STUDY OF EMULSION AND ASPHALTING CONTROL TO OPTIMIZE PRODUCT QUALITY & EQUIPMENT PERFORMANCE

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CERTIFICATE

This is to certify that work in this thesis has been completed by Mr. Hassaan Bader and Mr. Muhammad Hamza under the supervision of Dr. Salman Raza Naqvi at School of Chemical and Materials Engineering (SCME), National University of Sciences and Technology, H-12, Islamabad, Pakistan.

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DEDICATION
TO

OUR PARENTS

Who are always be with us no matter how tough the situation is.

AND TEACHERS

For motivating and directing us to help us reach our best.

ACKNOWLEDGEMENTS

First of all, we bow before Allah Almighty to thank for His countless blessings and rewards who gave us the strength and ability to complete the thesis.

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ABSTRACT

The project regarding the solution of the two major concerns of the upstream oil refineries. One is the formation of the water in oil emulsions at the well head of the oil extraction. The water containing the salt concentration is needed to be removed from the crude oil. The presence of BS&W (Basic Sediments and water) in the crude oil must be within the contractual agreement. Usual range of BS&W in crude varies in the range of 0.2-3 %. Generally, the chemical demulsification technique is used by different oil refineries but the lack of efficiency of the system due to the variation in the supply of the surfactant is the point of concern for the most of the industries as they are lacking the system that can inject the stoichiometric amount of the de emulsifier in the system by taking into account the variation in the concentration of the emulsifier in the feed at well head. The second major problem is that of the precipitation of the asphaltene which is the most heavy class of the compound present in the crude with the average molecular mass of around 700 kg/kmol. The precipitation of asphaltene which is also known as asphalting is highly undesirable as it further cause disturbance in the running plants. Enhanced pressure drop and poor heat transfer of different equipment due to the asphalting is inevitable. The project was much more focused on the chemical solution of the former and the physical solution of the later problem

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LIST OF SYMBOLS

C: Flow rate of the crude oil

μ_c: Viscosity of crude oil

ρ: Density of crude oil

P_p: flow rate of propane

σ: Interfacial tension

g: Acceleration of gravity

g_c: Conversion factor

H: Enthalpy

U: Internal Energy

T: Temperature

T_{pr}: Pseudo Reduced critical temperature

P_{pr}: Pseudo Reduced critical Pressure

F: Flow rate

A: Area

C: Cost

CHAPTER 1

INTRODUCTION

For the sake of simplicity, the crude oil extraction process has been divided into three major operations i.e. upstream, midstream and downstream operations. Downstream operations are much related to the refining of the crude oil. While the rest of the two processes are pertaining to the crude oil extraction from the well and its pretreatment. The crude oil at the well head is in raw form and does contain a variety of lighter hydrocarbons as well. In this regard it is important to understand the gas to oil ratio of the well. Depending upon the GOR wells are classified into three categories:

- Gas well (GOR> 100,000 scf/stb)
- Condensate well (5000<GOR< 100,000 scf/stb)
- Oil well (GOR<5000 scf/stb)

The phenomenon of the formation of crude oil is well known. The crude oil is composed of both hydrocarbons and non-hydrocarbons. The carbon compounds ranging from C1-C120 are there in their composition. Besides the useful constituents, there are other compounds in crude oil as well which are undesirable. In them, the nitrogenous compounds are the major ones. The nitrogen can be present in the aromatic ring structure like pyridine, pyroles, indoles and quinolines, etc. These compounds are detrimental to various unit operations in oil refineries by causing catalyst poisoning and gum formation in fuel oil. Sulfur-containing compounds like thiophenes and sulfides are also present.

1.1 Quality of crude oil:

It is to be noted here that physio-chemical properties of the crude oil are liable to vary from region to region and even within the well the composition of the crude can also show variation. While talking about the quality of crude oil the most important term to be considered is the specific gravity of the crude. In this regard, the crude oil with higher specific gravity is the least desirable. While the crude with lower specific gravity is the most desirable as it contains the higher fractions of much useful low boiling hydrocarbons. If we talk about the oil wells of Pakistan then we can divide them into two main portions on the basis of region. These are the south wells and the north wells. Studies and surveys show that the well the north of Pakistan have lower specific gravity ranges (0.7910-0.850). The wells in Punjab are included in the North region wells. Toot, Khaur, Adhi and Dakhni crude oil wells are famous. While the south region majorly consists of the well at Sindh i.e. Tando Adam, Kunnar, Dabhi, Badin and Umer well have much higher sp. gr. ranges (0.8088-0.9320).

1.2 Oil Extraction in Pakistan:

The process of petroleum exploration on the subcontinent started back in 19th century when a well was drilled at Kundal in district Mianwali in 1866. This was drilled a few years after the drilling of the first oil well in the world in 1859. After 1885 all control of the drilling activities was over taken by the government of India.

Attock Oil Refinery showed successful drilling practice in 1915. But the control of these kind of exploration activities could not remain in the hands of government for the longer run and finally in 1947 private companies took charge and control of these activities.

After the separation of the subcontinent, there was the need of the proper legislation needed to be established for the oil and gas sector. In the upcoming years the drilling activities were also compromised by foreign private companies. Then government of Pakistan decided to developed OGDCL in 1961.

Pakistan Petroleum limited (PPL) was established in 1950s by the incorporation of the assets of Burmah Oil Limited.

