CO}_2 \text{ Captured} &= 445.56 \text{ tons/h (from Aspen HYSYS simulation)} \end{aligned}$$

$$\begin{aligned} \text{Power Cost} &= 0.05 \text{ USD/kW.h} * 4,002 \text{ kW} * (1\text{£}/1.571\text{USD}) * (550.8/468.2) \\ \text{Power Cost} &= 149.84 \text{ £/h} \\ \text{Power Cost} &= 0.336 \text{ £/ton CO}_2 \text{ Captured} \end{aligned}$$

5. Microfilter Design

This step is necessary to remove the solids coming from the flue gas that were scrubbed during the absorption and those formed during the degradation process of the amine. This equipment is suitable for treatments with amines at temperatures over those present in the simulation and therefore complies with the necessary specifications to fit the purpose.

The flow rate across the filter will be the recycled stream with the following characteristics (taken from the simulation in Aspen-HYSYS):

$$\begin{aligned} \text{Total Q} &= 16,449 \text{ m}^3/\text{h} \text{ (72430 gpm / 104.3 mgd / 4345.8 kgal/h)} \\ \text{Temperature} &= 47 \text{ °C} \\ \mu &= 1.544 \text{ cP} \\ \text{Specific Gravity (SG)} &= 1.014 \end{aligned}$$

Using HDC II filter,

$$\begin{aligned} Q_{\max} &= 30500 \text{ gpm} \\ \text{Housing } \Delta P &= 0.3 * \text{SG [bar]} - \text{ Pall, 19 HDC II Medium 80 in length cartridges} \\ \text{Cartridges } \Delta P &= 0.12 * \mu \text{ [mbar/(m}^3/\text{h)]} - \text{ Pall, HDC II Medium 80 in length} \\ \text{Maximum Pressure Drop Allowed} &= 3.44 \text{ bar @ } 82 \text{ °C} \end{aligned}$$

Then,

$$\begin{aligned} \text{Number of Microfilters Required} &= \text{Total Q} / Q_{\max} \\ \text{Number of Microfilters Required} &= 2.37 \sim 3 \end{aligned}$$

$$\begin{aligned} \text{Actual Number of Microfilters} &= \text{Number of Microfilter Required} + 1 \text{ (Spare Equipment)} \\ \text{Actual Number of Microfilters} &= 4 \text{ systems} \end{aligned}$$

$$\begin{aligned} \text{Number of cartridges} &= 19 \text{ (cartridges/housing)} * 4 \text{ systems} = 76 \text{ Cartridges} \\ \Delta P \text{ Housing} &= 30.41 \text{ kPa} \\ \Delta P \text{ Cartridges} &= 304.8 \text{ kPa} \end{aligned}$$

$$\text{Total } \Delta P = 335.2 \text{ kPa (3.35 bar } \rightarrow \text{ ok!)}$$

Cost estimation:

According to literature [29], cost estimation can be made depending on the necessary flow rate according to Figure 8:

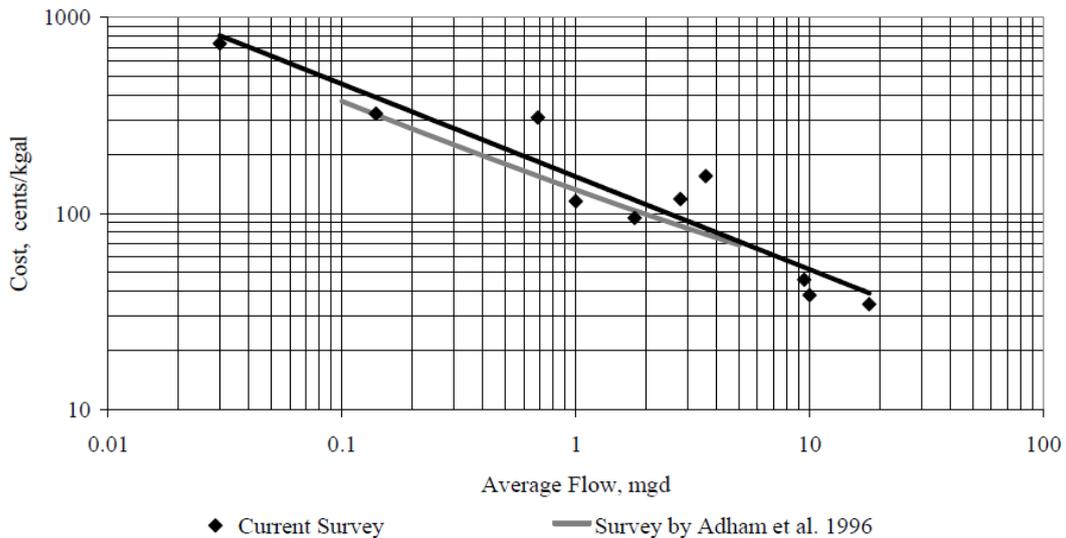


Figure 8: Cost estimation for microfiltration according to flow rate [29] pp. 3.

According to the flow rate (Q) from the simulation and assuming the last available cost according to the curve in Figure 8,

$$\begin{aligned} \text{Cost of microfiltration} &\sim 33 \text{ cents/kgal (read)} \\ \text{Cost of microfiltration} &= 0.33 \text{ USD/kgal} * 4345.8 \text{ kgal/h} * (590.8/394.3) / 1.571 \text{ USD/£} \\ &\text{(CEPCI taken from Table AIII)} \\ \text{Cost of Microfiltration} &= 1368 \text{ £/h} \\ \text{CO}_2 \text{ Captured} &= 445.56 \text{ tons/h (from Aspen HYSYS simulation)} \\ \text{Cost of Microfiltration} &= 3.07 \text{ £/Ton CO}_2 \end{aligned}$$

6. Heat Exchangers

The calculations were exported from variables in Aspen Hysys.

Table 7: Heat exchanger conditions

Shell side conditions			
Parameters		Inlet	Outlet
Shell side flow	kmol/hr	703849.23	703849.23
Temperature	°C	59.09	110.00
Pressure	kPa	548.80	200.00

Tube side conditions			
Parameters		Inlet	Outlet
Tube side flow	kmol/hr	693251.16	693239.26
Temperature	°C	123.72	71.15
Pressure	kPa	215.00	150.00

The type of heat exchanger is U-tube bundle.

The overall heat transfer coefficient, U and the parameter UA where A is the area:

Table 8: U and UA

Overall U	kJ/h m ² °C	4888565.07
UA	kJ/h °C	294871298.39

The area for heat transfer is 60.32 m².
 The shell and tube material is stainless steel.

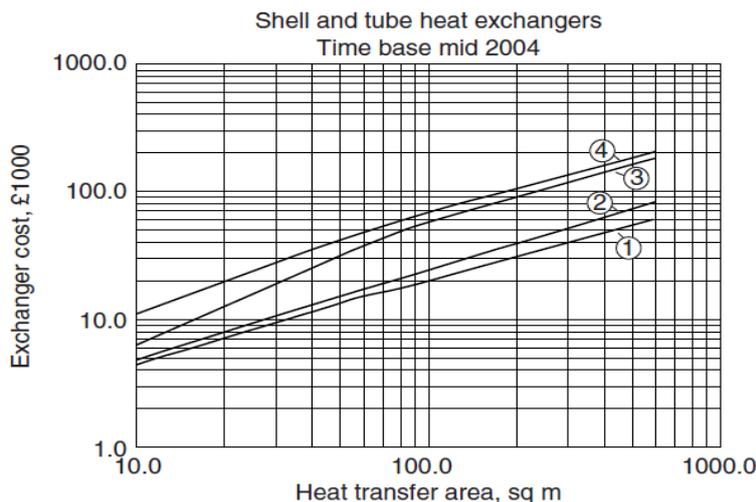


Figure 9: Correlation for heat exchanger cost [18] pp. 254

The purchase cost of the heat exchanger is given by the formula:

$$\text{Purchased cost} = (\text{bare cost from figure}) \times \text{Typefactor} \times \text{Pressure factor}$$

