Conversion of Existing Bubble Cap Column to Packed Bed Column

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Certificate

This is to certify that work in this thesis has been carried out by **Mr. Hassan Nadeem, Mr. Ali Qayum and Mr. Syed Minhas Mehmood and** completed under my supervision in School of Chemical and Materials Engineering, National University of Sciences and Technology, H-12, Islamabad, Pakistan.

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DEDICATION

TO

OUR PARENTS

without whom none of this would have been possible and for their support throughout our lives

AND TEACHERS

for inspiring us and supporting us throughout the entirety of the project

ACKNOWLEDGEMENT

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Author

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Nomenclature

- Cp Specific heat capacity at constant pressure
- Cv Specific heat capacity at constant volume
- P Pressure
- P Power
- T Temperature
- v volumetric flowrate
- n Number of moles
- R Universal gas constant
- ρ Mass density
- z Gas compressibility factor
- M Relative molecular mass
- H Enthalpy
- H Head
- m mass flowrate
- Q Heat flow
- Δ change in
- d Diameter
- A Area
- π pi
- Re Reynolds number
- μ Viscosity
- Pr Prandtl number
- k Thermal conductivity
- Nu Nusselt number
- Jh Chilton and Colburn factor
- h Heat transfer coefficient
- U Overall heat transfer coefficient
- l Length
- G Mass velocity

- f Fanning friction factor
- g Gravitational constant
- ΔFp Pressure drop factor
- γ Specific gravity
- Cd Drag force coefficient
- u Velocity
- NLL Normal liquid level
- t time
- Q Fluid flowrate
- H Height
- G Flowrate of gas
- L Flowrate of liquid

Abstract

ICI Soda Ash is the major producer of Soda Ash in the country and the absorption of ammonia into the brine is one of the major processes for the manufacturing of Soda Ash. The company is using Absorber with single Bubble Cap Trays, which is an obsolete technology and has is being eventually replaced with several other different packing in chemical industries around the world.

This project is based on the feasibility analysis and conversion of the bubble cap trays to packed bed column. After literature review, consideration of current industrial practices and consultation with both supervisors (industrial and project) and calculated the performance of the process on the existing parameters. The simulation of the process is also done on the existing parameters and the design and costing of packed bed column and a comparison is drawn within different packing. After performing the proposed modification, it has been concluded that by using packed bed column not only the efficiency can be increased but space, which the column occupies, can also be saved. This conversion helps to minimize losses and improve the unit's yield.

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Chapter-1

Introduction

1.1. Background

ICI Soda Ash is one of the biggest Soda Ash producing Industry in Pakistan with a yield of 250 tons/day. The vitality prerequisites of the plant are met by their own steam control plants which are as of now being kept running on the coal imported from South Africa. Absorption is one of the main unit operations involved in the Solvay's process for Soda Ash Manufacturing. Brine purification plant is sent to the absorbers for ammoniation. The purpose of the absorber is to absorb ammonia into the brine in the desired amount to produce ammoniated brine before its reaction with carbon dioxide in carbonating towers.

1.2. History of ICI

The company, ICI (Imperial Chemical Industries) Pakistan Limited, is a well-known Chemical Manufacturing Industry with its headquarters in Karachi. The company was founded back in 1952 and has four major businesses; Soda, Polyester, Life sciences and Chemicals. The project mentioned in this thesis is being done on one of the sections of the Soda Ash Industry, which is ICI's Soda Ash Plant located in Khewera, Pakistan. It is one of the largest Soda Ash production plant in the country with over 250 tons/day. Light and dense Soda Ash is produced on the plant through Solvay's process. There are several coal burning boilers present at the plant in order to meet the energy requirements for the processes.

1.3. Solvay's Process

The process is widely used to manufacture Calcium Carbonate and this is also known as Ammonia-Soda Ash process. The raw materials used in this process are brine, limestone and ammonia. The overall equation of the process is as following:

Step 1

$$2NaCl + CaCO_3 \rightarrow Na_2CO_3 + CaCl_2$$

The total process is described in four different steps. During the first step the carbon dioxide passes through the concentrated solution of sodium chloride and ammonia. In the Industry, this step is done by passing brine over the two towers. In the first tower the ammonia gas is dissolved in the brine and carbon dioxide bubbling through the ammoniated brine forms sodium bicarbonate precipitates in the second tower. This is presented in the reaction below:

Step 2

$$NaCl + CO_2NH_3 + H2O \rightarrow NaHCO_3 + NH_4Cl$$

The ammonia, thought of as a catalyst, is regenerated in later step and is very little consumed for the process. When the limestone is heated, also known as calcination reaction, the carbon dioxide is produced at 950-1050 ℃.

Step 3

$$CaCO_3 \rightarrow CO_2 + CaO$$

During step 3, the sodium bicarbonate produced in step 1 after being filtered out of chloride solution is reacted with Calcium Oxide produced in step 2. This is presented in the following reaction:

Step 4

$$2\,NH_4Cl + CaO \rightarrow 2\,NH_3 + CaCl_2 + H_2O$$

The Carbon dioxide produced is recycled so that it can be used in step 1. The process should have minimal loses of ammonia so that it may be used over and over again. Case salt, limestone and thermal energy are the major inputs of this process and Calcium chloride is the major by product. Calcium chloride is currently not being used at the plant and several beds are buried throughout the plant. There are several coil boilers present which meets the energy requirements of the plant.

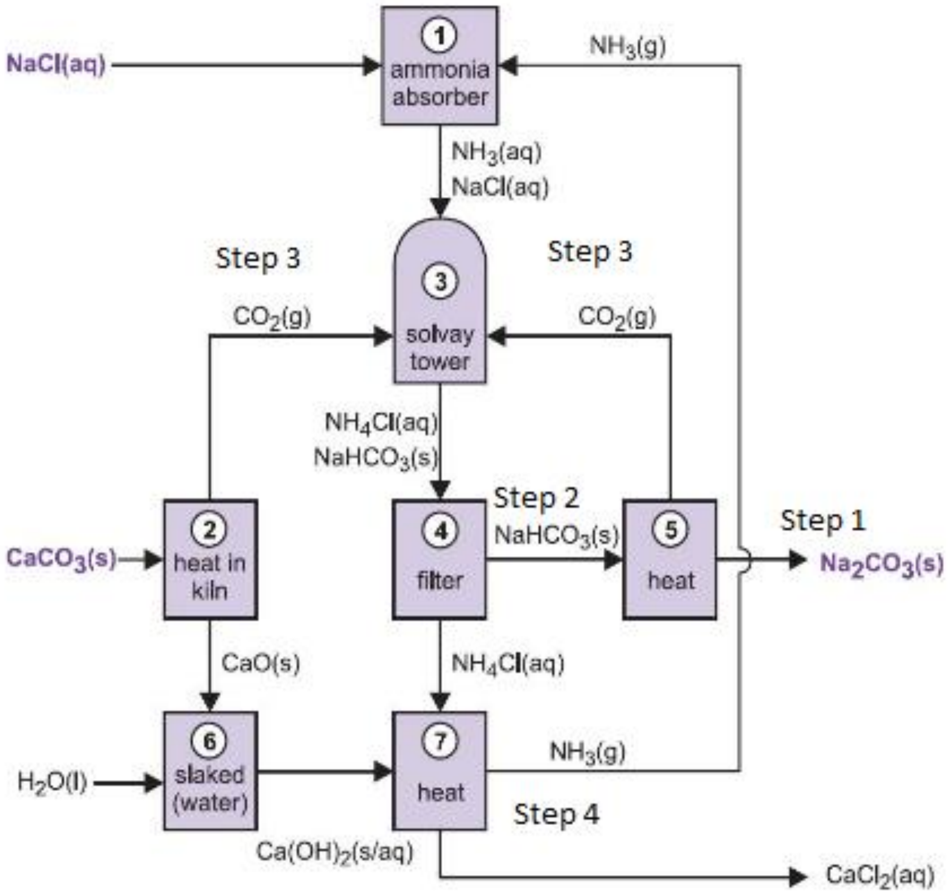


Figure 1 Solvay's Process

1.4. Project/Study Objective

The purpose of this project was to propose different packing of the absorber in order to increase the absorption of ammonia gas into the brine. ICI Soda Ash is using bubble cap packing since 1940 and the technology has become obsolete and by changing bubble cap column to packed bed column the absorption of ammonia can increase and hence increase the efficiency of overall system which in turn will increase the yield of the Soda Ash produce. Just by changing the packing of the column overall economics can be increased.

1.5. Organization of remainder of the study

The authors of this thesis have done their best in arranging each and every detail in the process outlined. It is trusted that the substance of this report is adequate for anybody hoping to actualize this procedure on a modern scale.

Chapter-2

Literature Review

2.1. Absorber:

In the Absorber, the vent gases are dissolved in the counter current way. The vast majority of NH_3 is dissolved, while other is expelled with fumes gasses. In ingestion significant mass exchange happens from gas stage to fluid stage. The following are the primary reasons for absorption:

- To separate undesirable compound from the system
- To prepare a valuable compound
- To separate a compound for economic purposes

2.2. Physical Absorption:

In physical absorption, the significant mass exchange occurs from the gas stage to the fluid stage exclusively by the diffusion and physical absorption.

The following are the two main types of absorbers present:

- Packed Absorbers segment
- Plate Absorbers segment

The following is the comparison between packed and plate absorber:

- The plate absorbers are utilized for vast fluid and gas stream rates.
- Whereas the packed absorbers are utilized for little width, in light of the fact that for small measurements plates are hard to introduce and clean.
- The productivity of plates segments are anything but difficult to foresee than in the case of packed absorbers.
- Plate absorbers are anything but difficult to work in light of the fact that in the packed one it is troublesome to keep up fluid conveyance over the packing.
- Cooling can be effortlessly controlled in the plate sections with the use of cooling jackets.
- In the plate sections, it is simpler to expel the side-streams.

- For fluid containing suspended solids, plate absorbers are normally utilized. Since they are all the more effortlessly to clean then the pressed sections. For small measurements, fluid having suspended particles, pressed segments are utilized. The packing is supplanted when they are fouled.
- For destructive fluids in the packed ones are less expensive than the plate ones.
- Packed absorbers have vast fluid hold up than the plate segments. Fluid hold-up is essential to study when fluid is dangerous or combustible.
- For frothing frameworks, packed ones should be utilized.
- Pressure drop per packing is brought down in the packed one than in the equal plate column.

2.3. Problems in the Absorber:

2.3.1. Flooding:

As the gas reaches the maximum velocity in the column at that point it is said to be flooded. During this phenomenon the gas gives high resistance to the liquid, which is coming from the top of the column, and thus little mass transfer from gas phase to liquid phase. High pressure of the gas damages the tray, so actual velocity of the gas should be fifty percent of the flooding velocity to have effective mass transfer from the gas to the liquid phase.

2.3.2. Weeping:

It is the phenomenon during which the liquid drops down the holes of the tray column. The tray is unable to hold the liquid on the trays and liquid falls down the tray. It often occurs in the tray column.

2.3.3. Channeling:

In the packed columns the liquid that is coming from the top makes some pathways and leave few areas of packing dry, due to the low liquid flow rates. Due to this phenomenon, there is very low amount of heat and mass transfer. So the fluids are evenly distributed throughout the packing in order avoid channeling.

2.4. Process Description

After the brine is being purified, it is pumped from the reservoir to top most compartment of the tower washer section of the absorber. At the top of the tower washer, there is after washer where the cold purified water is used to scrub the final traces of salt and ammonia that rises up the column.

The brine is branched of Absorber Vacuum Washer (AVW), absorber proper and Mono-Carbonating Tower washer (MCT), before entering the tower washer. Within the tower washer the brine flows down the column due to gravity and maintains the certain bubbling depth of brine in each compartment. The waste gases coming out of the carbonating tower enters in the tower washer through the gas compartment. The liquor droplets carried over with these gases from the towers are separated out in the gas compartment and then drained into the CVL stocks through "Tower Washer run-off main."

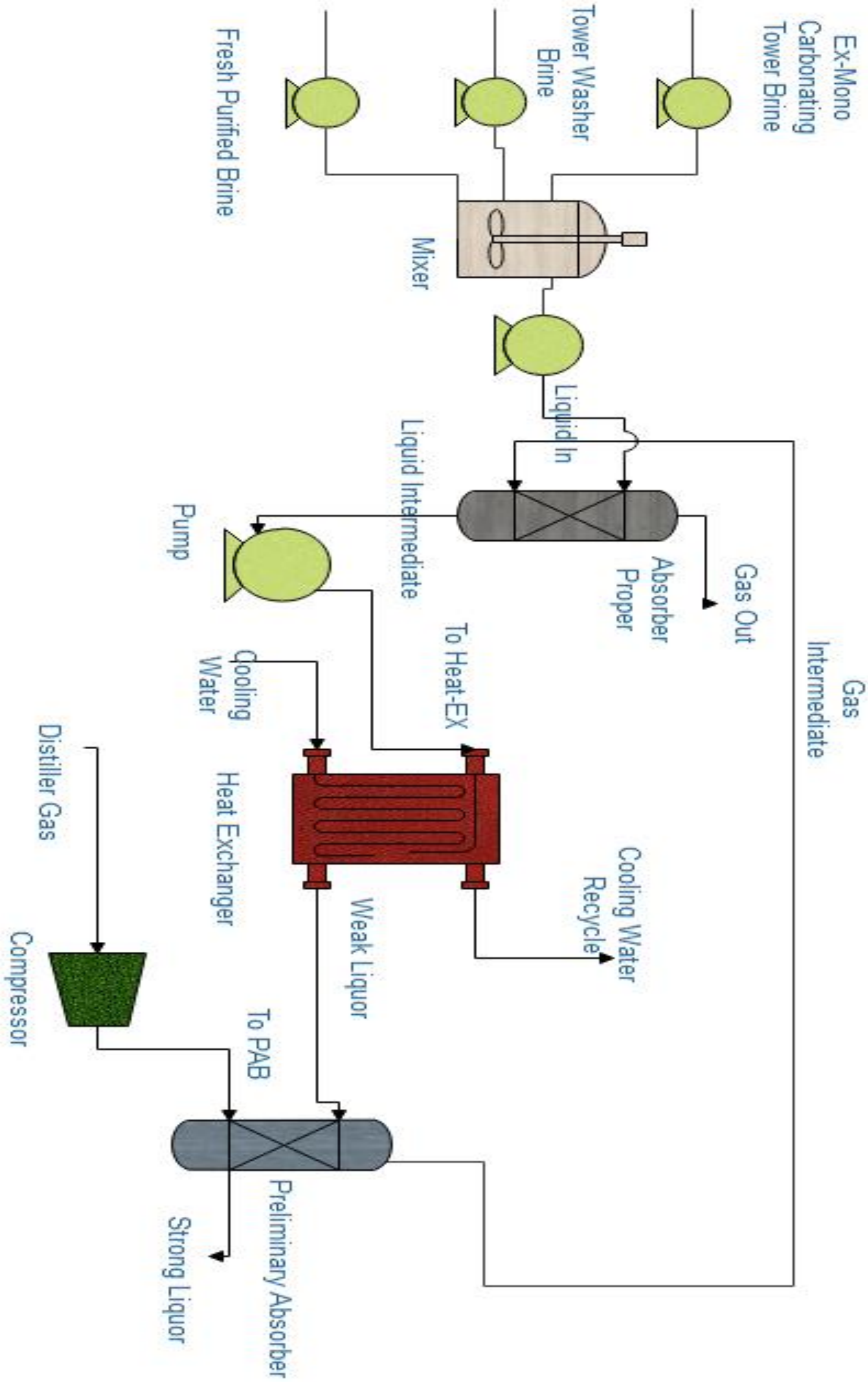
The gases rising up through the bubble cap are absorbed within the brine coming down. The brine that is coming down absorbs ammonia and the remaining gases vent to the atmosphere through a knock-out box. The exit gas temperature is maintained by regulating the flow of the brine.

As the absorption process is exothermic and as it comes in contact with the hot gases within the tower, the brine leaving the tower washer becomes hot. In order to carry out further absorption, the brine needs to be cooled down and for this purpose it is passed through the plate heat exchangers.

The vacuum pull is applied in order to pull the distiller gases, which are being fed to the gas compartment, in the 3 compartments of AVW on the top of the absorber proper. As the brine absorbs ammonia the heat is generated. The ammoniated brine known as weak liquor coming out of the absorber is cooled down by passing through the plate heat exchangers. After the weak liquor is pumped back to the preliminary absorber (PAB), further absorption again causes the temperature to rise. This liquor known as strong liquor is further sent to the stock tanks and is known as VAT Liquor.

The column consists of tower washer, AVW, absorber proper and PAB and this is known as Absorber Column. From AVW to tower washer there is no passage of gases and tower washer is a separate column placed at the top of the absorber. The operation of absorber is chemically simple as it is an operation of temperature, pressure and vacuum controls. The absorber operation depends upon the gas supply from distillers and any upsetting in the distiller will directly affect the absorber working and vice versa.

2.5. Process Flow Diagram



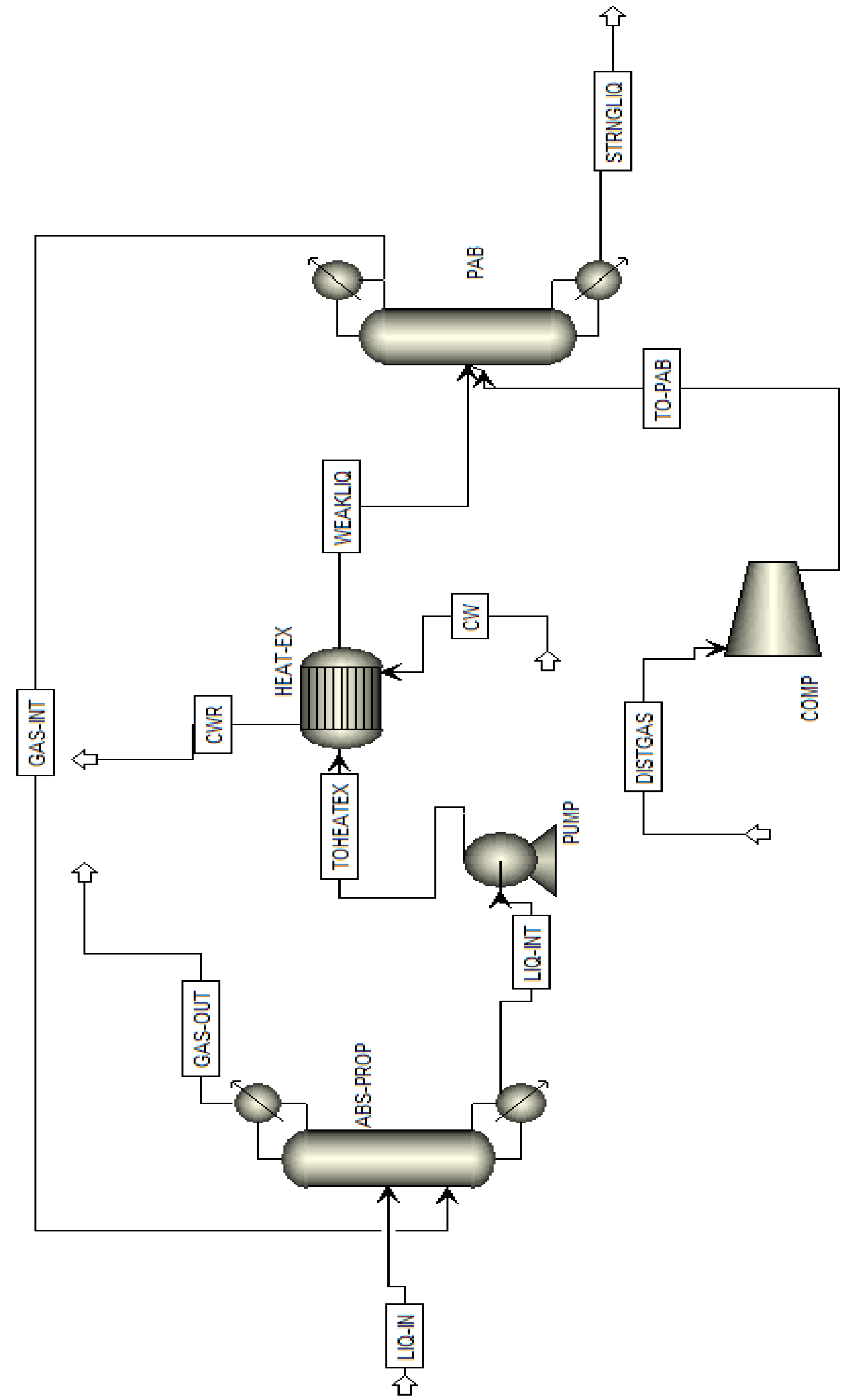


Figure 2 Absorber Column for Ammoniation of Brine (Aspen Plus)

2.6. Absorber Column

2.6.1. PURPOSE

The purpose of the absorber is to absorb ammonia into brine in the desired amount to produce ammoniated brine, before its treatment with carbon dioxide in carbonating towers. As carbon dioxide is sparingly soluble in brine, ammonia is first dissolved in brine.

2.6.2 PARTS OF ABSORBER COLUMN

- After Washer (AW).
- Tower Washer (TW).
- Absorber Vacuum Washer (AVW).
- Proper Absorber.
- Preliminary Absorber (PAb.).
- Plate Heat Exchanger (PHE).
- Re-ammoniator.
- Vat. Liquor Stock tanks.

The absorber column is about 78ft. high and it consists of 20 compartments, each 6ft. in diameter. Each compartment is fitted with a single bubble cap and over flow boxes.

The carbonating towers waste gas rises up the bubble cap and makes contact with the brine on the plate and then passes through the gas riser to the above plates.

Brine flows across the plate in the zig zag path through the over flows and on to the plates below. A weir plate at the entrance of the overflows maintains the liquid level over the plate.