Pakistan Petroleum has currently 325 operating wells in which most of them are associated in nature and contain Oil along with NGL/NG. According to the statistics of 2020 PPL holds 1793.5 bcf (billion cubic feet) of NG while in the rest of two sectors the production capacity is 13.3 MMbbl (Million Barrel) and 543.1 Ktons (Thousand tonnes) in oil/ condensate and in LPG respectively. This production carries huge percentage of total oil and gas production of Pakistan i.e. 22%.

1.3 Stability of crude oil:

It is said that the potential of the crude oil to exhibit the phenomenon of asphalting determines the stability of that sample. Asphaltenes are present as a solid component in the oil due to the adsorption of absorption with the resins which make them dissolve into the oil. Hence, we can say that resins are responsible for refraining asphaltenes form precipitation. Eventually, making the system stable. So, to determine the stability of the crude oil it is necessary that it should have greater concentration of resins in it so that the asphaltene remain dissolved in the bulk without asphalting.

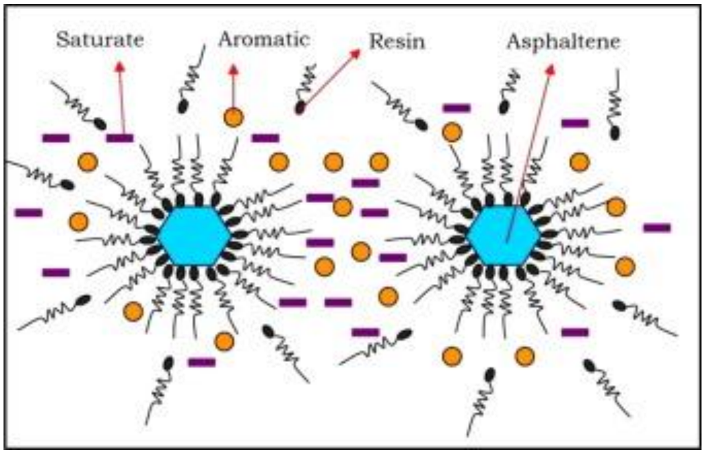


Figure 1 SARA classes

There is an index which is used to determine the stability of the crude oil which is known as Colloidal Instability Index (CII). CII is the ratio of the sum of saturates and asphaltene to the sum of aromatics and resins. The status of the crude oil in the respective range of CII is shown in the table.

1.4 Classes of Compounds:

We generally divide the crude oil into four basic classes of compounds:

1. Saturates
2. Aromatics
3. Resins
4. Asphaltenes

1.4.1 Saturates:

Saturated compounds are those that do not contain double or triple bonds. The presence of only the single bonds shows that they are completely saturated. Alkanes and their derivatives are included in this class of compounds. They can be present in the form of gases as well. For example, the lighter hydrocarbons like methane, ethane, propane and butane, etc. are comprised of C-H single bonds so these are saturated gases. While other alkanes i.e. C10-C20 etc. liquid in nature.

Status	Range
Stable	CII ≤ 0.7
Unstable	CII ≥ 0.9
Unclear stability status	0.7 < CII < 0.9

Table 1 CII ranges indicating crude stability

1.4.2 Aromatics:

Hydrocarbons can be classified into aliphatic and aromatic hydrocarbons regarding the presence of the ring structure in them. Aromatic hydrocarbons have the C-H ring in their structure. They are called aromatics due to their pleasant smell. Even the word aromatic has Latin origin i.e. “Aroma” which means “smell”. The carbon-hydrogen ring is unsaturated and non-polar as well. There can be the other functional groups attached to the ring. For example Toluene and chlorobenzene etc.

1.4.3 Resins:

Resins have complex structure than that of the aforementioned classes of compounds. Also, the molecular weight of this class of compound is higher than that of the other two classes. Resins contain both the polar and the non-polar parts in their structure. This kind of structure is responsible for the stability of the asphaltene in the bulk of other hydrocarbons present in the form of a pool.

1.4.4 Asphaltenes:

This class is the most complex and has the highest molecular weight in all four types of constituents of crude oil. Asphaltenes are highly polar and very high molecular weight i.e. ~ 700 kg/kmol and extremely high boiling point ~ 525 °C. These are present in the solid phase and are homogenized in the crude with the help of the linking agents or resins. Asphaltenes are highly soluble in aromatic compounds but in n-Alkenes, they don't show any solubility. This property can be exploited for the classification of compounds during SARA analysis. Also, several heteroatoms are present in the asphaltene like O,N and S. Unlike the other compounds there is not a single generalized structure defined for asphaltene. Therefore, researchers have proposed a number of models for the description of the structure of asphaltene which will be discussed later.

1.5 SARA Analysis:

To characterize the crude oil into its four fractions and then quantify them and use the separated entity for further SARA analysis is done. The sample of the crude oil is dissolved in liquid propane in the very first step of the analysis. Asphaltene and resins will not show dissolution in the solvent so they will be precipitated while the rest of the two classes will be soluble in it. The distinction between aromatics and saturates can be made by using different separation techniques like chromatography. Then asphaltenes and resins can be treated further by dissolving them into any of the n-Alkane as a solvent. So, to dissolve the resins and precipitate asphaltene from it.

The tree diagram ahead depicts the whole analysis briefly.

