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Modelling and Analysis of the CO_2 Post-Combustion Capture Process.

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ABSTRACT

This project is all about carbon capture and storage.

The project aims at setting-up a flow-sheet model of the post-combustion capture of carbon dioxide in order to assess the energy consumption and costs of the process. This should also demonstrate how effective the process is on reducing carbon dioxide in the atmosphere.

In addition to this, an exploration of the other capture techniques was conducted in order to be able to see which of the techniques is more viable and feasible economically.

The modelling of the CO_2 capture process that contains a variety of different unit operators was carried out in PRO-II. The Amine thermodynamic was used, since it is the one that is suitable for the absorption of carbon dioxide by the solvent made up of a mixture of water (96%) and monoethanol amine (4%).

The results showed that, from a flue gas composed of carbon dioxide (CO_2 ; 14%), nitrogen (N_2 ; 80%) and oxygen (O_2 ; 6%), 65.46% pure CO_2 saturated with water was captured. Thus, the temperature within the absorber varied from 32 to 43.4 °C, with a pressure of 18 Atm; while in the regenerator, temperature moved from 70 to 106.1°C with a 1 Atm pressure. Moreover, it was also seen from the overall heat balance of both the absorber and stripper that the process produced more energy than it consumed. This is essential as such a process is energetically expensive.

Preface

This report entitled **Modelling and Analysis of the CO_2 Post-Combustion Capture Process** is the result of a project started on July 2008 and ended on December 28th 2008.

The report was written on the 10th semester of the Master of Science in Engineering, in the Chemical Engineering field.

This project would never have been a success without the help of my supervisor Tron Solberg, lecturer at Aalborg University Esbjerg who was always present to answer my questions. I would not forget to mention Rolf Hansen lecturer who was of great help.

As a guide to the reader, references are indicated by [x, page y] in the report, where x represents the number of a specific book listed in the reference list and y represents the number of a specific page in the book x. Figures, tables and equations are numbered in succession according to the chapters they appear.

Aalborg University Esbjerg, February 6th 2009

Serge Duval Cheungoue

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1. INTRODUCTION

All over the world it is realized that the use of fossil resources or fossil fuels in the past and currently as well is causing a tremendous problem on the ecosystem, such as climate changes and acid rains. Fossil fuels account for more than 85% of the world's energy which is used, for the electricity production, for transportation and for heating purposes and industrial processes. Throughout all these processes, many billion tonnes of greenhouse gases are released into the atmosphere and carbon dioxide (CO_2) is the one with the greatest impact on climate change; in fact there are large natural flows of CO_2 in and out of the atmosphere, but due to human activities the amount of CO_2 present in the atmosphere has drastically increased by 40% within the past 200 years [1] and even today CO_2 concentrations continue to rise simply because all of the CO_2 produced from combustion of fossil fuels is emitted to the atmosphere.

Other greenhouse gases include methane (CH_4), nitrous oxide (NO_x) and chlorofluorocarbons (CFC_s), which are also known to be damaging the ozone layer.

In order to stop the damaging effect of CO_2 in the atmosphere and the big effect it has on climate change, it has to be absorbed right from the point of source (as it would rather be difficult to absorb the one already accumulate in the atmosphere) and discard or put it somewhere else than just realising it into the atmosphere. Hence, one of the techniques appropriate to this is Carbon Capture and Storage (CCS).

Carbon capture and storage is emerging as a promising potential greenhouse gas mitigation technique in certain circumstances, but studies are still under way to determine its potential in terms of both cost and effectiveness. Basically, CCS involves the separation of CO_2 from flue gases or off-gases coming from large stationary sources such as coal-fired power plants, hydrogen production plants or cement factories plants. The CO_2 is then transported by pipeline to a storage site where it is injected into suitable geological formations such as depleted oil and gas reservoirs, deep saline water, deep unmineable coal seams or injected to be further used to enhance oil recovery.

Additionally, organizations as International Energy Agency (IEA) and Intergovernmental Panel on Climate Change (IPCC) suggest that fossil fuels will continue to be the world's dominant source of energy in the coming decades, as such, CCS technology if it proves economically practicable on a large enough scale, could play a useful role in ensuring that the world's continued reliance on fossil fuels does not further contribute to the accumulation of CO_2 in the atmosphere. Thus, from a purely financial perspective, capturing and storing carbon could worth the expense only if avoided emissions have value. It follows that CCS systems are not likely to be deployed on a large scale in the absence of explicit policy directives that put limits, and hence a price tag, on greenhouse gas emissions.

Moreover, there is a hope that as with other industrial by-products and municipal waste that are no longer discharge into lakes, streams, and atmosphere, CO_2 emissions from fossils fuels can also be put somewhere besides the atmosphere or treated to reduce or eliminate emissions. As for example, sulphur dioxide and nitrous oxide emissions from power plants were found to cause air pollution, thus technologies were developed to remove them from the smoke stacks of power plants.

Additionally, biological pathogens from municipal sewage were damaging the water quality of lakes and rivers, so sewage treatment plants that remove the pathogens were developed. As such, many more examples of these demonstrate that the management of industrial by-products can be improved to reduce or eliminate their environmental impacts. And it is in this regard that many countries and regions around the world are taking initiatives such as regional agreements and unilateral plans in order to reduce their emissions of green house gases (CO_2) more or less by encouraging the use of renewable energy and or by developing new technologies to capture CO_2 from industrial processes and store it somewhere rather than to just release it into the atmosphere. Although these technologies may be costly to be achieved, but through intensive research and efforts, alternatives may be proposed at an acceptable cost

1.1 Problem Analysis

This project is all about carbon capture and storage, and thus will consist of two parts. The first part will outline the various ways or processes that could be used in order to remove CO_2 from the flue gases coming from the most stationary sources of CO_2 such as from power plants, petroleum refineries, gas processing facilities, and cement factories. The second part is a PRO-II simulation of carbon capture from flue gases emitted from a power plant production and this simulation is based on the post-combustion capture of CO_2 .

Basically carbon dioxide capture and storage involves three stages: carbon dioxide capture, transport and storage. And in carbon capture and storage, there may be two ways of absorbing or capturing CO_2 :

CO_2 from all origin could be offset naturally by plants through what is known as photosynthesis, but this process is somehow limited simply because the amount of CO_2 present is by far superior to the number of plants and trees.

The other way is to capture CO_2 directly from the industrial source, then concentrate it into a nearly pure form and store it somewhere. The little disadvantage with this second choice is that it is only directed towards large sources of emissions of CO_2 and does not take into account small sources of CO_2 such as automobiles as these would be difficult to capture directly at the source.

1.2 SOURCES OF CO_2

Though CO_2 is one of the gases in our atmosphere being uniformly distributed over the earth's surface at a concentration of about 330ppm or 0.033%, carbon dioxide emissions mainly arise from a number of different sources, like the combustion of fossil fuels. As a result of the tremendous worldwide consumption of such fossil fuels, the amount of CO_2 in the atmosphere has increased over the past century, now rising at a rate of about 1ppm or 0.001% per year.

For instance the combustion of coal produces 36% of the total as coal is mostly used in large industrial plants, especially for power generation. There are also significant amounts released from transportation, industrial processes such as iron and steel manufacture and cement production.

Another important source is the combustion of natural gas. In terms of energy released, the burning of natural gas only releases about half as much CO_2 as would come from the burning of coal to produce the same amount of energy. Nevertheless, natural gas is a major source of CO_2 and global combustion of natural gas amount to over 22% of the total from all fossil fuels. Natural gas is used in large power stations but also to provide heating in buildings, as well as in some industrial processes such as chemical production and glassmaking.

However, the most important fossil fuel in terms of CO_2 emissions is oil, contributing more than 40% of the total. Oil is particularly important for powering cars and other vehicles, so the bulk of its use is in small engines. World demand for transport continues to increase, almost all of this being fuelled by petrol and diesel oil. [1]

The above is depicted on the following diagram

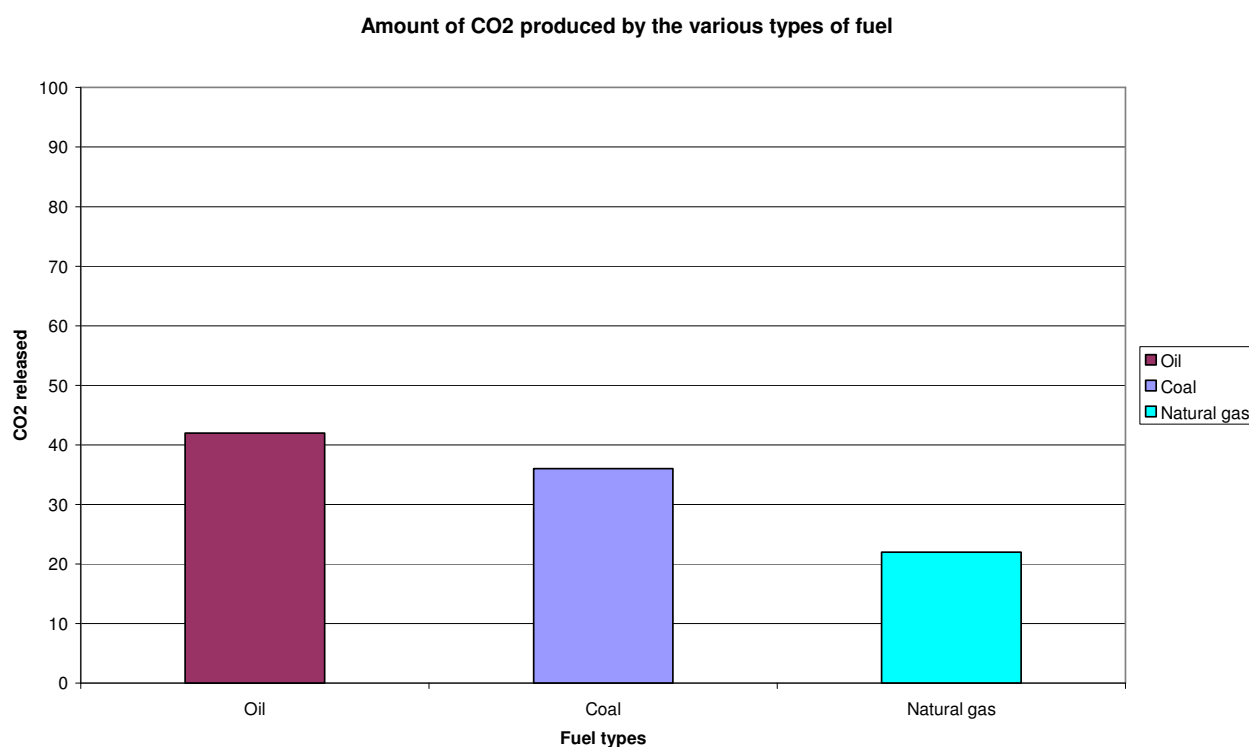


Figure 1.2.1: Amount of CO_2 produced by the various fuel's types.

On the other hand, the diagram shown above does not have the same configuration when it comes to the production of a specific amount of energy. This means that, in the course of the production of say 1 GJ of energy, the amount of carbon dioxide released by coal will be higher to the amount of carbon dioxide released by oil because of the fact that coal contains less energy and has higher carbon content than oil. Therefore more coal will be used than oil for the production of the same amount of energy. Consequently more CO_2 will be released too.

This is illustrated on the diagram shown below.

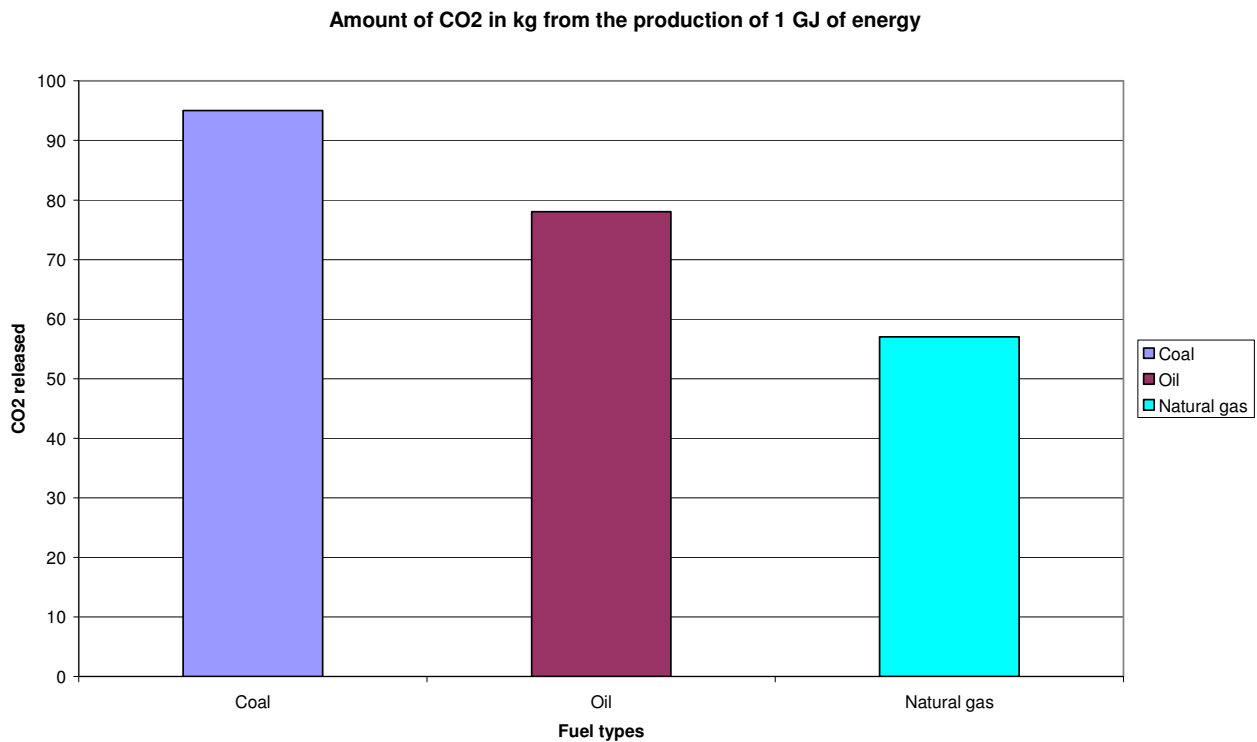


Figure 1.2.2: Amount of CO₂ released from the production of 1 GJ of energy. [1]

Figure 1.2.2 shows how the various fossil fuels emit different amounts of carbon dioxide in producing the same amount of energy. Hence for the production of 1 GJ of energy, 95 kg of CO₂ is released through the use of coal as fuel while 78 kg is released through the use of oil and finally 57 kg via the use of natural gas as fuel.

2. CARBON DIOXIDE CAPTURE

In carbon dioxide capture, the carbon dioxide is separated from emissions and concentrated into a nearly pure form, then compressed into liquid and transported through pipelines or tanks to the location site where it is to be stored.

2.1 Capture Techniques

Basically there are three ways of capturing CO₂ from the flue gas emitted from fossil-fired power plants.

These include:

- Post-combustion capture

- Pre-combustion capture
- Oxy-fuel combustion capture

An overview of the three capture techniques is shown on the following figure.

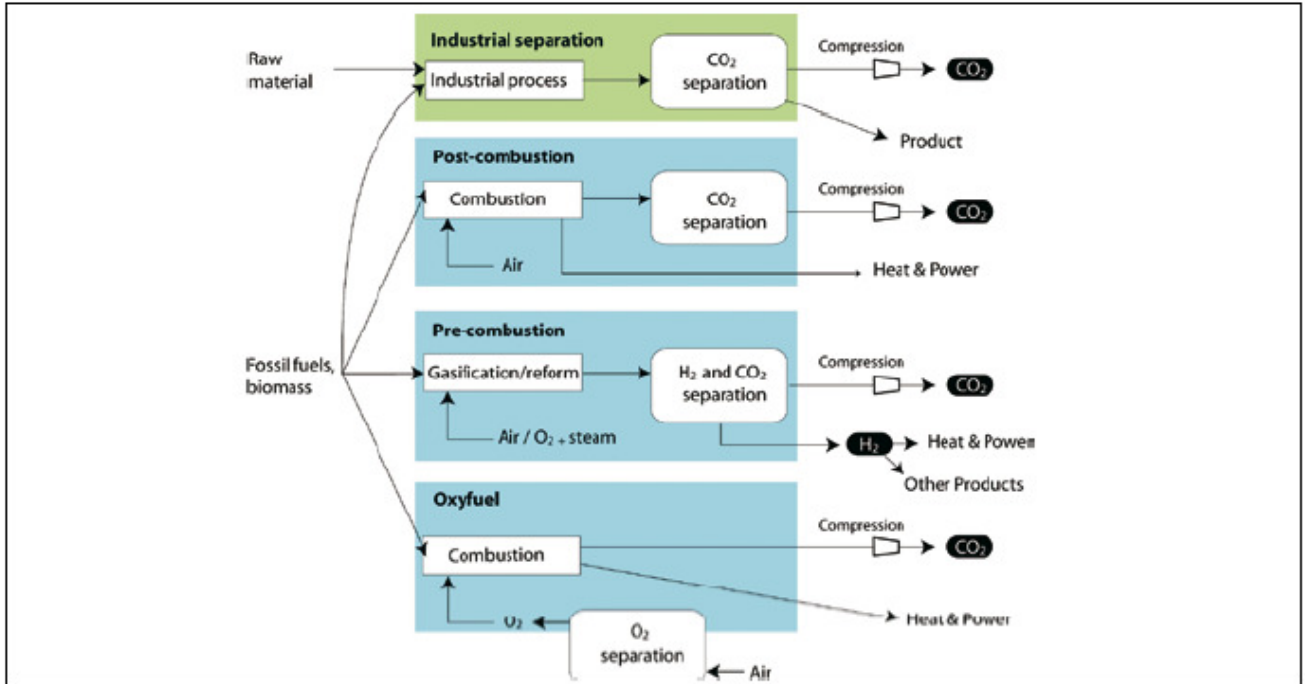


Figure 2.1: Schematic representation of capture systems [4]

Though they all follow different path ways, yet they have the same achievement that is the capture of carbon dioxide.