The column is divided into two main sections with 10 compartments each, and the main sections are as following:

- Tower Washer for the absorption of carbonating tower waste gas
- Proper Absorber for the absorption of distiller gas

Further division according to compartments is as:

After Washer ------------ 1 compartment

Tower Washer ------------ 9 compartment

Absorber Vacuum Washer ------------ 3 compartment

Proper Absorber ------------ 5 compartment

Preliminary Absorber ------------ 2 compartment

Chapter-4

Instrumentation and Process Control

4.1. Parameters Recordings:

Following are the recordings for the control parameters:

- Total purified brine flow, Tower washer brine flow and totalizer
- Tower washer brine leaving and entering temperature
- Tower washer temperature and Absorber top temperature
- Weak Liquor leaving and entering temperature
- Strong liquor leaving PAB temperature
- Distiller gases entering PAB temperature
- Ammonia Injection

4.2. Control Loops

4.2.1. Emergency Brine Flow Control

After the tower washer brine is cooled enters from the top compartment of Absorber Proper and on the other hand Caisse Cooler gases, which are usually ammonia with small amount of carbon dioxide, enters from the bottom. In the absorber section, majority of the ammoniation is carried out. This reaction being exothermic raises the temperature of the liquor; this could also mean inadequacy of brine flow for the reaction with gases. Therefore, an emergency brine flow to this section is provided which is linked with third compartment temperature, so that brine flow may be increased whenever temperature of weak liquor increases from set limit.

However, the loop is not being used as it resulted in operational difficulties.

4.2.2 Purified Brine Flow Control

Gases ex-carbonating towers are washed in the Tower Washer by means of purified brine. Any change in TW pressure has an impact on the performance of Carbonating Towers. Sharp changes in TW pressure cause rapid change in the rate of drawing, which in turn causes problems in magma handling and reduced conversion. A control scheme comprising of differential temperature of tower washer exit gas (vent) to purified brine and inlet gas pressure with some compensation, as designed to regulate the purified flow to the tower.

4.3. The Control System

In order to make provisions for adequate safety, as well as for the efficient handling and automation of the project, a control system needs to be in place. As the conditions of the process are quite dynamic, there are a lot of deviations in several parameters. Therefore, a fast acting control system will not only ensure safety but also smooth operations in the plant. It will counter the possible deviations to keep process variables within acceptable limits. The diagram below shows the overall design of the control system, and what follows is an up-close look at how controllers are placed around each of the process equipment.

4.4. The Heat Exchanger Control

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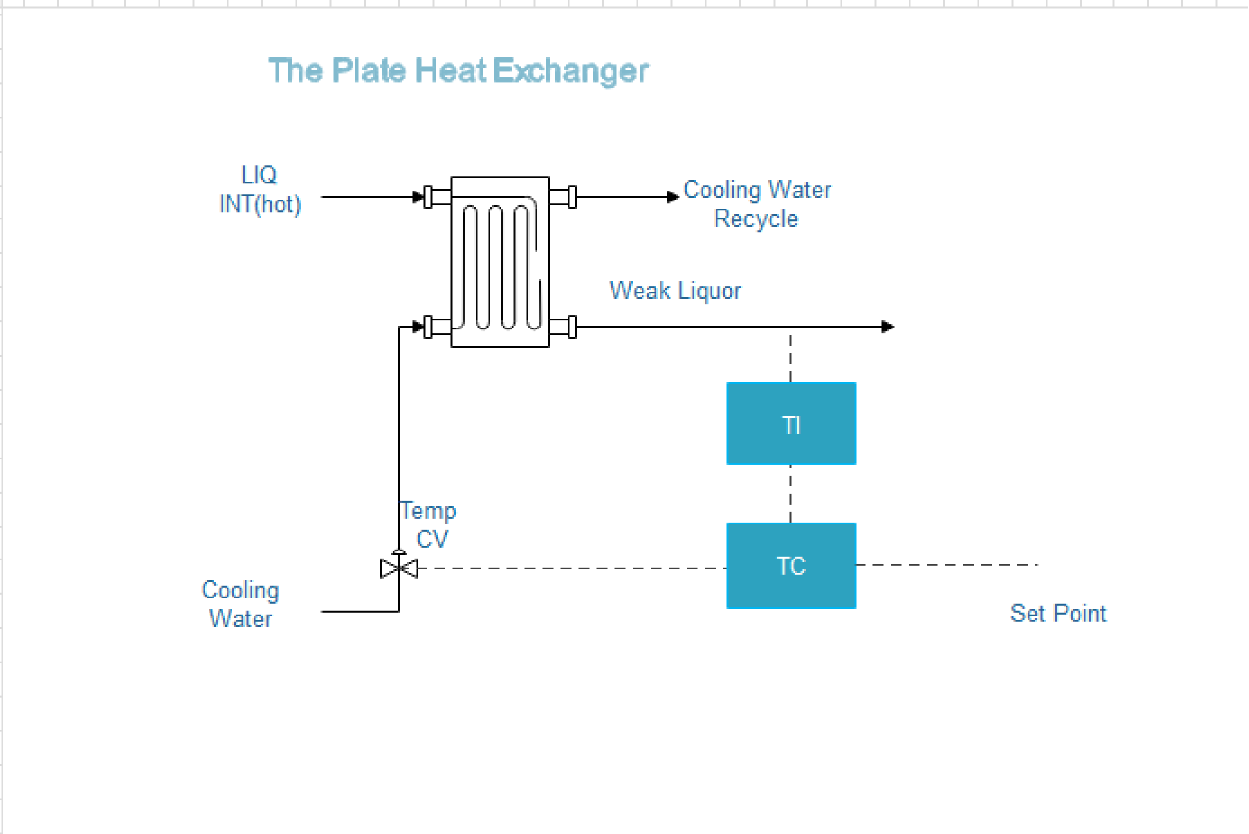


Figure 3 The Heat Exchanger Control Loop

- A plate heat exchanger is fitted with temperature control loop.
- The controlled variable is the temperature of the “Weak Liquor”.

- The manipulated variable is the flowrate of “Cooling Water” stream.

4.5. The Centrifugal Pump Control

- The pump is fitted with a flow control loop.
- The Controlled variable is discharge flow rate i.e. flow rate TO HEAT EX.
- The manipulated variable is the flow rate of LIQ INT.

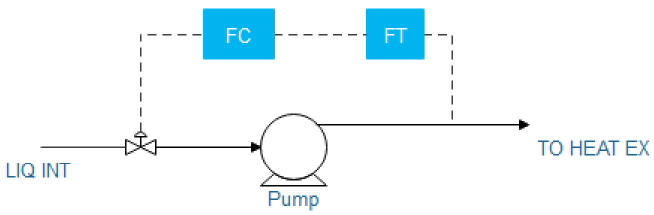


Figure 4 Centrifugal Pump Control

4.6. Distiller Gas Compressor Control

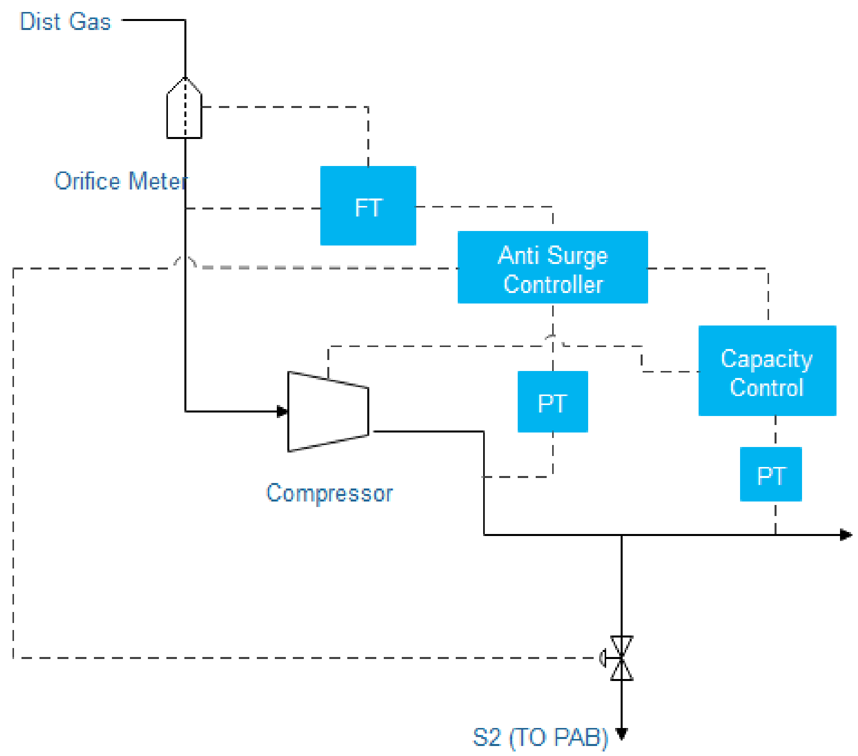


Figure 5 Compressor Control System

- The compressor is fitted with capacity and antisurge control loops.
- For antisurge control loop, the controlled variable is the surge tendency of the compressor.
- The manipulated variable is the flowrate of the 'TO PAB' stream
- For the capacity control loop the controlled variable is the input flowrate to the compressor.

10.7. The Absorber Control Loop

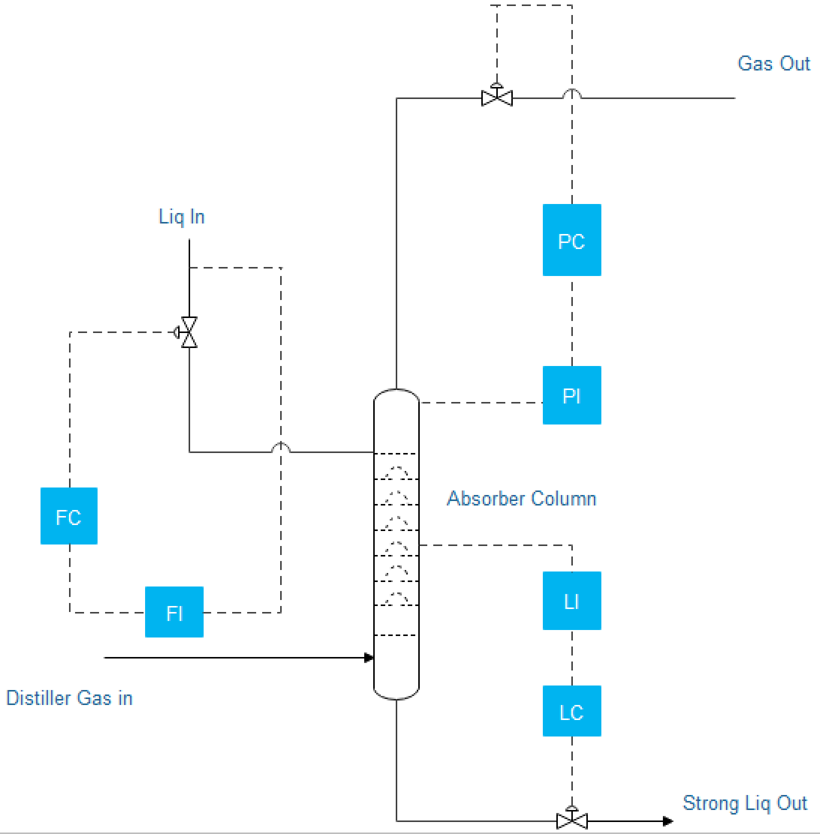


Figure 6 Absorber Control System

- The absorber column is fitted with pressure, level and flow control loops.
- For the level control loop, the controlled variable is the absorber fluid level
- The manipulated variable is the flowrate of the 'Strong Liq Out' stream
- For the pressure control loop the controlled variable is the absorber pressure
- The manipulated variable is the 'Gas Out' flowrate
- For the flow control loop the controlled variable is the flowrate of gas into the vessel

The manipulated variable is the 'Liq In 'flowrate.

Chapter-5

Equipment Description

The operation of absorber depends upon the control of temperatures, pressure and vacuum. It also depends upon the gas supply from distillers and carbonating towers, any abnormality in there operation will in turn effect the absorbers operation and vice versa.

5.1. Tower washer and After washer

Upper portion of the absorber is called Tower washer. Its purpose is to recover the ammonia from the waste gases of the carbonating towers. The topmost compartment of this portion is called After Washer. Purified water is used to wash (scrub) the final traces of ammonia. The reason for using water in this compartment is to avoid deposition of salt in the exit gas main, which would occur if brine were used and also to protect the environment from ammonia. Ammonia tester measures ammonia loss in exit gas.

Purified brine is pumped and sprayed in the tower washer i.e., in the second compartment from top. The purified brine main line before entering the tower washer is branched off to supply a certain portion of brine to absorber vacuum washer and a small line to the sealing of TLC pump.

The waste gases from the carbonating towers enter the gas compartment of the tower washer, where the liquor droplets carried over by the waste gases from towers separates out and is drained into the CVL line through Tower Washer run-off line. The gases travel up the column through bubble cap plates and are finally wasted to atmosphere. The brine as it travels down absorbs ammonia and some carbon dioxide from the waste gases of the carbonating towers. The brine leaving the tower washer becomes hot, because it comes in contact with the hot gases from the tower and also due to the absorption of ammonia. Therefore it is necessary to cool the brine before it is sent to absorber. The brine leaving is called Tower Washer Brine, which is then passed through PHE and is cooled by cooling water.

5.2. Absorber Vacuum Washer

Absorber vacuum washer is fed with purified brine, which was branched off from the main header, to scrub ammonia from the gases coming from the proper absorber. The gases leaving the absorber vacuum washer are strong gases having carbon dioxide concentration of 90-92%. Vacuum is applied by means of vacuum engine (absorber vacuum engine) to the top most compartment of absorber vacuum washer, which sucks the gases from the absorber and delivers into FMC with other gases.

5.3. Proper Absorber

Tower Washer Brine after being cooled enters the proper absorber top. Here it absorbs ammonia from the distiller gases; as reaction is exothermic and the gases coming from distiller are hot the temperature of the liquor rises. This partly ammoniated brine leaving from the proper absorber proper bottom is called Weak Liquor is cooled in PHE and fed back to the preliminary absorber.

5.4. Preliminary Absorber

Weak Liquor after being cooled is fed to the first compartment of preliminary absorber, where further ammoniation takes place the temperature rise and the liquor leaving from the second compartment of preliminary absorber is called Strong Liquor, which is then sent to the Vat Liquor stock tanks. This ammoniated brine from stocks is called Vat Liquor. From the storage, Vat. Liquor is pumped to Mono carbonating Tower after cooling where it is partially carbonated with kiln gas and enters the reammoniator for the final pickup of ammonia concentration to the desired extent.

The gases from the distiller containing ammonia and carbon dioxide are first cooled in Caisse Cooler. The main line carrying the gases from the Caisse Cooler branches off to re-ammoniator before entering the preliminary absorber. A valve (chain) on the main line between preliminary absorber and the point from where the line branches off diverts the gases to re-ammoniator and unused gases from re-ammoniator are carried through a line which then joins the distiller gases main line before entering the gas compartment below preliminary absorber (also called preliminary absorber gas compartment). At this point make up ammonia line from tankers is also added to the distiller gases main line.

In the gas compartment, condensate from distiller gas main, which is strong ammonia solution is collected and drained into the Vat. Liquor stock tanks through a line called PAB run-off.

5.5. OPERATING PARAMETERS:

5.5.1. Pressure

Absorber Vacuum at Control Valve	7.5 ± 1 in. Hg
Absorber Top	5.5 ± 1 in. Hg
Absorber Bottom	2.5 ± 1 in. Hg
Preliminary Absorber	0.0 ±0.5 in. Hg
Tower washer	7.0 ± 1 in Hg

5.5.2. Temperature

Absorber Top	35 – 45 C
Tower Washer Brine Leaving	35 – 45 C
Tower Washer Brine Entering Absorber	35 – 40 C
Weak liquor Entering PAB.	35 – 40 C
Strong Liquor Leaving PAB	65 – 68 C
Gases Entering PAB	> 60 C
Vat. Liquor to MCT	34 – 40 C
Gases Ex - RTT	45 – 50 C

5.5.3. Liquor Tests

Tower Washer Brine Direct Test	16 – 22 mls N/20 ml
Absorber Vacuum Washer Direct test	3 – 4 mls N/20 ml
Absorber Direct Test	35 – 45 mls N/20 ml
Vat. Liquor Direct Test	98 – 100 mls N/20 ml

Filter Vacuum Washer 6 – 8 mls N/20 ml

5.6. Special Precautions

Tower Washer

- Turbid brine is not fed
- TW bottom pressure should not rise – normal working pressure varies between 5
- 7 in.Hg., depending upon flow rate of brine
- There should be no interruption of water supply to topmost compartment (AW
- Gas compartment is properly drained.

Absorber Vacuum Washer

- Vacuum applied should be such that there is 1in.Hg. Pressure on the heater top and slight vacuum on PAB.
- Keep a watch on the vacuum difference i.e., vacuum readings of AVE or top of AVW, absorber top, absorber bottom and PAB. These readings are very important as they help in detecting the obstruction in any section of the absorber.
- Keep sufficient brine flow to the AVW so that the direct test is controlled.
- Flush the AVW with purified water once per shift. Flushing of water should be strictly controlled to avoid unnecessary dilution of brine in the absorber.

Proper Absorber and PAB:

- Vacuum control.
- Absorber top temperature should be controlled between 40 – 45 C
- Weak liquor temperature of 69 C. gives the desired test in liquor and should be controlled between 68 – 70 C
- Weak liquor should be cooled to 38 – 40 C. before feeding it to PAB.
- Strong liquor leaving temperature should be controlled between 68 – 70 C. at 69 C. it gives the desired test.
- Control the temperatures of liquor leaving absorber and PAB by adjusting the flow of brine.
- Steam the vent system regularly.

- Make sure that the PAB lute is in proper working order.

5.7. Common Problems and Troubleshooting

Probable Causes	Remedial Solutions
Obstruction in tower waste gases inlet	Steam the main thoroughly
Obstruction in tower washer exit gases main	Increase the flow of purified water in the after washer
Obstruction in tower washer brine outlet	Check TWB pumps running load, check sieves and clean/change if clogged

Table 1High Pressure in Tower Washer

Probable Causes	Remedial Solutions
Partial blockage in TW/WL sieves	Change-over or clean the sieves
Inefficient working in TWB or WL pumps	Check operating loads and put the stand by pump on duty
Obstruction in AVW compartments	Carryout thorough steaming
Obstruction in vacuum main from AVW to AVE	Carryout steaming
Air ingression due to drawing/sucking of lute	Check the absorber vacuum lute and refill if empty

Table 2 Gradual and Steady rise in Pressure

Probable Causes	Remedial Solutions
Chocking of necks/vents	Regular steaming of vents
Blockage/channeling in vent washers	Steam/Flush the vent washers

Table 3 Pressure in Liquor Stocks

Probable Causes	Remedial Solutions
Greater amount of gases going to PAB	Notch up the AVE to match the gases being fed to the PAB
inefficient Absorber Vacuum Engine	If the vacuum doesn't increase with the above action, change the engine
Choking of Absorber Vacuum Washer compartments	The vacuum gradient in the AVE, absorber top, absorber bottom and PAB will be abnormal. The vacuum at the top and bottom drops. Flushing and steaming of AVW and Absorber top will ease the situation
Liquor holding in the Proper Absorber	The pump tripping will cause reduction in the quantity of distiller gases being generated and hence vacuum on the PAB will be observed even with the same set of AVE. Get the tripped pump started as soon as possible, meanwhile reduce the vacuum to avoid holding of liquor.

Holding of liquor in the PAB compartments	Blockage in the PAB runoff. Can be removed by knocking and through steaming

Table 4 High Pressure on PAB

Probable Causes	Remedial Solutions
Choking/scaling up of distiller gas main to PAB	Blockage in the drains of caisse cooler gas or condensate compartment will result in excessive pressure on the distiller column due to gas holding. PAB will come under vacuum. The system will normalize only after the drains are cleared.
Obstruction in Caisse Cooler gas or condensate drain or Caisse Cooler tube leakage	Liquor quantity in the caisse cooler condensate compartment will increase according to the nature of the leakage of caisse cooler tube. Bursting of the tube will result in building in liquor in the condensate compartment and holding of distiller gases. The set of caisse cooler boxes with leaking tubes has to be isolated as soon as possible to overcome the problem
Tripping of feeder liquor pump	The pump tripping will cause reduction in the quantity of distiller gases being generated and hence vacuum on the PAB

	will be observed even with the same set of AVE. Get the tripped pump started as soon as possible, meanwhile reduce the vacuum to avoid holding of liquor.

Table 5 High Vacuum on PAB

Probable Causes	**Remedial Solutions**
Absorber high top	-Check absorber vacuum -Check steam leakage in AVW absorber, ammonia injection main and optimize use of steam in distillers -Check WL entering temperature and acid/manual clean the PHE.