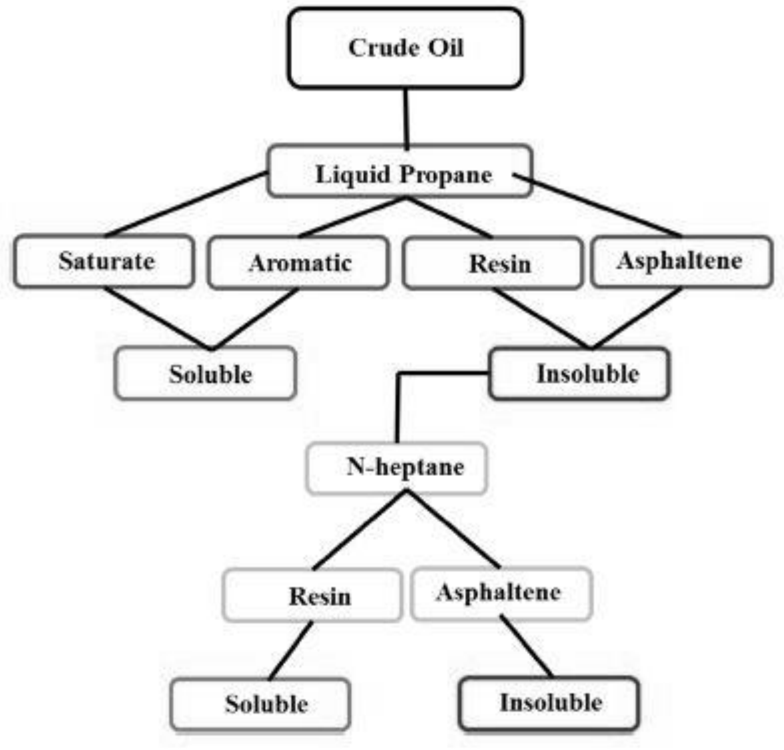


Figure 2 Solubility of the four classes

1.6 Emulsification:

The liquid when dissolved into another liquid or solvent in a minute quantity and it becomes stable thereby dispersing itself into the solvent. Then the kind of mixture forms is known as an emulsion. Crude oil also contains emulsions in it which are highly undesirable. The water in oil emulsions is the most common in which water is made stable in the crude oil with the help of the binding agent that is resins or asphaltenes generally. The binding agent is known as "Surfactant" or "Emulsifier".

As discussed earlier, resins have polar heads with a non-polar tail which makes them capable of forming emulsions by contacting the polar head with water droplets from the reservoir and non-polar tail that remains dissolved in the bulk of crude due to its lyophilic nature. So, there at the interface of the water and oil, there is a sort of balance is generated which is known as hydrophilic lyophilic balance (HLB). This is a very important factor to be kept in mind while choosing a de-emulsifier.

On the basis of the continuous and dispersed phase of the emulsion, we can divide the emulsions of crude oil into the following four categories:

- Water in Oil (W/O) emulsions
- Oil in Water (O/W) emulsions
- Oil in Water in Oil (O/W/O) emulsions
- Water in Oil in Water (W/O/W) emulsions

Water in Oil emulsions is the most common among these. The continuous media in this case is the crude oil while water is the dispersed phase. This is highly undesirable due to multiple reasons:

- The presence of chloride ions in the aqueous phase makes it detrimental to the pipeline
- The pressure drop is very high. So, pumping becomes much energy-intensive.
- The viscosity of the crude becomes much higher. This is due to the stability of the small droplets of the water in crude oil in case of tight emulsion. While in loose emulsions, due to the higher droplet size of the water, the dispersed media is not stable which helps to reduce the viscosity.
- The specific gravity becomes higher which is not desirable.
- Corrosion and the scaling of the equipment is the problem.
- It can cause catalyst poisoning in the downstream processing.

They understand the phenomenon of emulsification we need to know that how does this occur. Surface tension (SF) and Interfacial tension (IFT) are the two terms used for the liquid-gaseous interaction and liquid-liquid interaction respectively. The cohesive forces in the bulk of the liquid create the situation in which the top layer of the molecules faces a lot of inward pull or the cohesion from the sides molecules but not from everywhere as in the case of the molecule present in the bulk medium. This limits the movement of the molecules at the surface of the interface of the two liquids. That strain created at the interface is known as interfacial tension in the case of two liquids that are not miscible in each other. And “surface tension” in case of the liquid in contact with the gas.

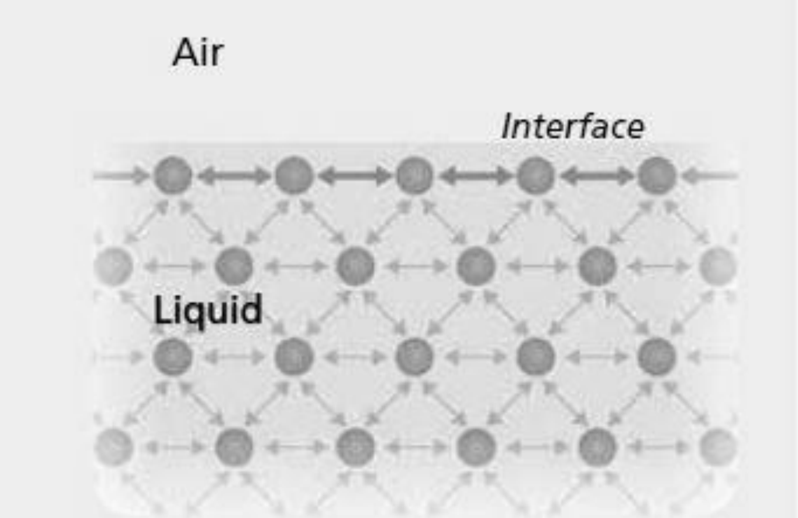


Figure 3 Demonstration of interfacial tension

In the case of ‘free water’, the interfacial tension between the water and oil is relatively high as compared to the case when W/O emulsion is formed. The interfacial tension is weakened by the surfactant itself. Due to weak IFT, the emulsions get stability.