Materials		Pressure factors		Type factors	
Shell	Tubes				
① Carbon steel	Carbon steel	1–10 bar	× 1.0	Floating head	× 1.0
② C.S.	Brass	10–20	× 1.1	Fixed tube sheet	× 0.8
③ C.S.	Stainless steel	20–30	× 1.25	U tube	× 0.85
④ S.S.	S.S.	30–50	× 1.3	Kettle	× 1.3
		50–70	× 1.5		

From the Figure 9, the capital cost is £50,000.

Capital cost as of 2004 is £42,500.

To update the cost, the annual index:

As of 2004	As of Nov 2011
444.2	686.6

Therefore, the cost of the heat exchanger in 2012 is £65,692.3. To convert the cost into £/h, one year with an operational period of 8000 hours is assumed. Therefore, cost per hour in a year is 8.2 £/h.

7. Cooler Design

Table 9: Cooler conditions

Shell side conditions (CO ₂ lean solvent)			
Parameters		Inlet	Outlet
Shell side flow	kmol/hr	893694.71	893694.71
Temperature	°C	25	25
Pressure	kPa	101.3	101.3
Tube side conditions (Cooling water)			
Parameters		Inlet	Outlet
Tube side flow	kmol/hr	693239.26	693239.26
Temperature	°C	71.153	46.86
Pressure	kPa	150	100

The type of heat exchanger is U-tube bundle.

The overall heat transfer coefficient, U and the parameter UA where A is the area:

Overall U	1448758.50
UA	87387053.92

The area for heat transfer is 60.32 m².

The shell and tube material are carbon steel and stainless steel respectively.

From Figure 9, the purchase cost of the cooler is £37,000.

The cost as of 2004 is £31,450. The updated cost of the cooler as of 2012 is £48,612.30. To convert the cost into £/h, one year with an operational period of 8000 hours is assumed. Therefore, cost per hour in a year is 6.1 £/h.

8. Stripper Condenser Design

The condenser operates at full reflux with a negligible pressure drop.

Table 10: Flows involving condenser

Flow into condenser	kg/h	960514.75
Flow out of condenser	kg/h	454175.65
Reflux	kg/h	506339.10

The heat transferred, Q is 1.308E+09 kJ/h.

The UA parameter is exported from Aspen Hysys and was found to be 360000 kJ/ °C h. The estimated U was found from Coulson & Richardson's Chemical Engineering and Design, from the following table:

Table 11: Estimated U values [18]

Shell and tube exchangers		
Hot fluid	Cold fluid	U (W/m ² °C)
<i>Heat exchangers</i>		
Water	Water	800–1500
Organic solvents	Organic solvents	100–300
Light oils	Light oils	100–400
Heavy oils	Heavy oils	50–300
Gases	Gases	10–50
<i>Coolers</i>		
Organic solvents	Water	250–750
Light oils	Water	350–900
Heavy oils	Water	60–300
Gases	Water	20–300
Organic solvents	Brine	150–500
Water	Brine	600–1200
Gases	Brine	15–250
<i>Heaters</i>		
Steam	Water	1500–4000
Steam	Organic solvents	500–1000
Steam	Light oils	300–900
Steam	Heavy oils	60–450
Steam	Gases	30–300
Dowtherm	Heavy oils	50–300
Dowtherm	Gases	20–200
Flue gases	Steam	30–100
Flue	Hydrocarbon vapours	30–100
<i>Condensers</i>		
Aqueous vapours	Water	1000–1500
Organic vapours	Water	700–1000
Organics (some non-condensables)	Water	500–700
Vacuum condensers	Water	200–500
<i>Vaporisers</i>		
Steam	Aqueous solutions	1000–1500
Steam	Light organics	900–1200
Steam	Heavy organics	600–900

The overall heat transfer coefficient, U was found to be 3060 kJ/h m² °C.