Table 6 Temperature Related Up-Setting

Chapter-6

Material & Energy Balance

6.1. Material Balance

The Equipment Used

These are the following equipment used:

- Absorber
- Plate Heat Exchanger
- Pumps
- Compressor

6.1.1. Overall Material Balance

Liquid In

- Total Liquid In: 80977.6 kg/hr
- NH_3 = 825.6 kg/hr
- NaCl = 20640 kg/hr
- **H_2O = 59512 kg/hr**

Liquid Out

- Total Liquid Out = 85348 kg/hr
- NH_3 = 9000 kg/hr
- NaCl = 20640 kg/hr
- **H_2O = 55680 kg/hr**
- **$C0_2$ = 28kg/hr**

Gas In

- Total Gas In = 52500 kg/hr
- NH_3 = 9074.4 kg/hr
- CO_2 = 14700 kg/hr
- **H_2O = 27995.4 kg/hr**

- **Inert = 730.2 kg/hr**

CO_2 Balance

$$Mass\ In\ =\ Mass\ Out$$

(0.21)(70000) = y + 28

y = 14672 kg/hr

NH_3 Balance

$$Mass\ In\ =\ Mass\ Out$$

825.6 + x = 9000 + 900

x = **9074.4 kg/hr**

*Assume NH_3 loss = 900 – 825.6 = 74.4 kg/hr

Water Balance

$$Mass\ In\ =\ Mass\ Out$$

59512 + 27995.4 = 55680 + z

z = 31827.4 kg/hr

Provided that G_{out} is 90% CO_2 on dry basis (wt%)

14672/0.9 = 16302 kg/hr

CO_2 + NH_3 + I = 16302

I = 16302 -14672 – 900 = **730.2 kg/hr**

G_{out} = 16302.2 kg/hr

Water = 31827.4 kg/hr

Total = **48129.6 kg/hr**

Total Material Balance

$$Mass\ In\ =\ Mass\ Out$$

80977.6 + 52500 = 48157.6 + 85320

133477.6 kg/hr = 133477.6 kg/hr

6.2. ENERGY BALANCE

Reference Temperature = 25 °C

$$Cp\ =\ A\ +\ BT\ +\ CT^2\ +\ DT^3\ +\ E/T^2$$

NH_3

Cp (25 °C) = 35.5 J/k.mol (g)

Cp (33 °C) = 83 J/k.mol (l)

Cp (70 °C) = 88.5 J/kmol (l)

Cp (60 °C) = 36.7 J/k.mol (g)

Cp (out) = 26.5 J/k.mol (g)

<u>Water</u>

75.3 J/k.mol (l) (NIST)

36.5 J/k.mol (g) (NIST)

<u>CO_2</u>

At 60 °C = 38.7 J/k.mol

At 25 °C = 37.12 J/k.mol

<u>NaCl (aq)</u>

Cp = 71.9 J/k.mol (Dortmund Data Book)

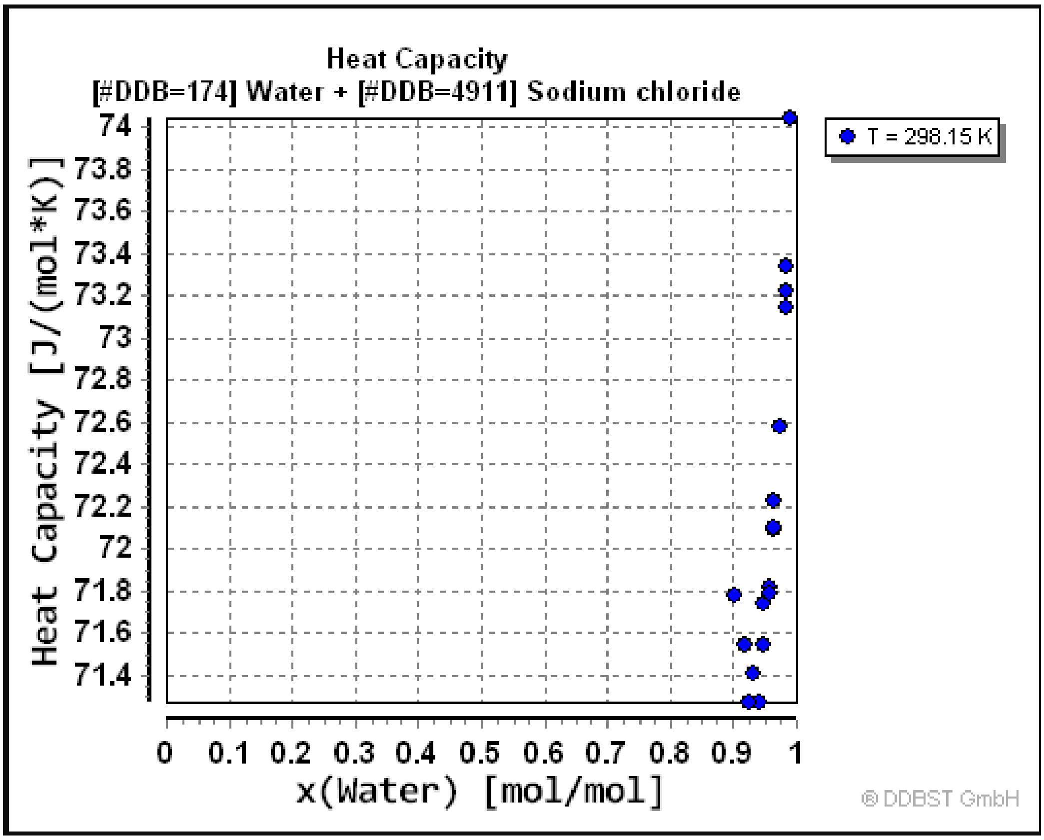


Figure 7 Heat Capacity of Brine Solutions

6.2.1. STREAM ENTHALPIES

<u>Liquid In</u>

NH_3 = 48.56 kmol/hr

NaCl = 352.8 kmol/hr

H_2O = 3306.2 kmol/hr

Total = 3707.56 kmol/hr

$\mathbf{Cp_{mix}}$ = (48.56/3707.56) × (83) + (352.8/3707.56) × (71.9) + (3306.2/3707.56) × (75.3)

$\mathbf{Cp_{mix}}$ = 75 KJ/k.mol

$$Q = mCp_{mix}\Delta T$$

Q: Enthalpy of stream (KJ/hr)

m: Molar Flow Rate (KMol/hr)

$Cp_{mix:}$ Heat Capacity of Mixture(KJ/k.mol)

ΔT: Temperature Difference (^{o}C)

Q = (3707.56) × (75) × (33 – 25)

Q = 2226829.3 KJ/hr

<u>Liquid Out</u>

NH_3 = 529.4 k.mol/hr

NaCl = 352.8 k.mol/hr

H_2O = 3093.3 k.mol/hr

CO_2 = 0.636 k.mol/hr

Total = 3976.1 k.mol/hr

Cp_{mix} = (529.4/3976.1) × (88.5) + (352.8/3976.1) × (71.9) + (3093/3976.1) × (75.3) + (0.636/3796.1) × (38.7)

$\mathbf{Cp_{mix}}$ **= 76.7 KJ/k.mol**

$$Q = mCp_{mix}\Delta T$$

Q: Enthalpy of stream (KJ/hr)

m: Molar Flow Rate (KMol/hr)

$Cp_{mix:}$ Heat Capacity of Mixture(KJ/k.mol)

ΔT: Temperature Difference (ºC)

Q = 3976.1 (76.7) × (70 – 25)

Q = 13731558 KJ/hr

<u>Gas In</u>

NH_3 = 533.8 k.mol/hr

H_2O = 1555.4 k.mol/hr

CO_2 = 334.1 k.mol/hr

Inert = 21.56 k.mol/hr

Total = 2444.6 k.mol/hr

Cp_{mix}= (533.8/2444.6) × (36.7) + (334.1/2444.6) × (38.7) + (1555.4/2444.6) × (36.5) + (21.56/2444.6) × (30)

Cp_{mix} = 36.8 KJ/k.mol

$$Q = mCp_{mix}\Delta T$$

Q: Enthalpy of stream (KJ/hr)

m: Molar Flow Rate (KMol/hr)

Cp_{mix}: Heat Capacity of Mixture (KJ/k.mol)

ΔT: Temperature Difference (ºC)

Q = 2444.6 × (36.8) × (60 - 25)

Q = 31475339 KJ/hr

<u>Gas Out</u>

* Gas Out Temperature =58 ℃

NH_3 = 52.9 k.mol/hr

H_2O = 1768 k.mol/hr

CO_2 = 333.45 k.mol/hr

Inert = 21.56 k.mol/hr

Total = 2175.91 k.mol/hr

Cp_{mix} = (333.45/2175.9) × (387) + (21.56/2175.91) × (30) + (52.9/2175.91) × (36.5) + (1768/2175.91) × (36.5)

Cp_{mix} = 36.8 KJ/k.mol

$$Q = mCp_{mix}\Delta T$$

Q: Enthalpy of stream (KJ/hr)

m: Molar Flow Rate (KMol/hr)

Cp_{mix}: Heat Capacity of Mixture(KJ/k.mol)

ΔT: Temperature Difference (ºC)

Q = 36.8× (2175.91) × (58 - 25)

Q = 2640467 KJ/hr

6.2.2. Heat rejected, cooling flow rates and compositions

Solute Free Basis

$$\boldsymbol{G_s = G\,[1 - y]}$$

G: Gas Flow rate Kmol/hr

Gs: Solute Free Gas Flow Rate Kmol/hr

y: molar fraction

G_s = 2444.6 [1 -0.218]

G_s = 1910.8 k.mol/hr

$$\mathbf{L}_{\boldsymbol{s}} = \boldsymbol{L}[\mathbf{1} - \boldsymbol{x}]$$

L: Liquid Flow rate Kmol/hr

Ls: Solute Free Liquid Flow Rate Kmol/hr

x: molar fraction

L_s = 3707.56[1 -0.0131]

L_s = 3659 k.mol/hr

$$Y_{N+1} = \frac{y}{1-y}$$

Y_{N+1} : Ammonia Free Fraction in Gas In

y: Ammonia Fraction in Gas In

Y_{N+1} = 0.279

$$X_0 = \frac{x}{1 - x}$$

X_0 : Ammonia Free Fraction in Liquid In

x: Ammonia Fraction in Liquid In

X_0 = 0.0133

$$\boldsymbol{G_s[Y_N + 1 - Y_i] = L_s[X_i - X_0]}$$

0.279 – Y_i = [36.59× (0.1255 - .0133)]/[2444.6]

Y_i = 0.111 **X_i = 0.1255**

X_N = 0.133/(1 -0.133)

X_N = 0.1536

Stage = 8

[0.1536 – 0.133]/10 = 0.014

i = 8

X_8 = [0.0133 + 8 (0.014)]

X_8 = 0.1255

G_i = 463.6 (NH_3)

334.1 (CO_2)

1555.14 (H_2O)

21.56 (Inert)

Total = 2374.4 kmol/hr

6.2.3. Stream enthalpy for cooling [Heat Rejection]

Cp_{mix} = (459.2/3905.9) × (88.5) + (352.8/3905.9) × (71.9) + (3093.8/3905.9) × (75.3) + (0.636/3905.9) × (38.7)

Cp_{mix} =76.7 KJ/k.mol

$$Q = m\mathrm{Cp}_{\mathrm{mix}}\Delta T$$

Q: Enthalpy of stream (KJ/hr)

m: Molar Flow Rate (KMol/hr)

$Cp_{mix:}$ Heat Capacity of Mixture(KJ/k.mol)

ΔT: Temperature Difference (ºC)

Q = (76.5) × (3905.9) × (70 - 40)

Q = 10159245 KJ/hr

6.2.4. Heat Generation (Absorption)

Reaction:

$$NaCl.H_2O + NH_3 \rightarrow NaCl.NH_4OH$$

ΔH_{sol} = 44085 KJ/k.mol (Dortmund Data Bank)

$$NH_3\ Out - NH_3\ In$$

= 480 Kmol/hr

ΔH_{sol} = (480) × (44085)

ΔH_{sol} = 21197000 KJ/hr

6.2.5. Overall Energy Balance

<u>Energy In</u>

Liquid In = 22.27×10^5 KJ/hr

Gas In = 31.48×10^5 KJ/hr

Total = 53.74×10^5 KJ/hr

<u>Generation</u>

212×10^5 KJ/hr

<u>Energy Out</u>

Liquid Out = 137.3×10^5 KJ/hr

Gas Out = 26.4×10^5 KJ/hr

Total = 163.7×10^5 KJ/hr

<u>Intermediate Cooling</u>

101.6×10^5 KJ/hr

Energy In + Generation = Energy Out + Intermediate Cooling + Miscellaneous Losses by Radiation

$(53.74 + 212) \times10^5 = (264 + 137.7 + 101.6 + x) \times10^5$

$x = 0.41\times10^5$ KJ/hr *Heat Loses from column to surrounding by radiation

Chapter-7

Modelling and Simulation

The program was simulated by using given data of Bubble Cap Tray column on Apsen Hysys and Apsen Plus V8.8. Aspen Plus V8.8 is the up-to-date version of this extensively used process optimization software. With a bunch of brand-new novelties in modeling, V8.8 guarantees an enhanced user experience, latest property data, and a much precise modeling.

The simulation was done for same throughputs for Packed bed absorption column.

Simulation of Packed bed absorber Column

Initially these components were selected with H_2S as Inert.

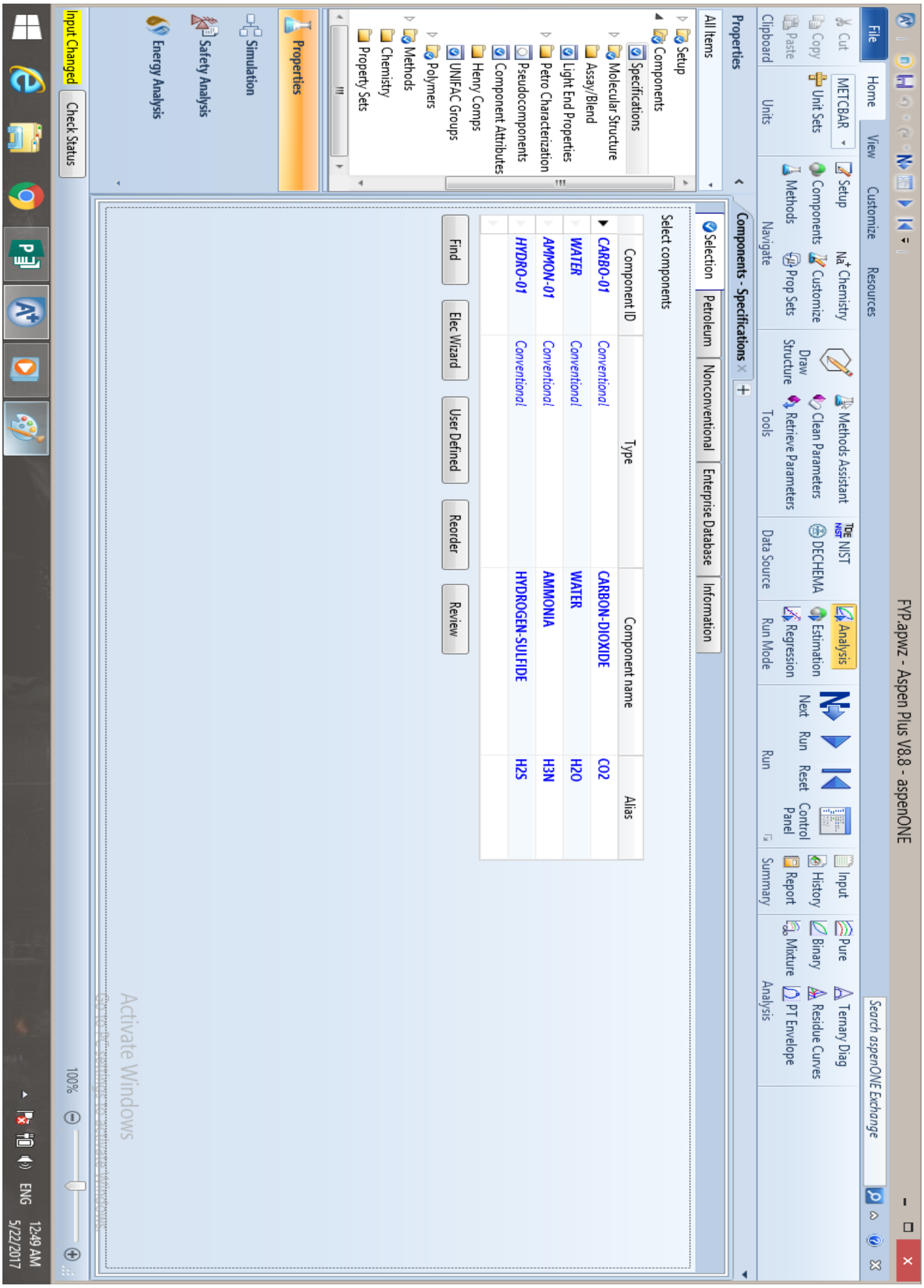


Figure 8 Initial Components

Liquid In Stream

Liq in stream is the mixture of Ex-Mono carbonating tower brine, Tower Washer Brine and Fresh Brine.

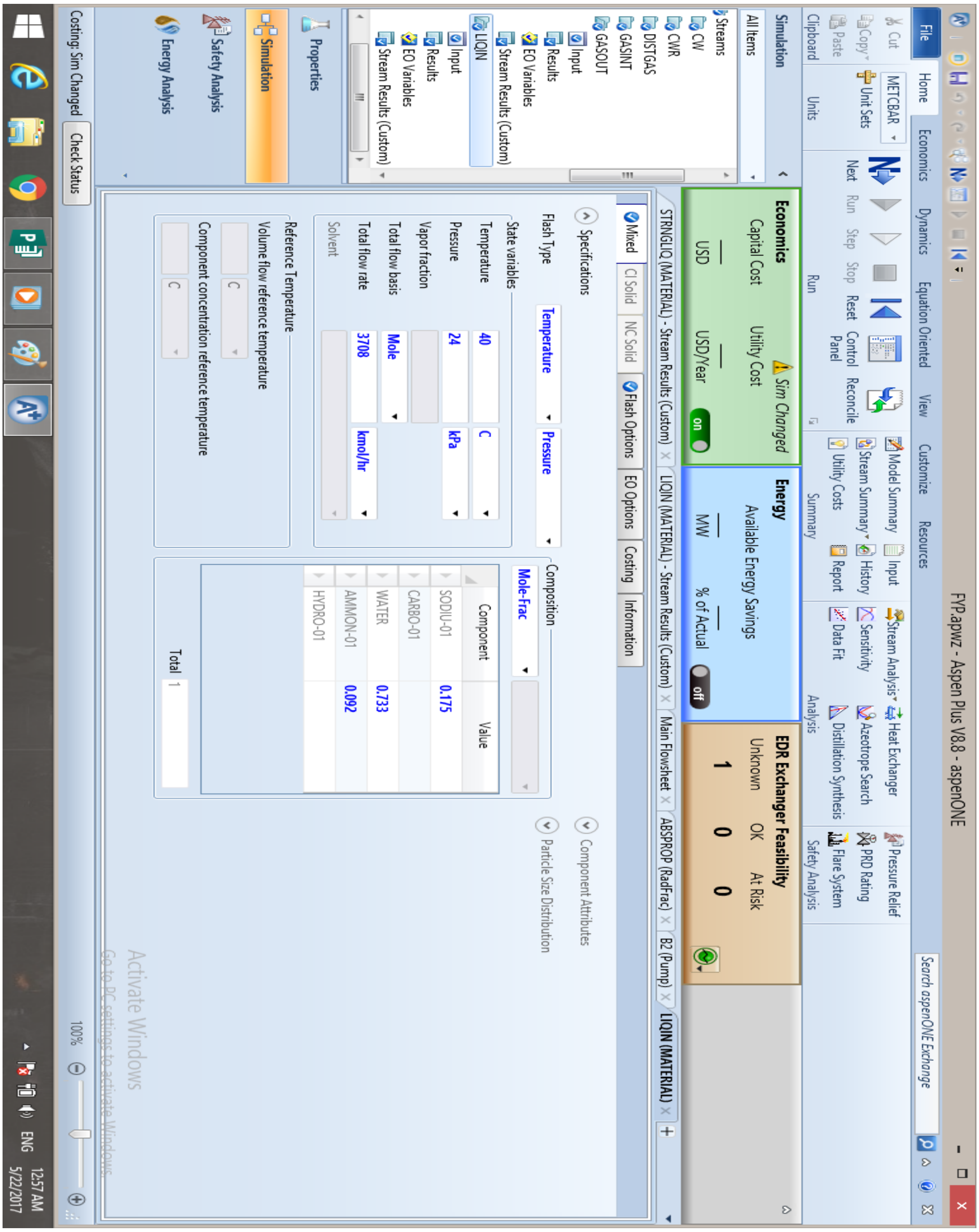


Figure 9 Liquid In Compositions

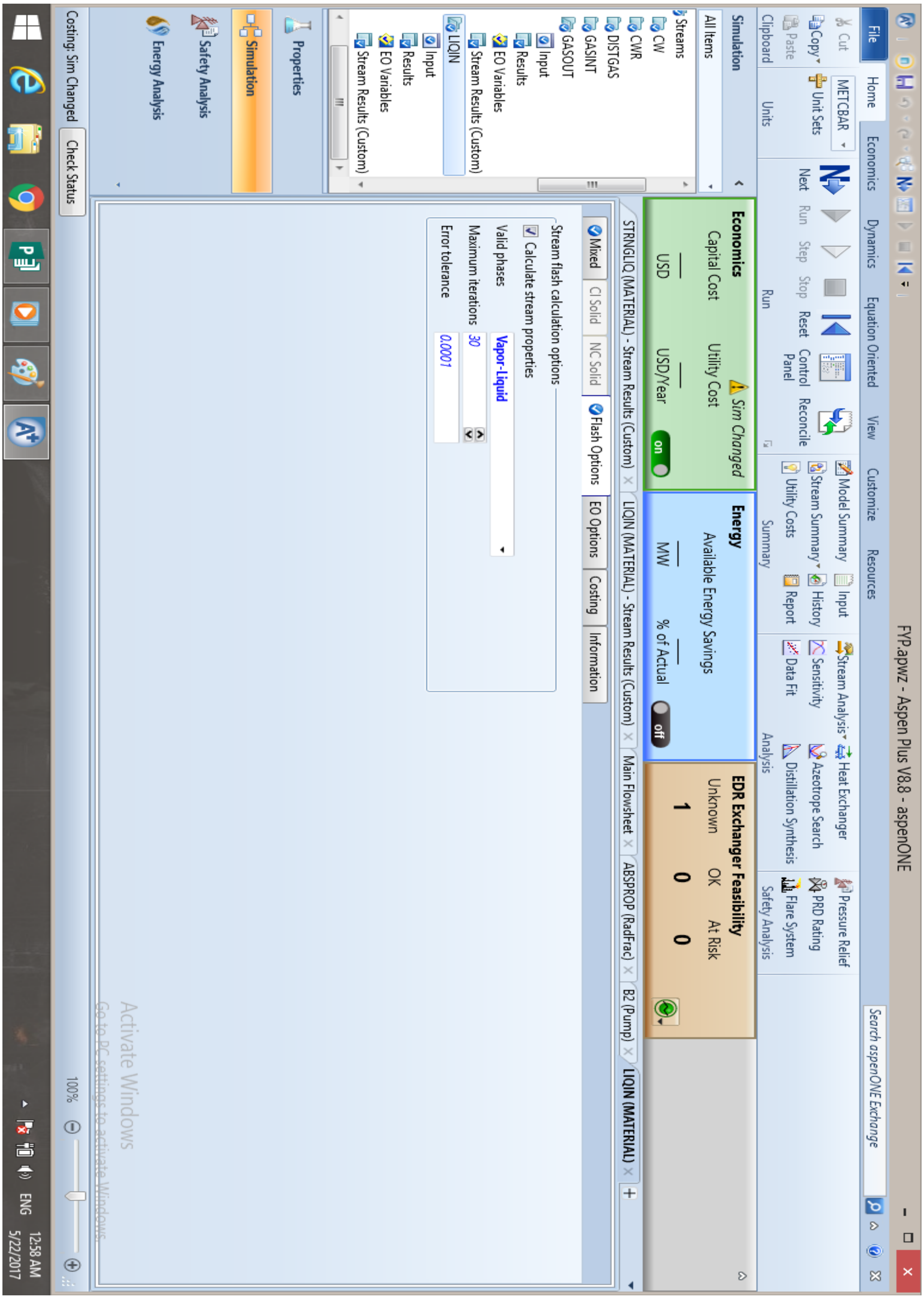


Figure 10 Liquid In Compositions 2

Absorber Proper

Absorber Proper is first absorption column. It is basically the combination of two columns, absorber vacuum washer and proper absorber.