To break that emulsion, we need to invade at the interface such that the surfactant molecule undoes its role of stabilization of the emulsions. There are several methods for that for example, heating, agitation, and chemical addition. Among these, chemical demulsification is considered to be the best due to its economic viability easy to use method. The choice of suitable chemical is the crucial task here. So, HLB comes to play its role. For the case of water in oil emulsions such de-emulsifier is chosen that has a greater value of HLB. It is to be noted here that HLB is none other than one-fifth of the molar percentage of the polar compounds in the de-emulsifier. So, we can have the HLB value in the range of 0-20. For the greater value of HLB, the de-emulsifier will be more inclined towards the polar or hydrophilic nature and vice versa for the lower HLB which will be favorable in the case of O/W emulsions.

1.7 Mechanism of Demulsification:

There are basically three steps of the demulsification:

1. The reaction of de-emulsifier with the surfactant to displace it.
2. Agitation, creaming, or sedimentation.
3. Phase separation due to coalescence of water droplets.

The coalescence of the water droplets occurs due to the higher interfacial tension of the droplets of the water than before due to the breakage of the thin film. This step of phase separation always occurs after the step of sedimentation of flocculation.

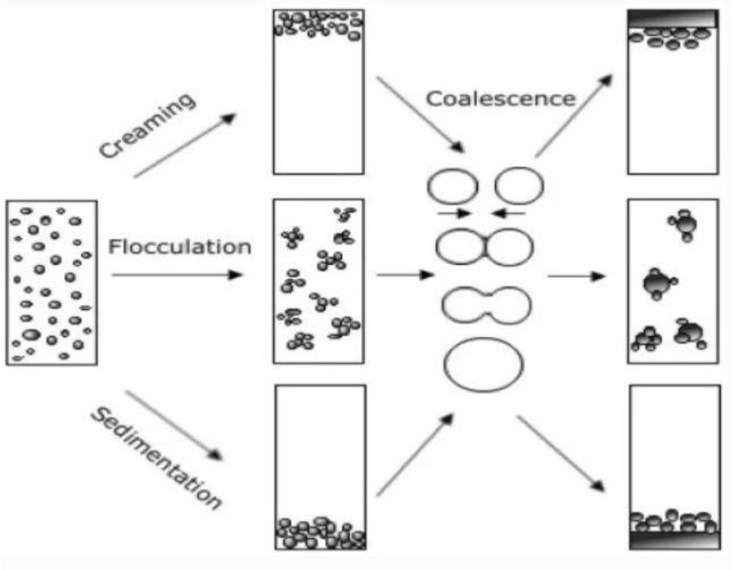


Figure 4 mechanism of de-emulsification

1.8 Methods of Demulsification:

1.8.1 Heating:

Heating the most common method of the de emulsification. The process of heating adds kinetic energy to the molecules of the liquid which decreases the viscosity of the liquid crude which is due to the intermolecular forces between the water molecules. The weak intermolecular forces due to enhanced kinetic energy of the bulk tends to decrease the viscosity of the liquid crude. By beating it becomes easy for the emulsions to break apart due to their deactivation.

The collision of the water molecules as increased, the chances of the coalescence of the separated water phase due to the deactivation of the emulsifier. The causes the formation of the two phase whose separation is required.

The increased heating of the system can raise the density difference between the separated water and oil which aids the separation of the two phase form one another.

1.8.2 Chemical method:

The chemical method of the demulsification is widely used due to its less energy consumption. The de-emulsifier used in this method must have the strong interaction with W/O interface. Also, there must be the ability to flocculate the water molecules by bringing them together to form the large clusters. The water is being separated by the chemical reaction though it's the common notion it is directly attracted by the de-emulsifier.

For the case of weak surfactant the function of coalescence is performed by the flocculation itself but for the safe side the action of the emulsifier must be such that that it should promote the rupture of the water and oil interface. The task of de-emulsifier is make emulsifier inefficient by converting it into the ineffective product as a result of chemical reaction.

Agitation:

The purpose of agitation is to support the coalescence by raising the turbulence in the emulsion and increasing the chance of the collision and settling down the water oil emulsions. However, the introduced agitation must generate the Reynold number within the range of 5000-100000 as higher turbulence can result in the stabilization of the emulsions again.

1.8.3 Coalescing plates:

Another method is to introduce the coalescing plates in the vessel to aid demulsification process. The plates acting as a baffle introduce the turbulence in the chamber. These plate are generally made perforated to provide the settling space for the water droplets to coalesce. Moreover, the perforations cause the formation of the eddies which causes further turbulence needed for coalescence.

Electrostatic Coalescence:

The electrically charged plates can do the thing as well. The water oil emulsion when put under the influence of the two electrically charged plates, give rise to the coalescence of the water droplet due to many reasons:

The water becomes polarized in the electric field. Due to the presence of the poles the water droplet can built affinity for each other and ending up in coalescence to form larger droplet.

The impurities present in water molecules and the polar nature of the water itself encourages it to be close to one of its respective charged plates.

The provision of alternative current supply breaks apart the water oil interface by doing the work against surface tension the water droplets are elongated in one cycle and they are retained back to its droplet shape into the other cycle of the sinusoidal current. This makes the interface weak and ultimately breaks the emulsion.

This process seems to be very useful but it comes with the challenges as well. The major challenge is that of “chaining”. This is the phenomenon in which the polar droplets joins to form the long chain that connects both the grids. Hence, the short circuiting occurs that loses a lot of energy. This generally happens if the water in emulsions is present in high concentration.