2.1.1 Post-combustion capture

The post-combustion capture of CO₂ involves the removal of CO₂ from the flue gas of a power plant using chemical solvents, mostly amines. The flue gas is run through a high pressure gas/liquid contactor, as a tower for instance where the washing solvent and the gas have counter current directions (the solvent run downwards and the gas goes upwards) inducing a perfect mixing. This process is somehow similar to distillation where the rate of separation or purity of the product is subject to the size of the tower or contactor. In the course of the mixing, CO₂ partitions into the amine solvent which is then passed to another tower and heated to release nearly pure CO₂. While the CO₂ is compressed for transportation to the storage site, the solvent is recycled back to the first tower.

It should be noted that prior to the capture of carbon dioxide, all solid particles, nitrogen oxides and sulphur dioxide (for a coal-fired plant) are normally removed and based on current practice, 85% to 90% of the CO₂ would be removed through the above-mentioned technique. [1]

The post-combustion capture technique of carbon dioxide has the advantage that existing electrical generation plants could be retrofitted with a post-combustion capture unit and the technology is well established. But as a disadvantage, there is the large energy requirement to regenerate the amines and to compress the carbon dioxide from near-atmospheric pressure to pipelines pressures. For newly constructed plants with post-combustion capture, it may be possible to reduce costs and energy requirements significantly by more efficient energy integration and pre-concentrating the CO_2 before separation. [2]

Basically this process could be represented as shown below.

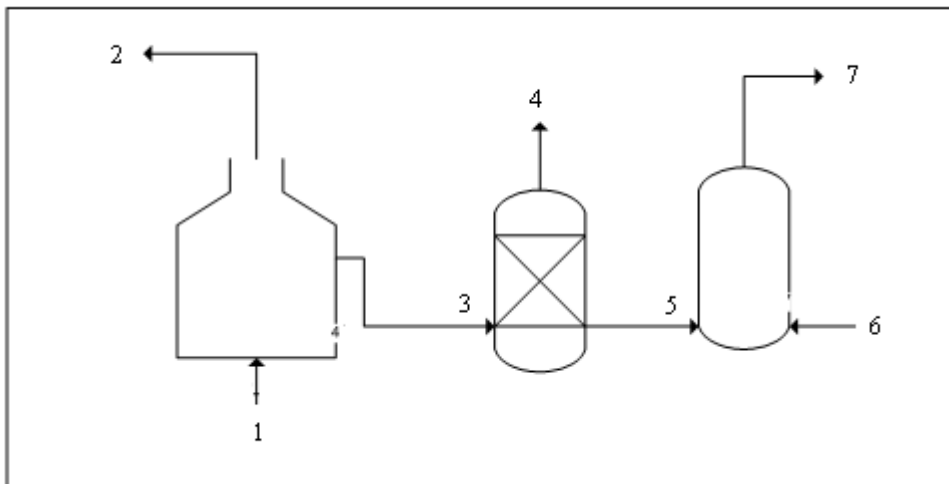


Figure 2.1.1: Post-combustion CO_2 capture using gas fuel.

1. Air and fuel getting in the boiler
2. Rotation energy to be used for electricity production
3. Flue gas sent to the tower for CO_2 absorption
4. Clean flue gas released into the atmosphere
5. Amine solvent rich in CO_2 to be heated for CO_2 desorption
6. Heat supplied to the system
7. CO_2 captured and compressed for storage

As said earlier, for a coal-fired station, there should be a particle and sulphur removal equipment just after the boiler in order to set the flue gas free of particles and sulphur before it is sent to the first tower.

2.1.2 Pre-combustion capture

It involves the conversion of fossil fuel into a gas consisting of hydrogen (H_2) and CO_2 prior to combustion. That is before the combustion, the fossil fuel reacts with steam and air or oxygen to create a gas mixture consisting primarily of CO and H_2 . The CO is then reacted with water to produce CO_2 and more H_2 , which can be used for electricity generation or other purposes, and CO_2 for storage. Actually, the capture of CO_2 here is done through the use of a solvent that physically absorbs CO_2 at an elevated pressure and upon reduction of pressure, the CO_2 is released from the solvent. Thus, after the CO_2 has been separated out, the H_2 is mixed with nitrogen and burnt in a gas turbine to generate electricity.

Advantages for this method include the reduced equipment size, the ability to produce hydrogen, lower compression requirements and lower energy use for the separation process.

Disadvantages include the lack of experience of the electrical utility industry with advanced chemical processing facilities such as the Integrated Gasification Combined Cycle (IGCC) station and lack of widespread application for the purpose of power production.

The gasification station has not yet been accepted as commercially competitive compared to conventional coal-fired power stations. But Gasification is used today to produce ammonia for fertilizers and H_2 for petroleum processing. Although the initial fuel conversion steps of pre-combustion are more elaborate and costly, the higher CO_2 concentrations in the gas stream and the higher pressure make the separation easier as earlier-mentioned.

An attempt of the process representation is shown on figure 2.1.2.

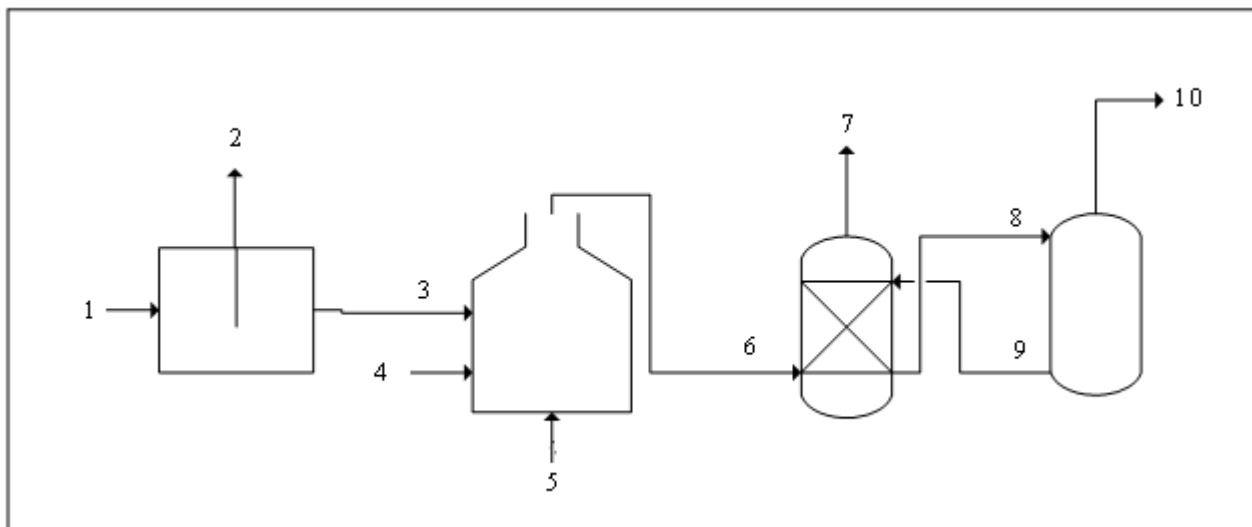


Figure 2.1.2: Pre-combustion CO_2 capture using gas fuel.

1. Air entering the air separator
2. Nitrogen released to the atmosphere
3. Oxygen sent into the gasifier

4. Fuel (gas in this case) sent into the gasifier
5. Steam sent into the gasifier
6. Mixture of CO_2 , steam and H_2 sent into the CO_2 absorber
7. H_2 sent into the gas turbine to be used for electricity production
8. Desorption of CO_2
9. Steam recycled to optimize the separation process
10. CO_2 captured and compressed for storage

It should also be noted here that for a coal-fired plant, particles and sulphur are removed prior to the CO_2 absorption and desorption process.

2.1.3 Oxy-fuel combustion capture

The oxygen fuel combustion capture is one of the promising technologies that utilize oxygen rather than air for the combustion of fossil fuels. It involves changing the combustion conditions so as to avoid the mixing of nitrogen and CO_2 . Thus, nitrogen is kept out of the combustion chamber by removing the oxygen needed for combustion from the air. The advantage is that the only gaseous emissions are water vapour, carbon dioxide, and small amounts of SO_2 . These can be easily separated and CO_2 can then be captured. The disadvantage includes the fact that there are energy and economic constraints or penalties in the production of oxygen as part of the process.

However, burning fossil fuel in pure oxygen produces very high temperatures which would require the use of special or even yet-to-be-developed materials. Instead, in order to moderate the temperatures to levels more typical of conventional boilers, some of the exhaust gases are re-circulated to the combustion chamber.

Additionally, oxygen fuel combustion capture is less attractive with natural gas due to the molecular nature of its main component, which is methane. The molecule of methane consists of one carbon atom and four hydrogen atoms, thus for combustion in pure oxygen, much more of oxygen is needed in order to oxidise the hydrogen than in the case using coal. Therefore, a more fruitful application of oxy-fuel technology may be using it with coal or possibly heavy oil. [1]

A representation of the process is shown on figure 2.1.3 below.

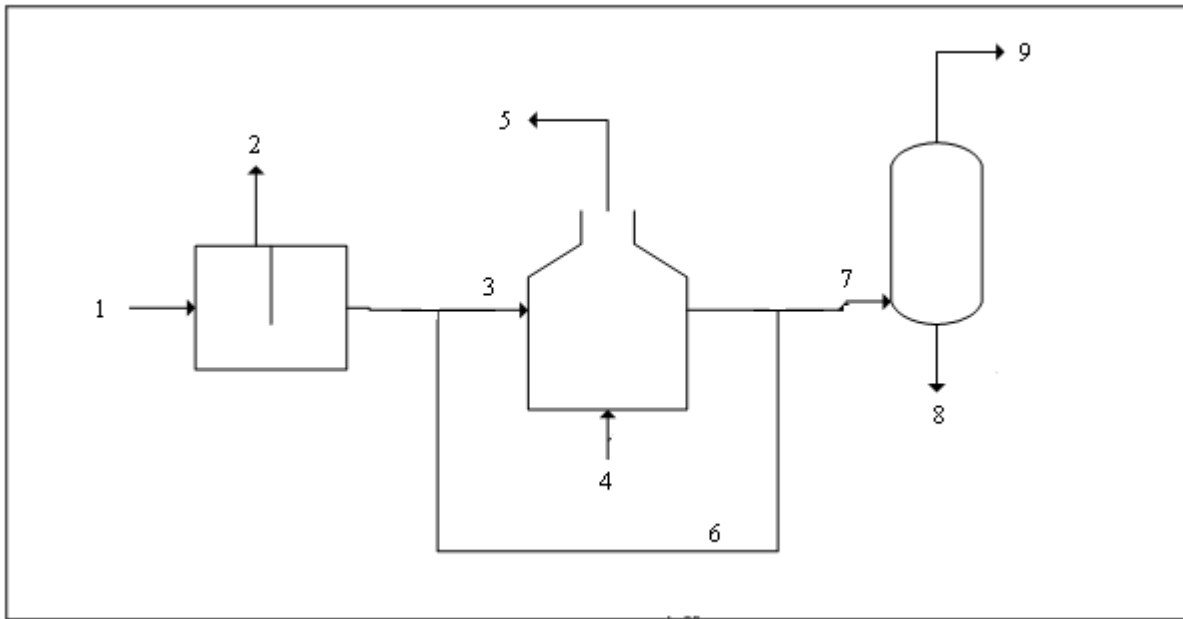


Figure 2.1.3: Oxy-fuel combustion CO_2 capture using gas fuel.

1. Air entering the air separator
2. Nitrogen released to the atmosphere
3. Oxygen entering the boiler
4. Fuel entering the boiler
5. Rotational energy to be used for electricity production
6. Recycled flue gas of carbon dioxide and water vapour
7. Stream of water vapour and CO_2 sent into the condenser
8. Water collected at the bottom
9. CO_2 captured and compressed for storage

As already said for the other two processes, it is necessary to remove all solid particles and sulphur dioxide before the capture of CO_2 in the case of a coal-fired power station. This is normally done for the ease of the process.

2.2 Comparative benefits of the three capturing techniques

The above three capturing techniques comes as a result of the significant scientific research and engineering development that has been invested in all of the approaches over the past decade. Each of the techniques has its advantages and disadvantages as summarized by the table below.

Technology	Advantages	Disadvantages
Post-combustion	<ul style="list-style-type: none"> - Mature technology - Standard retrofit of existing power generation capability - Ability to bypass post-combustion capture allows plant to operate with its full original reliability and provide peaking power in excess of capture capacity 	<ul style="list-style-type: none"> - High energy penalty lowers plant efficiency by 30% - High cost compared to electricity production without CCS
Pre-combustion	<ul style="list-style-type: none"> - Higher CO_2 concentrations lower the costs and energy penalties compared to post-combustion capture - Combine with H_2 production for transportation sector 	<ul style="list-style-type: none"> - Gasification technology is immature for power production - Repowering of existing capacity is needed - Large capital investment needed for repowering
Oxygen-combustion	<ul style="list-style-type: none"> - Minimal post-combustion separation compared to air-fired power plants - Potentially higher generation efficiencies 	<ul style="list-style-type: none"> - High cost of oxygen separation - High cost compared to electricity production without CCS

Table 1: Comparative benefits of the various combustion types

3. Transport and storage options of captured CO_2

Following the capture and compression processes, carbon dioxide needs to be transported for storage in geological formations, the ocean, in mineral carbonates or for use in industrial processes as carbonated soft drinks. Actually, there may be many options for transportation and storage areas as well.

3.1 Transport of carbon dioxide

Generally there are two ways of transporting carbon dioxide. It could be done through the use of pipelines and through the use of ships.

But based on the cost and the amount of CO_2 that could be transported, pipelines at the least expensive means of CO_2 transport over distances up to several hundred kilometres.

Additionally, the transport of oil over many years proves that CO_2 can also be transported safely and reliably providing certain conditions are met. Thus, the gas must be dried before it is introduced into the pipeline in order to avoid corrosion of the steel because dry CO_2 is not corrosive to pipelines, even if the CO_2 contains contaminants.

Shipping of CO_2 , analogous to shipping of liquefied petroleum gases, is economically feasible under specific conditions but is currently carried out on small scale due to limited demand. This situation occurs when the volume to be transported is low relative to the distance to be covered. Here, the CO_2 is compressed and liquefied before being loaded onto the ship.

To finish, CO_2 can also be carried by rail and road tankers, but this option is less attractive for a large scale transport.

3.2 Storage options of CO_2

Many ideas have been proposed so far concerning the storage of the captured carbon dioxide. These include making artificial containers to hold it on land, injecting it into deep ocean or in deep geological formations. It is important to note that, the captured CO_2 can also be used as feed stock in industrial processes.

Considering the making of artificial containers to hold CO_2 on land, it involves the construction of an artificial repository on land where the CO_2 would be refrigerated and pressurised. But the cost of such an installation, not to mention its size, would be such that this approach would not be able to compete with the geological storage of carbon dioxide.

As far as the geological storage of carbon dioxide is concerned, it is known that the earth's surface is made up primarily of two types of rocks:

- Those formed by cooling magma either from volcanic eruptions or from magmatic intrusions far beneath the land surface
- Those formed as thick accumulations of sand, clay, salts, and carbonates over millions of years.

The second types occur primarily in what are termed sedimentary basins. Geographic locations overlying sedimentary basins are best suited for geological storage of CO_2 and fortuitously the majority of CO_2 sources are located in or near to sedimentary basins.

Sedimentary basins often contain many thousands of meters of sediments where the tiny pore spaces in the rocks are filled with salt water (saline formations) where oil and gas reservoirs are found. Sedimentary basins consist of many layers of sand, silt, clay, carbonate, and evaporite (rock formations composed of salt deposited from evaporating water). The sand layers provide storage space for oil, water, and natural gas. The silt, clay, and evaporate layers provide the seal that can trap these fluids underground for periods of millions of years and longer. Geological storage of

CO_2 would take place deep in sedimentary basins trapped below silt and clay layers, much in the same way that oil and natural gas are trapped today. Recall that oil and natural gas come from the process of formation of sedimentary rocks, which result from the compilation of mud, silt and other sediments. That is in the course of the formation of sedimentary rocks, bacterial activity, heat, and pressure change the plant and animal remains into oil and natural gas.

Hence, possible storage formations include oil reservoirs, gas reservoirs, saline formations and even coal beds as indicated on Figure 4.1 below.