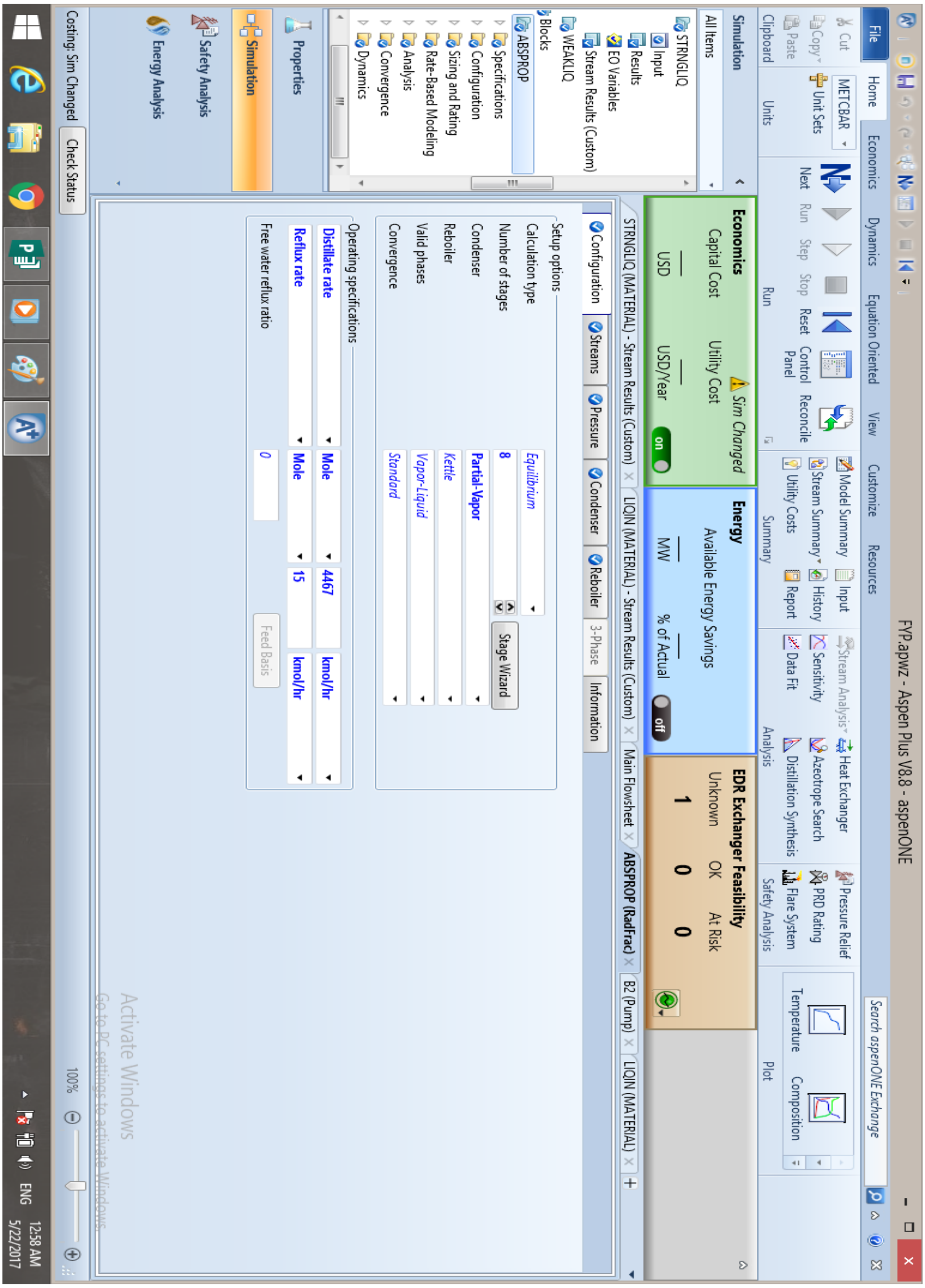


Figure 11 Absorber Proper Column 1

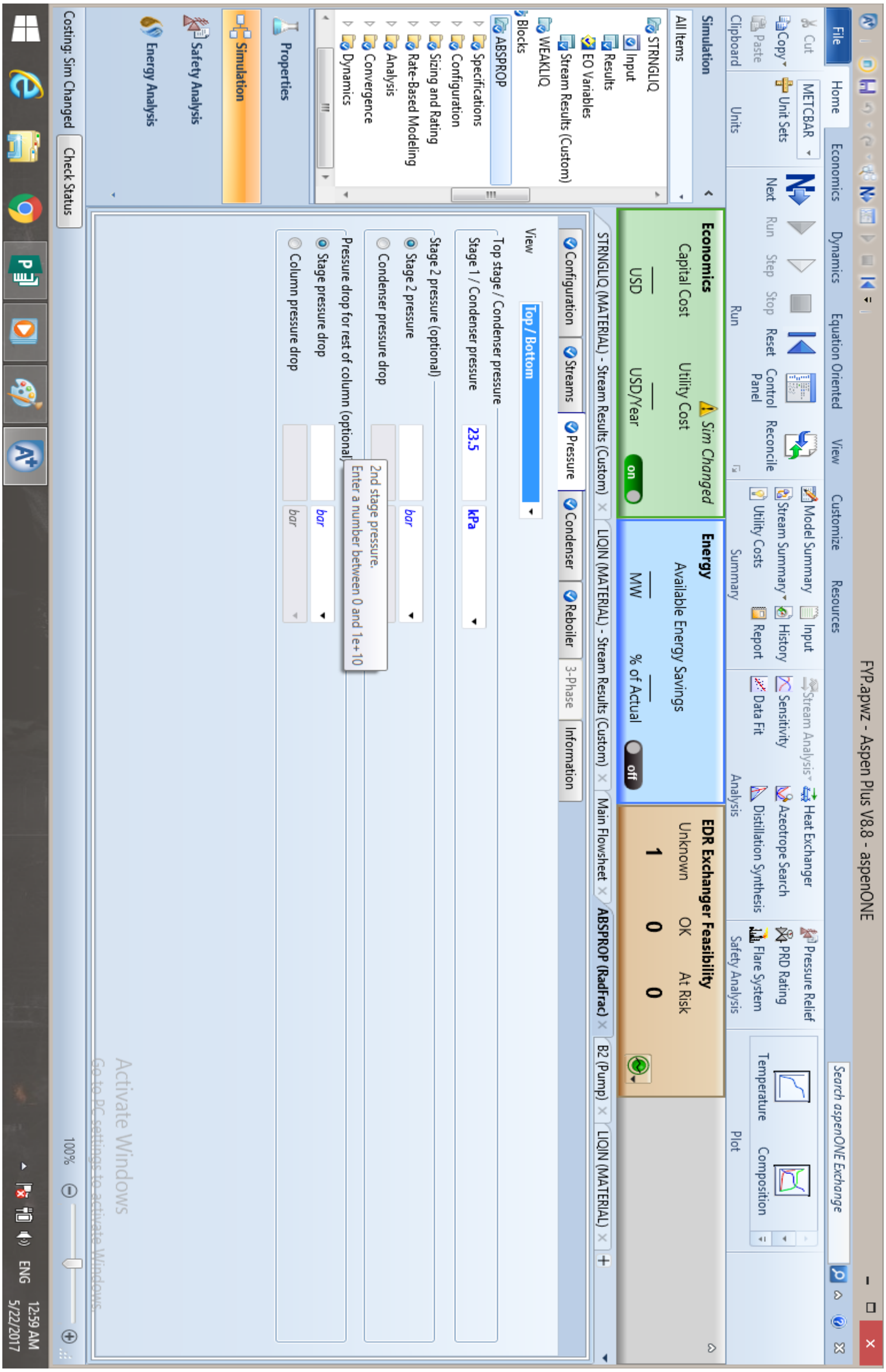


Figure 12 Absorber Proper Column 2

Absorber Proper Packing Characteristics

Absorber Proper has been changed to a Packed bed column with following packing characteristics

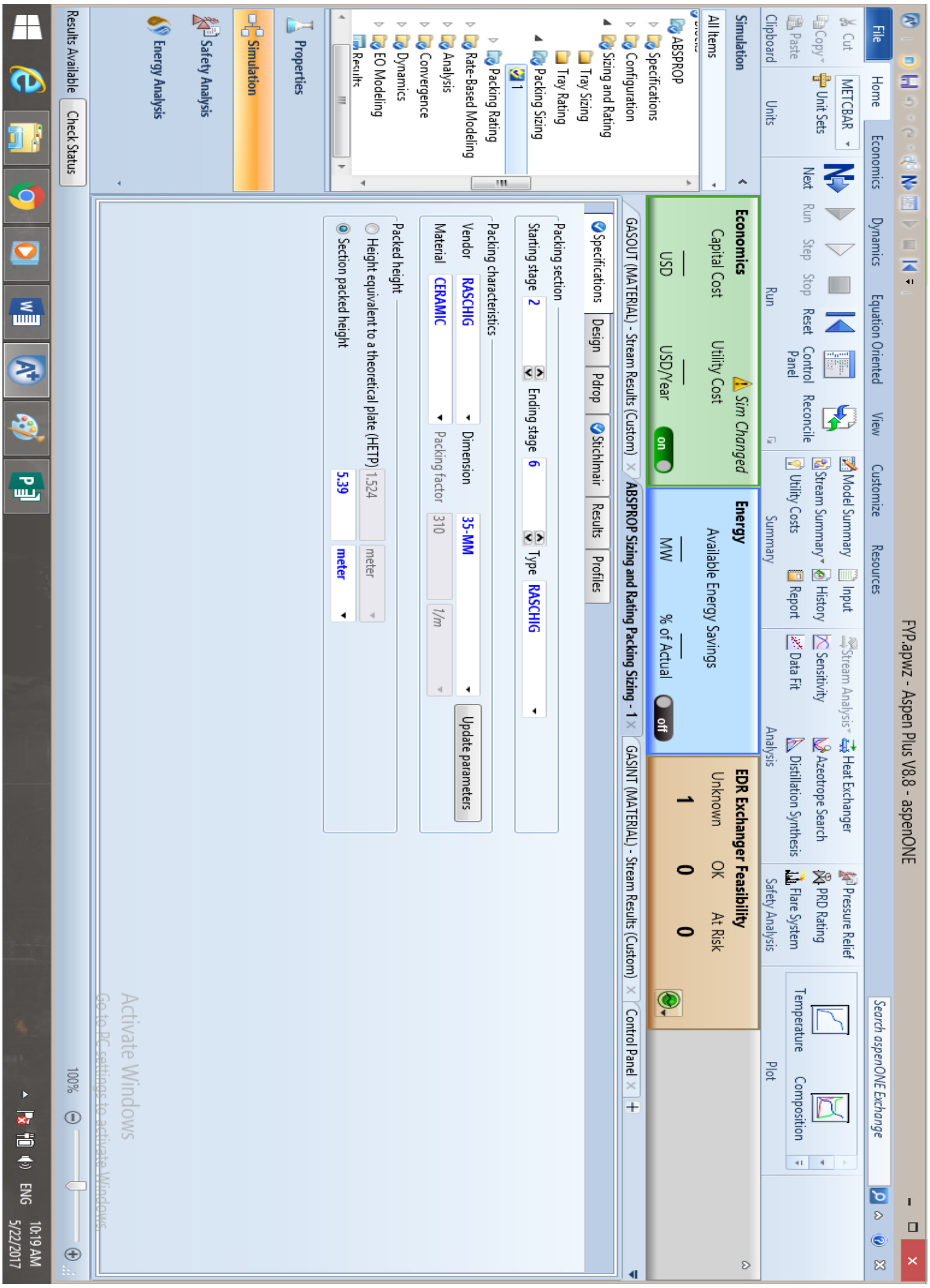


Figure 13 Absorber Proper Packing

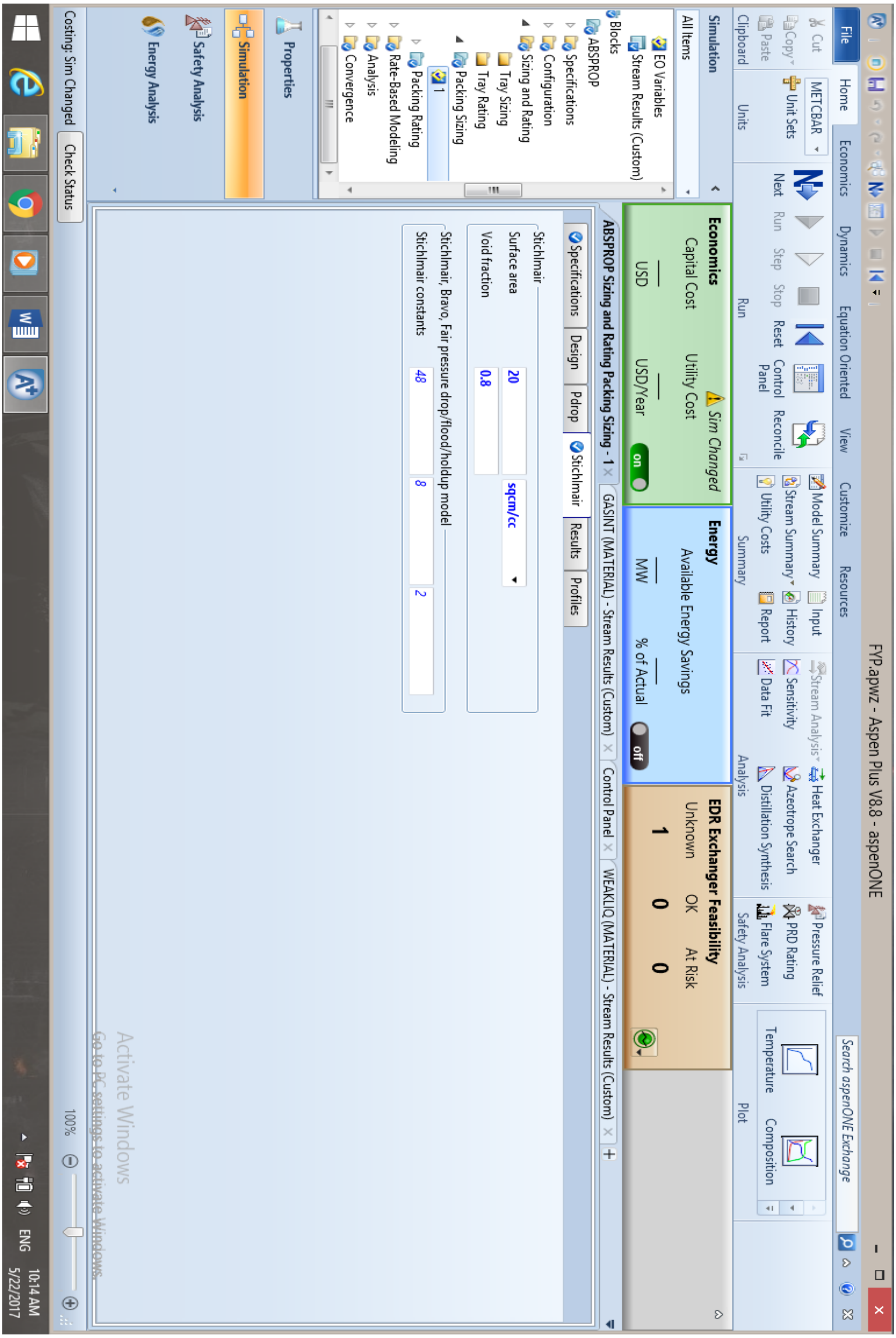


Figure 14 Absorber Proper Packing 2

Heat Exchanger

A plate heat exchanger is placed through which weak liquor obtained from the bottom stream of Absorber proper passes. This weak liquor is at high temperature due to exothermic absorption process. Heat exchanger is placed to reduce it to optimum temperature of 42 °

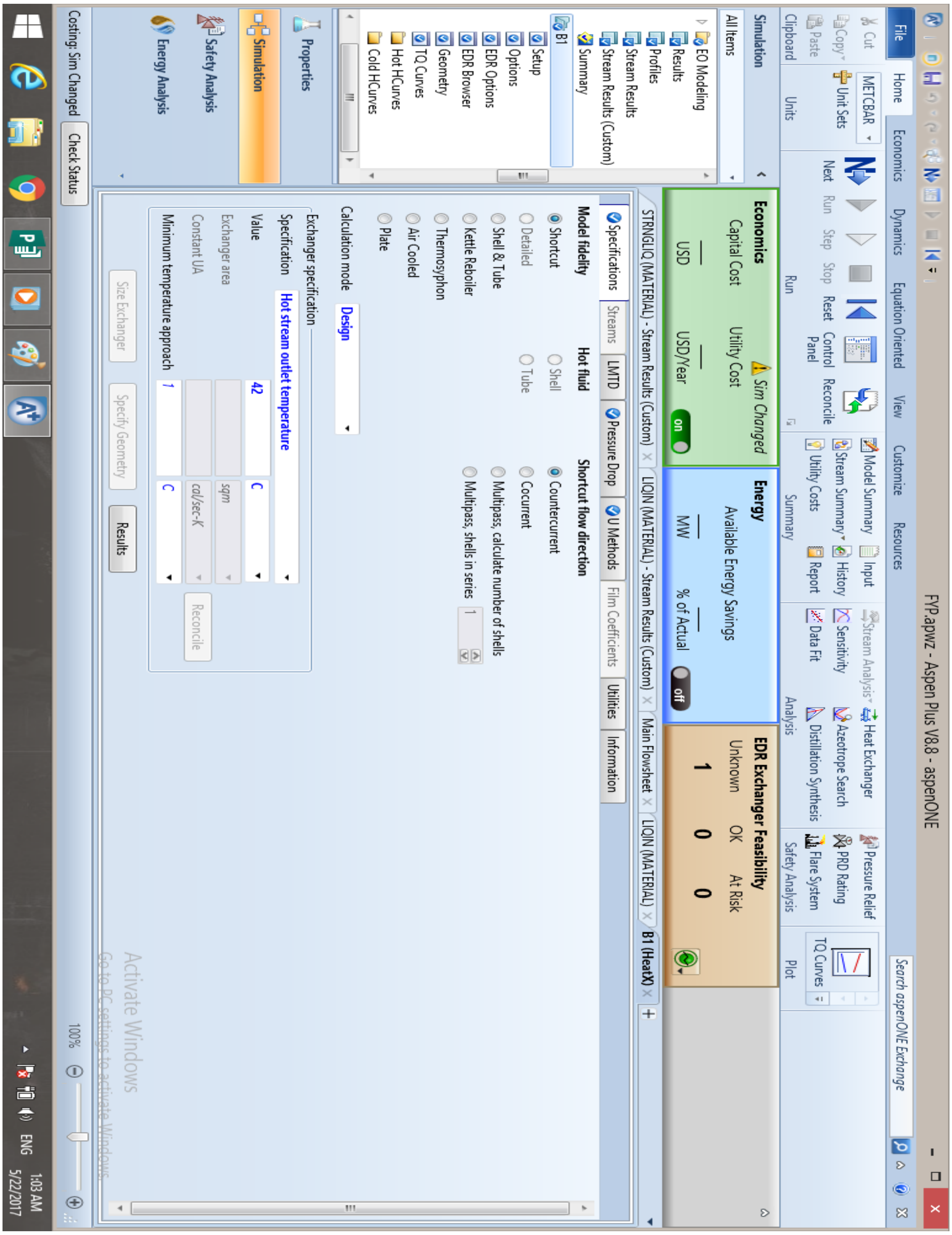


Figure 15 Heat Exchanger (Aspen Plus) 1

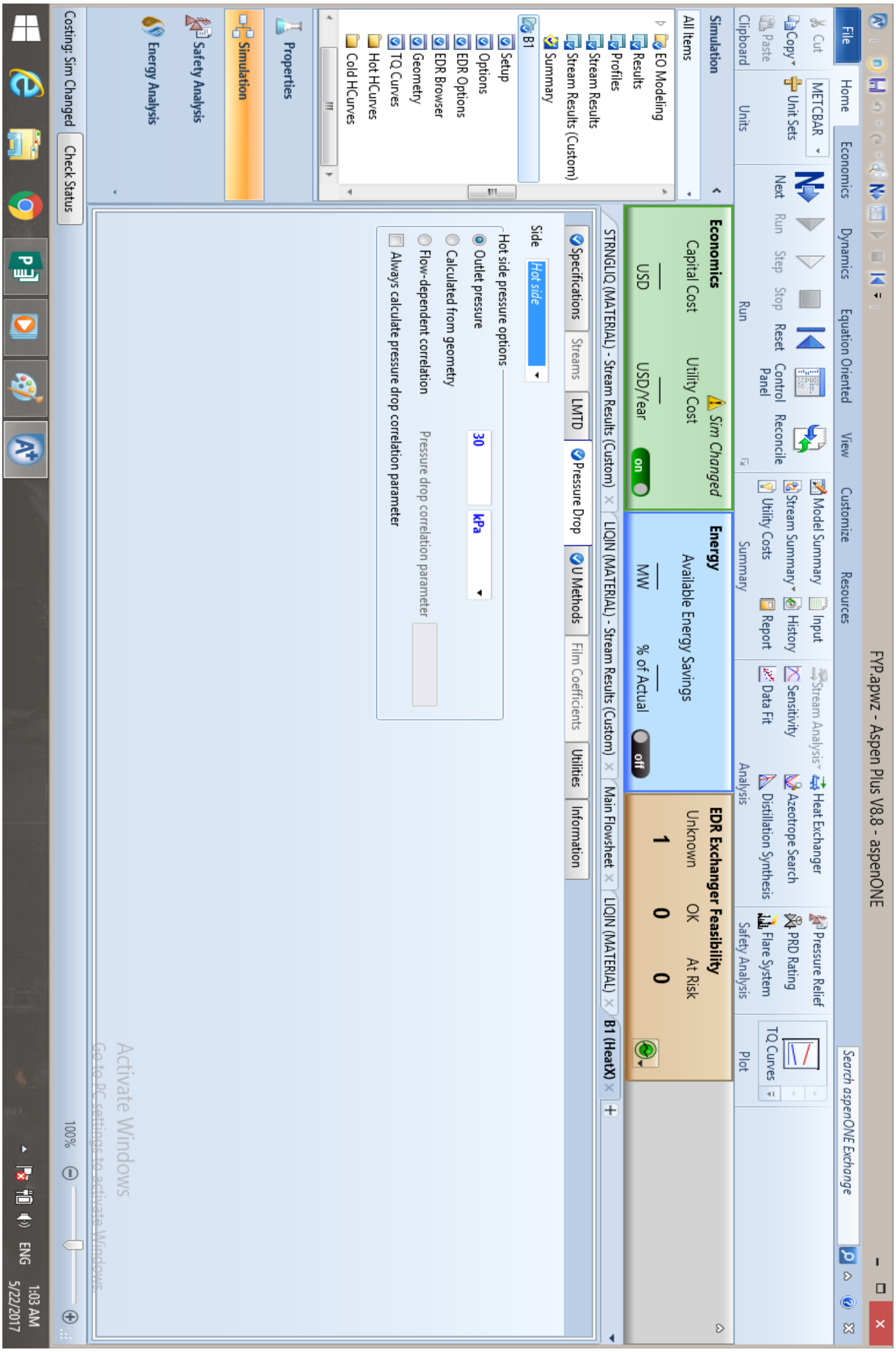


Figure 16 Heat Exchanger (Aspen Plus) 2

Distiller Gas Stream

Distiller gases are entering into Preliminary Absorber after passing through a compressor. Distiller gas stream composition