1.8.4 Centrifugation:

Due to the difference in the densities of the crude oil and water. The centrifugal forces can be utilized to separate the water and oil emulsions. The centrifuges are installed to take care of that problem in most of the upstream oil refineries but their application is made limited by many of the factors including high initial cost and tedious maintenance with high operating cost.

CHAPTER 2

PROCESS DESIGN METHODOLOGY

Designing is an action meant at giving a complete explanation of a manufacturing system or part of a system. For successive engineering and utilization these explanations characterize an explicit specification of size, performance and other characteristics are important.

2.1 Process Flow Diagram

The process flow diagram for the production of Chlorine Dioxide is developed so that the material and energy balances on the equipment can be applied.

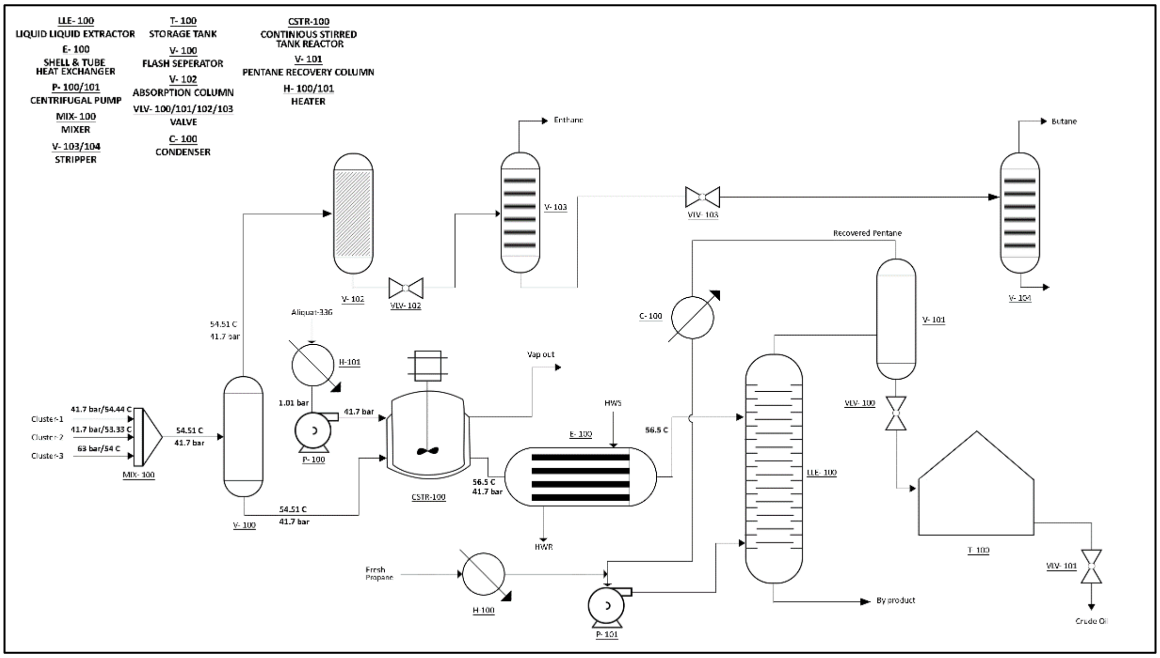


Figure 5 Process Flow Diagram

MATERIAL BALANCE AND ENERGY BALANCE

Material and Energy balance is applied on every equipment of the whole unit to calculate the flow rates and other parameters for every stream. Following are the assumptions we have taken for our material balance:

Steady State System

Basis 1h

In= out

Molar mass of Crude oil=1924.55 kg/kmol

Molar mass of NG= 22.50

Density of crude oil= 1164.13 kg/m3

Density of NG= 346

2.2 MATERIAL BALANCE

2.2.1 Mixing Vessel (MIX-100):

In this vessel crude oil from 29 working wells is collected at three clusters from where it is send to mixing vessel and with product crude at 54.51C and 41.7 bar.

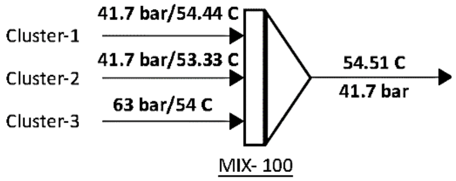


Figure 6 Mixer

Molar flow rate I

Molar flow rate Plant-02= 2285.4 kg

Molar flow rate Plant-03= 28560.3 kg

Mole Balance on Mixer:

=P-01 + P-02 +P-03

=2285.4 + 2285.4 +28560.3

=74268.6 kg

2.2.2 Flash Tank (V-100):

Flash tank separates gaseous components from crude oil before it goes for further treatment of De-emulsification & de-asphalting.