On the other hand, storage of CO_2 in deep, onshore or offshore geological formations uses many of the same technologies that has been developed by the oil and gas industry and has been proven to be economically feasible under specific conditions for oil and gas fields. In addition, the combination of CO_2 storage with enhanced oil recovery (EOR) or potentially enhanced coal bed methane recovery (ECBM) could lead to additional revenues from the oil or gas recovery. Indeed, by injecting carbon dioxide under high pressure in oil reservoirs can help push some of the oil that has been left behind toward producing wells, thereby increasing recovery rates while sequestering carbon at the same time. Well-drilling technology, injection technology, computer simulation of storage reservoir performance and monitoring methods from existing applications are being developed further for utilization in the design and operation of geological storage projects. [4]

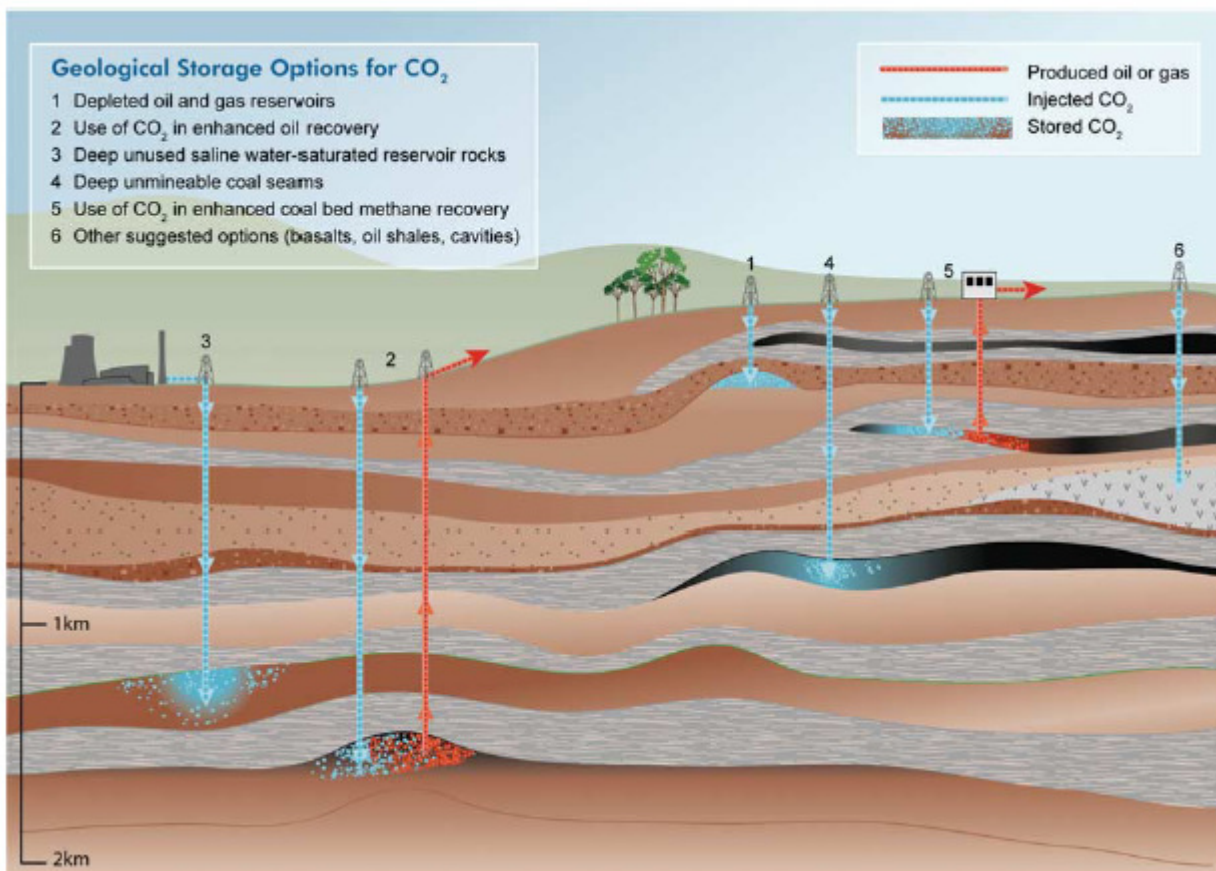


Figure 3.1: Options for geological storage of CO_2 [4]

The presence of an overlying, thick, and continuous layer of silt, clay, or evaporite is the single most important feature of a geologic formation that is suitable for geological storage of CO_2 . These fine-textured rocks physically prevent the upward migration of CO_2 by a combination of viscous and capillary forces. Oil and gas reservoirs are found under such fine-textured rocks and more presence of the oil and gas demonstrates the presence of a suitable reservoir seal. [3]

In saline formations, where the pore space is initially filled with water, after the CO_2 has been underground for hundreds to thousands of years, chemical reactions will dissolve some or all of the CO_2 in the salt water, and eventually some fraction of the CO_2 will be converted to carbonate minerals, thus becoming a part of the rock itself.

Another use of the captured CO_2 could be as feedstock in chemical processes that produce valuable carbon-containing products such as mineral carbonates and carbonated drinks. But usually, the potential for industrial uses of CO_2 is small because the processes that use captured CO_2 as feedstock do not always achieve net lifecycle emission reductions as the amount of captured CO_2 used is much less.

As per the formation of mineral carbonates, although the technology is still immature, it involves basically the reaction of CO_2 with metal oxide bearing materials such as calcium and magnesium to form insoluble carbonates. Such a reaction exists in nature and takes place on a geological time scale.

When CO_2 reacts with metal oxides (indicated here as MO, where M is a divalent metal, for instance, calcium, magnesium or iron) the corresponding carbonate is formed and heat is released according to the following chemical reaction:



Thus considering calcium or magnesium, we have the following equations:



Moreover, it should be noted that such direct gas-solid reactions are too slow to be practical and are only feasible at reasonable pressures for refined, rare materials like the oxides or hydroxides of calcium and magnesium.

3.3. Cost and economic considerations

As mentioned earlier, from a purely financial perspective, capturing and storing carbon is worth the expense only if avoided emissions have value. It follows that CCS systems are not likely to be developed on a large scale in the absence of explicit policy directives that put limits, and hence a price tag on greenhouse gas emissions.

Nowadays, few fully integrated industrial carbon capture and storage projects are currently in operation, but interest is growing around the world. In the fall of 2005, the Intergovernmental Panel

on Climate Change (IPCC) released an exhaustive special report on CCS which, amongst other things, modelled the cost and economic potential of CCS, based in part on early commercial experiences with all the major components of CCS. The report shows that approximately 90% of the total CCS costs relate to the capture of CO_2 , a process that requires additional energy; about 10-40% more in the case of new coal-fired or gas-fired power plants that are retrofitted to capture CO_2 . The costs of transport and storage are less onerous in comparison, and can be minimized by achieving economics of scale and/or by sitting emitting plants close to potential storage sites such as oil and gas reservoirs.

Overall, it is estimated that capturing, transporting and storing CO_2 from a new coal-fired or gas-fired power plant would increase the cost of electricity generated by that plant by between 37% and 91%. This makes CCS a comparatively high-cost method for mitigating carbon emissions, at least for the time being.

4. PRO-II simulation of carbon capture.

PRO-II is a comprehensive computer simulation system targeted for process engineers in the chemical, petroleum, natural gas, solids processing, and polymer industries. It combines the data resources of a large chemical component library and extensive thermodynamic property prediction methods, with the most advanced and flexible unit operations techniques.

It can be summarized as follows:



Figure 4.1: A representation of PRO-II.

Indeed, the input is the specifications we put into PRO-II in order to obtain the interested product or output. Thus, the input language is an extension of the universal language of chemical engineering. Experts systems, extensive input processing, and error checking allow the user to be instantly productive. And this is done through colour display.

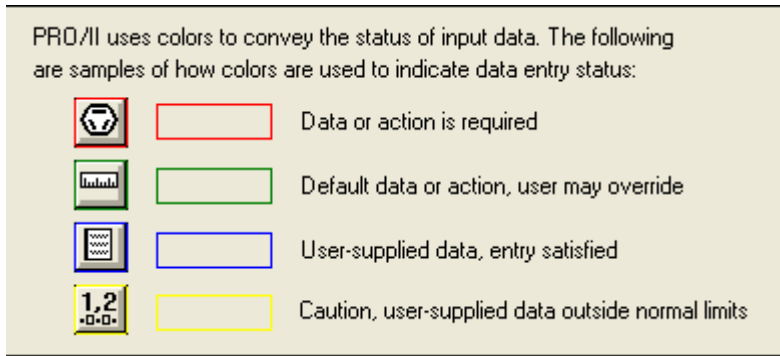


Figure 4.2: Sample colours used in PRO-II. [Simsci-essor]

However, a negative feature of PRO-II is that the simulation has to be saved continuously as one proceed, otherwise at a certain extend it will crash.

4.1 Simulation basics

As stated earlier in problem analysis, the objective here is the simulation in PRO-II of the capture of carbon dioxide from flue gases emitted at a power plant using the amine mono-ethanolamine (MEA) mixed with water in an average of 4% MEA and 96% H_2O .

But the simulation does not include the power plant itself and it is assumed a flue gas composed of carbon dioxide (CO_2 ; 14%), nitrogen (N_2 ; 80%) and oxygen (O_2 ; 6%).

Having in mind that carbon capture is an expensive process as it consumes enormous amounts of energy, we tried as much as possible to reduce the duties of the process by using a calculator to regulate both the exact amount of water and MEA used by the process and a recycling stream of MEA/ H_2O solvent was also included. Additionally, the heat generated by the process was used again within the process so as to minimize any loss of energy.

Before that, one needed to know how a simulation is set, ran and results viewed in PRO-II. Thus in setting up a simulation, we need to supply problem data and this can be done in a number of ways. The colour codes in PRO-II alert us when data is required (as indicated in Figure 4.2 above), marking the pathway towards a completed simulation.

Moreover, it should be noted that data should be entered in a routine, logical manner since some options depend on other options. For example one cannot enter the stream composition or composition-based process specifications before listing the components in the process. Or, one may want to change the input set of units of measure before entering user defined components and streams. All calculations depend on the choice of thermodynamic methods.

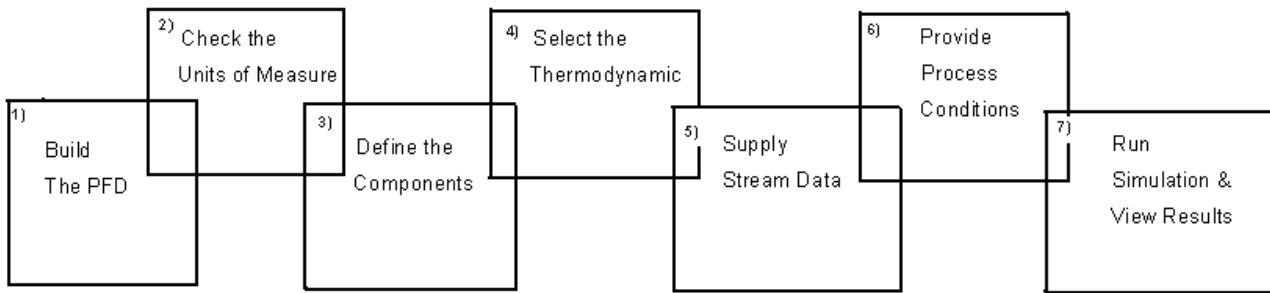


Figure 4.3: Indications on how to develop a PRO-II simulation.

To develop a PRO-II simulation, we should follow the steps shown in Figure 4.3, above as outlined in the following:

1) Build the PFD.

The process flow diagram (PFD) is drawn by simply selecting and positioning the unit operations in the main window. Next, the feed and product streams for each unit are drawn. Often the product streams from one unit are feed streams to other units.

After selecting the unit operations and putting everything together, the process flow diagram of our simulation looks as shown below.

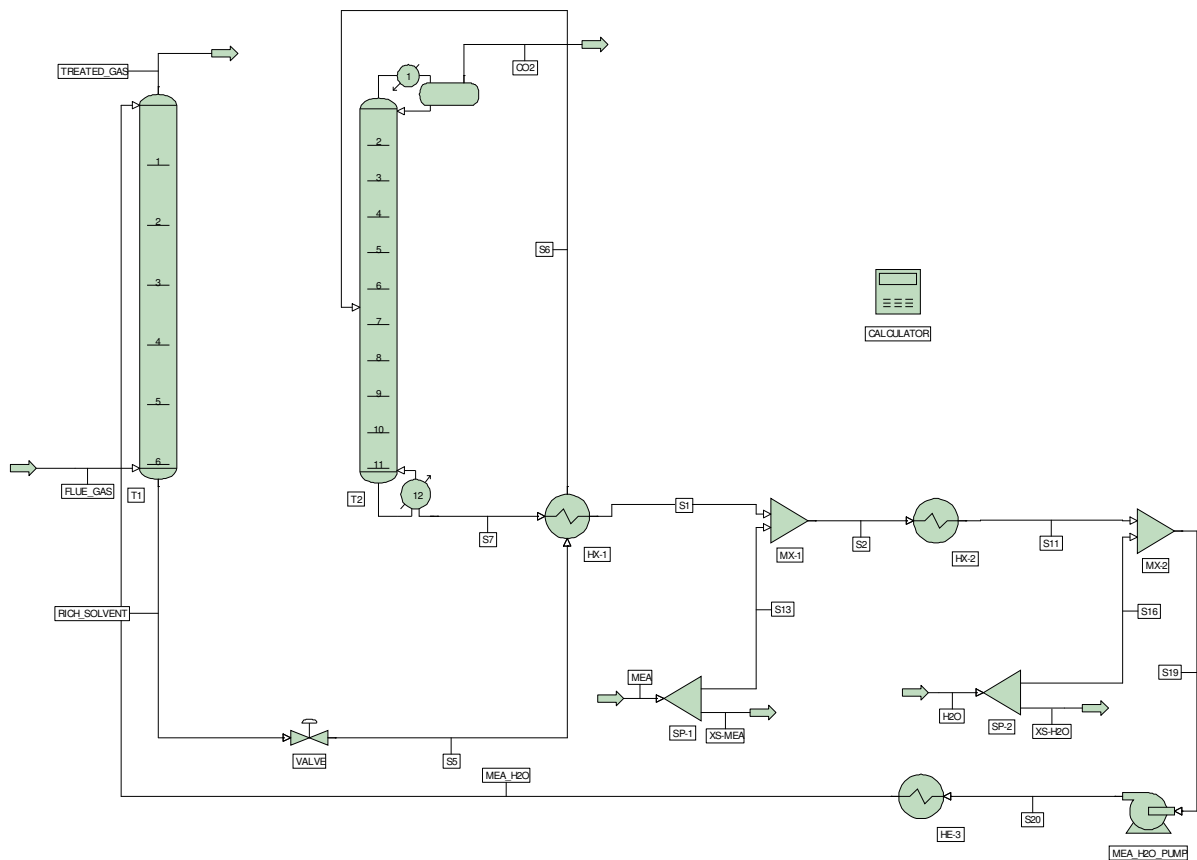


Figure 4.4: Process flow diagram for the capture of CO_2

As one can see, the above PFD is made of two distillation columns used for the absorption and regeneration of CO_2 respectively, two mixers for the water and MEA make-up respectively, two splitters to split the exact amount of water and MEA needed by the process, three heat exchangers, one valve for the lowering of pressure, one calculator to ensure that the system operates at a steady state, a pump to increase the pressure and a number of streams to connect the various unit operations.

2) Check the Input Unit of Measure.

The English unit of measure set is the default used by PRO-II. This set can be changed in the default Unit of Measure for Problem Data Input window, locally override individual dimensional units within data entry windows. As indicated on Figure 4.5, the units of measure were changed from the English units to the SI units where temperature and pressure were giving in Celsius and Atmosphere respectively.

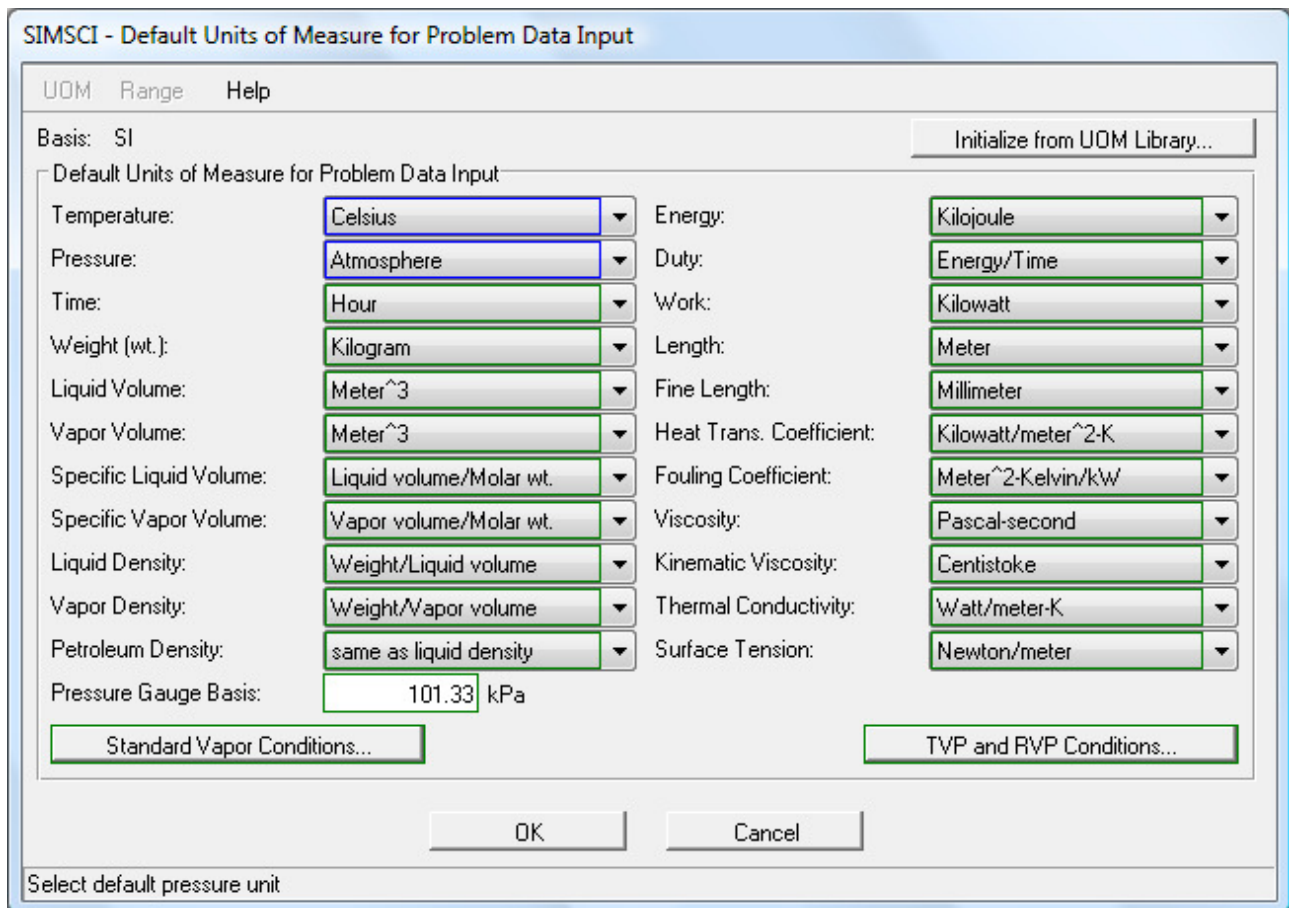


Figure 4.5: A view of how to set units in PRO-II.

3) Define The Components.

Components can directly be defined by typing their library component names, by selecting from lists of chemicals, or by entering user-defined and petroleum components. It is best to list the components in volatility order, starting with the highest volatile component.

In fact components were entered from the list of “most commonly used components”.