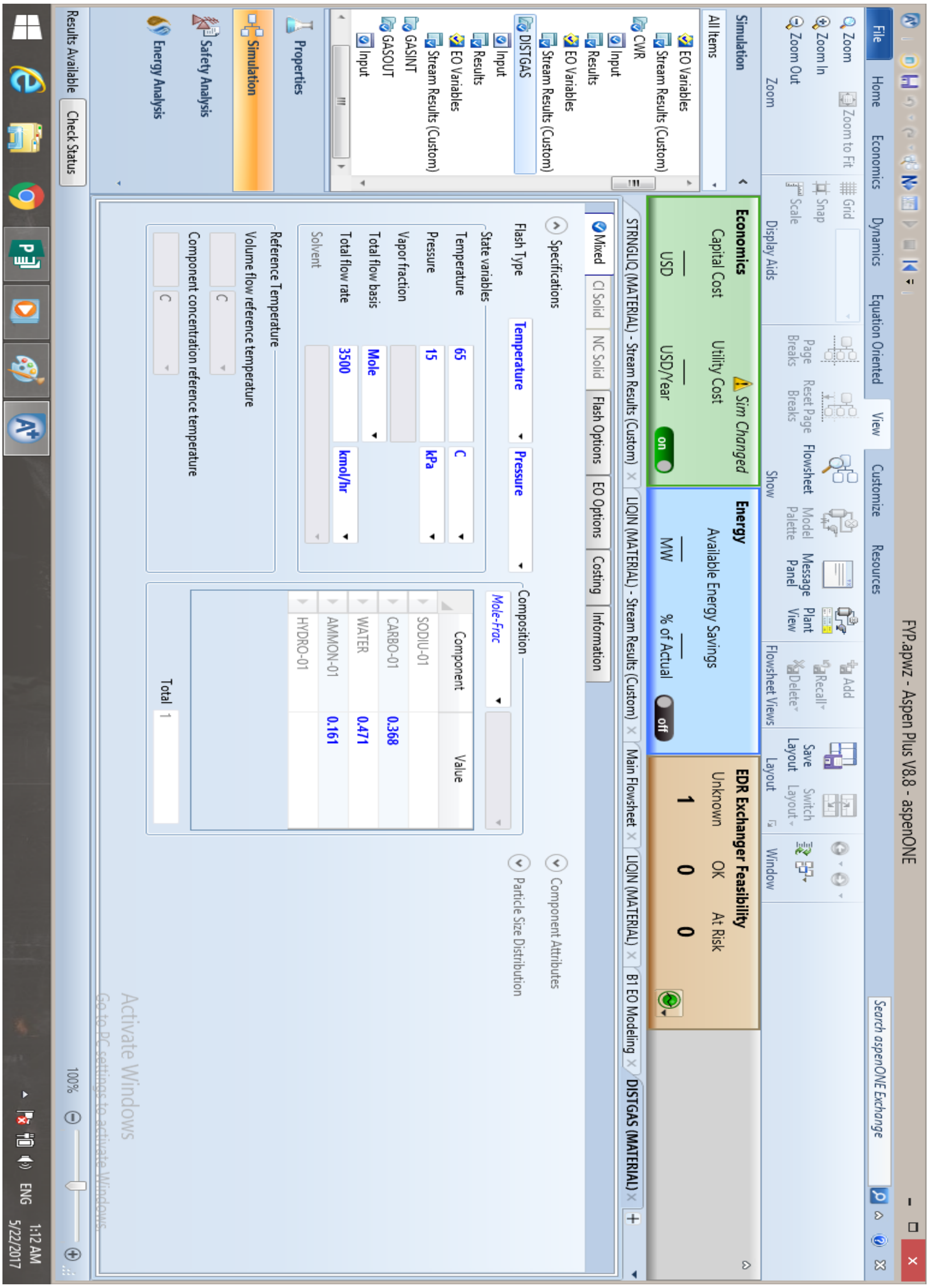


Figure 17 Gas Stream In

Compressor

A centrifugal compressor is used to pump distiller gas to Preliminary absorber at required pressure.

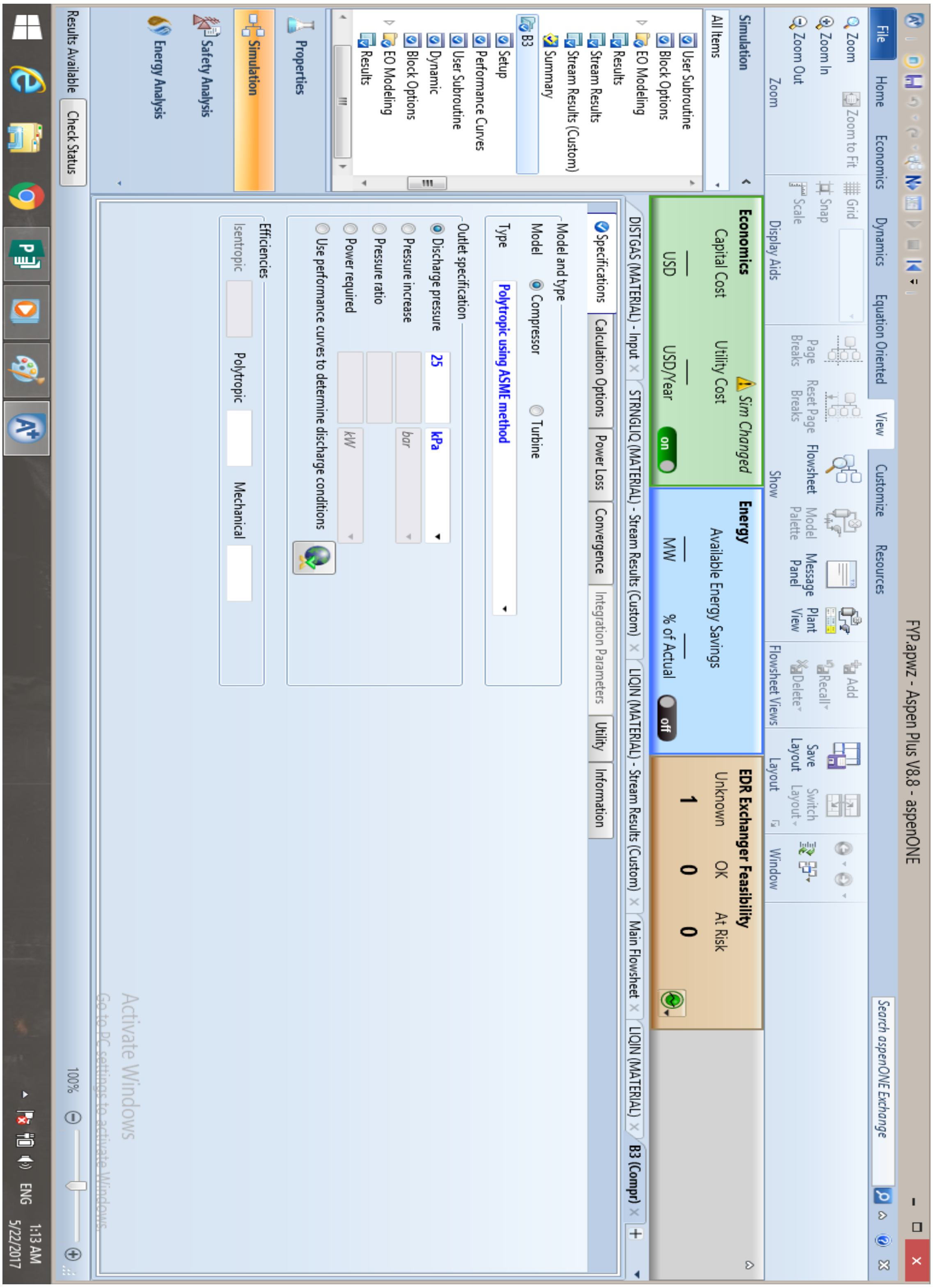


Figure 18 Compressor (Aspen Plus)

Preliminary Absorber

Weak liquor enters from the top and compressed distiller gases enter from the bottom. Maximum absorption takes place in this column.

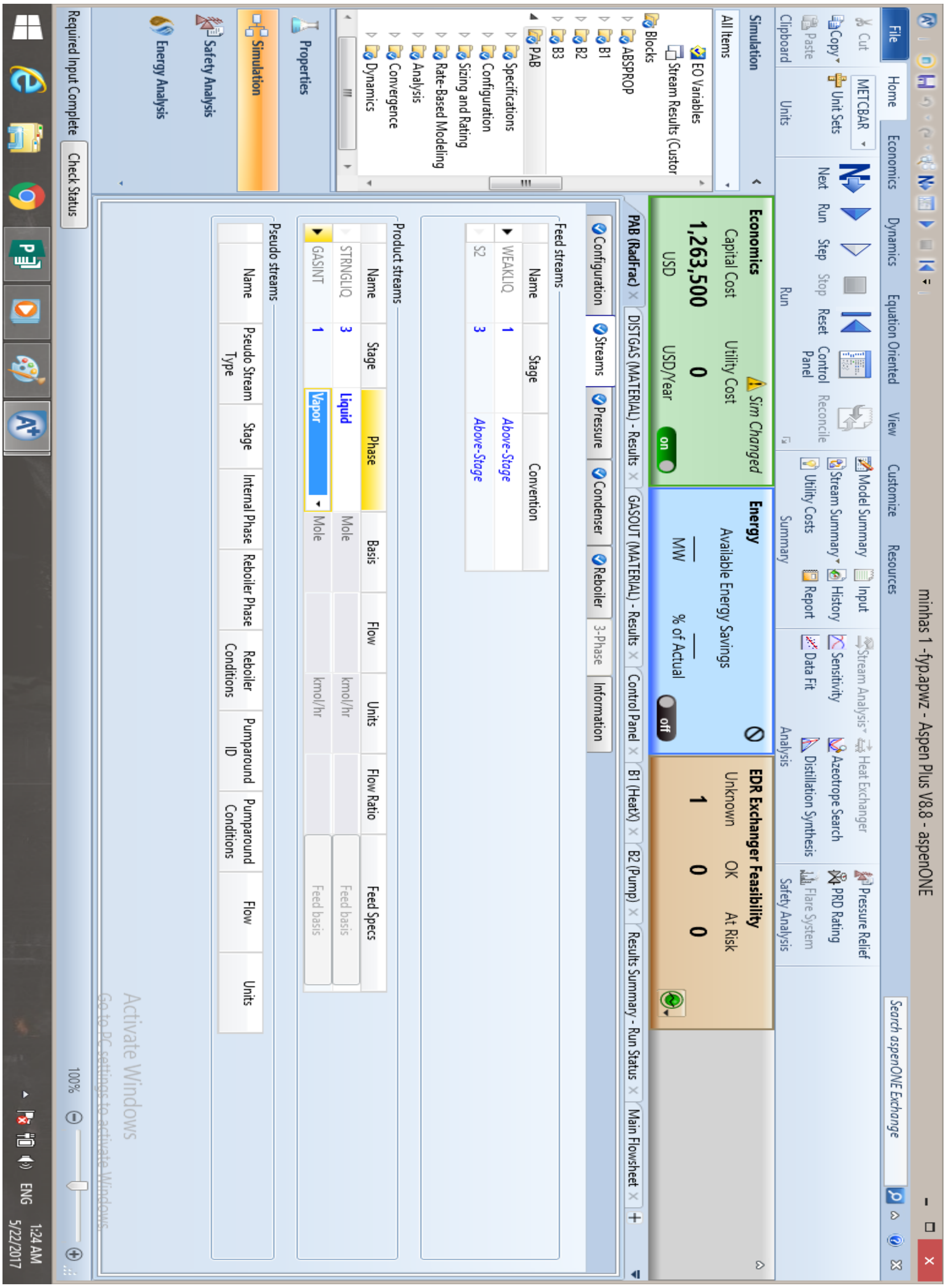


Figure 19 Preliminary Absorber (Aspen Plus) 1

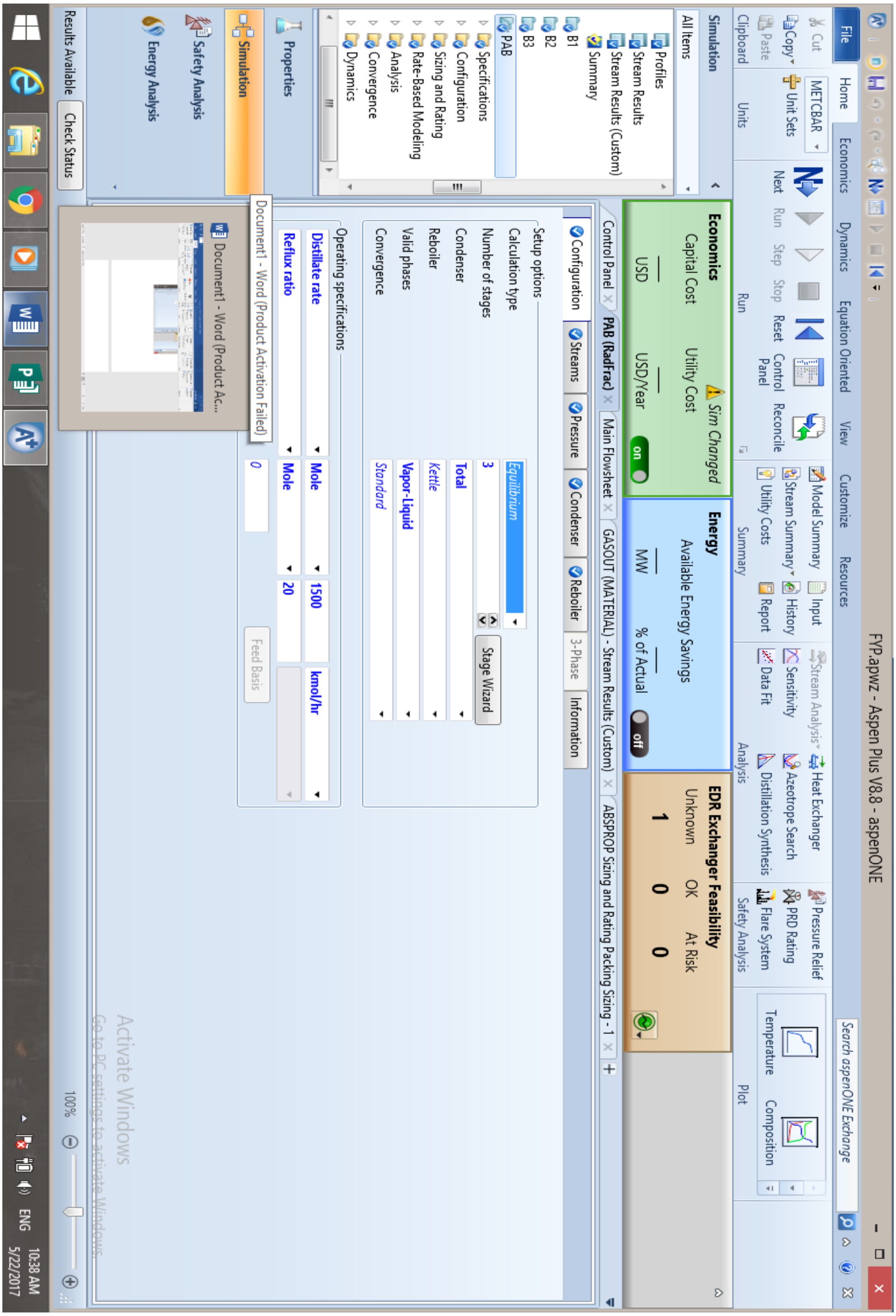


Figure 20 Preliminary Absorber (Aspen Plus) 2

Strong Liquor

After absorption in Preliminary absorber Liq out is obtained which is called strong Liquor.

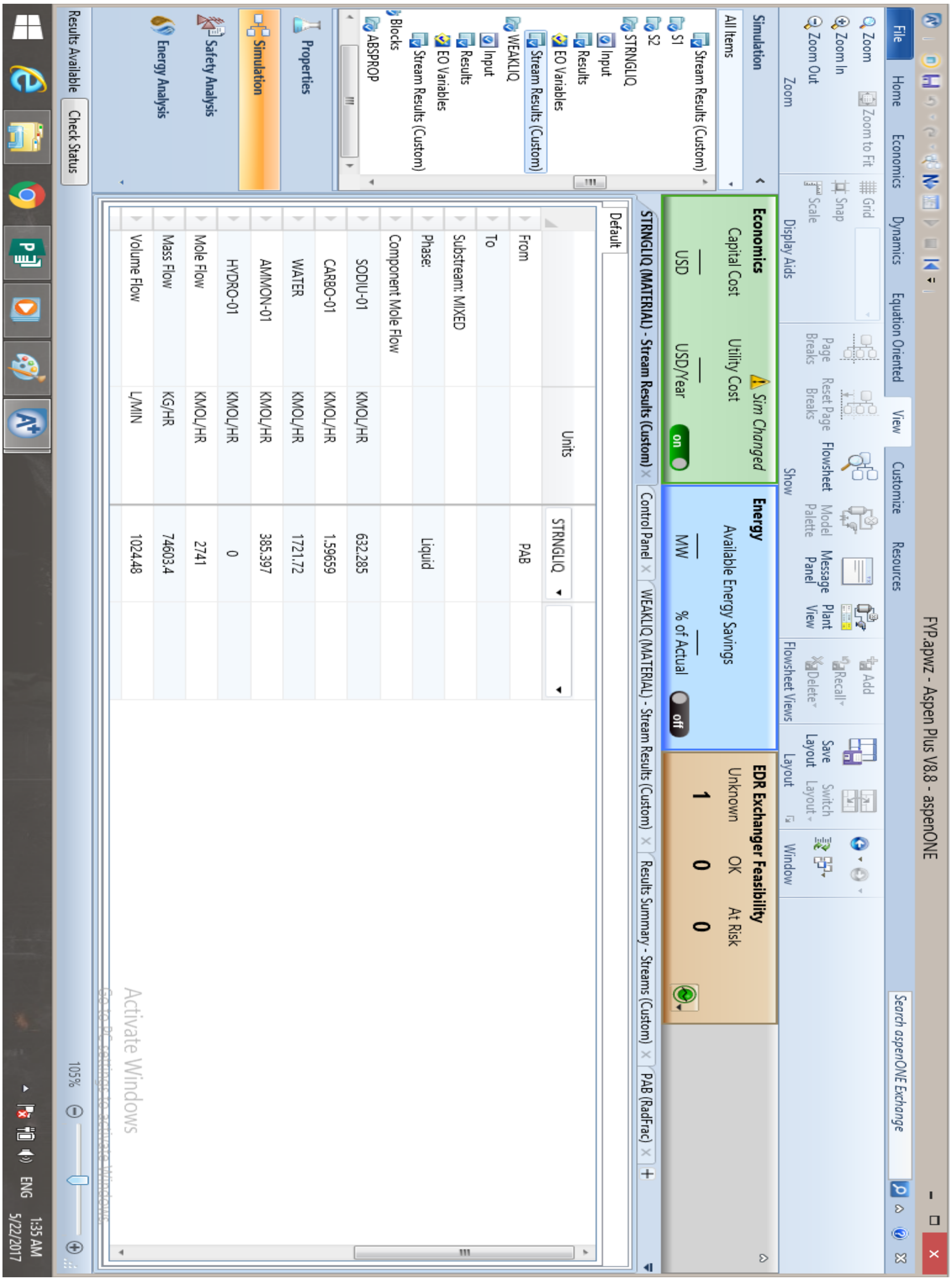


Figure 21 Strong Liquor Stream

Gas Intermediate

Gas stream obtained from the Preliminary absorber then moves upward towards Absorber Proper, this stream is called intermediate gas stream.