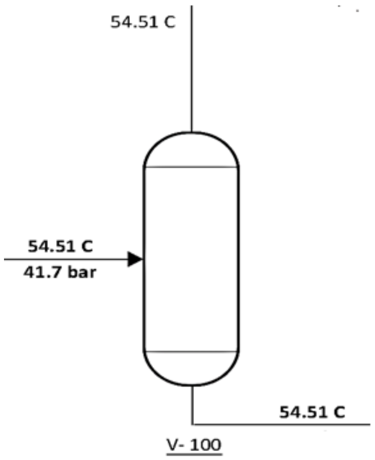


Figure 7 Flash Separator

$\sum_{i=1}^{n} \frac{(ki-1)zi}{1+\frac{(ki-1)V}{F}} = 0$ (Richford Rice equation)

$$xi = \frac{zi}{1 + \frac{(ki-1)V}{F}}$$

$$yi = \frac{ki\ zi}{1 + \frac{(ki-1)V}{F}}$$

$$\frac{F}{V} = 0.000234$$

F=3.28V

$Z = V + F$

3237= 3.28V +V

3237=(1.000234)V

V= 3237/ 1.000234

V= 3236.25 kmole

V= 3236.25*22.5

V= 72815.625 kg

F= 3237-3236.25

F= 0.755 kmole

F= 0.755 * 1924.55

F=1453.03 kg

2.2.3 Reactor (CSTR-100):

In emulsion formation, a polar compound not only assist the colloid formation but also help it sustain for a long period.

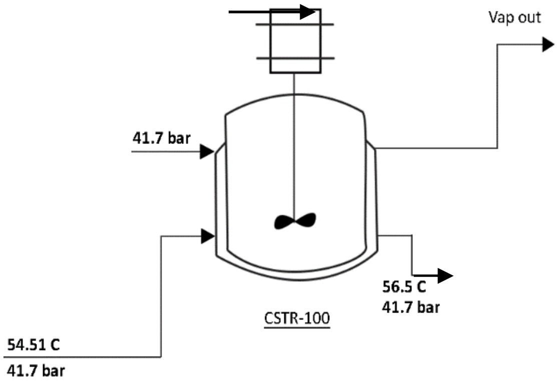


Figure 8 CSTR

C25H54COO-Na+ + C25H54Cl-N+ → (c25H54)2COO-N+ + NaCl

nA-336 = 0.1044 kgmol

ncrude= 0.755 kgmol

nresins = 0.755(0.0154) = 0.1163 kgmol

Limiting Reactant is Aliquat-336

conversion = 90%

Reacted nA-336 = 0.90 * 0.1044 =0.09396 kgmol

Unreacted nA-336 = 0.1044 – 0.09396 = 0.01044 kgmol

n(C25H54)2COO-N+ = 0.09396 kgmol

n(C25H54)2COO-N+ = 0.094 kmol

n(C25H54)2COO-N+ = 766* 0.094

n(C25H54)2COO-N+ = 72 kg

nNacl = 0.094 kmole

nNacl = 0.094*58.5

nNacl = 5.5 kg

2.2.3 Extraction Column (LLE-100):

Propane is used as a solvent for de-asphalting of crude in a Liquid-Liquid Extraction Column.

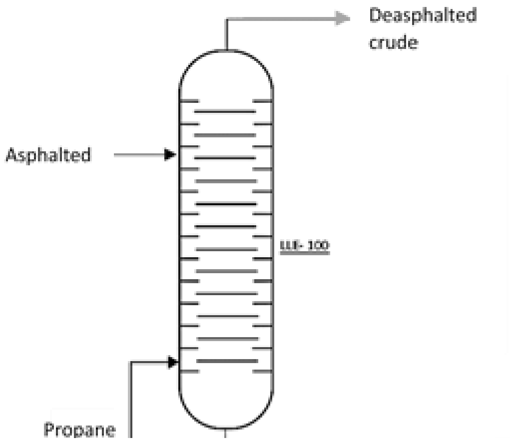


Figure 9 Extraction Column

Propane to the extraction column must be more than twice mC5 = 3000 kg

mCO = 1453 kg

Overall Balance

C-5 + crude = Raffinate+ Extract

3000 + 1453 = Rn +E1

Rn +E1 = 4453 kg

E1= 4453 - Rn

Asphaltene Balance

NCO (xA) + nC5(xA)= Rn(xA) +E1(xA)

1453(0.4318) + 3000(0) = E1 (0.00125) + Rn (0.43)

627.40 +0 = 0.00125(4453 - Rn) + 0.43Rn

326.6 =0.4287Rn + 5.566

Rn= 321.0/0.4287

Rn= 748.68 kg

Rn +E1 = 4453 kg

E1 = 4453- Rn

E1 = 4453- 748.68

E1 = 3704.31 kg

2.2.4 Propane Recovery Column(V-101):

Propane along with de-asphalted crude is obtained from top of LLE column from where a propane recovery column is attached to recover propane and to be reused in the operation.

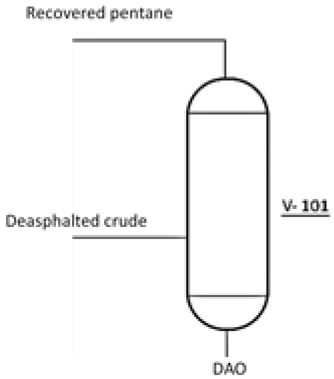


Figure 10 Propane recovery column

$$\sum_{i=1}^{n}\frac{(ki-1)zi}{1+\frac{(ki-1)V}{F}}=0$$

$$xi=\frac{zi}{1+\frac{(ki-1)V}{F}}$$

$$yi=\frac{ki\ zi}{1+\frac{(ki-1)V}{F}}$$

$$\frac{V}{F}=0.908$$

V=0.908F

$$Z=V+F$$

3704.31= 0.908F +F

3704.31=1.908F

F= 3704.31/ 1.908

F= 1941.46 kg

$$V=Z-F$$

V= 3704.31 -1941.46

V= 1762.85 kg

2.2.5 Recycle:

Recycle + Fresh C5 = C5 Fed

1763 + mC5 = 3000

mc5 = 3000- 1763

mc5 = 1237 kg

2.3 ENERGY BALANCE

Some important data which is to be used to calculate energy balance is listed here.