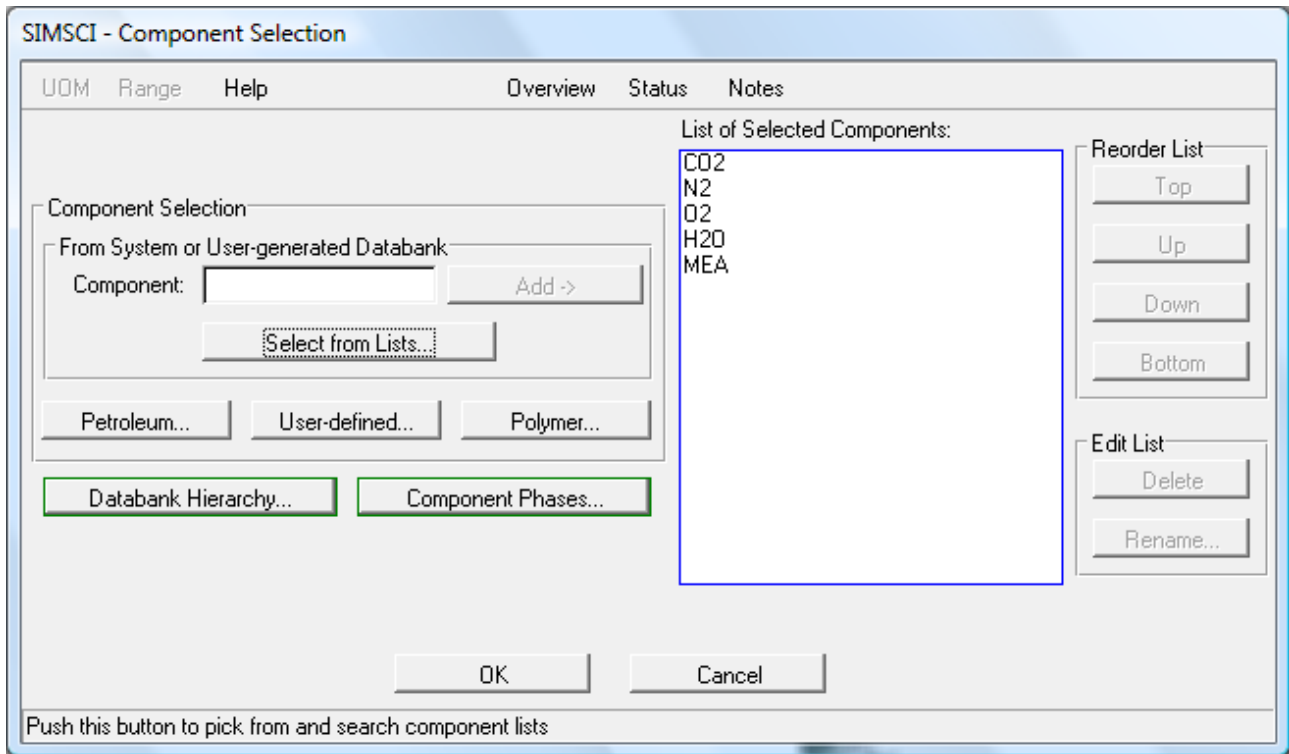


Figure 4.6: Selection of components.

4) Select the Thermodynamic Calculation Methods.

For most problems in PRO-II, a thermodynamic system may be selected from the list of Most Commonly Used thermodynamic methods. Selecting the proper thermodynamic methods is a critically important step in the solution of a simulation problem.

Therefore, out of the several thermodynamic methods, special packages' thermodynamic system with amine as primary method and a single liquid phase (VLE) was chosen because it gives not only the best representation of the flow sheet, but it is the one that gives the best agreement with reality.

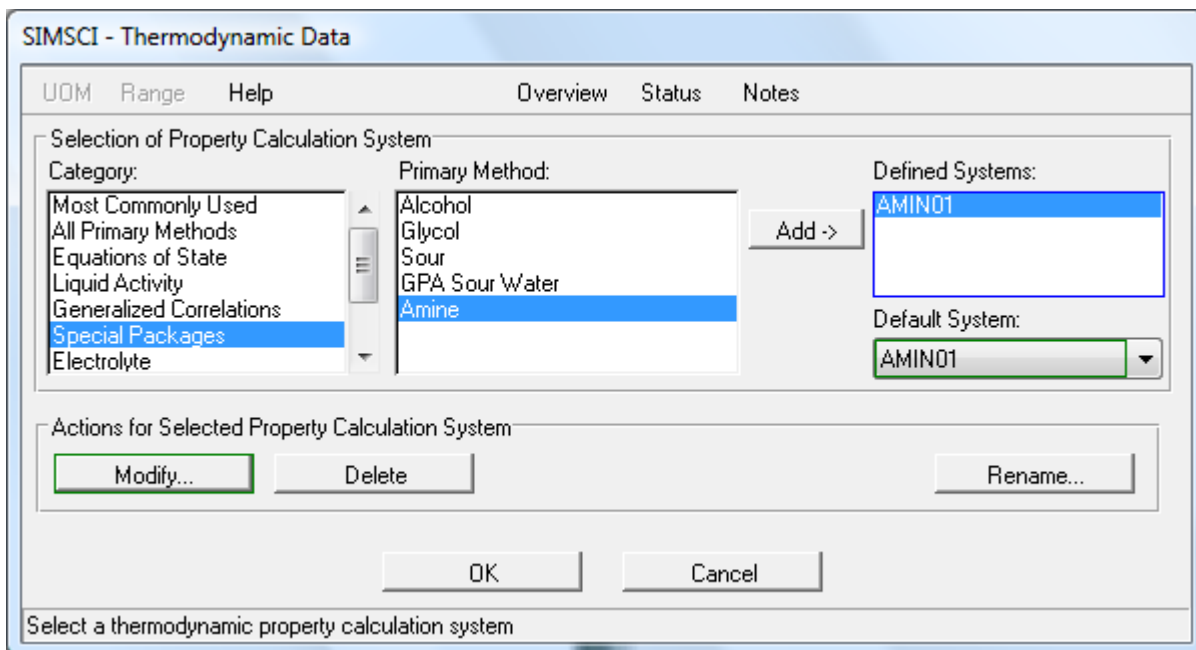


Figure 4.7: Setting of thermodynamic.

5) Supply Data for the Feed and Recycle Streams.

For all external feed streams and recycle estimate to the flow sheet, thermal conditions, flow rates and compositions must be supplied. This is done as seen below.

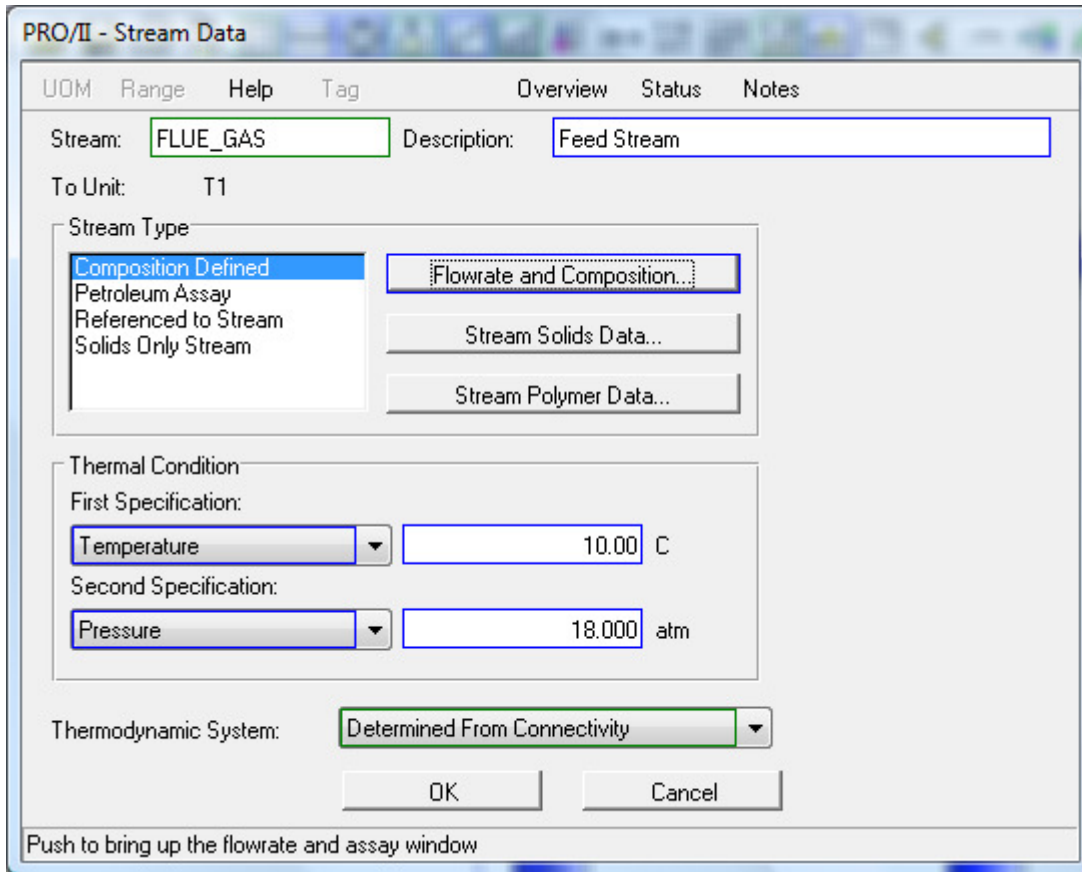


Figure 4.8: Stream data: thermal conditions of the feed stream.

Stream Data - Flowrate and Composition

UOM Range Help Tag

Specify flowrate and composition for stream FLUE_GAS

Fluid Flowrate Specification

Total Fluid Flowrate: 892.30 kg-mol/hr

Individual Component Flowrates

Copy	Component	Composition
Paste	CO2	0.14000
	N2	0.80000
	O2	0.06000
	H2O	
	MEA	

Clear Compositions Total: 1.0000 Normalize Component Flowrates Based on Specified Fluid Flowrate

OK Cancel

Exit the window after saving all data

Figure 4.9: Stream data: flow rate and composition of the feed stream.

It should be noted that, the above data were supplied for all streams involved in the PFD.

6) Provide the Process Conditions.

The required process data for each unit icon on the flow sheet are supplied by double clicking the icon for each unit operation to access the data entry windows. Within the data entry windows for each unit operation, the colour codes indicate which data must be supplied and which data have default values. This could be for example, the feed tray location, the pressure within unit or the flow rate.

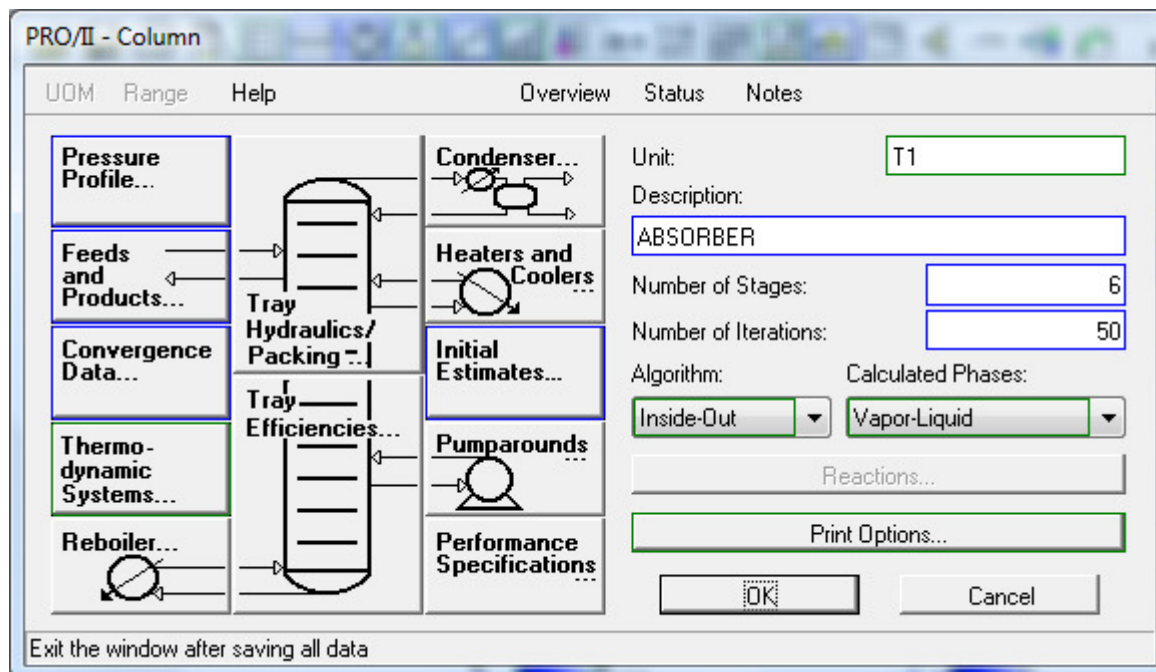


Figure 4.10: Process conditions of the unit Absorber

The same requirement was fulfilled for all the twelve unit operations involved in the process. A quick review is also a good idea at this point. Do the thermodynamic methods support the unit operation calculations?

7) Run the Simulation and View the Results.

Prior to executing the simulation, one should verify that there are no red-bordered fields or red linked text. If all the borders are blue, green, or black on the ribbon bar buttons, unit operation labels, and stream labels, then enough information to run the flow sheet have been supplied. At this point, one may click the Run icon.

Results may be viewed in a variety of ways ranging from plots and output reports to pop-up windows with values for each stream and unit.

Programmer's File Editor - [C:\Users\Serge\AppData\Local\Temp\VW8FED.tmp]

File Edit Options Template Execute Macro Window Help

THERMODYNAMIC SYSTEM AMIN

STREAM 'TREATED_GAS'

	TOTAL	VAPOR
RATE, KG-MOL/HR	767.3101	767.3101
TEMPERATURE, C	32.00	32.00
PRESSURE, ATM	18.00	18.00
MOLECULAR WEIGHT	28.2652	28.2652
FRACTION		1.0000
ENTHALPY, KJ/KG-MOL	82.6086	82.6086
CP, KJ/KG-C	1.0712	1.0712
MOLAR FLOWRATES, KG-MOL/HR		
1 - CO2	1.2189E-04	1.2189E-04
2 - N2	712.1090	712.1090
3 - O2	53.2798	53.2798
4 - H2O	1.9200	1.9200
5 - MEA	1.2000E-03	1.2000E-03
MOLAR COMPOSITIONS		
1 - CO2	1.5886E-07	1.5886E-07
2 - N2	0.9281	0.9281
3 - O2	0.0694	0.0694
4 - H2O	2.5023E-03	2.5023E-03
5 - MEA	1.5639E-06	1.5639E-06

Ln 1 Col 1 | 27 | WR | Rec Off No Wrap DOS INS NUM

Figure 4.11: Report of one of the products (treated gas from the absorber) of the process.

4.2 Simulation proper

4.2.1 Process description

The capture of CO_2 from a flue gas composed of 14% CO_2 , 80% N_2 and 6% O_2 by the solvent MEA/H_2O was conducted in PRO-II.

We first started by drawing the PFD consisting of the following:

- A distillation column with neither a condenser nor a reboiler that acted as the absorber where the flue gas and the solvent MEA/H_2O have a counter current contact so as to facilitate the transfer of the vapour from the vapour phase (flue gas) to the liquid phase (MEA/H_2O). This was carried out at a high pressure and low temperature, because being just a physical phenomenon not involving no change in the chemical species present in the system.

- A second distillation column having a condenser and a reboiler that was used like the regenerator or stripper, to regenerate the carbon dioxide captured from the first tower. Thus, the rich solvent flowed to the top of the regenerator where the absorbed CO_2 is stripped from the solvent and passed overhead. Here the recovery of CO_2 and MEA/H_2O was enhanced by a high temperature and low pressure.
- A valve connecting the two towers, used to lower the pressure of the rich solvent out from the absorber before it got into the regenerator.
- Three heat exchangers to make use of the energy generated by the process.
- Two splitters used to send into the process the exact amount of water and MEA respectively needed for the absorption of CO_2 .
- Two mixers used for the water and MEA make-up for the recycling stream.
- A pump to recirculate the solvent back to the absorber.
- A calculator to make sure that the system operated at a steady state.
- A number of streams connecting the various unit operations.

The system was operated by entering the flue gas at the bottom of the absorber consisting of a tower of six trays in a counter current flow with the solvent.

The rich solvent from the absorber was passed through a valve to lower the pressure before it was heat exchanged with the lean solvent from the regenerator and sent into the regenerator made of a tower with 12 trays, a reboiler and a condenser. The reboiler supplied the heat input to the regenerator. Thus, the product from the condenser was refluxed to the regenerator and the resultant gas, which was essentially pure CO_2 saturated with water was collected, ready to be compressed and stored.

The lean solvent from the regenerator went through a series of heat exchangers to be cooled and a series of mixers for the water and MEA make-up, then pumped back to the absorber.

4.2.2 Process flow diagram

After the simulation was ran, the PFD of the CO_2 capture system looked as in Figure 4.12.

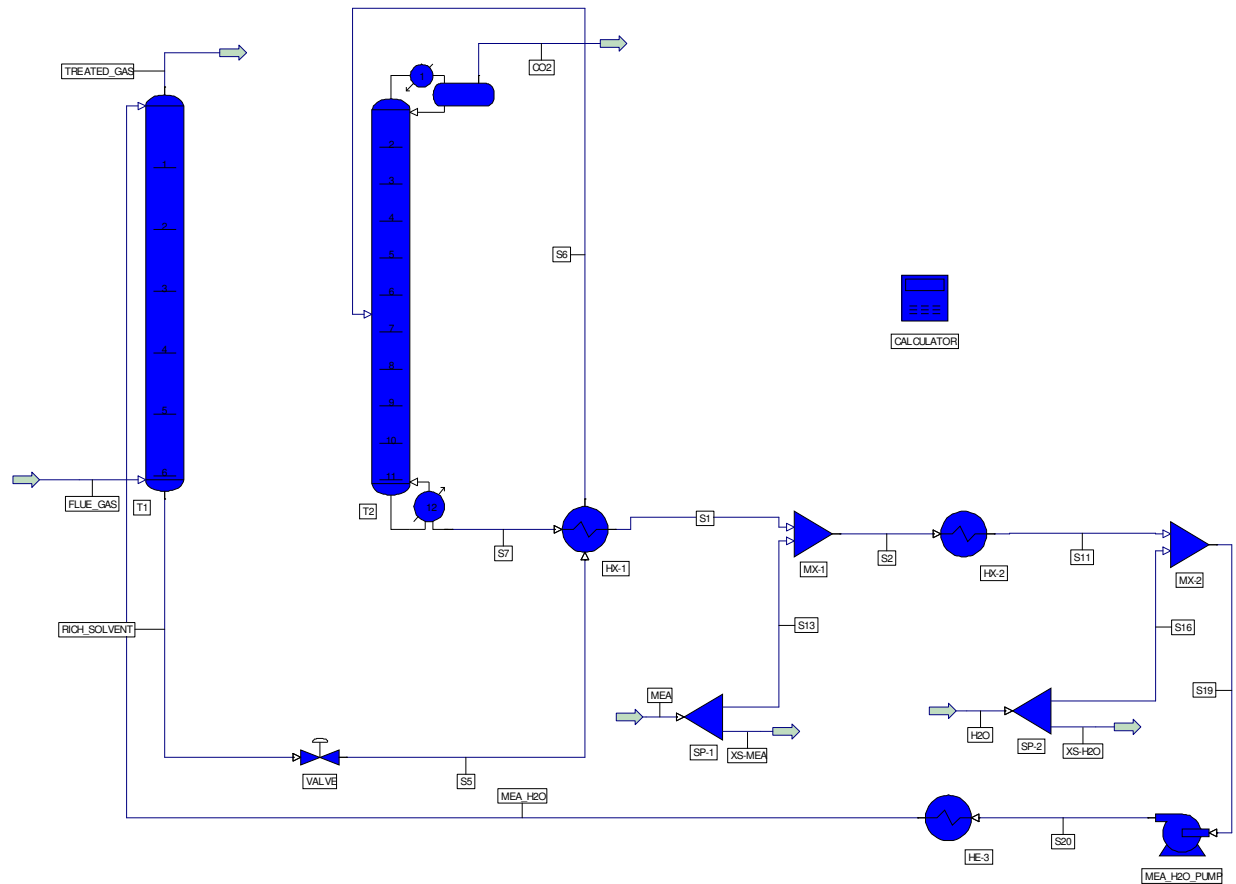


Figure 4.12: Process flow diagram for the capture of CO_2 .