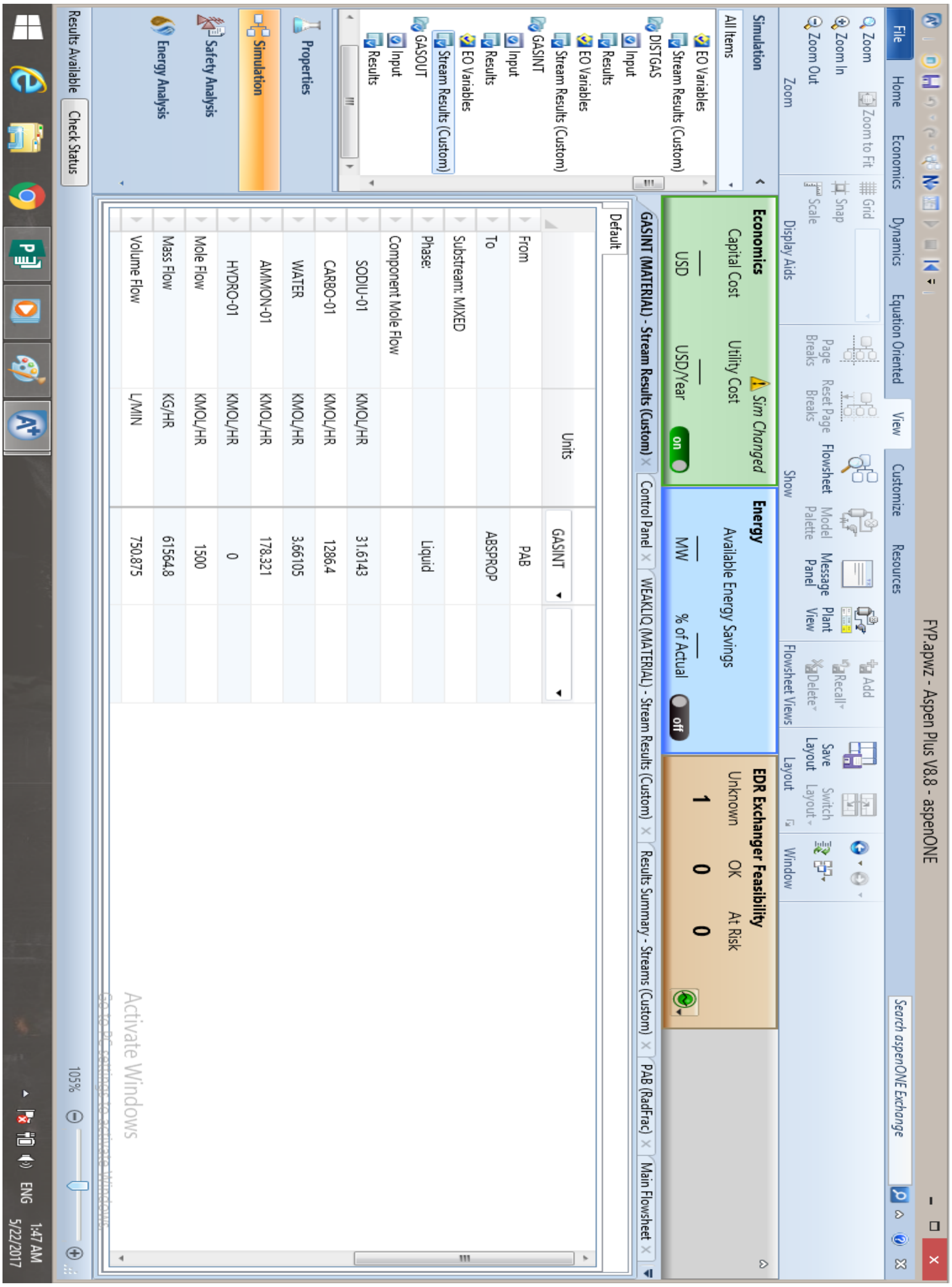


Figure 22 Gas Intermediate (Aspen Plus)

Gas Out Stream

Gas out stream obtained from Preliminary absorber is final stream that is vent to atmosphere.

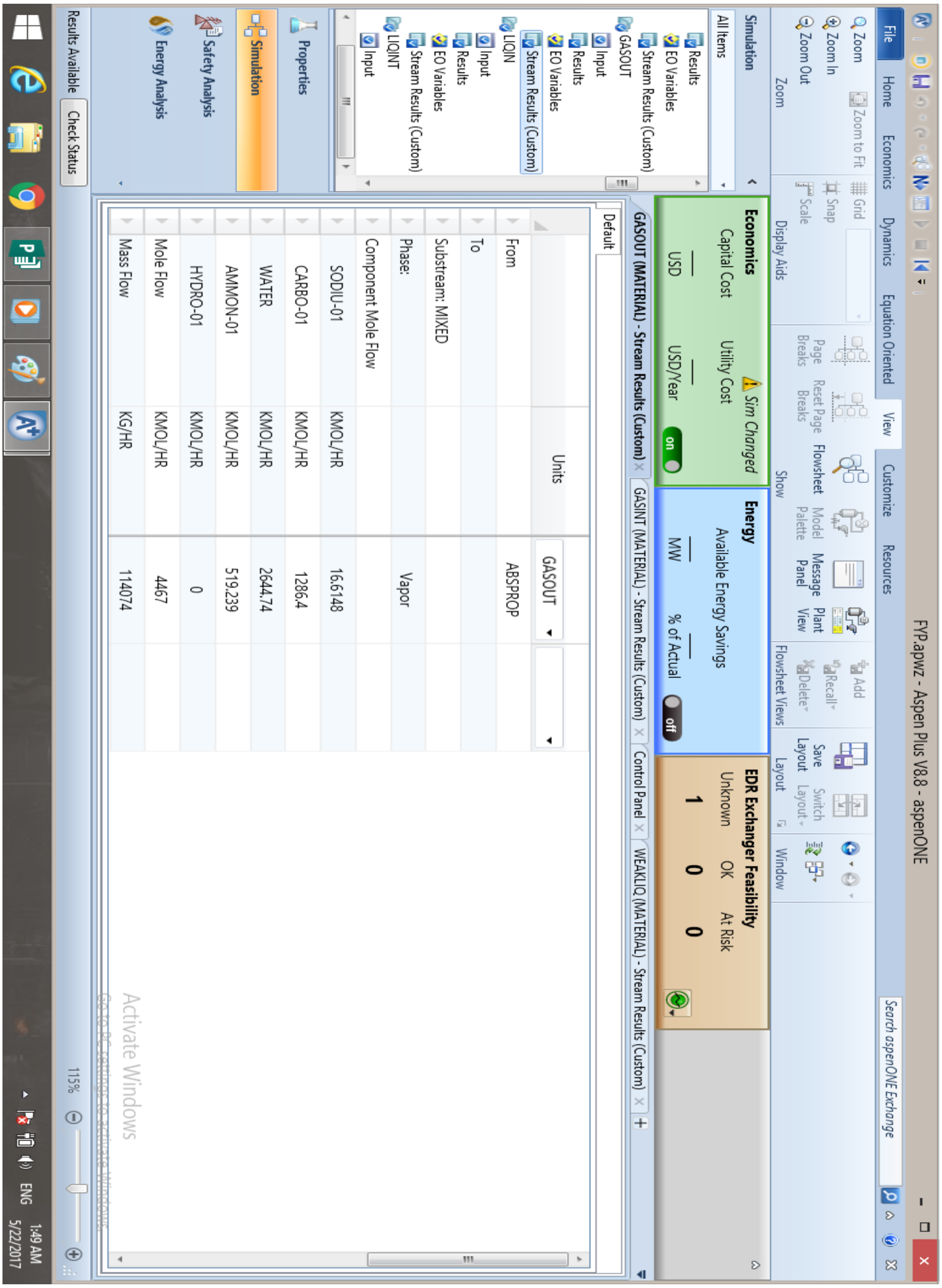


Figure 23 Gas Out Stream

Chapter-8

Design and Costing

8.1. Absorber Design

Stream	Flow (kmol/hr)	Density (kg/m3)
Gas In	3444.6	0.75
Liquid In	3707.56	1140
Gas Out	3175.91	0.75
Liquid Out	3976.1	1190

Table 7 Absorber Flow Rates

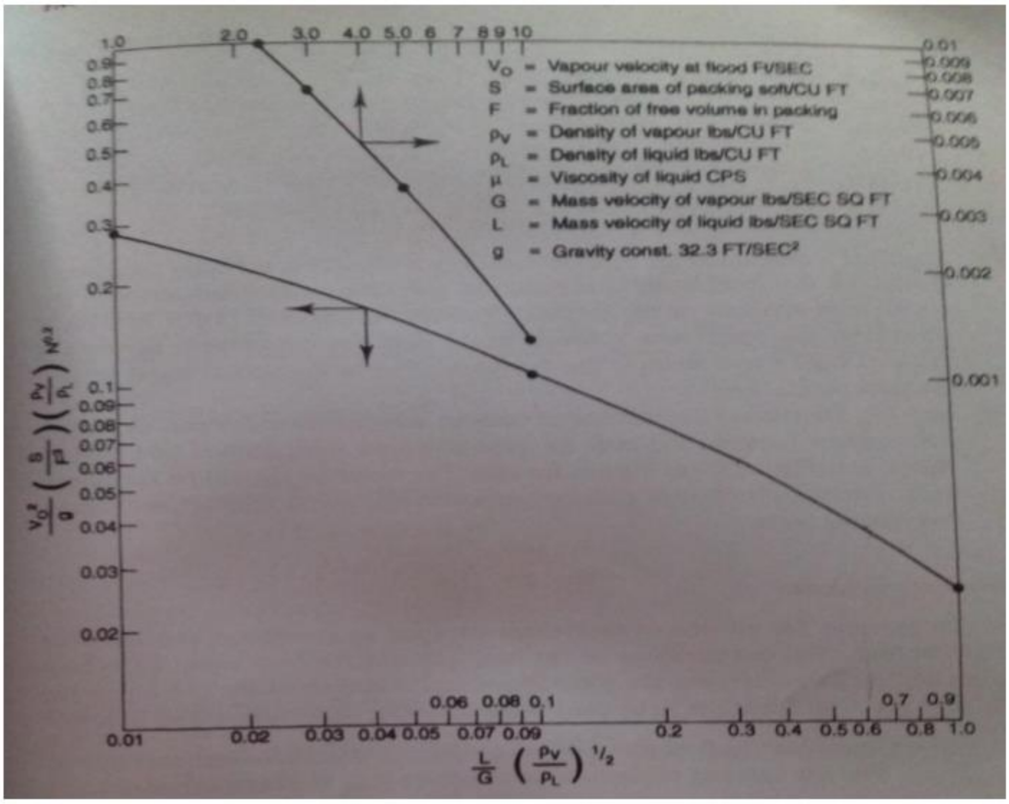


Figure 24 Superficial Velocity Correlation Curve (NPTEL Chemical Engineering Design II)

<u>Superficial Gas Flow Velocity</u>

Packing

Bottom: $\left(\frac{L_{out}}{G_{in}}\right)\left(\frac{p_g}{p_l}\right)^{0.5}$

= 0.0408

Top: $\left(\frac{L_{out}}{G_{in}}\right)\left(\frac{p_g}{p_l}\right)^{0.5}$

=0.0432

Choose 0.0432 [Greater value]

From graph read off corresponding Y axis value,

$$\left(\frac{{v_f}^2}{g}\right)\left(\frac{s}{F^3}\right)\left(\frac{p_v}{p_l}\right) u^{0.2} = 0.18$$

$${v_f}^2 = \left[\frac{0.18 \times g \times F^3 \times p_l}{s \times p_v \times u^{0.2}}\right]$$

V_f: Superficial Gas Velocity (ft/s)

g: gravitational constant 32.3 ft/s^2

F: Free Volume Factor

P_l : Liquid Density (kg/m^3)

P_v: Gas Density (kg/m^3)

u : Liquid Viscosity (Centipoise)

Type	*Ceramic Pall Ring*
Size (inches)	*3*
Free Volume	*80%*
Surface Area (ft^2/ft^3)	*20*

Table 8 Packing Data

u: 1.68cp (NIST)

P_{l}: 1140 kg/m^3

P_{v}: 75 kg/m^3

V_f = 14.25 ft/s

8.1.1. Design Velocity

To find tower C.A, operate below flooding

$$V_{design} = 0.8\, v_f$$

V_{design} = 0.8(14.25)

V_{design} = 11.4 ft/s = 3.47 m/s

$$Area = \frac{[volumetric\ flow\ rate]}{[design\ velocity]}$$

Area = [7.6/3.47]

Area = 2.19 m^2

$$Diameter = \left[\frac{4A}{3.142}\right]^{0.5}$$

Diameter = 1.67m

8.1.2. HETP

HETP(ft) = $1.5D_p + 0.5$ *D_p = packing dia

HETP(ft) = 1.5(3) + 0.5

HETP(ft) = 5 ft

HETP(m) = 1.524 m

TABLE 1

EXPERIMENTAL DATA. TOTAL PRESSURES

t Deg. C.	T Deg. C.	p mm.	Θ Deg. C.	Θ/T Exp.	Θ/T Calc.
Concentration: 12.1 per cent Molal					
0.0	273.1	15.45	......	.....	
10.2	283.3	45.90	195.50	0.690	
19.9	293.0	89.40	204.90	0.698	
30.0	303.1	159.76	212.52	0.703	
40.3	313.4	268.28	220.33	0.704	
50.0	323.1	409.47	229.08	0.710	0.707
60.0	333.1	622.43	235.35	0.707	
70.0	343.1	889.48	242.27	0.707	
81.0	354.1	1280.40	250.69	0.708	
91.0	364.1	1777.00	259.04	0.709	

Figure 25 Vapor Liquid Equilibrium of Ammonia Saline System (Nakazawa 1984)

P_{sat} = Vapor pressure of NH_3 in Brine at 60°C & 12.3% molality

P_{sat} = 622 mmHg

P_{sat} = 0.818atm

Total Pressure = 2.5 atm

Y_i = Liquid phase activity coefficient for NH_3 in brine

$$Y_i = \frac{[Actual\ Vapor\ Pressure]}{[Roults\ law\ Vapor\ Pressure]}$$

Y_i = [0.818]/ [0.133(14.4)]

Y_i = 0.4266

$$K_i = \frac{Y_i P_{sat}}{P}$$

K_i = 0.139

$$A = \frac{L}{KV}$$

A = 13.14

$$N_{sp} = \left(\frac{A}{1-A}\right)\ln\left[\frac{mx_{in} - y_{out}}{\left(\left(1-\frac{1}{A}\right)\left(1-\frac{mx_{in} - y_{out}}{mx_{in} - y_{in}}\right)\right]}\right]$$

x_{in} = 0.013 m=Ki= 0.139

x_{out} = 0.133

y_{in} = 0.218

y_{out} = 0.0243

N_{sp} = 3 stages

In the absence of coordination parameters Cornell's Method cannot be used but a decent approximation can be made.

HETP = H_{og}

$$z = H_{og} \times N_{og}$$

z = 1.524*3

z = 4.527 m

8.1.3. Actual Height

Vatavuk & Neveil Method

$Total\ Height\ (ft) = 2 + z + 25\%$ (Dia of Packed Bed)

Total Height (ft) = 17.6 ft

Total Height (m) = 5.39 m

8.1.4 Pressure Drop Calculations

Using Equation:

$$\frac{\Delta P}{L} = \frac{150\mu(1-\varepsilon)^2 u_o}{\epsilon^3 d_p^2} + \frac{1.75(1-\varepsilon)p u_o^2}{\epsilon^3 d_p}$$

L= Height (m) L=4.532 m

ε= Bed voidage ε= 0.8

D_p = Packing Diameter (m) D_p = 0.0762 m

p= Gas phase density (kgm^{-3}) p = 75 kgm^{-3}

u_o= Superficial velocity (m/s) u_o= 4.34 m/s

μ= kinematic viscosity (Pa.s) μ = 1.68 x 10^{-3} Pa.s

ΔP= Pressure drop in column (Pa) **ΔP= 57.5 kPa**

Absorber Specification Sheet	
Type	Packed Bed
Packing	3 inch Ceramic Pall Ring
Diameter	1.67 m
Height	5.39 m
Packed Height	4.532 m
Absorbent	Brine

Table 9 Absorber Specification Sheet

8.2. Plate Heat Exchanger

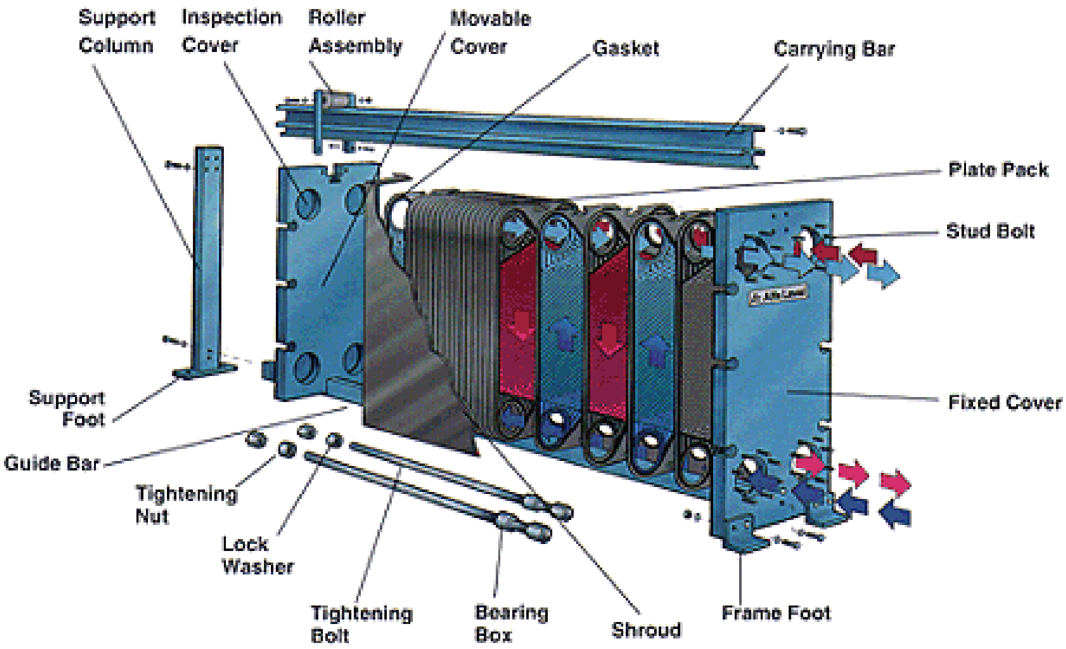


Figure 26 Plate Heat Exchanger

8.2.1. Assumptions

- All the physical properties are constant at standard pressure (1 atm)
- There is negligible heat loss
- The change in kinetic and potential energy is also negligible
- The operating conditions of the exchanger are steady-state
- The fluid streams have no phase change
- The thermal resistances through the wall are distributed uniformly
- The velocity and temperature are also uniform at each side
- The heat transfer area on each side is also taken uniform
- The mass flow rates, cold and hot, are same

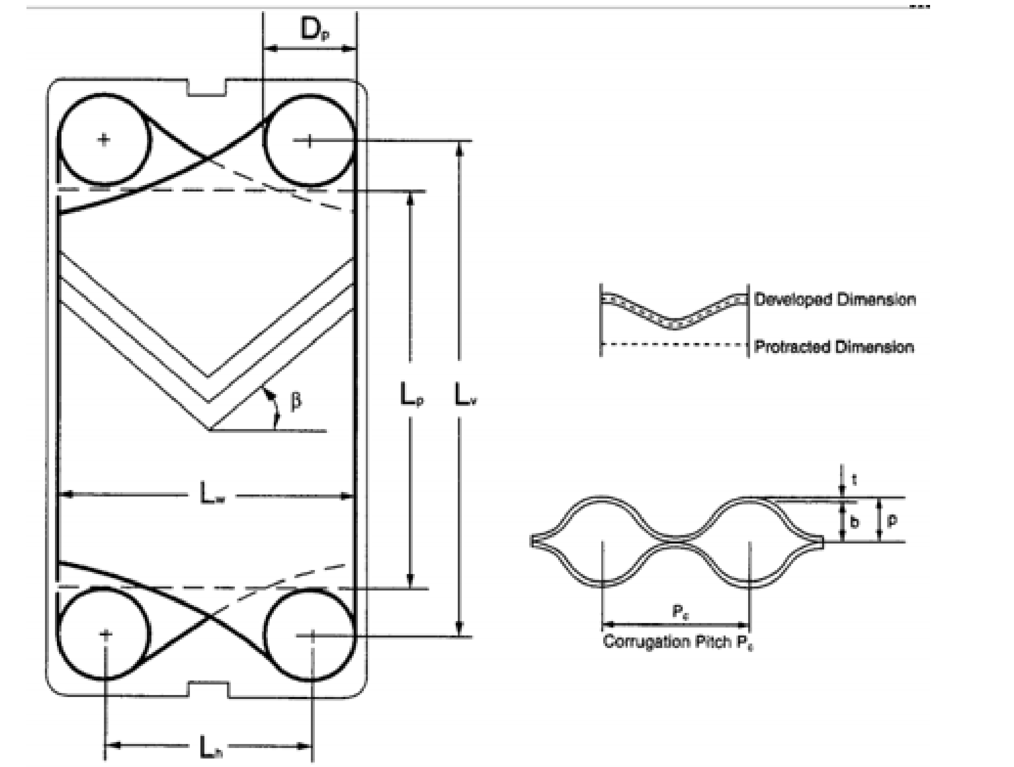


Figure 27 Inside view of a Plate Heat Exchanger

8.2.2. For Liquid Out:

T_{hi}	: inlet hot stream temperature	^0C
T_{ho}	: outlet hot stream temperature	^0C
T_{ci}	: inlet cold stream temperature	^0C
T_{co}	: outlet cold stream temperature	^0C
m_c	: cold stream mass flow rate	kg/s
m_h	: hot stream mass flow rate	kg/s
G_c	: The cold channel mass velocity	kg/m^2s
G_h	: The hot channel mass velocity	kg/m^2s
C_h	: steam heat capacity	J/kg.K