cp =$\frac{1}{\sqrt{d}}(0.338 + 0.00045t)$

T= temperature in Fahrenheit

D= specific gravity of the feed

c= specific heat Capacity of petroleum (kCal/gK)

Density of liquid phase=979.9 kg/m3

Density of vapor phase= 346.05 kg/m3

Feed density= 185 kg/m3

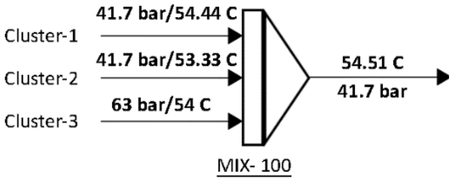
Molar mass of crude= 1381 kg/kgmole

cp= 1061 kJ/kgmol K

cp NG= 44

2.3.1 Mixing Vessel (MIX-100):

Crude from three well clusters are collected and mixed in a vessel. These streams come at different temperature and pressure. The enthalpy change occurred is calculated below.



T1= 54.44 °C

n= 996.1 kmole

H1= 996.1*1061*54.44

H1= 1.59 * 104 kW

T2= 53.33 °C

n= 996.1 kmole

H2= 996.1*1061*53.33

H2= 1.56 * 104 kW

T3= 54 °C

n3= 1245 k moles

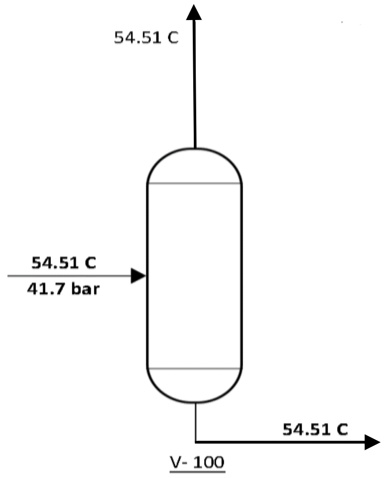
H3= 1245*1061*54.00

H3= 1.98 * 104 kW

Hfeed= 1.59 * 104 + 1.56 * 104 +1.98 * 104

Hfeed =5.13 * 104 kW

2.3.2 Flash Tank(V-100):



HF=nF cp ΔTF

HF= 0.755*1061.2*54.51

HF= 43673 kW

HV= nv cp ΔTv

HV= 2482* 44* 54.51

HV= 5.95* 105 kW

2.3.3 Aliquat-336 Heater (H-101):

ΔHA-336=(nCp ΔT)A-336

ΔHA-336=0.1044*626.2*(55-25)

ΔHA-336= 1961.25 kW

2.3.4 Aliquat-336 Pump (P-100):

Isentropic work= H2 – H1

=0.1044(3.4-1.56)*104

= 1920.96 kW

Adiabatic Efficiency= Ƞ = 75%

Actual work = WA= $\Delta Hs/\eta$

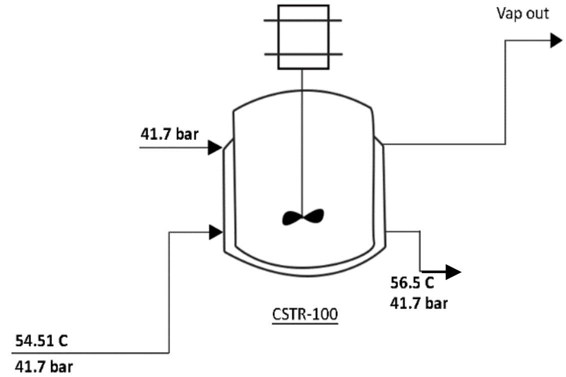
Actual work = 1920.9/0.75

Actual work =2561.2 kW

2.3.5 Reactor (CSTR-100):

The cold mixture is separated in this tank based on density mixture. Oxygen is removed from the top in gaseous form and chlorine dioxide is taken from the bottom and stored in a storage tank.

Exothermic reaction:



Hrx = -3970 kW

HA-336= 0.1044*626.2*55

HA-336= 3.595*104 kW

HCO =0.755* 924.2*54.51

HCO =3.80*104 kW

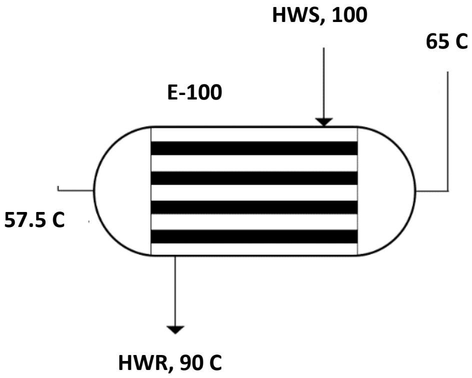
HCRV-Liq =HCO+ HA-336 + Hrx

HCRV-Liq = 3.80*104+3595.64-3970

HCRV-Liq = 3.43*103 kW

2.3.6 Heat Exchanger (E-100):

De-emulsified crude is heated with shell & tube heat exchanger to meet the require temperature for extraction process.