As one can observe, there is a colour change of the PFD from green to blue. And this blue colour indicates that the simulation was reached.

4.2.3 Feed stream

The feed or flue gas stream properties are given in table 1.

Table 1: Feed stream compositional report.

THERMODYNAMIC SYSTEM AMIN

STREAM 'FLUE_GAS'

TOTAL

VAPOR

RATE, KG-MOL/HR	892.3000	892.3000
TEMPERATURE, C	10.00	10.00
PRESSURE, ATM	18.00	18.00
MOLECULAR WEIGHT	30.4921	30.4921
FRACTION		1.0000
ENTHALPY, KJ/KG-MOL	1121.2659	1121.2659
CP, KJ/KG-C	1.0438	1.0438
 MOLAR FLOWRATES, KG-MOL/HR		
1 - CO ₂	124.9220	124.9220
2 - N ₂	713.8400	713.8400
3 - O ₂	53.5380	53.5380
4 - H ₂ O	0.0000	0.0000
5 - MEA	0.0000	0.0000
 MOLAR COMPOSITIONS		
1 - CO ₂	0.1400	0.1400
2 - N ₂	0.8000	0.8000
3 - O ₂	0.0600	0.0600
4 - H ₂ O	0.0000	0.0000
5 - MEA	0.0000	0.0000

Table 1 is all about the thermodynamic (Amin) used in the process, the composition and properties (flow rate, temperature, pressure, molecular weight, mass fraction, enthalpy and heat capacity) of the flue gas. In addition to that, the flow rate of each of the species that composes it is also given and as observed, the feed is made up only of vapour.

4.2.4 Treated gas

When carbon dioxide is absorbed by the solvent *MEA/H₂O*, the remaining gas, which is not harmful to the nature anymore, is then released into the atmosphere. This gas is what termed “treated gas” and its composition and properties are given in table 2 below.

Table 2: Treated flue gas composition and properties.

THERMODYNAMIC SYSTEM	AMIN	
STREAM 'TREATED_GAS'	TOTAL	VAPOR
	-----	-----
RATE, KG-MOL/HR	767.3101	767.3101
TEMPERATURE, C	32.00	32.00
PRESSURE, ATM	18.00	18.00
MOLECULAR WEIGHT	28.2652	28.2652

FRACTION		1.0000
ENTHALPY, KJ/KG-MOL	82.6086	82.6086
CP, KJ/KG-C	1.0712	1.0712
MOLAR FLOWRATES, KG-MOL/HR		
1 - CO2	1.2189E-04	1.2189E-04
2 - N2	712.1090	712.1090
3 - O2	53.2798	53.2798
4 - H2O	1.9200	1.9200
5 - MEA	1.2000E-03	1.2000E-03
MOLAR COMPOSITIONS		
1 - CO2	1.5886E-07	1.5886E-07
2 - N2	0.9281	0.9281
3 - O2	0.0694	0.0694
4 - H2O	2.5023E-03	2.5023E-03
5 - MEA	1.5639E-06	1.5639E-06

Table 2 gives basically the same information as table 1. The treated flue gas is predominantly made up of nitrogen, N_2 (92.81%) and oxygen, O_2 (6.94%).

4.2.5 Absorber

The table below gives the absorber's summary including the temperature, pressure, net flow rates and heat rates of the various vapour and liquid compositions. In addition, the feed and product location or tray number are given.

Table 3: The absorber's compositional report.

Rigorous Column 'T1', 'ABSORBER'

ITERATIONS, MAXIMUM PER TRIAL	50
TOTAL ALL TRIALS	45

COLUMN SUMMARY

TRAY	----- NET FLOW RATES -----				HEATER		DUTIES
	TEMP DEG C	PRESSURE ATM	LIQUID KG-MOL/HR	VAPOR KG-MOL/HR	FEED	PRODUCT M*KJ/HR	
1	32.0	18.00	6417.7		6416.0L	767.3V	
2	32.0	18.01	6417.7	769.1			
3	32.0	18.02	6417.7	769.1			
4	32.1	18.03	6417.9	769.1			
5	33.9	18.04	6425.8	769.3			

6 43.4 18.05 777.2 892.3V 6540.9L

FEED AND PRODUCT STREAMS

TYPE	STREAM	PHASE	FROM TRAY	TO TRAY	LIQUID FRAC	FLOW RATES KG-MOL/HR	HEAT RATES M*KJ/HR
FEED	MEA_H2O	LIQUID		1	1.0000	6415.96	14.4083
FEED	FLUE_GAS	VAPOR		6	0.0000	892.30	1.0005
PROD	TREATED_GAS	VAPOR	1			767.31	0.0634
PROD	RICH_SOLVENT	LIQUID	6			6540.95	15.3454

OVERALL MOLE BALANCE, (FEEDS - PRODUCTS) 0.00
 OVERALL HEAT BALANCE, (H(IN) - H(OUT)) -6.4862E-06
 Currently using rigorous calculations

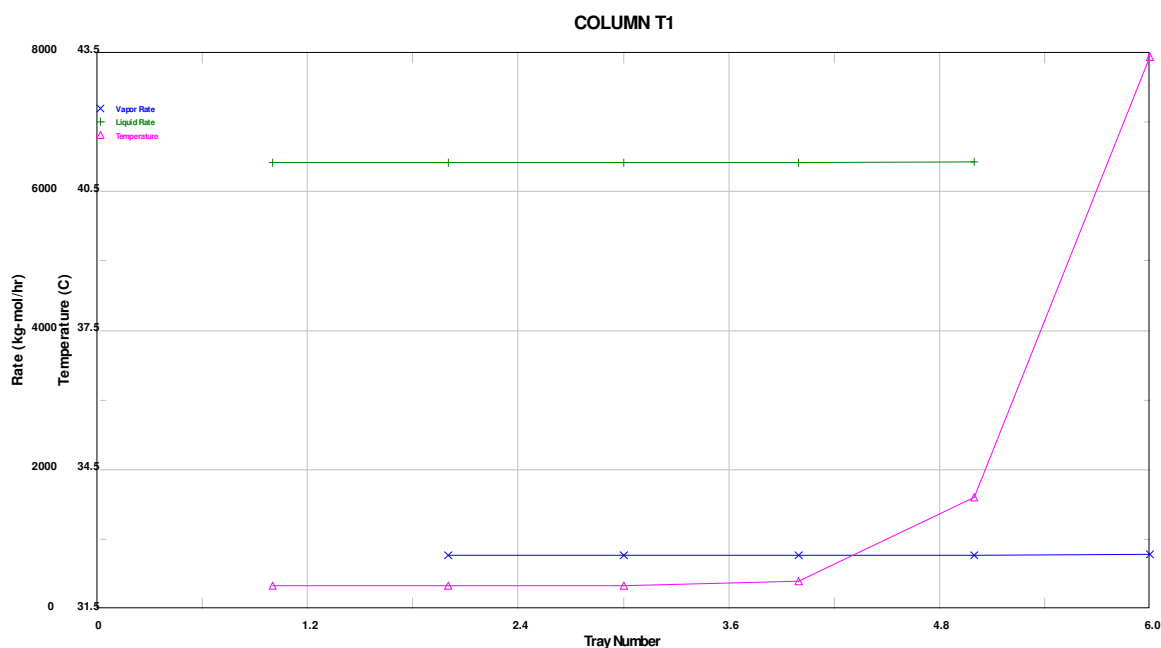


Figure 4.13: Overview of temperature and flow rates within the absorber.

Figure 4.13, gives the evolution of temperature and flow rates inside the absorber. While temperature reduces upwards in the tower, the vapour and liquid flow rates remain somehow constant. The reduction in temperature is explained by the fact that the flue gas and the solvent have different temperature, thus their interaction causes a temperature drop and the more these two interact, further the drop is until the point of perfect mixing (tray number 3) where temperature becomes constant.

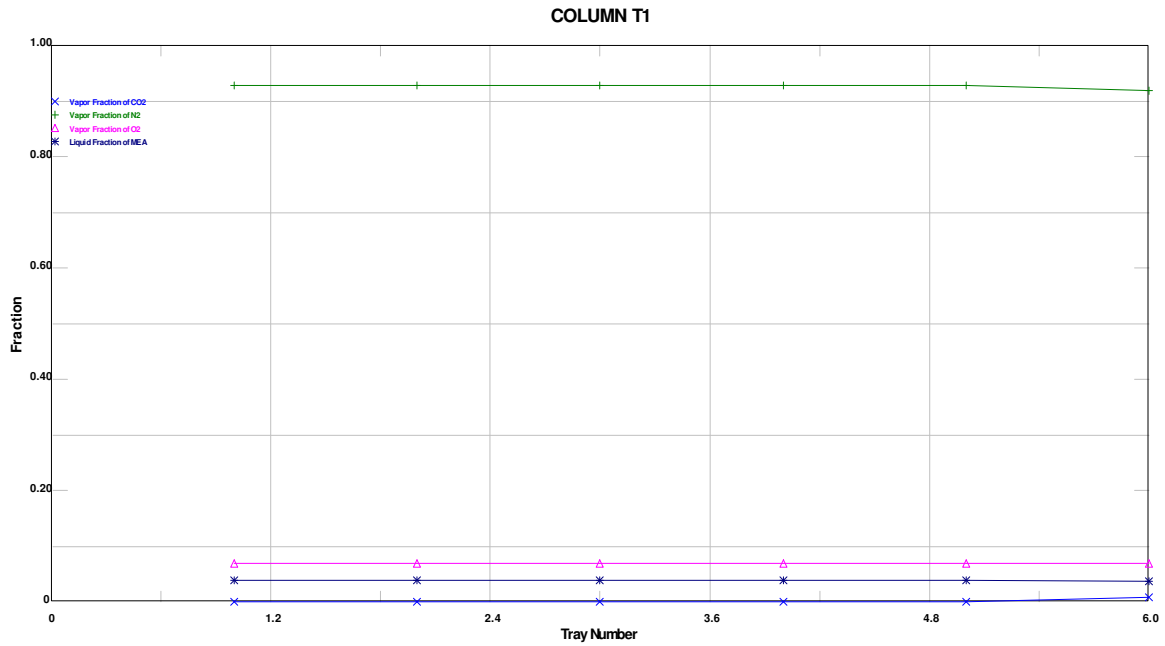


Figure 4.14: Liquid and vapour compositions of the absorber.

The components included figure 4.14 are CO_2 , O_2 , N_2 and MEA .

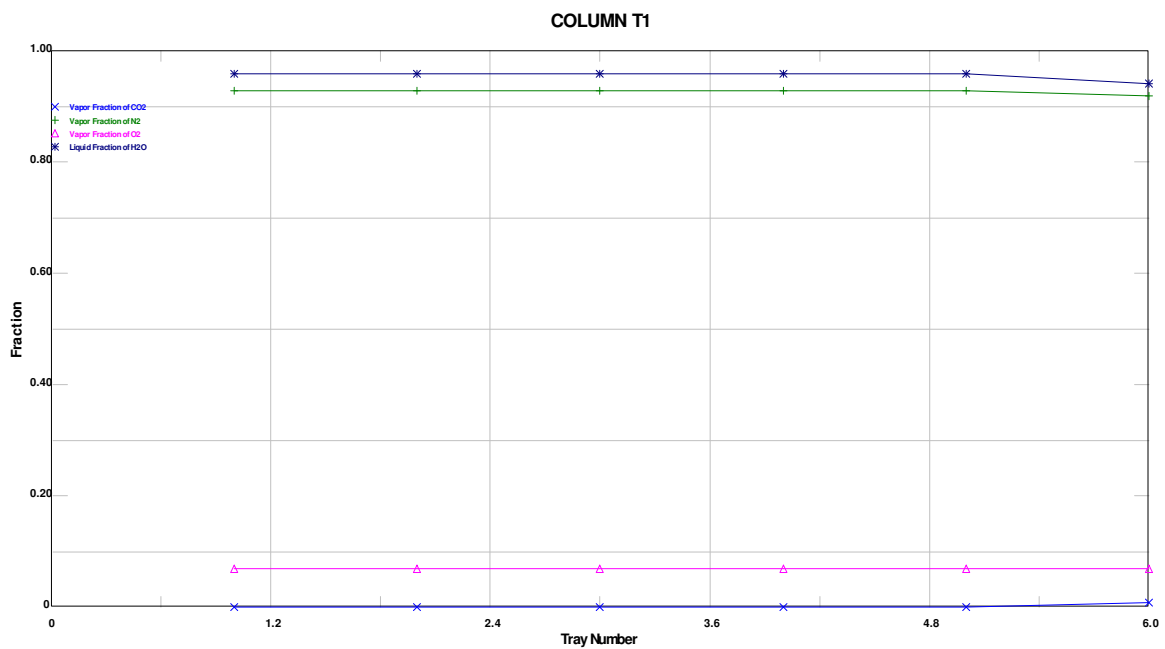


Figure 4.15: Liquid and vapour compositions of the absorber.

Components considered on the above figure are CO_2 , O_2 , N_2 and H_2O .

The importance of having two figures describing the liquid and vapour compositions of the absorber, is because only four components can be represented at a time on a figure and there is a need to see the evolution of all components in the tower.

These figures show that as the flue gas enters the absorber on tray number 6, at the bottom, the fraction of carbon dioxide reduces as the gas moves upwards. This explains why there is no carbon dioxide on the absorber's top tray or tray number 1, because all of it has been transferred to the solvent. Also, the other components of the flue gas N_2 and O_2 remain unchanged in terms of their compositions right up to the top tray, where they are evacuated.

4.2.6 Overhead product from the regenerator

The overhead product of the regenerator is simply pure carbon dioxide (65.46%) saturated with water (33.49%).

Table 4: The compositional report of the overhead product.

THERMODYNAMIC SYSTEM AMIN

STREAM 'CO2'

	TOTAL	VAPOR
	-----	-----
RATE, KG-MOL/HR	190.8185	190.8185
TEMPERATURE, C	70.00	70.00
PRESSURE, ATM	9.1862E-01	9.1862E-01
MOLECULAR WEIGHT	35.1420	35.1420
FRACTION		1.0000
ENTHALPY, KJ/KG-MOL	25879.6018	25879.6018
CP, KJ/KG-C	1.0710	1.0710

MOLAR FLOWRATES, KG-MOL/HR

1 - CO2	124.9178	124.9178
2 - N2	1.7310	1.7310
3 - O2	0.2582	0.2582
4 - H2O	63.9115	63.9115
5 - MEA	1.1588E-14	1.1588E-14

MOLAR COMPOSITIONS

1 - CO2	0.6546	0.6546
2 - N2	9.0712E-03	9.0712E-03
3 - O2	1.3533E-03	1.3533E-03
4 - H2O	0.3349	0.3349
5 - MEA	6.0727E-17	6.0727E-17

4.2.7 Bottom product from the regenerator

The bottom product of the regenerator is actually the lean solvent composed of 95.87% H_2O and 3.84% MEA which has to be concentrated again, in terms of water and MEA contents, through their respective make-up (water and MEA make-up).

Table 5: Lean solvent's compositional report.

THERMODYNAMIC SYSTEM AMIN

STREAM 'S7' "lean solvent" from the regenerator

	TOTAL	LIQUID
	-----	-----
RATE, KG-MOL/HR	6350.1286	6350.1286
TEMPERATURE, C	106.08	106.08
PRESSURE, ATM	1.19	1.19
MOLECULAR WEIGHT	19.7446	19.7446
FRACTION		1.0000
ENTHALPY, KJ/KG-MOL	8103.2172	8103.2172
CP, KJ/KG-C	4.0487	4.0487
MOLAR FLOWRATES, KG-MOL/HR		
1 - CO2	18.0245	18.0245
2 - N2	1.1636E-26	1.1636E-26
3 - O2	2.8840E-26	2.8840E-26
4 - H2O	6088.0120	6088.0120
5 - MEA	244.0921	244.0921
MOLAR COMPOSITIONS		
1 - CO2	2.8384E-03	2.8384E-03
2 - N2	1.8324E-30	1.8324E-30
3 - O2	4.5417E-30	4.5417E-30
4 - H2O	0.9587	0.9587
5 - MEA	0.0384	0.0384

4.2.8 Regenerator

As the noun says, the regenerator is where carbon dioxide gets regenerated from the rich solvent by an elevation of temperature and a reduction of pressure. Thus, table 6 gives a summary of this process.

Here, we have among other things, all that has been mentioned earlier for the absorber, but also the reboiler and condenser duties and the specifications set in the regenerator.

Table 6: Regenerator's summary.