C_c	: ipa-water mixture heat capacity	J/kg.K
Q_c	: Amount of heat transfer under clean condition	W
Q_f	: Amount of heat transfer under fouled condition	W
U_f	: Fouled overall heat transfer coefficient	W/m^2K
U_c	: overall heat transfer coefficient	W/m^2K
A_e	: Actual effective area	m^2
A_1	: Single plate efective area	m^2
A_{1p}	: Single plate projected area	m^2
N_t	: total number of plates	
N_e	: The effective number of plates	
N_p	: Number of passes	
Ncp	: the total number of channels per pass	
Lv	: Vertical distance	m
L_h	: Horizontal distance	m
t	: Plate thickness	m
L_c	: Plate pack length	m
L_w	: Effective channel width	m
p	: The plate pitch	m
b	: the mean channel spacing	m
D_h	: The hydraulic diameter of the channel	m
$\emptyset$	: The enlargement factor	
β	: Chevron angle	^{o}
μ_h	: viscocity of hot fluid	$N.s/m^2$
μ_c	: viscocity of cold fluid	$N.s/m^2$

Pr	: prandalt number	
Re	: reynolds number	
Nu	: nusselt number	
h_c	: convective heat transfer coefficient on clod fluid	$W.m^2/K$
h_h	: convective heat transfer coefficient on hot fluid	$W.m^2/K$
R_{fh}	: fouling factor for hot fluid	$m^2.K/W$
R_{fc}	: fouling factor for cold fluid	$m^2.K/W$
k_w	: thermal conductivity of the plate material	W/m.K

Specification Data

A_c : 100 m^2

L_v : 1.45m

L_h : 0.35m

t : 0.8mm

L_c : 0.40m

L_w : 0.70m

Φ : 1.5

β : 45

N_p : 1

Chevron Angle (degree)	Heat Transfer			Pressure Loss		
	Reynolds Number	C_h	n	Reynolds Number	K_p	m
≤ 30	≤ 10	0.718	0.349	< 10	50.000	1.000
	> 10	0.348	0.663	10–100	19.400	0.589
				> 100	2.990	0.183
45	< 10	0.718	0.349	< 15	47.000	1.000
	10–100	0.400	0.598	15–300	18.290	0.652
	> 100	0.300	0.663	> 300	1.441	0.206
50	< 20	0.630	0.333	< 20	34.000	1.000
	20–300	0.291	0.591	20–300	11.250	0.631
	> 300	0.130	0.732	> 300	0.772	0.161
60	< 20	0.562	0.326	< 40	24.000	1.000
	20–400	0.306	0.529	40–400	3.240	0.457
	> 400	0.108	0.703	> 40	0.760	0.215
≥ 65	< 20	0.562	0.326	50	24.000	1.000
	20–500	0.331	0.503	50–500	2.800	0.451
	> 500	0.087	0.718	> 500	0.639	0.213

Figure 28 Chevron Angle Correlations for Plate Heat Exchangers (D.Q Kern)

<u>Projected Plate Area</u>

$$L_p = L_v - L_w + L_H$$

L_p = 1.17m

$A_p = L_p \times L_w$ **A_p= 0.819m^2**

<u>Single Plate Heat Transfer Area</u>

$A_{sp} = \Phi \times A_p$

A_{sp} = 1.2285 m^2

$$N_{eff} = \frac{A_e}{A_{sp}}$$

N_{eff} = 81.4 plates

Total Plates = 2 + N_{eff} = **83.4**

Plate Pitch

$$P = \frac{L_c}{N_{Total}}$$

P = 0.40/83.4

P = 4.914×10^{-3} m

Mean Channel Flow Gap

$$b = p - t$$

b = 4.114×10^{-3} m

One Channel Flow Area

$$A_{ch} = b \times L_w$$

A_{ch} = 2.89×10^{-3} m^2

Channel Hydraulics

$D_h = \frac{2b}{\Phi}$ D_h= 5.49x10^{-3}m

$$N_{cp} = \frac{[N_t - 1]}{2N_p}$$

N_{cp} = 41.2

Q_H = (3976.1)(76.7)(70-40)

Q_H = 91.5x10^5 KJ/hr

$$Q_{cp} = mCp\Delta T$$

$91.5\times10^5 = m(4.814)(55-25)$

m = 20.25 kg/s

<u>Mass Flow Rate per Channel</u>

$$m_{ch} = \frac{m_w}{N_{cp}}$$

m_{ch} = 20.25/41.2

m_{ch} = 0.49 kg/s

$$G_{ch} = \frac{m_{ch}}{A_{ch}}$$

G_{ch} = **169 kg/m^2/s**

$G_{cc} = G_{ch}$ = 169 kg/m^2s

8.2.3. Water

C_p = 4.18 KJ/kg.K

$\mu = 8.4\times10^{-4}$ Pa.s

k = 0611 W/m.k

Pr = 5.748

$K_{f\ water}$ = 0.0000069 m^2.K/W

8.2.4. Product Stream

C_p = 3.57 KJ/kg.K

K = 0.650 W/m.K

μ = 1.68×10^{-3} Pa.s

P_r = 9.2

m = 85348 kg/hr

n = 0.66

$$R_{ch} = G_{ch} * \frac{D_h}{u_{product}}$$

R_{ch} = 551.58 (hot fluid)

$$R_{ec} = G_{ch} \times \frac{D_h}{u_w}$$

R_{ec} = 1103.5 (cold fluid)

$$H_{hot} = (\frac{k_{hot}}{D_h})(c_h)(R_{eh})(Pr^{1/3})$$

H_{hot} = 4440 W/m^2.K

$$H_{cold} = (\frac{k_{cold}}{D_h})(c_h)(R_{eh})(Pr^{1/3})$$

H_{cold} = 5755 W/m^2.K

$$U_c = \frac{1}{\left[\left(\frac{1}{H_{cold}}\right) + \left(\frac{1}{H_{hot}}\right) + \left(\frac{1}{k_w}\right)\right]}$$

U_c = 2234.8 W/m^2.K

U_f (*Fouled Overall Heat transfer coefficient*)

$$U_f = \frac{1}{\left[\left(\frac{1}{U_c}\right) + \left(R_{f\ water}\right) + \left(R_{product}\right)\right]}$$

U_f = 2168 W/m^2.K

8.2.5. LMTD

$$\Delta T_2 = T_{H\ out} - T_{c\ in}$$

ΔT_2 = 40 – 25 = 15

$$\Delta T_1 = T_{H\ in} - T_{out}$$

ΔT_1 = 70-55=15

LMTD = 15

8.2.6. Actual Heat Duties

Clean Surface

$$Q_c = U_c \times A_e \times LMTD$$

$Q_c = 3.35\times10^6$W

Fouled Surface

$Q_f = 2168\times100\times15$

$Q_f = 3.25\times10^6$ W

Required Heat

$Q_H = 2.54\times10^6$ W

$Q_c = 2.54\times10^6$ W

Fouled surface duty is greater than requirement. Hence design is feasible

Heat Exchanger Specification Sheet	
Type	Plate Type
Effective Area	$100m^2$
Number of plates	84
Mass Flow Per Channel	0.49 kg/s
Hot side Fluid	Ammoniated Brine
Cold side Fluid	Cooling Water
Heat Duty Required	91.5 kJ/hr
Heat Exchanger Duty	3.25×10^6 W

Table 10 Heat Exchanger Specification Sheet

8.3. Compressor

<u>Power to compress Gas_{in}</u>

ΔP required for Gas Flow = 6 inch Hg

Account for pressure loss in column : (0.1 atm/ meter column height)

$= 0.1 \times 4.572 = 0.457$ atm

$P_2 = 1+6(0.0334) + 0.457$

$P_2 = 1.658$ atm

T=333K M:2444.6 kmol/hr

Compound	Molar Fraction
NH_3	0.218
CO_2	0.137
H_2O	0.636
I	0.0087

Table 11 Gas in Compositions

Compound	Critical Temperature
NH_3	405.5K
CO_2	304.2K
H2O	647K

Table 12 Component Critical Temperatures

Compound	Critical Pressure
NH_3	111.3 atm
CO_2	72.9 atm
H2O	218 atm

Table 13 Component Critical Pressures

$T_{c\ mix}$:

$$X_{NH3}T_{c\ NH_3} + x_{CO_2}T_{c\ CO_2} + x_{H_2O}T_{c\ H_2O}$$

= 541 K

$P_{c\ mix}$:

$$X_{NH3}P_{c\ NH_3} + x_{CO_2}P_{c\ CO_2} + x_{H_2O}P_{c\ H_2O}$$

= 172.9 atm

Flow rate = 70000 m^3/hr = 19.4 m^3/s

From figure:

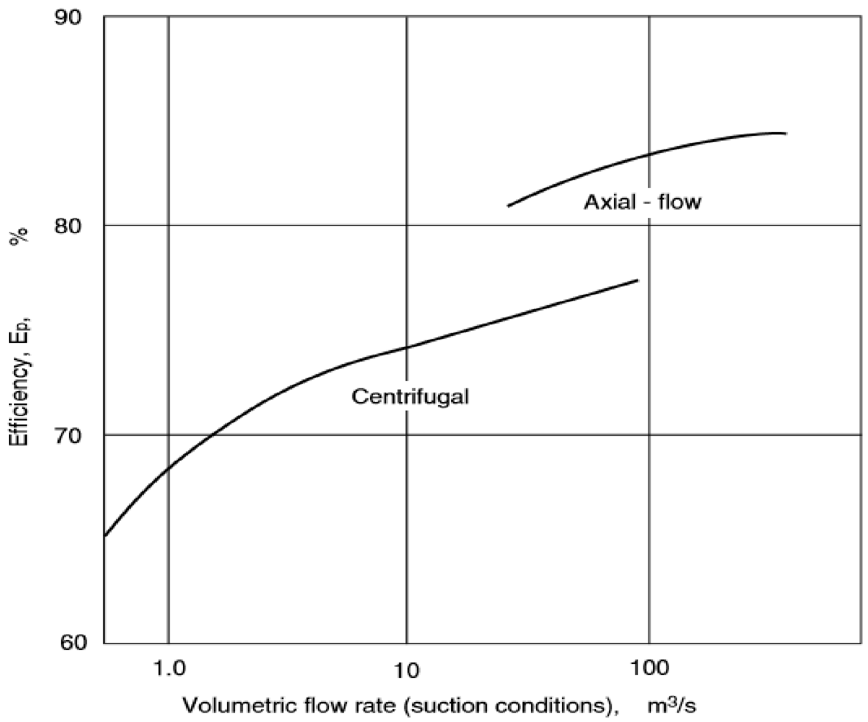


Table 14 Compressor Efficiency Curve (Coulson & Richardson's Vol. 6)

Ep = 0.74

For triatomic gas (NH_3, H_2O, CO_2)

y = 1.33

$$m = \frac{\gamma - 1}{\gamma E_p}$$

m = (1.33 - 1)/(1.33×0.74)

m = 0.338

$$\left(\frac{T_2}{T_1}\right) = \left(\frac{P_2}{P_1}\right)^m$$

T_1 = 280K

$T_{r\ (mean)}$ = 0.567

$P_{r\ (mean)}$ = 0.0077

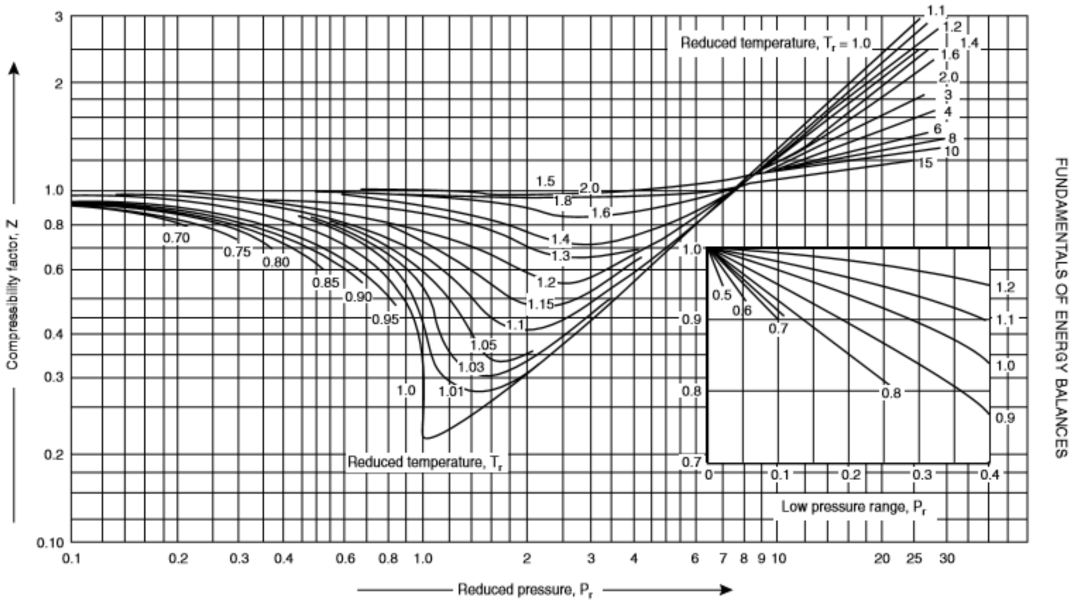


Figure 3.8. Compressibility factors of gases and vapours

Table 15 Compressibility Chart (Coulson & Richardson's Vol. 6)

From Graph : **z = 0.94**

$$n = \frac{1}{1-m}$$

n = 1.51

$$W_p = \left[\frac{RT}{n} - 1\right].n.\left[\frac{\left(\frac{P_2}{P_1}\right)(n-1)}{n} - 1\right]$$

Compressor Specification Sheet	
Type	Centrifugal
Polytrophic Efficiency	74%
Cp/Cv	1.33
Compressibility Factor	0.94
Compressibility Ratio	1.658
Power	1.14 MW

Table 16 Compressor Specification Sheet

W_p = 1240 KJ/kmol

Actual Work = 1240/0.74 = 1676.7 KJ/kmol

$$Power \ = \left[\frac{Actual\ Work}{3600}\right] \times [Molar\ flow\ rate]$$

Power = 1138.6 KW

Power = 1.14 MW

8.4. Pump Design

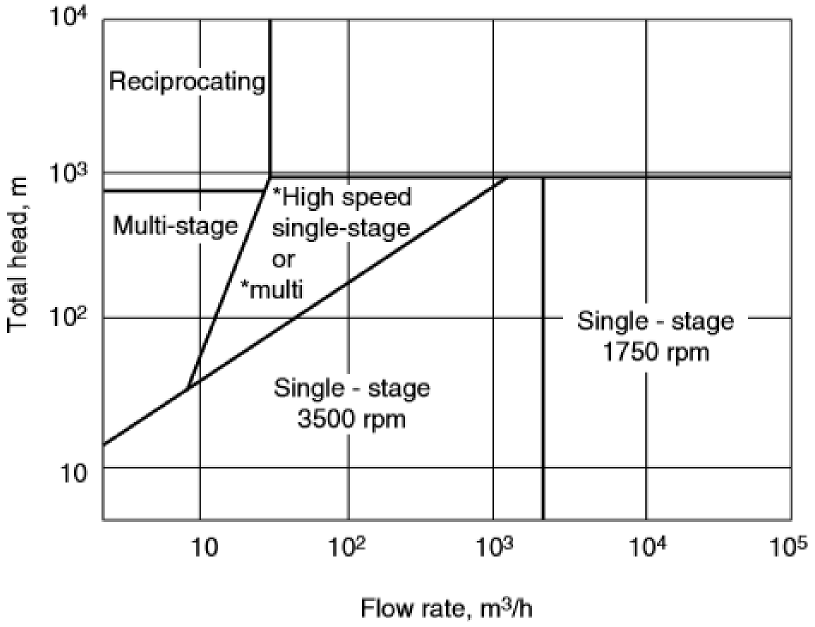


Figure 29 Pump Selection Chart (Coulson & Richardson's Vol. 6)

Liquid In: 3707.56 kmol/hr

Pressure: 30kPa

Density: 1140 kg/m^2

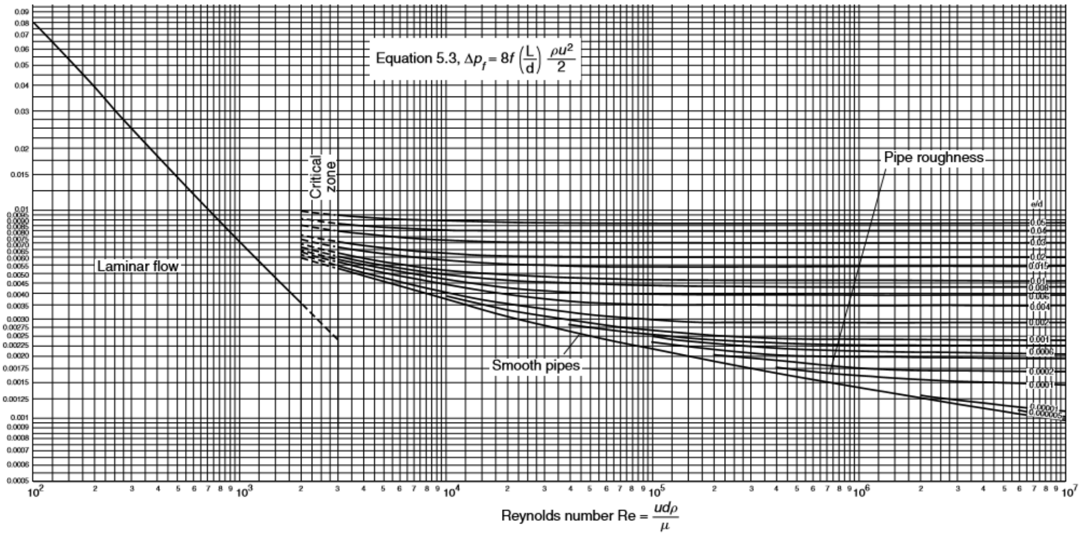


Figure 5.7. Pipe friction versus Reynolds number and relative roughness

Figure 30 Figure 26- Pipe Friction Factor versus Reynolds Number and Relative Roughness

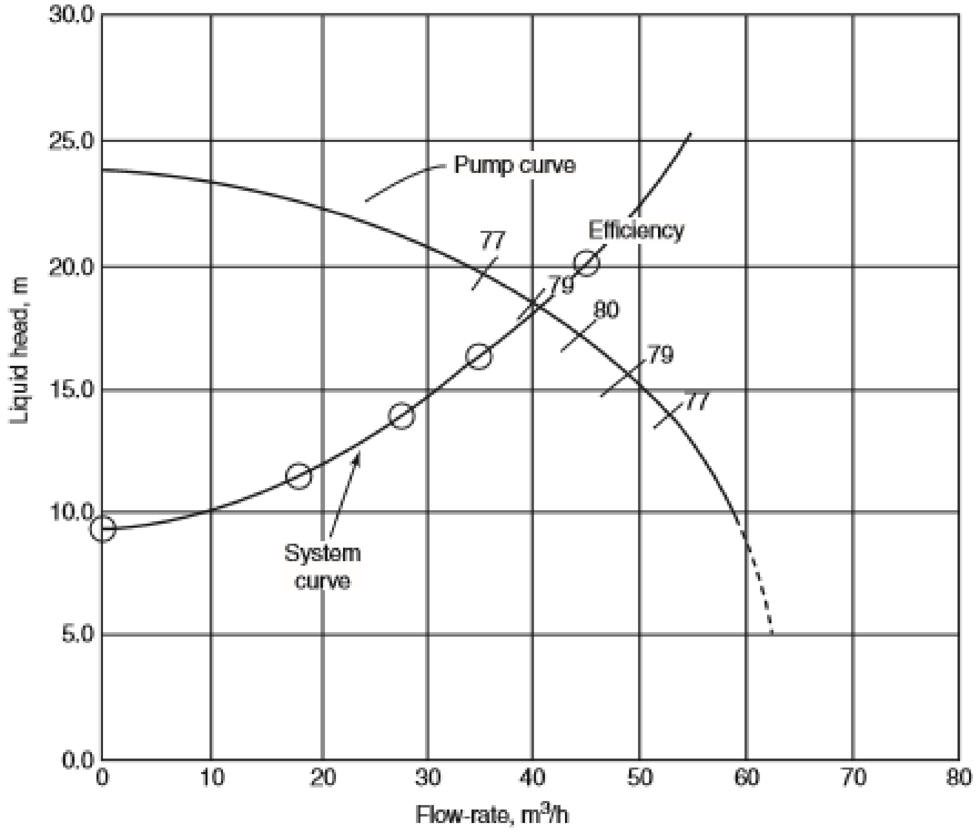


Figure 31 Pump Efficiency Curve (Richard & Coulson Vol.6)

Assume D_i = 6`` = 0.1524m

$$A = (\frac{\pi}{4})D_i{}^2$$

=0.01824m

F = 69m^3/hr

V_L = 1.05m/s

u = 1.68×10^{-3} Pa.s

R_e = [(1140)(1.05)(0.1524)]/[1.68×10^{-3}]

R_e = 108585

Roughness

Comm. Steel Pipe = 0.046

$$Relative\ roughness\ =\ \left(\frac{absolute\ roughness}{D_i}\right)$$

Relative roughness = $\mathbf{3.02 \times 10^{-4}}$

From friction factor chart

f = 0.00215

Pipe length = 30m (assume)

Equivalent Length

2 elbow: 15(0.1524)

1 gate valve (3/4): 40(0.1524)

1 gate valve (FO): 7.5(0.1524)

E` = 9.525m

Total Length = 39.525m

$$\Delta P_f\ =\ 8f\left(\frac{L}{D_i}\right)\left(\frac{p_v{}^2}{2}\right)$$

ΔP_f = 3.135 kPa

Total Head = 9m

Total Energy Needed

$$g\Delta z + \frac{\Delta P}{p} - \frac{\Delta P_f}{p} - W = 0$$

$9.81(6) + (30\times10^3)/(1140) + (2\times10^3)/(1140) = W$

W = 88J/kg

$$Power = W.\frac{m}{n}$$

n = 0.71

Power = 2.78 kW

Pump Characterization

Single Stage = 3500rpm

$$N_s = \frac{NQ^{0.5}}{(gh)^{0.75}}$$

N_s = 0.28

Manufacturer Specification speed

= $N_s \times 1.73 \times 10^4$

= 4850

Impeller Classification

Mixed Flow Impeller

Range = 1500~7000

$NPSH_{available}$

Assume pipeline height from TW = 2.5m

Vapor pressure of brine ~ Vapor pressure of lean/water

@60°C = 18.9kPa

$$NPSH = \frac{P}{pg} + \Delta z - \frac{P_f}{pg} - \frac{P_v}{pg}$$

NPSH = 9.7m

$$NPSH_{available} > NPSH_{required}$$

Pump Specification Sheet	
Type	Centrifugal
Centrifugal Type	3500 RPM Single Stage
Impeller	Mixed Flow Type
Manufacturer (Dimensionless) Specification Speed	4850
NPSH Available	9.7m
NPSH Required	9.0m
Power	2.78 KW

Table 17 Pump Specification Sheet

8.5. Costing

Methodology

CAPCOST for Chemical Engineering Process Plant was used to determine costs for equipment for which accurate estimations were not available. In the case of this project, CAPCOST was used to determine the purchase cost for compressor and pumps.