The inlet and outlet temperatures of the condenser are given by:

Condenser Inlet	103.43
Condenser Outlet	38.28

Assuming cooling water as the cooling agent with an inlet temperature of 25 °C and a maximum temperature increase of 20°C, the mass of water required is:

$$m = \frac{Q}{C_p \Delta T}$$

The specific heat capacity of 4.2 kJ/kmol °C was used. Hence, mass of water is 6.92E+06 kg/h. The driving force of the heat exchanger was found to be 30.48°C and therefore, the area for heat transfer was found to be 117.6 m².

The type of heat exchanger is a shell and tube heat exchanger where the material of the shell and tube are carbon steel and stainless steel respectively.

The purchase cost found was and as of 2004, £160,000. The estimated cost in 2012 is £247,300 and cost per hour in a year of 30.91 £/h.

9. Reboiler Design

The type of reboiler is a kettle reboiler.

The heat transferred, Q is $2.87E+09$ kJ/h.

Table 12: Flows involving reboiler

Flow into reboiler	kg/h	17972649.79
Flow out of stripper bottoms	kg/h	16675110.21
Total flow out	kg/h	34647759.99

From Table 11, estimated U is 5400 kJ/h m^2 $^{\circ}C$.

The temperatures involving the boiler are:

Temperature into boiler	$^{\circ}C$	123.07
Temperature in boiler	$^{\circ}C$	123.72

Steam enters the system at 3 bar. The saturation temperature is 133.5 $^{\circ}C$.

The LMTD found was $10.10^{\circ}C$. Therefore, the area for heat transfer is $52,640$ m^2 . The capital cost found as of 2012 is $\pounds 26,440,000$ with a cost/hour in a year of 3305 \pounds/h . This can be improved by using steam at a higher pressure and temperature.

ENERGY REQUIREMENTS

Heating

The only piece of equipment with calorific energy requirement in the process flow diagram is the reboiler in the stripper, the variables associated are (taken from the simulation in Aspen-HYSYS):

$$\text{Reboiler Heat} = 2'870,174 \text{ MJ/h}$$

$$\text{Reboiler Temperature} = 123.7^{\circ}C$$

The easiest way to supply this amount of energy in a reboiler is by using saturated steam and the appropriate temperature should assure a high heat transfer coefficient. The temperature approach assumed for the process will be used to calculate the minimum temperature and therefore pressure for the steam required as follows,

$$\text{Temperature Approach} = 7^{\circ}C \text{ (Assumed)}$$

$$\text{Min. Temperature of steam} = \text{Temperature of reboiler} + \text{Temperature Approach}$$

$$\text{Min. } T_{\text{Steam}} = 123.7^{\circ}C + 7^{\circ}C = 130.7^{\circ}C$$

$$P_{131^{\circ}C}^{\text{sat}} = 279.6 \text{ kPa} \sim 3 \text{ bar (Taken from Aspen-HYSYS)}$$

$$\lambda_{3 \text{ bar}}^{\text{vap}} = 2191.4 \text{ MJ/ton (Taken from Aspen-HYSYS)}$$

$$\text{Steam Flow rate} = \text{Reboiler Heat} / \lambda_{3 \text{ bar}}^{\text{vap}}$$

$$\text{Steam Flow rate} = 1310 \text{ ton/h}$$

Cost Estimation

By using the data found in the literature [27], the cost of this steam will be:

$$\text{Type of Steam} = \text{Low Pressure Steam (L.P. Steam 3 bar)}$$

$$\text{Price of L.P. Steam} = 0.17 \text{ USD/ton (2005)}$$

Now, by using the CECPI in Table 3 and the exchange rate for USD:

$$\text{Cost of Steam} = \text{Steam Flow rate} * \text{Price of Steam} * \text{Correction factors}$$

$$\text{Cost of steam} = 1310 \text{ ton/h} * 0.17 \text{ USD/ton} * (590.8/468.2) / (1.571 \text{ USD/}\pounds)$$

$$\text{Cost of Steam} = 178.8 \text{ }\pounds/h$$

$$(\text{CO}_2 \text{ Captured} = 445.56 \text{ ton/h})$$

$$\text{Cost of Steam} = 0.4013 \text{ }\pounds/\text{ton CO}_2$$

Cooling

There are three streams that need to be cooled down during the process but as it will be seen, only one coolant stream is necessary due to the possibility of heat exchanger nets (HEN) that can be applied and the difference in temperature of the different processes.

a. Cooling for condenser in Stripper

Taking the values of the process variables from the simulation (Aspen-HYSYS) and assuming a water temperature between 25-30°C (25 °C used in this case),

$$\begin{aligned} \text{Heat to remove} &= 1'307,881 \text{ MJ/h} \\ \text{Condenser Temperature} &= 38.4 \text{ °C} \\ \text{Inlet Water Temperature} &= 25 \text{ °C (from cooling tower)} \\ \text{Cp (water)} &= 4.126 \text{ kJ/kg.°C} \\ \text{Temperature Approach} &= 7 \text{ °C} \\ \text{Water Outlet Temperature} &= \text{Condenser Temperature} - \text{Temperature Approach} \\ \text{Water Outlet Temperature} &= 31.4 \text{ °C} \\ \text{Water } \Delta T &= (31.4 - 25) \text{ °C} \\ \text{Heat} &= \text{Flow rate} * \text{Cp} * \Delta T \\ \text{Water Flow rate} &= \text{Heat} / (\text{Cp} * \Delta T) \\ \text{Water Flow rate} &= 49,873 \text{ ton/h} \end{aligned}$$

b. Cooling for recycled DEA stream

Once again, the variables in the process are (from Aspen-HYSYS simulation),

$$\begin{aligned} \text{Heat to remove} &= 1'540,199 \text{ MJ/h} \\ \text{DEA inlet Temperature} &= 71.15 \text{ °C} \\ \text{DEA outlet Temperature} &= 46.9 \text{ °C} \\ \text{Inlet Water Temperature} &= 31.4 \text{ °C (from condenser)} \\ \text{Cp (water)} &= 4.126 \text{ kJ/kg.°C} \\ \text{Temperature} &= \text{DEA outlet Temperature} - \text{Temperature Approach} \\ \text{Water Outlet Temperature} &\sim 39.9 \text{ °C} \\ \text{Water } \Delta T &= (39.9 - 31.4) \text{ °C} \\ \text{Heat} &= \text{Flow rate} * \text{Cp} * \Delta T \\ \text{Water Flow rate} &= \text{Heat} / (\text{Cp} * \Delta T) \\ \text{Water Flow rate} &= 43,916 \text{ ton/h} \end{aligned}$$

c. Cooling for inter-stage Compression for carbon storage

The heat to be removed to improve the compression efficiency is (from Aspen-HYSYS simulation),

$$\begin{aligned} \text{Heat to remove} &= 141,604 \text{ MJ/h} \\ \text{Inlet Water Temperature} &= 31.4 \text{ °C (from condenser)} \\ \text{Cp (water)} &= 4.126 \text{ kJ/kg.°C} \\ \text{Water flow rate still available} &= \text{Water for condenser} - \text{Water for DEA cooling} \\ \text{Water flow rate still available} &= (49,873 - 43,916) \text{ ton/h} = 5957 \text{ ton/h} \\ \Delta T &= \text{Heat} / (\text{Flow rate} * \text{Cp}) = 5.8 \text{ °C} \end{aligned}$$

$$\text{Water outlet Temperature} = \text{Water Inlet Temperature} + \Delta T = (31.4 + 5.8) \text{ °C}$$

$$\text{Water outlet inter-stage temperature} = 37.2 \text{ °C} \rightarrow \textit{(Suitable!)}$$