Rigorous Column 'T2'

ITERATIONS, MAXIMUM PER TRIAL	40
TOTAL ALL TRIALS	235

COLUMN SUMMARY

TRAY	TEMP DEG C	PRESSURE ATM	----- NET FLOW RATES -----			HEATER DUTIES M*KJ/HR
			LIQUID	VAPOR	FEED PRODUCT	
			----- KG-MOL/HR -----			
1C	70.0	0.92	2471.6		190.8V	-107.2524
2	100.1	1.05	2607.7	2662.4		
3	100.6	1.07	2609.0	2798.5		
4	100.9	1.08	2610.0	2799.8		
5	101.3	1.10	2610.8	2800.8		
6	101.7	1.11	2608.4	2801.7		
7	102.9	1.12	9454.6	2799.3	6540.9M	
8	104.1	1.14	9487.1	3104.5		
9	104.6	1.15	9503.4	3137.0		
10	105.0	1.16	9515.7	3153.2		
11	105.4	1.18	9525.2	3165.6		
12R	106.1	1.19		3175.1	6350.1L	130.4587

FEED AND PRODUCT STREAMS

TYPE	STREAM	PHASE	FROM	TO	LIQUID FRAC	FLOW RATES KG-MOL/HR	HEAT RATES M*KJ/HR
			TRAY	TRAY			
FEED	S6	MIXED		7	0.9978	6540.95	33.1886
PROD	CO2	VAPOR	1			190.82	4.9383
PROD	S7	LIQUID	12			6350.13	51.4564

OVERALL MOLE BALANCE, (FEEDS - PRODUCTS) -3.9968E-12

OVERALL HEAT BALANCE, (H(IN) - H(OUT)) -7.5448E-04

SPECIFICATIONS

SPECIFICATION NUMBER	PARAMETER TYPE	TRAY NO	COMP NO	SPECIFICATION TYPE	SPECIFIED VALUE	CALC VALUE
1 (ACTIVE)	TRAY LIQ	1		TEMPERATURE	7.000E+01	7.000E+01
2 (ACTIVE)	TRAY VAP	12		MOL RATIO	5.000E-01	5.000E-01

Currently using rigorous calculations

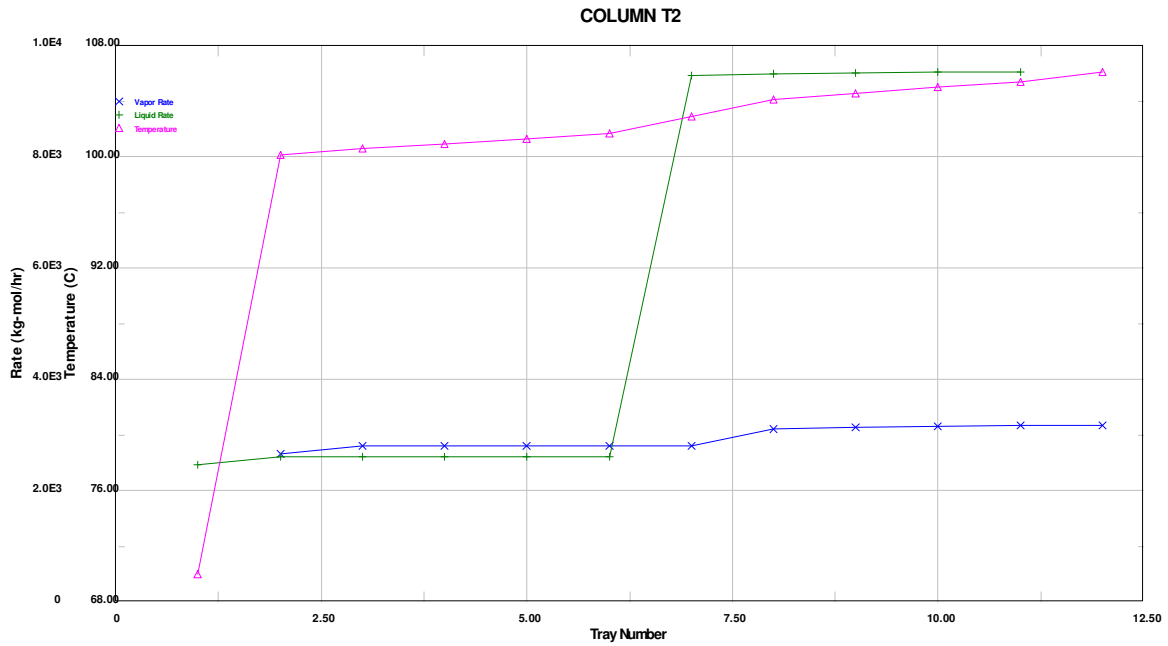


Figure 4.16: Overview of temperatures and flow rates within the regenerator.

Contrary to the absorber, the regenerator is operated at elevated temperatures that further increase as one goes down the tower as seen on the above figure. It can also be seen that the separation of CO_2 from the rich solvent takes place as soon as it gets inside the tower on tray number 7.

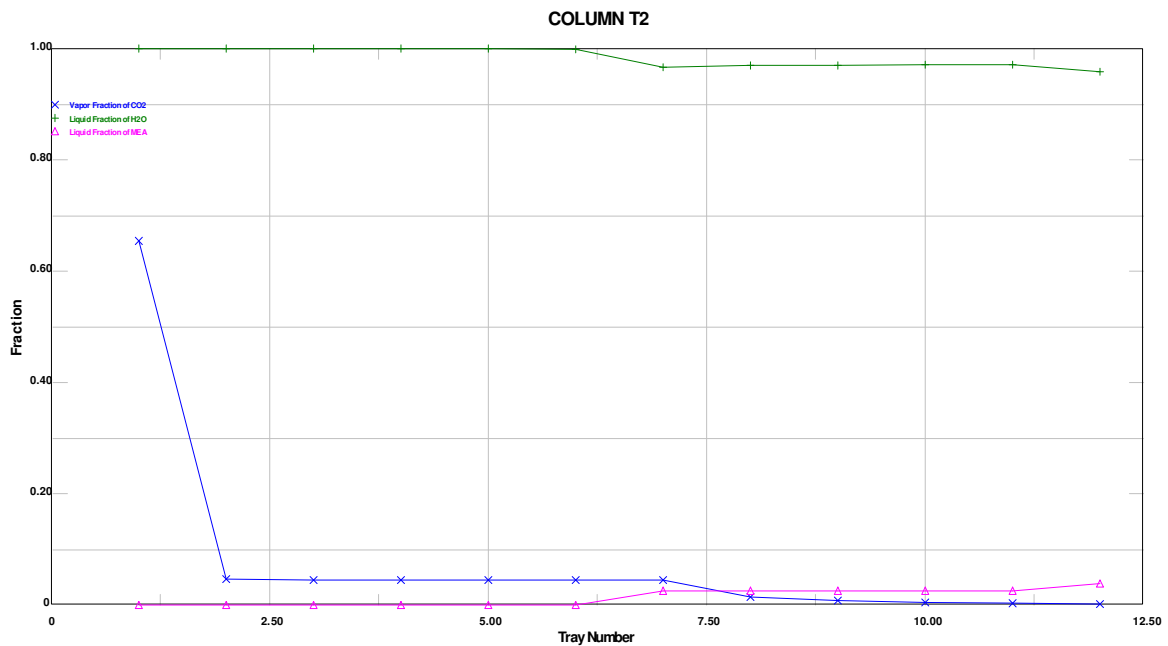


Figure 4.17: Liquid and vapour compositions of the regenerator.

Figure 4.17 gives a distinct picture of the separation between CO_2 , H_2O , and MEA within the generator. That is, as one moves upwards in the column, the fraction of MEA reduces while CO_2 fraction increases up to a fraction of 0.6546 pure CO_2 mixed with water.

4.2.9 Calculator

The calculator is the unit operator that ensures that the capture process operates at a steady state. That is the used amount of water and MEA should be the same as the one entering the process through the splitter, for their make-up ready for the next capture. And in doing that, a number of parameters were set in it.

Recall that mass balance is defined by:

$$[flow, in] - [flow, out] - [accumulation] = 0 \quad (4)$$

It follows that:

$$[acunulation] = [flow, in] - [flow, out] \quad (5)$$

And for a steady state $[accumulation] = 0$ (6)

Therefore $[flow, in] = [flow, out]$ (7)

This means that, the flow rates of water and MEA getting out through the two outflows (treated gas and CO_2 streams), should be the same as those getting in through the streams H_2O and MEA .

Thus, we have the following equations:

$$R_1 = P_1 + P_2 \quad (8)$$

$$R_2 = P_3 + P_4 \quad (9)$$

Where P_1 and P_3 are the respective flow rates of water and MEA in the stream “treated gas”.

P_2 and P_4 the respective flow rates of water and MEA in the stream “ CO_2 ”.

R_1 and R_2 are just the results 1 and 2.

Hence, the results from the calculator are used to define the flow rates of water and MEA in the splitter as follows:

Stream MEA: $\frac{F_{S13}}{R_2} = 1$ (10)

With

F_{S13} the flow rate of MEA entering the process through the splitter and R_2 the result from the calculator.

$$\text{Stream } H_2O: \quad \frac{F_{S16}}{R_1} = 1 \quad (11)$$

With

F_{S16} the flow rate of water entering the process through the splitter and R_1 the result from the calculator.

At this point, one is more than sure that the process is carried out at a steady state, that is the flow in is the same as the flow out.

Table 7: Calculator's overview.

UNIT 12, 'CALCULATOR'

Result	Name	Value	Result	Name	Value
1	R1	6.58316E+01	3-	200	Undefined
2	R2	1.20001E-03			

Parameter	Value	Parameter	Value
1	1.92009E+00	4	1.15879E-14
2	6.39115E+01	5-	50
3	1.20001E-03		Undefined

5. Discussion

The purpose of this project was to examine the various possibilities of capturing and storing carbon dioxide from the flue gases exhausting from power plants, petroleum refineries, gas processing facilities and cement factories. In addition to that, a simulation of CO_2 capture using the software Pro-II was also conducted.

It was seen that the process of carbon capture and storage involved the following stages: carbon dioxide capture, transport and finally storage. Basically, there are three ways through which CO_2 could be captured notably, pre-combustion capture, post-combustion capture and oxy-fuel combustion capture.

The pre-combustion capture technique, as its name indicates, consists of removing CO_2 from the fossil fuel before it is combusted. This means that, fossil fuel is being separated into CO_2 and H_2 where CO_2 gets captured, then H_2 mixed with nitrogen and burnt, resulting in a non hazardous flue gas released into the atmosphere.

But this technique is still immature, though it theoretically gives high CO_2 concentrations through.

Post-combustion capture of carbon dioxide involves the absorption of CO_2 from the flue gas (from the combustion of fossil fuels) using chemical solvents. This technique is the most advanced nowadays, with a well established technology, though it has a large energy requirement.

In capturing CO_2 , the flue gas is passed through a high pressure gas/liquid contactor, where the solvent and the gas have counter current directions, so as to maximize their interaction, resulting in a high absorption of carbon dioxide to the solvent. The lean solvent containing carbon dioxide is then heated in another tower to release CO_2 which is compressed and ready to be stored, while the resulting solvent is recycled back to the absorption tower.

Oxy-fuel combustion capture on its own involves using oxygen rather than air to combust fossil fuels. Thus, the combustion conditions are changed in order to avoid mixing nitrogen and carbon dioxide by keeping nitrogen out of the combustion chamber. This means that, the oxygen needed is removed from air and used to combust fossil fuels.

This technique has the advantage producing only gaseous emissions of carbon dioxide, water and small amounts of sulphur dioxide, which can be easily separated. But the production of oxygen needed for the combustion is rather an expensive step for the viability of the oxy-fuel combustion technique.

The second part of the project which was the simulation carbon capture was done using the post-combustion capture technique. As such, it was assumed a flue gas from the combustion of fossil fuels with the following composition: carbon dioxide (CO_2 ; 14%), nitrogen (N_2 ; 80%) and oxygen (O_2 ; 6%). Thus, the chemical solvent used for the capture or absorption of CO_2 was the amine mono-ethanolamine (MEA) in water, in a proportion of 4% MEA and 96% H_2O .

Having in mind that carbon capture is an expensive process as it consumes enormous amounts of energy, we tried as much as possible to reduce the duties of the process by using a calculator to regulate both the exact amount of water and MEA used by the process and a recycling stream of MEA/ H_2O was also included. Additionally, the heat generated by the process was used again within the process so as to minimize any loss of energy.

After the process flow diagram was constructed with the various unit operators, the system was operated by entering the flue gas at the bottom of the absorber consisting of a tower of six trays in a counter current flow with the solvent (MEA/ H_2O).

The rich solvent from the absorber was passed through a valve to lower the pressure before it was heat exchanged with the lean solvent from the regenerator and sent into the regenerator made of a tower with 12 trays, a reboiler and a condenser. The reboiler supplied the heat input to the regenerator. Thus, the product from the condenser was refluxed to the regenerator and the resultant gas, which was essentially pure CO_2 (65.46%) saturated with water (33.49%) was collected, ready to be compressed and stored.

The lean solvent from the regenerator went through a series of heat exchangers to be cooled and a series of mixers for the water and MEA make-up, then pumped back to the absorber as mentioned earlier.

Furthermore, it should be noted that the result (65.46% CO_2 captured) obtained here is based on the current model. Though it was entered an absorption fraction of 0.93, we ended up having but an absorption fraction of 0.6546. This could be explained by the fact that, in PRO-II, a simulation is

reached at a steady state, meaning that the values which are entered might not be obtained at the end of the simulation if they do not allow a steady state.

Typically, the captured amount of carbon dioxide strongly depends on factors like the height of the towers, including the number of trays, as well as the flow rates of the flue gas and the capturing solvent.

Hence, considering the same capturing process, but this time with an absorber of three trays and a stripper of 6 trays, only 48.35% CO_2 was captured. This clearly indicates that the rate of capture depends on the number of stages in the column, as the higher the column, the more it gives or favours a perfect mixing between the solvent and the flue gas.

The same observation was done concerning the flow rates of both the solvent and flue gas.

On the energy point of view, the process is found viable because of its low consumption of energy related to the MEA consumption. As indicated in Table 5.1, only a fraction of 0.0019 MEA (0.19%) is sent in (stream S13) for its make-up. It follows that, for a single cycle, 0.19% MEA is used for the capture of 65.46% CO_2 .

Table 5.1: Overview of the MEA splitter.

SIMULATION SCIENCES INC. R PAGE P-5
 PROJECT PRO/II VERSION 8.1 ELEC V6.6
 PROBLEM OUTPUT
 SPLITTER SUMMARY 01/10/09

UNIT 10, 'SP-1'				
STREAM ID	FRACTION	-----	RATES	-----
-----	-----	-----	KG-MOL/HR	KG/HR
-----	-----	-----	-----	-----
FEED	MEA		0.620	37.872
PRODUCTS	S13	0.0019	0.001	0.073
	XS-MEA	0.9981	0.619	37.799
TEMPERATURE, C		28.0000		
PRESSURE, ATM		1.0000		
PRESSURE DROP, ATM		0.0000		
MOLE FRAC VAPOR		0.0000		
MOLE FRAC TOTAL LIQUID		1.0000		
MOLE FRAC MW SOLID		0.0000		

Additionally, the overall heat balance of the absorber and regenerator is respectively negative, meaning that more energy is produced in the course of this process than actually used. Thus, this energy is used back in the process, so as to further cut down the energetic expenses.

6. Conclusion

The task given at the beginning of the project was to examine the various possibilities to capture carbon dioxide and store it somewhere, including the simulation of a post-combustion CO_2 -capture process.

Out of the tree techniques of capturing carbon dioxide, the post-combustion capture technique is the one more feasible as the existing power plants and heavier industries can easily be retrofitted with this technique without any major change.

Moreover, the operation of MEA based CO_2 -capture simulation was successfully done. From a flue gas composed of carbon dioxide (CO_2 ; 14%), nitrogen (N_2 ; 80%) and oxygen (O_2 ; 6%), 65.46% pure CO_2 saturated with water was captured. Thus, it points out the effectiveness of the carbon dioxide capture technique.

Finally, the design and performance variables for the simulation are shown in Table 6.1 below.