For the Absorber Column, the cost of packing was estimated for 2017 using CEPCI Cost indices. Similarly, the correlations for vessel erection were used to estimate Column Costs.

For Plate Heat Exchangers, Heat Transfer Engineering, (Hewitt & Pugh) provided the Area to Cost Correlations that enabled costing to be done. Base price from 2007 was extrapolated to 2017 using CEPCI Indices

Once the Purchase Cost of Equipment was calculated, the Lang Factorial Method (Coulson & Richardson's Vol. 6) was used to calculate Physical Plant Cost, Fixed Cost, Working Capital, eventually and the Total Investment Cost.

Since the project is based on a sub-unit of the plant, the payback period could not be calculated. However, the yearly operating cost was calculated and the total investment to Operating Cost Ratio was considered.

Year	CEPCI Index Value
2004 (basis)	444.2
2017	570

Table 18 CEPCI Cost Index

8.5.1. Absorber

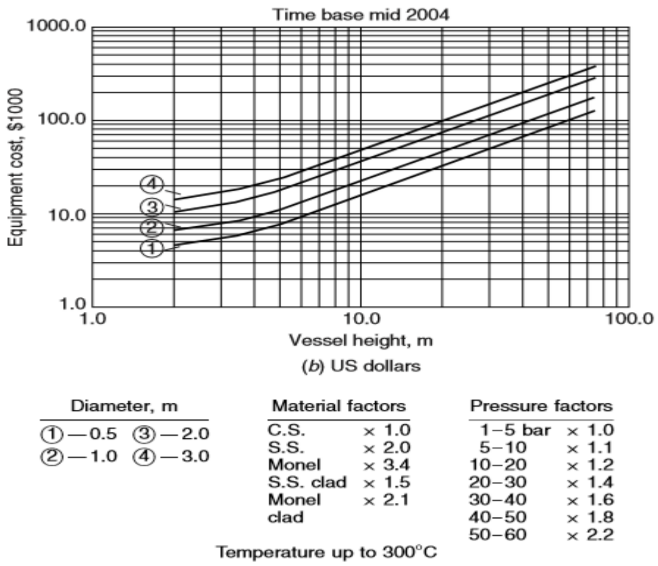


Figure 32 Vessel Cost Curve (Coulson & Richardson's Vol. 6)

Purchase cost (2004) = (48000)(2)(1.1)

Purchase cost (2004) = $ 105600

Purchase cost (2017) = 105600*(570/444.2)

Purchase cost (2017) = $135,500

Packing = $400/$m^3$

Volume of Packing = [(π/4)(2.67)^2(4.537)]

Volume of Packing = 25.4 m^3

Purchase cost [2004] = 25.4(400)

Purchase cost [2004] = $10,161

Purchase cost [2017] = $13,040

Redistributor Cost = $2000

Total Cost = $150,540

8.5.2. Plate Heat Exchanger

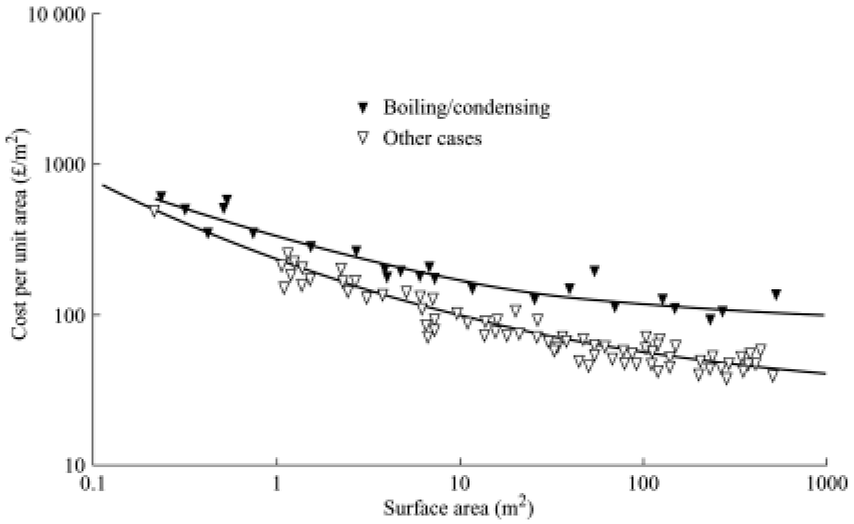


Figure 3 Cost per unit area as a function of area for plate-and-frame heat exchangers [4].

Figure 33 Area to Cost Correlation for Plate Heat Exchangers (Heat Transfer Engineering, Hewitt & Pugh)

Exchange cost per unit area from the curve

£ 100/m^2 (2007)

=100/PPP(2007)

= \$142.5/m^2

[2007]

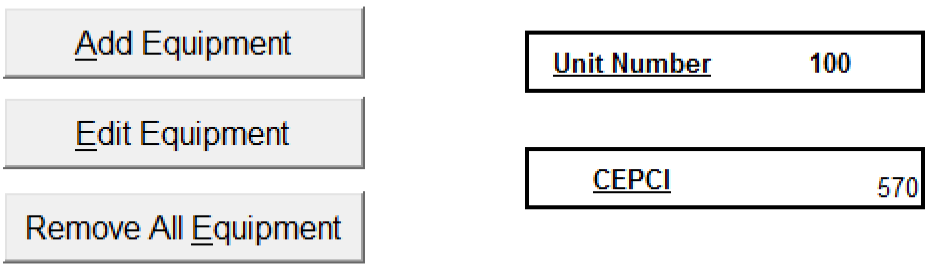
CPI : 525.4

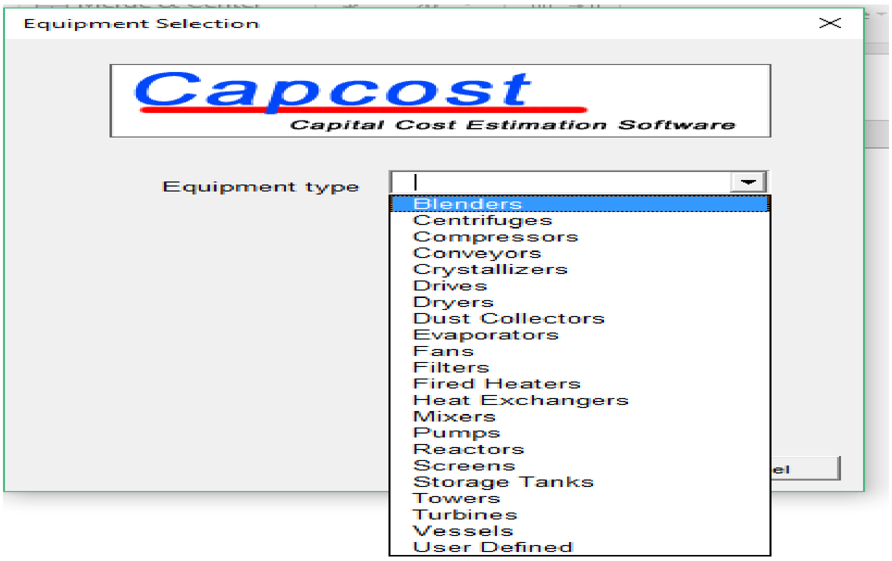
[2017]

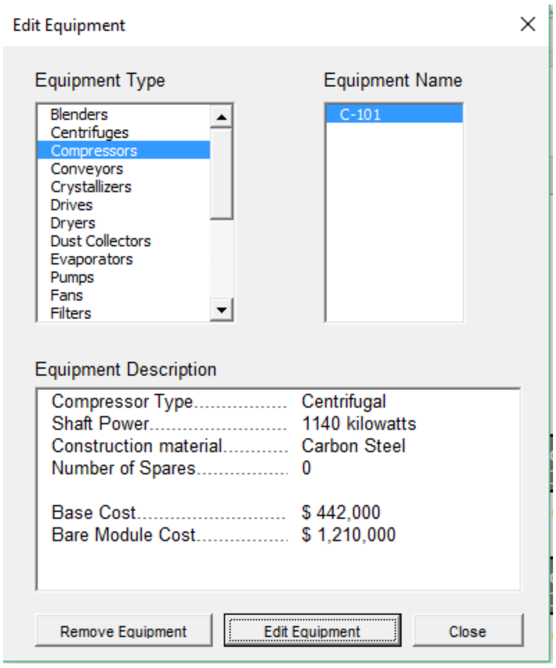
142.5(570/525.4)(100)

=$15,500

Compressor Cost

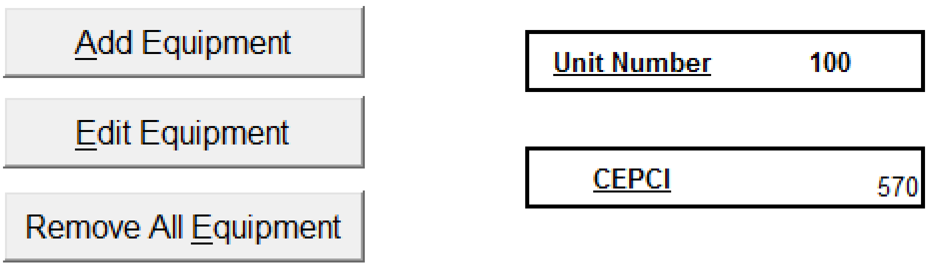


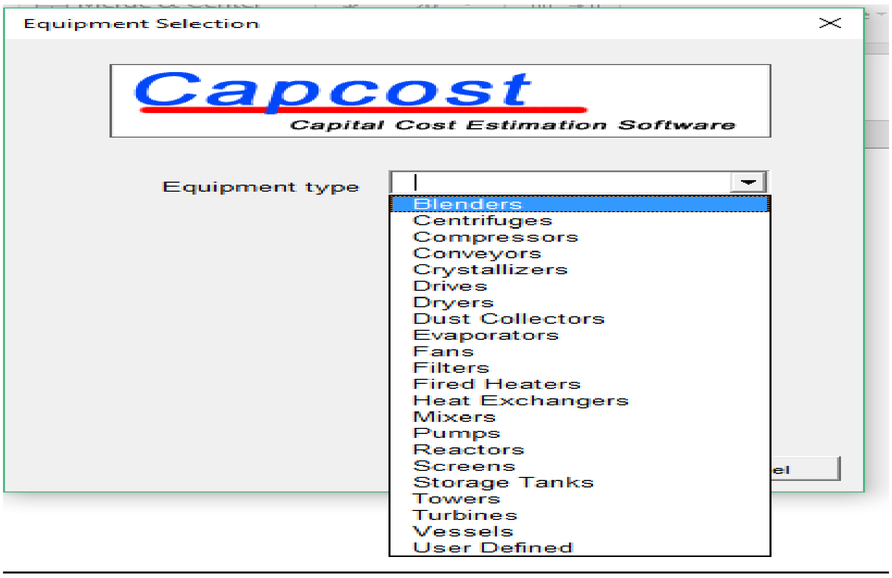


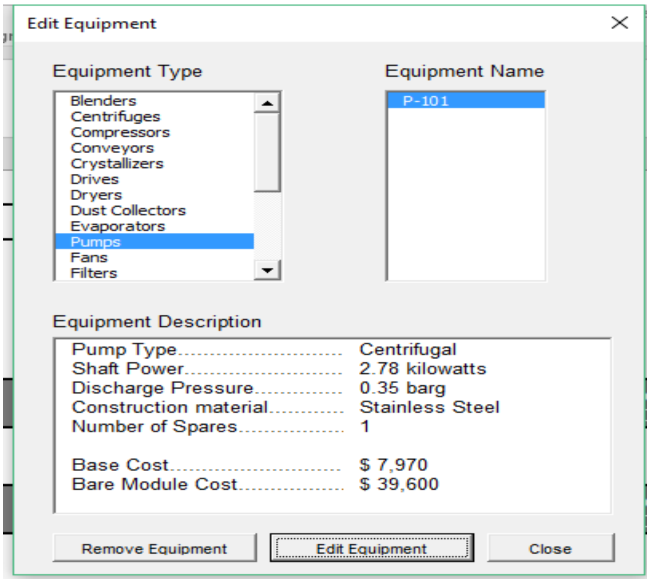


CAPCOST : $442,000 [2017]

8.5.3. Pump







CAPCOST : $8000 [2017] (Spare 1)

8.5.4. Duty Cost

<u>Cooling water for PHE</u>

Duty: 2.54×10^{6}W

Cooling Water: 20.25 kg/s

Assume working hours of 8400/p.a

Cooling water = 20.25(3600)(8400)

Cooling water = 6.12×10^8 kg/year

Cooling water cost = \$0.1/ton

Cooling water cost = \$61,200 p.a

8.5.5. Pump operating cost

Power : 2.78W

Yearly usage : 23,352 KWH

Electricity Price : \$0.1/KWH [NEPRA]

Electricity Price = \$2340 p.a

8.5.6 Compressor operating cost

\$20,000

8.5.7. Purchase Cost Equipment

EQUIPMENT	$	$
Absorber	150,540	
Compressor	442,000	
PHE	15,500	
Pump	8000	
PCE		616,040

Table 19 PCE

8.5.8. Physical Plant Cost

PCE		616,040
F_1	0.4	
F_2	0.7	
F_3	0.2	
F_4	0.1	
F_{total}	1.4[1+1.4][616,040]	
PPC		1,478,496

Table 20 PPC 1

Factor	**Value**
F1-Equiment Erection	0.45
F2-Piping	0.45
F3-Instrumentation	0.15
F4-Electrical	0.1
F5-Buildings	0.1
F6-Utilities	0.45
F7-Storages	0.2
F8-Site Development	0.05
F9-Ancillary Buildings	0.2
Total	3.15

Table 21 PPC 2

Since the project classifies as a minor extension, storage, site development is omitted. Utilities are cost separated separately.

8.5.9. Fixed Capital

PPC		147,849,6
F_{10}	0.3	
F_{11}	0.05	
F_{12}	0.1	1
F_{total}	0.45	[1+0.45][1478496]
Fixed Capital		2,143,820

Table 22 Fixed Capital

Factor	**Value**
F10-Design and Engineering	0.25
F11-Contractor's fee	0.05
F12-Contingency	0.1
Total	1.40

Table 23 Fixed Capital

Working Capital 16.6% of Fixed Capital: **$356,180**

Total investment for startup = **$2,500,000**

Cost of Major Equipment ($)	
Compressor	442000
Plate Heat Exchanger	15500
Packed Bed Absorber	150540
Pump	8000

Table 24 Cost of Major Equipment

8.5.10. Variable Costs

Raw Materials

NH_3 Price: $100/t

Brine : $13/t

NH_3 (5 % make up stream) : **$381,108 p.a**

Brine:

13x 69x8400

$7,534,800 p.a

Utilities

Cooling Water: $61,200

Pump: $2340

Compressor: $20000 p.a

Miscellaneous Material: $10,800

T.V.C: $7,999,448

8.5.11. Fixed Cost

Maintenance: $107,191

Labor: $50,000

Unit Overheads: $10,000

Capital Charges(7%): 175,000

T.F.C: $350,000

Annual Operating Cost: $8,349,448

8.5.12. Unit Analysis

[Operating Cost]/[Investment] = 3.33

From an economic stand point, the change from tray to packed column is feasible as it is smaller than the yearly operating cost by a factor of 3.33.

Chapter-9

HAZOP Analysis

NODE 1 – The Ammoniated Brine Absorber						
ID	**Guide Word**	**Deviation**	**Causes**	**Consequences**	**Safeguards**	**Actions**
A	**Level**	High	• Reduced Gas In Temperature • Reduced Gas In Pressure • Increased Water Flow • Level Controller Malfunction	• Liquid Entrainment in Gas Out	• Level Recorder and Controller	• Adjust Level Control Valve to drain the separator • Adjust T &P Upstream
B	**Level**	No/Low	• Increase in temperature of the Gas in • Increase in the pressure of the Gas in • Level controller malfunction • Leakage/ damage to the separator vessel	• Gas entrainment in the Liquid Out • Damage to the pump due to cavitation. • Absorber running over capacity resulting in	• Level recorder and controller	• Adjust the level controller to retain liquid in the separator. • Adjust the temperature and pressure controllers upstream

				gas slippage		
C	**Temperature**	High	Process deviation upstream	• Increase in gas out flow • Decrease in Liquid out and water out flow	• Temperature Controller Upstream	• Rectify Temperature Control upstream
D	**Temperature**	Low	Process deviation upstream	• Decrease in gas out flow • Increase in Liquid out and water out flow	• Temperature Controller Upstream	• Rectify Temperature Control upstream
E	**Pressure**	High	• Process upset upstream • Pressure recorder and controller malfunction	• Decrease in gas out flow • Increase in water out and Liquid out flow • Increased risk of vessel rupture	• Pressure Recorder and Controller • Pressure safety valve	• Adjust the pressure control valve • Rectify Control System
F	**Pressure**	Low	• Process upset upstream • Pressure recorder and	• Increase in gas out flow • Decrease	• Pressure recorder and controller	• Adjust the pressure control valve

			controller malfunction	in Liquid out and water out flow • Increase risk of Vessel collapsing.	• Pressure safety valve	• Rectify Control System

Node 2- Plate Heat Exchanger						
ID	**Guide Word**	**Deviation**	**Causes**	**Consequences**	**Safeguard**	**Action**
A	**Flow**	High	• Upset Process Flow Conditions Upstream	• Reduction in Temperature Difference for the gas • Increased cooling water requirements	• Flow transmitter and controller • Bypass	• Adjust the flow controller to maintain the flowrate • In case of uncontrollable flow, open bypass
B	**Flow**	Low/No	• Upset in process conditions upstream • Blockage in process flow lines	• Increase in ΔT of the gas Overcooled gas can disrupt the function of the separator	• Flow transmitter and controller • Bypass to flare stack • Shut off valve	• Adjust the flow controller to maintain the flowrate • In case of very low flow, turn on bypass
C	**Pressure**	High	• Process upset upstream • Pressure recorder and controller malfunction	• Increase in ΔT for the gas • Increased cooling water requirements • Increased risk of	• Pressure Recorder and Controller	• Adjust the pressure control valve • Rectify Control • Bypass a fraction of the incoming flow to the flare stack to

				vessel or pipe rupture		reduce the pressure
D	**Pressure**	Low	• Process upset upstream • Pressure recorder and controller malfunction	• Decrease in ΔT for the gas • Increased chances of back flow generation within the system,	• Pressure Recorder and Controller	• Adjust the pressure control valve • Rectify the control • If the pressure is too low, bypass the stream in its entirety
E	**Temperature**	High	• Process upset upstream • Temperature controller failure	• Increase in ΔT for heat exchanger • Increase in cooling water requirement • Risk of damage to heat exchanger plates	• Temperature Recorder and controller	• Adjust the temperature control valve • Rectify the issue with the Temperature Control System

F	**Temperature**	Low	• Process upset upstream • Temperature controller failure	• Low temperature gas can disrupt the function of the separator	• Temperature Recorder and controller	• Adjust the temperature control valve • Rectify the issue with the Temperature Control System

Conclusion

The study in to the "Conversion of Bubble Cap Tray Column to Packed Bed Column for the Ammoniation of Brine" makes some interesting observations. With the developments in industrial design, recent years have seen rapid developments in the design of packed bed and packing, thus leading to an ever-increasing replacement of tray columns. Likewise, packed columns generally have greater efficiency due to greater area of contact as opposed to Tray Columns. Packed columns also have a lesser pressure drop (0.1-0.5 mbar/ stage) as opposed to Tray Columns (7mbar/ stage).

According to the design, the packed bed has a smaller diameter (1.67 m) and height (5.39 m) than the current bubble cap tray column (1.83 m and 7.8 m). These specifications hold for producing the same amount of ammoniated brine as is being produced by bubble cap tray column. In addition, intermediate cooling has been incorporated to limit the rise in temperature and maintain absorption.

In addition, the trouble shooting of the Absorber has been studied along with a reasonably detailed HAZOP Analysis. The economic analysis dictates that a total investment of $2,500,000 would be needed. The project was based on the Absorber, and since ammoniated brine is only an intermediate product the economic feasibility was done as a ratio of operating cost to the total investment needed. With the total annual operating cost coming around to be $8,349,448, this ratio stands at 3.33

Before deciding whether the packed bed column is a more feasible option, certain factors must be considered. Packings are generally more expensive than trays. Also, Tray Columns are much better at dirty service than packed ones. In this case, scaling would be a bigger problem in packed bed because the entire column packing must be cleaned. In a tray column, trays can be individually cleaned by operators via entrance in the side of towers.

Another factor that needs to be considered is the robustness and flexibility of the separation process needed. A packing column is ideal for a separation process that does not require much flexibility in operation. A tray column is better for more flexible and robust operation.

Overall, study shows that it is economically convenient to change the current bubble cap tray column to a packed bed. However, it needs to be ascertained if it is practically convenient to do so keeping in mind the advantages and disadvantages of each kind.

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