As seen in these calculations, the water from the condenser is enough for both DEA cooling and CO₂ compression inter-stage cooling, so the cost can be estimated as:

$$\text{Price of Water from Cooling Towers} = 0.015 \text{ £/ton (2004)}$$

$$\text{Water Flow rate required} = 49,873 \text{ ton/h}$$

$$\text{Cost of Cooling water from Towers} = 995 \text{ £/h (Nov 2011)}$$

$$\text{Cost of Cooling water from Towers} = 2.23 \text{ £/ton CO}$$

CAPITAL AND OPERATING COST CALCULATIONS:

Total Capital Investment:

$$\text{Total Capital investment (TCI)} = \text{Fixed capital investment (FCI)} + \text{Working capital (WC)}$$

$$(\text{TCI}) = \text{FCI} + 10\% \text{ TCI}$$

$$(\text{FCI}) = \text{Total equipment cost (TEC)} + \text{Direct cost (DC)} + \text{Indirect cost (IC)}$$

$$= \text{£}372,437,287.29$$

$$\text{Total Capital investment (TCI)} = \text{£}372,437,287.29 + 10\% \text{ of TCI} = \text{£}413,819,208.1053$$

$$\text{Adding the initial raw material cost (DEA and water): (TCI)} = \text{£}417,974,468.1053$$

Total Product Cost:

$$\text{Total product cost (TPC)} = \text{Manufacturing cost (MC)} + \text{General expenses (GE)}$$

$$(\text{TPC}) = (\text{MC}) + (9\% \text{ TPC})$$

$$(\text{MC}) = \text{Direct product cost (DPC)} + \text{Fixed charges (FC)} + \text{Plant overhead cost (POC)}$$

$$= (\text{Direct product cost (DPC)} + \text{Raw material cost}) + \text{Fixed charges (FC)} + (5\% \text{ TPC})$$

$$\text{£}111,512,670.4687$$

Total Annual Cost:

Total cost per year = Total product cost + Energy cost + Abatement cost + Geological sequestration cost = £147,258,623.7395

Total cost per year:

The capital recovery (equivalent annual cost for the total capital investment):

$$A = P \times \left[\frac{i(1+i)^n}{(1+i)^n - 1} \right]$$

Where P = £417,974,468.1053, i = 15% and n = 25 years

A = £55,957,911.88

Total cost per year = £196,987,006.8782

- The working capital includes the amount of money required for one month production (raw material, monthly cash on hand, account payable and other costs), it ranges from 10-20% of TCI [30]. It is assumed as 10% of the total capital investment.
- The plant overhead cost include the expenditure required for routine plant services it ranges from 5-15% of TPC [30]. It is assumed as 10% of the total capital investment.

Tables below summarize the percentages and costs used in the above calculations

Table 13: Total equipment cost

Equipment	Total cost
Absorber	4.11E+07
Stripping	5.45E+06
Compressor	8.78E+07
Heat exchanger	1.31E+05
Microfiltration	1.03E+07
Pump	2.55E+06
Sum	£152,096,904.00

Table 14: Breakdown of direct cost

Direct cost		Assumed	Comments and (Justifications)
% of fixed capital investment			
Purchased equip.	15-40	0	All equipment including spare parts, modifications during start-up. (Calculated separately in the table above)
% of purchased equipment cost			
Installation	25-55	0	Installation of all equipment with supports, insulation, paint (Included in equipment cost)
Instrumentation & control	8-50	10	Purchase, installation, calibration, computer tie-in (high control is required mainly on the columns)
Piping	10-80	20	Process piping, pipe hangers, fittings, valves and insulation
Electrical	10-40	20	Electrical equipment -switches, motors, conduit, wire etc. and electrical materials
Building	10-70	20	Process buildings - Auxiliary buildings - Maintenance shops - Building services
Service facilities & Yard improvement	40-100	40	Utilities – Facilities - Nonprocess equipment - Distribution and packaging, Yard: Site development
Land	4-8	4	Surveys and fees and property cost
Sum	–	114	