Table 6.1: Design specifications

Parameter	Design value
Flue gas conditions	T=10°C, P=18 Atm, 14% CO_2 , 80% N_2 , 6% O_2 , F= 892.3 Kg.mol/hr
Solvent MEA/H_2O	T=32°C, P=17 Atm, 4% MEA , 96% H_2O , F= 6415.9573 Kg.mol/hr
Absorber	T= 32 – 43.4°C, P≈ 18 Atm
Regenerator	T= 70 – 106.1°C, P≈ 1 Atm
Reboiler steam flow	F= 3175.1 Kg.mol/hr (1.19 Atm)

APPENDIX

A. Text Report of the simulation process.

\$ Generated by PRO/II Keyword Generation System <version 8.1>

\$ Generated on: Sat Jan 10 20:59:56 2009

TITLE

DIMENSION SI, TEMP=C, PRES=ATM, STDTEMP=0, STDPRES=1

SEQUENCE SIMSCI

CALCULATION RVPBASIS=APIN, TVP=37.778, RECYCLE=ALL

COMPONENT DATA

LIBID 1,CO2/2,N2/3,O2/4,H2O/5,MEA, BANK=SIMSCI,PROCESS

ASSAY CONVERSION=API94, CURVEFIT=IMPROVED, KVRECONCILE=TAILS

THERMODYNAMIC DATA

METHOD SYSTEM=AMIN, SET=AMIN01, DEFAULT

STREAM DATA

PROPERTY STREAM=FLUE_GAS, TEMPERATURE=10, PRESSURE=18, PHASE=M, &
RATE(M)=892.3, COMPOSITION(M)=1,0.14/2,0.8/3,0.06

PROPERTY STREAM=RICH_SOLVENT, TEMPERATURE=32, PRESSURE=17, PHASE=M,
& RATE(M)=6525.36, COMPOSITION(M)=1,0.13/4,0.84/5,0.03

PROPERTY STREAM=TREATED_GAS, TEMPERATURE=32, PRESSURE=17, PHASE=M, &
RATE(M)=766.958, COMPOSITION(M)=2,0.92/3,0.06/4,0.018/ & 5,0.002

PROPERTY STREAM=MEA_H2O, TEMPERATURE=32, PRESSURE=17, PHASE=M, &
RATE(M)=6400.01, COMPOSITION(M)=1,0.0013/4,0.96/5,0.0387

PROPERTY STREAM=S5, TEMPERATURE=59.21, PRESSURE=1.1, PHASE=M, &
RATE(M)=7360.2, COMPOSITION(M)=1,0.0125/4,0.96/5,0.0275

PROPERTY STREAM=S6, TEMPERATURE=88.35, PRESSURE=1.1, PHASE=M, &
RATE(M)=7358.8, COMPOSITION(M)=1,0.0125/4,0.96/5,0.0275

PROPERTY STREAM=S7, TEMPERATURE=122.62, PRESSURE=2, PHASE=M, &
RATE(M)=6372.4, COMPOSITION(M)=1,0.0033784/4,0.96/5,0.03662

PROPERTY STREAM=CO2, TEMPERATURE=70, PRESSURE=0.91862, PHASE=M, &
RATE(M)=152.98, COMPOSITION(M)=1,0.8745/4,0.1255/5,4.83E-13

PROPERTY STREAM=S1, TEMPERATURE=90, PRESSURE=1.1, PHASE=M, &
RATE(M)=6372.36, COMPOSITION(M)=1,0.0039536/4,0.9332/ & 5,0.06285

PROPERTY STREAM=S2, TEMPERATURE=77, PRESSURE=1, PHASE=M, &
RATE(M)=6372.36, COMPOSITION(M)=1,0.0039536/4,0.9332/ & 5,0.06285

PROPERTY STREAM=S11, TEMPERATURE=66, PRESSURE=1, PHASE=M, &
RATE(M)=6372.36, COMPOSITION(M)=1,0.0039536/4,0.9332/ & 5,0.06285

PROPERTY STREAM=MEA, TEMPERATURE=28, PRESSURE=1, PHASE=M, &
RATE(M)=0.619999, COMPOSITION(M)=5,1

PROPERTY STREAM=S13, TEMPERATURE=28, PRESSURE=1, PHASE=M, &
RATE(M)=0.02, COMPOSITION(M)=5,1

PROPERTY STREAM=XS-MEA, TEMPERATURE=28, PRESSURE=1, PHASE=M, &
RATE(M)=0.5, COMPOSITION(M)=5,1

PROPERTY STREAM=H2O, TEMPERATURE=70, PRESSURE=1, PHASE=M, &
RATE(M)=1E6, COMPOSITION(M)=4,1

PROPERTY STREAM=S16, TEMPERATURE=70, PRESSURE=1, PHASE=M, &
RATE(M)=0.0699998, COMPOSITION(M)=4,1

PROPERTY STREAM=XS-H2O, TEMPERATURE=70, PRESSURE=1, PHASE=M, &
RATE(M)=0.600001, COMPOSITION(M)=4,1

PROPERTY STREAM=S20, TEMPERATURE=66, PRESSURE=17, PHASE=M, &
RATE(M)=6372.36, COMPOSITION(M)=1,0.0039536/4,0.9332/ & 5,0.06285

PROPERTY STREAM=S19, TEMPERATURE=66.4, PRESSURE=1, PHASE=M, &
RATE(M)=6372.36, COMPOSITION(M)=1,0.0039536/4,0.9332/ & 5,0.06285

UNIT OPERATIONS

COLUMN UID=T1, NAME=ABSORBER

PARAMETER TRAY=6,IO=50 DAMPING=0.4, ERRINC=2

FEED FLUE_GAS,6/MEA_H2O,1

PRODUCT BTMS(M)=RICH_SOLVENT, OVHD(M)=TREATED_GAS,0.889999, &
SUPERSEDE=ON

PSPEC PTOP=18, DPTRAY=0.01

PRINT PROPTABLE=PART

ESTIMATE MODEL=CONVENTIONAL, TTEMP=30, BTEMP=33

TEMPERATURE 1,30

VALVE UID=VALVE

FEED RICH_SOLVENT

PRODUCT M=S5

OPERATION PRESSURE=1

HX UID=HX-1

HOT FEED=S7, M=S1

COLD FEED=S5, M=S6

CONFIGURE COUNTER

OPER CTEMP=77

COLUMN UID=T2

PARAMETER TRAY=12,IO=40 DAMPING=0.4, ERRINC=2

FEED S6,7
PRODUCT BTMS(M)=S7, OVHD(M)=CO2,190
CONDENSER TYPE=PART, PRESSURE=0.91862
DUTY 1,1,,CONDENSER
DUTY 2,12,,REBOILER
PSPEC PTOP=1.0547, DPCOLUMN=0.13609
PRINT PROPTABLE=PART
ESTIMATE MODEL=CONVENTIONAL, RRATIO=3
SPEC ID=COL2SPEC1, TRAY=1, TEMPERATURE(C), VALUE=70
SPEC ID=COL2SPEC2, TRAY=12, RATE(KGM/H), PHASE=V,WET, DIVIDE, &
STREAM=S7, RATE(KGM/H),TOTAL,WET, VALUE=0.5
VARY DNAME=REBOILER,CONDENSER
REBOILER TYPE=KETTLE

CALCULATOR UID=CALCULATOR

RESULT 1,R1/2,R2
DEFINE P(1) AS STREAM=TREATED_GAS, RATE(KGM/H), COMP=4,WET
DEFINE P(2) AS STREAM=CO2, RATE(KGM/H), COMP=4,WET
DEFINE P(3) AS STREAM=TREATED_GAS, RATE(KGM/H), COMP=5,WET
DEFINE P(4) AS STREAM=CO2, RATE(KGM/H), COMP=5,WET
PROCEDURE
R(1)=P(1)+P(2)
R(2)=P(3)+P(4)
RETURN

SPLITTER UID=SP-1

FEED MEA
PRODUCT M=S13, M=XS-MEA
OPERATION OPTION=FILL
SPEC STREAM=S13, RATE(KGM/H), COMP=5,WET, DIVIDE, &
CALCULATOR=CALCULATOR, R(2), VALUE=1

MIXER UID=MX-1, NAME=MEA-MAKE UP

FEED S13,S1
PRODUCT M=S2
OPERATION DP=0.01

HX UID=HX-2

HOT FEED=S2, M=S11
OPER HTEMP=56

SPLITTER UID=SP-2

FEED H2O
PRODUCT M=S16, M=XS-H2O
OPERATION OPTION=FILL
SPEC STREAM=S16, RATE(KGM/H), COMP=4,WET, DIVIDE, &
CALCULATOR=CALCULATOR, R(1), VALUE=1

MIXER UID=MX-2, NAME=H2O MAKE UP
FEED S11,S16
PRODUCT M=S19

PUMP UID=MEA_H2O_PUMP
FEED S19
PRODUCT M=S20
OPERATION EFF=80, PRESSURE=17

HX UID=HE-3
HOT FEED=S20, M=MEA_H2O
OPER HTEMP=32

RECYCLE DATA
ACCELERATION TYPE=BROYDEN

END

SIMULATION SCIENCES INC.	R	PAGE H-1
PROJECT	PRO/II VERSION 8.1 ELEC V6.6	
PROBLEM	CALCULATION	
	HISTORY	12/21/08

=====
*** PROBLEM SOLUTION BEGINS
FEED FLASH COMPLETE
*** PROBLEM SOLUTION REACHED

*** THIS RUN USED 185.13 PRO/II SIMULATION UNITS

*** RUN STATISTICS
STARTED 17:12:49 12/21/08 NO ERRORS
FINISHED 17:13:17 12/21/08 NO WARNINGS
RUN TIMES NO MESSAGES
INTERACTIVE 0 MIN, 28.31 SEC
CALCULATIONS 0 MIN, 0.42 SEC
TOTAL 0 MIN, 28.73 SEC

SIMULATION SCIENCES INC. R PAGE P-1
 PROJECT PRO/II VERSION 8.1 ELEC V6.6
 PROBLEM OUTPUT
 COMPONENT DATA 01/10/09

COMPONENT	COMP. TYPE	PHASE	MOL. WEIGHT	DENSITY KG/M3
1 CO2	LIBRARY	VAP/LIQ/SOL	44.010	816.427
2 N2	LIBRARY	VAP/LIQ/SOL	28.013	807.960
3 O2	LIBRARY	VAP/LIQ/SOL	31.999	1141.912
4 H2O	LIBRARY	VAP/LIQ/SOL	18.015	999.014
5 MEA	LIBRARY	VAP/LIQ/SOL	61.084	1019.393

COMPONENT	NBP C	CRIT. TEMP. C	CRIT. PRES. ATM	CRIT. VOLM. M3/KG-MOL
1 CO2	-78.480	31.060	72.865	0.0940
2 N2	-195.806	-146.950	33.555	0.0892
3 O2	-182.962	-118.570	49.771	0.0734
4 H2O	100.000	373.980	217.666	0.0559
5 MEA	170.000	405.050	70.308	0.2250

COMPONENT	ACEN. FACT.	HEAT FORM. KJ/KG-MOL	G FORM. KJ/KG-MOL
1 CO2	0.22362	-393510.00	-394370.00
2 N2	0.03772	0.00	0.00
3 O2	0.02218	0.00	0.00
4 H2O	0.34486	-241814.00	-228590.00
5 MEA	0.44674	-206700.00	-103300.00

SIMULATION SCIENCES INC. R PAGE P-2
 PROJECT PRO/II VERSION 8.1 ELEC V6.6
 PROBLEM OUTPUT
 CALCULATION SEQUENCE AND RECYCLES 01/10/09

CALCULATION SEQUENCE

SEQ	UNIT ID	UNIT TYPE	SEQ	UNIT ID	UNIT TYPE
1	T1	COLUMN	7	MX-1	MIXER
2	VALVE	VALVE	8	HX-2	HX
3	HX-1	HX	9	SP-2	SPLITTER
4	T2	COLUMN	10	MX-2	MIXER
5	CALCULATOR	CALCULATOR	11	MEA_H2O_PUMP	PUMP
6	SP-1	SPLITTER	12	HE-3	HX

RECYCLE LOOPS

LOOPS			TEAR STREAMS		
ID	FIRST UNIT	LAST UNIT	ID	FROM UNIT	TO UNIT
LOOP1	T1	HE-3	S7	T2	HX-1
			MEA_H2O	HE-3	T1

SIMULATION SCIENCES INC. R PAGE P-3
 PROJECT PRO/II VERSION 8.1 ELEC V6.6
 PROBLEM OUTPUT
 VALVE SUMMARY 01/10/09

VALVE ID	VALVE
NAME	
FEEDS	RICH_SOLVENT
PRODUCTS MIXED	S5
TEMPERATURE, C	43.373
PRESSURE, ATM	1.000
PRESSURE DROP, ATM	17.050
MOLE FRAC VAPOR	3.76233E-04
MOLE FRAC LIQUID	0.99962
MOLE FRAC MW SOLID	0.00000

SIMULATION SCIENCES INC. R PAGE P-4
 PROJECT PRO/II VERSION 8.1 ELEC V6.6
 PROBLEM OUTPUT
 MIXER SUMMARY 01/10/09

MIXER ID	MX-1	MX-2
NAME	MEA-MAKE UP	H2O MAKE UP
FEEDS	S13	S11
	S1	S16
PRODUCTS LIQUID	S2	S19
TEMPERATURE, C	70.761	56.137
PRESSURE, ATM	0.990	0.990
PRESSURE DROP, ATM	1.000E-02	0.000
MOLE FRAC VAPOR	0.00000	0.00000
MOLE FRAC LIQUID	1.00000	1.00000
MOLE FRAC MW SOLID	0.00000	0.00000

SIMULATION SCIENCES INC. R PAGE P-5
 PROJECT PRO/II VERSION 8.1 ELEC V6.6
 PROBLEM OUTPUT
 SPLITTER SUMMARY 01/10/09

UNIT 10, 'SP-1'

	STREAM ID	FRACTION	----- RATES -----
	-----	-----	-----
			KG-MOL/HR KG/HR
	-----	-----	-----
FEED	MEA		0.620 37.872
PRODUCTS	S13	0.0019	0.001 0.073
	XS-MEA	0.9981	0.619 37.799
TEMPERATURE, C		28.0000	
PRESSURE, ATM		1.0000	
PRESSURE DROP, ATM		0.0000	
MOLE FRAC VAPOR		0.0000	
MOLE FRAC TOTAL LIQUID		1.0000	
MOLE FRAC MW SOLID		0.0000	

UNIT 11, 'SP-2'

	STREAM ID	FRACTION	----- RATES -----
	-----	-----	-----
			KG-MOL/HR KG/HR
	-----	-----	-----
FEED	H2O		999999.976 1.802E+07
PRODUCTS	S16	0.0001	65.833 1185.993
	XS-H2O	0.9999	999934.143 1.801E+07
TEMPERATURE, C		70.0000	
PRESSURE, ATM		1.0000	
PRESSURE DROP, ATM		0.0000	
MOLE FRAC VAPOR		0.0000	
MOLE FRAC TOTAL LIQUID		1.0000	
MOLE FRAC MW SOLID		0.0000	

SIMULATION SCIENCES INC. R PAGE P-6
PROJECT PRO/II VERSION 8.1 ELEC V6.6
PROBLEM OUTPUT
PUMP SUMMARY 01/10/09

UNIT 7, 'MEA_H2O_PUMP'

Feeds S19
Products Liquid S20

OPERATING CONDITIONS

	INLET	OUTLET
	-----	-----
TEMPERATURE, C	56.14	56.66
PRESSURE, ATM	0.99	17.00
MOLE FRAC VAPOR	0.0000	0.0000
MOLE FRAC LIQUID	1.0000	1.0000
MOLE FRAC MW SOLID	0.0000	0.0000
WEIGHT FRAC MW SOLID	0.0000	0.0000
ACT FLOW RATE, M3/HR	129.5316	129.5660
EFFICIENCY, PERCENT		80.0000
HEAD, M		169.4100
WORK, KW		72.9611

SIMULATION SCIENCES INC. R PAGE P-7
 PROJECT PRO/II VERSION 8.1 ELEC V6.6
 PROBLEM OUTPUT
 HEAT EXCHANGER SUMMARY 01/10/09

UNIT 3, 'HX-1'

OPERATING CONDITIONS

DUTY, M*KJ/HR	17.843
LMTD, C	28.227
F FACTOR (FT)	0.661
MTD, C	18.652
U*A, KW/K	265.731

HOT SIDE CONDITIONS	INLET	OUTLET
	-----	-----
FEED	S7	
LIQUID PRODUCT		S1
VAPOR, KG-MOL/HR	0.539	
K*KG/HR	9.786E-03	
CP, KJ/KG-C	1.902	
LIQUID, KG-MOL/HR	6349.585	6350.124
K*KG/HR	125.371	125.380
CP, KJ/KG-C	4.049	4.002
TOTAL, KG-MOL/HR	6350.124	6350.124
K*KG/HR	125.380	125.380
CONDENSATION, KG-MOL/HR		0.539
TEMPERATURE, C	106.083	70.761
PRESSURE, ATM	1.191	1.191

COLD SIDE CONDITIONS	INLET	OUTLET
	-----	-----
FEED	S5	
MIXED PRODUCT		S6
VAPOR, KG-MOL/HR	2.461	18.641
K*KG/HR	7.367E-02	0.601
CP, KJ/KG-C	1.034	1.134
LIQUID, KG-MOL/HR	6538.488	6522.308
K*KG/HR	132.013	131.486
CP, KJ/KG-C	3.894	3.924
TOTAL, KG-MOL/HR	6540.949	6540.949
K*KG/HR	132.087	132.087
VAPORIZATION, KG-MOL/HR		16.180
TEMPERATURE, C	43.373	77.000

PRESSURE, ATM 1.000 1.000

SIMULATION SCIENCES INC. R PAGE P-8
 PROJECT PRO/II VERSION 8.1 ELEC V6.6
 PROBLEM OUTPUT
 HEAT EXCHANGER SUMMARY 01/10/09

=====

UNIT 4, 'HX-2'

OPERATING CONDITIONS

DUTY, M*KJ/HR 7.395

HOT SIDE CONDITIONS

	INLET	OUTLET
	-----	-----
FEED	S2	
LIQUID PRODUCT		S11
LIQUID, KG-MOL/HR	6350.125	6350.125
K*KG/HR	125.380	125.380
CP, KJ/KG-C	4.002	3.990
TOTAL, KG-MOL/HR	6350.125	6350.125
K*KG/HR	125.380	125.380
CONDENSATION, KG-MOL/HR		0.000
TEMPERATURE, C	70.761	56.000
PRESSURE, ATM	0.990	0.990

UNIT 5, 'HE-3'

OPERATING CONDITIONS

DUTY, M*KJ/HR 12.443

HOT SIDE CONDITIONS

	INLET	OUTLET
	-----	-----
FEED	S20	
LIQUID PRODUCT		MEA_H2O
LIQUID, KG-MOL/HR	6415.957	6415.957
K*KG/HR	126.566	126.566
CP, KJ/KG-C	3.992	3.985
TOTAL, KG-MOL/HR	6415.957	6415.957
K*KG/HR	126.566	126.566

CONDENSATION, KG-MOL/HR		0.000
TEMPERATURE, C	56.657	32.000
PRESSURE, ATM	17.000	17.000

SIMULATION SCIENCES INC.	R	PAGE P-9
PROJECT	PRO/II VERSION 8.1 ELEC V6.6	
PROBLEM	OUTPUT	
	CALCULATOR SUMMARY	01/10/09

UNIT 12, 'CALCULATOR'

Result	Name	Value	Result	Name	Value
1	R1	6.58326E+01	3-	200	Undefined
2	R2	1.19997E-03			

Parameter	Value	Parameter	Value
1	1.92003E+00	4	1.15561E-14
2	6.39126E+01	5-	50
3	1.19997E-03		Undefined

SIMULATION SCIENCES INC.	R	PAGE P-10
PROJECT	PRO/II VERSION 8.1 ELEC V6.6	
PROBLEM	OUTPUT	
	COLUMN SUMMARY	01/10/09

UNIT 1, 'T1', 'ABSORBER'

TOTAL NUMBER OF ITERATIONS

IN/OUT METHOD 43

COLUMN SUMMARY

TRAY	TEMP DEG C	PRESSURE ATM	----- NET FLOW RATES -----			HEATER DUTIES M*KJ/HR
			LIQUID	VAPOR	FEED PRODUCT	
			KG-MOL/HR	KG-MOL/HR	KG-MOL/HR	
1	32.0	18.00	6417.7		6416.0L	767.3V
2	32.0	18.01	6417.7	769.1		
3	32.0	18.02	6417.7	769.1		
4	32.1	18.03	6417.9	769.1		
5	33.9	18.04	6425.8	769.3		
6	43.4	18.05		777.2	892.3V	6540.9L