Table 15: Breakdown of indirect cost

Indirect cost		Assumed	Comments and (Justifications)
% of direct cost			
Eng. & supervision	5-30	5	Engineering costs and engineering supervision and inspection
% of fixed capital investment			
Legal expenses	1-3	2	Preparing and submission of forms required by regulatory agencies, contract negotiations
construction expenses	10-20	10	Construction, operation, tools and equipment, supervision, accounting, Safety, medical, Permits, field tests, special licenses, Taxes, insurance,

			interest
Contractor fee	10-20	5	(assume 5% because it is a large capital investment project)
Contingency	5-15	10	(High risk project)
Sum	-	27	

Table 16: Breakdown of direct production cost

Cost	%	Assumed	Comments and (Justifications)
Operating labour	10-20 of total product cost	12	Skilled and unskilled labour
Laboratory charges	10-20 of operating labour	1.8	Cost of laboratory tests and quality control. It depends on labour involved in these tests
Direct supervisory & clerical labour	10-20 of operating labour	1.8	For Direct supervisory which related to the operating labour
Utilities	10-20 of total product cost	5	Electricity, fuel, cooling and heating agent, waste treatment and disposal, process water
Patents and royalties	0-6 of total Product cost	2	Payment for the use of patent rights or royalty
SUM (1)	-	22.6	
Main & repair	2-10 of fixed capital investment	4	Annual cost for repair equipment (Usually known as 4% in capture plants)
Oper. Supp.	0.5-1 of fixed capital investment	0.75	
SUM (2)	-	5.75	

Table 17: Breakdown of fixed charges

Cost	% of fixed capital investment	Assumed	Comments and (Justifications)
Machine depreciation	10	10	
Building depreciation	2-3	2	Compromise the amount of money that must be paid back from the initial investment. Better to be calculated precisely.
local taxes	1-4	2	Depend on the regional laws and the locality (area population)
Insurance	0.4-1	0.7	Depend on the process and the available protection facilities
SUM	-	14.7	

Table 18: Breakdown of general expenses

Cost	% of total product cost	Assumed	Comments and (Justifications)
Administrative	2-5	3	executive salaries, clerical wages, computer support, legal fee, office supplies and communications
Distribution and marketing	2-20	2	costs for sale office, sales people, shipping and advertising
Research and development	5	5	Developing of new methods and all cost related to personnel, equipment, supplies etc.
SUM	-	10	

Table 19: Energy cost

Energy cost	
Material	Cost/year
Feed compressor	7577040
Storage compressor	1.61E+07
Stripper feed pump	706880
Microfiltration pump	1198720
makeup pump	2.40E+00
SUM	£25,547,842.40

Table 20: Raw material cost

Material Cost	
Material	Cost/year
DEA	849,600.00
cooling agent	7,960,000.00
steam	1,424,000.00
Makeup water	416,000.00
SUM	£10,649,600.00

Table 21: Abatement cost

Abatement cost		
Material	Amount	Cost
discharge cost	£1.2/hr	£9,600.00
geological sequestration cost	£3.18/ton CO ₂	£11,356,620.40

Net energy calculation:

Energy Cost = £25,547,842.40/year

Energy required = Total energy cost (£/year) / energy cost per (MW.year)

Energy cost per (MW.year) = £840,000.00/(MW.year)

Energy required = 30.41 MW

Net energy produced from the plant at 90% capacity = 500/0.9 = 450 MW

Net energy produced after CCS plant = 450 – 30.41 ≈ 420 MW