FEED AND PRODUCT STREAMS

TYPE	STREAM	PHASE	FROM TO		LIQUID FRAC	FLOW RATES KG-MOL/HR	HEAT RATES M*KJ/HR
			TRAY	TRAY			
FEED	MEA_H2O	LIQUID		1	1.0000	6415.96	14.4079
FEED	FLUE_GAS	VAPOR		6	0.0000	892.30	1.0005
PROD	TREATED_GAS	VAPOR	1			767.31	0.0634
PROD	RICH_SOLVENT	LIQUID	6			6540.95	15.3450
OVERALL MOLE BALANCE, (FEEDS - PRODUCTS)						0.00	
OVERALL HEAT BALANCE, (H(IN) - H(OUT))							-6.4862E-06

SIMULATION SCIENCES INC. R PAGE P-11
 PROJECT PRO/II VERSION 8.1 ELEC V6.6
 PROBLEM OUTPUT
 COLUMN SUMMARY 01/10/09

UNIT 1, 'T1', 'ABSORBER' (Cont)

TRAY NET VAPOR RATES AND DENSITIES

TRAY	MW	ACTUAL DENS KG/M3	Z FROM DENSITY	RATES		
				K*KG/HR	NORMAL K*M3/HR	ACTUAL K*M3/HR
1	28.265	20.21505	1.00514	21.688	17.198	1.073
2	28.266	20.22671	1.00514	21.738	17.238	1.075
3	28.266	20.23751	1.00514	21.738	17.238	1.074
4	28.266	20.24163	1.00516	21.738	17.238	1.074
5	28.263	20.12740	1.00538	21.742	17.242	1.080
6	28.377	19.60016	1.00605	22.054	17.420	1.125

TRAY NET LIQUID RATES AND DENSITIES

TRAY	MW	ACTUAL DENS KG/M3	Z FROM DENSITY	RATES		
				K*KG/HR	STD LIQ M3/HR	ACTUAL M3/HR
1	19.729	987.285	0.01437	126.617	126.630	128.248
2	19.729	987.284	0.01437	126.617	126.630	128.248
3	19.729	987.282	0.01438	126.617	126.630	128.248
4	19.729	987.241	0.01438	126.621	126.634	128.257
5	19.754	984.312	0.01437	126.933	127.010	128.956
6	20.194	943.927	0.01487	132.087	133.338	139.933

TRAY TRANSPORT PROPERTIES

TRAY	THERMAL CONDUCTIVITY -		VISCOSITY		SURFACE -- TENSION N/M
	W/M-K		PAS		
	LIQUID	VAPOR	LIQUID	VAPOR	
1	4.2236E-01	2.6041E-02	9.4282E-04	1.8236E-05	7.0297E-02
2	4.2235E-01	2.6041E-02	9.4282E-04	1.8237E-05	7.0297E-02
3	4.2235E-01	2.6041E-02	9.4270E-04	1.8237E-05	7.0296E-02
4	4.2240E-01	2.6048E-02	9.4058E-04	1.8242E-05	7.0276E-02
5	4.0939E-01	2.6175E-02	9.0209E-04	1.8322E-05	6.9855E-02

6 2.8182E-01 2.6752E-02 7.1509E-04 1.8710E-05 6.6791E-02

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UNIT 2, 'T2'

TOTAL NUMBER OF ITERATIONS

IN/OUT METHOD 230

COLUMN SUMMARY

TRAY	TEMP DEG C	PRESSURE ATM	----- NET FLOW RATES -----			PRODUCT M*KJ/HR	HEATER DUTIES M*KJ/HR
			LIQUID KG-MOL/HR	VAPOR KG-MOL/HR	FEED		
1C	70.0	0.92	2471.6			190.8V	-107.2521
2	100.1	1.05	2607.7	2662.4			
3	100.6	1.07	2609.0	2798.5			
4	100.9	1.08	2610.0	2799.8			
5	101.3	1.10	2610.8	2800.8			
6	101.7	1.11	2608.4	2801.6			
7	102.9	1.12	9454.8	2799.2	6540.9M		
8	104.1	1.14	9487.3	3104.7			
9	104.6	1.15	9503.4	3137.1			
10	105.0	1.16	9515.8	3153.3			
11	105.4	1.18	9525.2	3165.6			
12R	106.1	1.19	3175.1			6350.1L	130.4582

FEED AND PRODUCT STREAMS

TYPE	STREAM	PHASE	FROM TRAY	TO TRAY	LIQUID FRAC	FLOW RATES KG-MOL/HR	HEAT RATES M*KJ/HR
FEED	S6	MIXED		7	0.9978	6540.95	33.1879
PROD	CO2	VAPOR	1			190.82	4.9384
PROD	S7	LIQUID	12			6350.13	51.4563

OVERALL MOLE BALANCE, (FEEDS - PRODUCTS) -7.9936E-13
 OVERALL HEAT BALANCE, (H(IN) - H(OUT)) -7.4264E-04

SPECIFICATIONS

SPECIFICATION CALCULATED	PARAMETER	TRAY	COMP	SPECIFICATION	SPECIFIED	
NUMBER	TYPE	NO	NO	TYPE	VALUE	VALUE
1 (ACTIVE)	TRAY LIQ	1		TEMPERATURE	7.000E+01	7.000E+01
2 (ACTIVE)	TRAY VAP	12		MOL RATIO	5.000E-01	5.000E-01

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UNIT 2, 'T2' (Cont)

REFLUX RATIOS

	REFLUX RATIOS		
	MOLAR	WEIGHT	STD L VOL
REFLUX / FEED STREAM S6	0.3779	0.3372	0.3344
REFLUX / VAPOR DISTILLATE	12.9525	6.6417	5.6058

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UNIT 2, 'T2' (Cont)

TRAY NET VAPOR RATES AND DENSITIES

TRAY	MW	ACTUAL DENS KG/M3	Z FROM DENSITY	RATES		
				K*KG/HR	NORMAL K*M3/HR	ACTUAL K*M3/HR
1	35.142	1.15165	0.99549	6.706	4.277	5.823
2	19.247	0.66838	0.99154	51.244	59.676	76.669
3	19.183	0.67405	0.99144	53.685	62.726	79.646
4	19.183	0.68202	0.99135	53.709	62.755	78.749
5	19.182	0.68999	0.99127	53.726	62.777	77.864
6	19.184	0.69794	0.99118	53.746	62.796	77.006
7	19.236	0.70618	0.99114	53.846	62.742	76.250
8	18.456	0.68379	0.99085	57.300	69.589	83.798
9	18.287	0.68481	0.99072	57.368	70.316	83.772
10	18.215	0.68954	0.99061	57.438	70.678	83.299
11	18.172	0.69531	0.99051	57.524	70.954	82.732
12	18.171	0.70204	0.99042	57.693	71.166	82.179

TRAY NET LIQUID RATES AND DENSITIES

TRAY	MW	ACTUAL DENS KG/M3	Z FROM DENSITY	RATES		
				K*KG/HR	STD LIQ M3/HR	ACTUAL M3/HR
1	18.020	975.077	6.03E-04	44.538	44.586	45.676
2	18.016	957.430	6.48E-04	46.979	47.026	49.068
3	18.016	957.142	6.56E-04	47.003	47.049	49.107
4	18.016	956.902	6.64E-04	47.020	47.067	49.138
5	18.017	956.613	6.72E-04	47.040	47.087	49.173
6	18.073	955.177	6.82E-04	47.140	47.198	49.353
7	19.321	941.653	7.47E-04	182.681	183.193	194.000
8	19.263	945.122	7.48E-04	182.749	183.058	193.361
9	19.237	946.656	7.54E-04	182.819	183.039	193.121
10	19.221	947.566	7.61E-04	182.906	183.069	193.027
11	19.220	948.251	7.68E-04	183.074	183.190	193.065
12	19.745	947.185	7.98E-04	125.381	125.384	132.372

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 PROBLEM OUTPUT
 COLUMN SUMMARY 01/10/09

UNIT 2, 'T2' (Cont)

TRAY TRANSPORT PROPERTIES

	TRAY - THERMAL CONDUCTIVITY -		----- VISCOSITY -----		SURFACE --
	W/M-K		PAS		TENSION
	LIQUID	VAPOR	LIQUID	VAPOR	N/M
1	6.4638E-01	2.0660E-02	4.0768E-04	1.5714E-05	6.4001E-02
2	6.7572E-01	2.4352E-02	2.7907E-04	1.2935E-05	5.8185E-02
3	6.7595E-01	2.4394E-02	2.7776E-04	1.2929E-05	5.8101E-02
4	6.7610E-01	2.4424E-02	2.7668E-04	1.2943E-05	5.8031E-02
5	6.7454E-01	2.4454E-02	2.7562E-04	1.2956E-05	5.7961E-02
6	6.3585E-01	2.4486E-02	2.7491E-04	1.2971E-05	5.7836E-02
7	4.0360E-01	2.4579E-02	2.8722E-04	1.3015E-05	5.6788E-02
8	4.3531E-01	2.4752E-02	2.8459E-04	1.2778E-05	5.6689E-02
9	4.5178E-01	2.4812E-02	2.8350E-04	1.2735E-05	5.6647E-02
10	4.6316E-01	2.4853E-02	2.8255E-04	1.2723E-05	5.6604E-02
11	4.7230E-01	2.4889E-02	2.8182E-04	1.2719E-05	5.6554E-02
12	4.4325E-01	2.4947E-02	2.8840E-04	1.2727E-05	5.6189E-02

MOLE FRAC LIQUID 1.0000 1.0000 1.0000 1.0000

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 PROBLEM OUTPUT
 STREAM MOLAR COMPONENT RATES 01/10/09

STREAM ID	S5	S6	S7	S11
NAME				
PHASE	MIXED	MIXED	LIQUID	LIQUID
THERMO ID	AMIN01	AMIN01	AMIN01	AMIN01

FLUID RATES, KG-MOL/HR

1 CO2	142.9476	142.9476	18.0279	18.0204
2 N2	1.7310	1.7310	0.0000	0.0000
3 O2	0.2582	0.2582	0.0000	0.0000
4 H2O	6151.9079	6151.9079	6087.9954	6088.0110
5 MEA	244.1046	244.1046	244.1046	244.0933

TOTAL RATE, KG-MOL/HR 6540.9494 6540.9494 6350.1279 6350.1247

TEMPERATURE, C	43.3727	77.0000	106.0832	56.0000
PRESSURE, ATM	1.0000	1.0000	1.1908	0.9900
ENTHALPY, M*KJ/HR	15.3450	33.1879	51.4563	26.2406
MOLECULAR WEIGHT	20.1939	20.1939	19.7447	19.7446
MOLE FRAC VAPOR	3.7623E-04	2.8499E-03	0.0000	0.0000
MOLE FRAC LIQUID	0.9996	0.9972	1.0000	1.0000

STREAM ID	S13	S16	S19	S20
NAME				
PHASE	LIQUID	LIQUID	LIQUID	LIQUID
THERMO ID	AMIN01	AMIN01	AMIN01	AMIN01

FLUID RATES, KG-MOL/HR

1 CO2	0.0000	0.0000	18.0204	18.0204
2 N2	0.0000	0.0000	0.0000	0.0000
3 O2	0.0000	0.0000	0.0000	0.0000
4 H2O	0.0000	65.8326	6153.8436	6153.8436
5 MEA	1.2000E-03	0.0000	244.0933	244.0933

TOTAL RATE, KG-MOL/HR 1.2000E-03 65.8326 6415.9573 6415.9573

TEMPERATURE, C	28.0000	70.0000	56.1374	56.6573
PRESSURE, ATM	1.0000	1.0000	0.9900	17.0000
ENTHALPY, M*KJ/HR	5.3605E-06	0.3476	26.5883	26.8509
MOLECULAR WEIGHT	61.0837	18.0153	19.7268	19.7268
MOLE FRAC VAPOR	0.0000	0.0000	0.0000	0.0000
MOLE FRAC LIQUID	1.0000	1.0000	1.0000	1.0000

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 PROBLEM OUTPUT
 STREAM MOLAR COMPONENT RATES 01/10/09

STREAM ID	TREATED_GAS	XS-H2O	XS-MEA
NAME			
PHASE	VAPOR	LIQUID	LIQUID
THERMO ID	AMIN01	AMIN01	AMIN01

FLUID RATES, KG-MOL/HR

1 CO2	1.2189E-04	0.0000	0.0000
2 N2	712.1090	0.0000	0.0000
3 O2	53.2798	0.0000	0.0000
4 H2O	1.9200	999934.1430	0.0000
5 MEA	1.2000E-03	0.0000	0.6188

TOTAL RATE, KG-MOL/HR	767.3101	999934.1430	0.6188
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TEMPERATURE, C	31.9961	70.0000	28.0000
PRESSURE, ATM	18.0000	1.0000	1.0000
ENTHALPY, M*KJ/HR	0.0634	5280.4624	2.7643E-03
MOLECULAR WEIGHT	28.2652	18.0153	61.0837
MOLE FRAC VAPOR	1.0000	0.0000	0.0000
MOLE FRAC LIQUID	0.0000	1.0000	1.0000

K*KG/HR	N/A	N/A	18015.279	3.787E-02
M3/HR	N/A	N/A	18468.101	3.750E-02
GAL/MIN	N/A	N/A	81312.495	0.165
STD LIQ RATE, M3/HR	N/A	N/A	18033.060	3.715E-02
SPECIFIC GRAVITY (H2O=1.0)	N/A	N/A	1.0000	1.0204
MOLECULAR WEIGHT	N/A	N/A	18.015	61.084
ENTHALPY, KJ/KG	N/A	N/A	293.130	73.132
CP, KJ/KG-C	N/A	N/A	4.187	2.648
DENSITY, KG/M3	N/A	N/A	975.481	1010.034
Z (FROM DENSITY)	N/A	N/A	6.5587E-04	2.4473E-03
SURFACE TENSION, N/M	N/A	N/A	0.0640	0.0480
THERMAL COND, W/M-K	N/A	N/A	0.65670	0.23720
VISCOSITY, PAS	N/A	N/A	4.07817E-04	0.01788

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

----- LIQUID -----				
RATE, KG-MOL/HR	6415.957	6540.949	6350.124	6350.125
K*KG/HR	126.566	132.087	125.380	125.380
M3/HR	128.107	139.933	129.343	129.344
GAL/MIN	564.036	616.108	569.481	569.482
STD LIQ RATE, M3/HR	126.571	133.338	125.383	125.383
SPECIFIC GRAVITY (H2O=1.0)	1.0010	0.9916	1.0010	1.0010
MOLECULAR WEIGHT	19.727	20.194	19.745	19.745
ENTHALPY, KJ/KG	113.840	116.173	268.271	268.271
CP, KJ/KG-C	3.985	3.893	4.002	4.002
DENSITY, KG/M3	987.977	943.927	969.360	969.360
Z (FROM DENSITY)	0.0136	0.0149	8.5948E-04	7.1455E-04
SURFACE TENSION, N/M	0.0703	0.0668	0.0629	0.0629
THERMAL COND, W/M-K	0.42664	0.28182	0.43952	0.43952
VISCOSITY, PAS	9.43363E-04	7.15094E-04	4.59139E-04	4.59140E-04

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

----- LIQUID -----				
RATE, KG-MOL/HR	6538.488	6522.308	6350.128	6350.125
K*KG/HR	132.013	131.486	125.381	125.380
M3/HR	139.743	141.106	132.372	128.315
GAL/MIN	615.269	621.269	582.817	564.955
STD LIQ RATE, M3/HR	133.250	132.633	125.384	125.383
SPECIFIC GRAVITY (H2O=1.0)	0.9917	0.9923	1.0010	1.0010
MOLECULAR WEIGHT	20.190	20.159	19.745	19.745
ENTHALPY, KJ/KG	116.121	248.638	410.399	209.288
CP, KJ/KG-C	3.894	3.924	4.049	3.990
DENSITY, KG/M3	944.687	931.826	947.185	977.127
Z (FROM DENSITY)	8.2287E-04	7.5296E-04	7.9767E-04	7.4066E-04
SURFACE TENSION, N/M	0.0668	0.0606	0.0562	0.0657
THERMAL COND, W/M-K	0.28333	0.29192	0.44325	0.43547
VISCOSITY, PAS	7.16033E-04	4.03700E-04	2.88404E-04	5.86050E-04

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

----- LIQUID -----

RATE, KG-MOL/HR	1.200E-03	65.833	6415.957	6415.957
K*KG/HR	7.330E-05	1.186	126.566	126.566
M3/HR	7.257E-05	1.216	129.532	129.566
GAL/MIN	3.195E-04	5.353	570.310	570.461
STD LIQ RATE, M3/HR	7.190E-05	1.187	126.571	126.571
SPECIFIC GRAVITY (H2O=1.0)	1.0204	1.0000	1.0010	1.0010
MOLECULAR WEIGHT	61.084	18.015	19.727	19.727
ENTHALPY, KJ/KG	73.132	293.130	210.074	212.149
CP, KJ/KG-C	2.648	4.187	3.992	3.992
DENSITY, KG/M3	1010.034	975.481	977.108	976.849
Z (FROM DENSITY)	2.4473E-03	6.5587E-04	7.3970E-04	0.0127
SURFACE TENSION, N/M	0.0480	0.0640	0.0657	0.0656
THERMAL COND, W/M-K	0.23720	0.65670	0.43662	0.43679
VISCOSITY, PAS	0.01788	4.07817E-04	5.83765E-04	5.78439E-04

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

GAL/MIN	N/A	81307.142	0.165
STD LIQ RATE, M3/HR	N/A	18031.873	3.708E-02
SPECIFIC GRAVITY (H2O=1.0)	N/A	1.0000	1.0204
MOLECULAR WEIGHT	N/A	18.015	61.084
ENTHALPY, KJ/KG	N/A	293.130	73.132
CP, KJ/KG-C	N/A	4.187	2.648
DENSITY, KG/M3	N/A	975.481	1010.034
Z (FROM DENSITY)	N/A	6.5587E-04	2.4473E-03
SURFACE TENSION, N/M	N/A	0.0640	0.0480
THERMAL COND, W/M-K	N/A	0.65670	0.23720
VISCOSITY, PAS	N/A	4.07817E-04	0.01788

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

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Here follows the list of books and symposiums consulted in the course of writing this project.

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3.

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Title: The Intergovernmental Panel on Climate Change special report on CO_2 capture and storage.

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