Feed temperature= 57.7 °C

Outlet temperature= 65 °C

Water in = 100 °C

Water out= 90 °C

ΔHF = 1453*2.275*(65-57.7)

ΔHF = 24130 kW

ΔHwater = m*4.18*(100-90)

ΔHwater= ΔHF

m*4.18*(100-90) = 24130

m= 24130 / (4.18*18*10)

m= 547.12kg

2.3.7 Propane Heater (H-100):

C-5 Inlet temperature= 25 °C

Required outlet = 57 °C

ΔHc5 = 3000(0.168)((57-25)

ΔHc5= 3.31 *103 kW

2.3.8 Propane Pump (P-101):

Isentropic compression work= ΔHS= H2-H1

= (9.57-4.2) *3000

ΔHS= 1.61* 104 kW

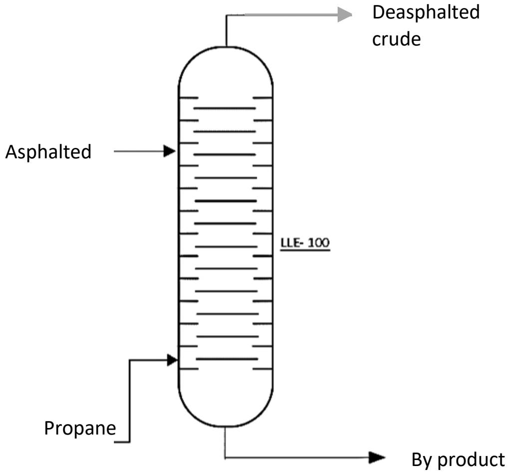
Adiabatic efficiency=η= 75 %

Actual Work= Wc =1.61 *104 /0.75

Wc = 2.15 * 104 kW

2.3.9 Extraction Column (LLE-100):

Propane is used as a solvent for de-asphalting of crude in a Liquid-Liquid Extraction Column.



H_{c5} = (mcpΔT) c5

= 3000(0.168)(61.5)

= 3.099*104 kW

HDAO= (mcpΔT)DAO

= 3400(95.86)(61.5)

= 5.57*104 kW

HP = (mcpΔT)P

= 356.1(1835)(61.5)

= 1.12 * 104 kW

2.3.10 Condenser (C-100):

Hc= Heat of condensation

H= 26.4*3000

H= -7.92 * 104kJ/hr

H= 22 kW

Water in= 50 °C

Water out= 70 °C

Water needed= m= 22/(4.18*20)

m= 947.368 kg/hr

2.3.11 Propane Recovery Unit (V-101):

H_F= m_F cp ΔT_F

H_F= 1941*168*65

H_F= 5887.7 kW

H_V= mv cp ΔTv

H_V= 1762*120*65

H_V= 3817 kW

Components	Feed crude	Flash tops	Flash bottoms	CRV-out	Raffinate	Extract
C1-C5	0.5318	0.8765	0.1085	0.1042	0.0021	0.9752
C6+	0.2682	0.1098	0.4167	0.4074	0.00101	0.0211
H2O	0.0012	0.0137	0.0020	0.020	0.0446	0.0007
Asphalt	0.095	0.00003	0.4318	0.432	0.9057	0.00002
Resins	0.038	0.00002	0.0153	0.0016	0.0034	0
Mass	74268.6 kg	72815.6 kg	1453.03 kg	1453.26 kg	356.02 kg	4096.98 kg

Table 2 Overall Material Balance

Equipment	Inlet Temperature (°C)	Outlet temperature (°C)	Molar Enthalpy (kcal/hr)
Mix-101	-	54.51	1.92 *104
V-100	54.51	54.51	7.19* 108
CRV-100	54.51	55.22	7.168* 108
E-100	55.22	65.1	7.242* 108
T-100	65.12	65.54	6.40* 106

Table 3 Overall Energy Balance

CHAPTER 3

DESIGN CALCULATIONS

3.1 Flash Separator:

GOR of well= 6790 scf/scb

NG flow rate= 775 MMscfd

Condensate flow rate= 11.40 bbl/MMscf NG

Which suggest that it's a condensate well.

Pressure= 41.69 bar =604.7 psia

Temperature= 54.51 °C

Gas SGFG = 0.68

Oil SGFG = 0.9

GOR of the well is within the range of 5000- 10000 scf/scb hence it's a condensate well.

qst=$\frac{2.4D^2Kp}{z(T+460)}\sqrt{\frac{\rho L-\rho g}{\rho L}}$

qL=$\frac{1440\,VL}{t}$

30//×10/

Total capacity of oil= 11.40×775

Total capacity of oil = 8838 bbl

Calculating z factor using Hall-Yarborough method:

$$p_{pc} = 709.604 - 58.718\gamma_g$$

$$T_{pc} = 170.491 + 307.344\gamma_g$$

Tpc=169.0+314.0*0.68

= 382.52 oR

Ppc=708.75-57.5*0.68

= 669.514 psia

Pseudo Reduced Critical temperature and pressure

Tpr=(T+460.0)/Tpc

Tpr=(130+460.0)/382.52

=1.54 oR

t=1/Tpr

=1/1.54

=0.6483 oR

Ppr=p/Ppc

Ppr=604.7/669.514

Ppr =0.90

Temperature dependent terms:

A = 0.06125*t*EXP(-1.2*(1-t2))

= 0.06125*0.6483*EXP(-1.2*(1-0.64832))

A = 0.03423

B =t*(14.76-9.76*t+4.58*t2)

= t*(14.76-9.76*0.6483+4.58*0.64832)

B = 6.7150

C =t*(90.7-242.2*t+42.4*t2)

=0.6483*(90.7-242.2*0.6483+42.4*0.64832)

C = -31.44

D =2.18+2.82*t

=2.18+2.82*0.6483

D = 4.008

F =-0.03423*0.9+(Y+Y2+Y3-Y4)/(1-Y)3-B*Y*Y+C*D*YD-1

Assumption:

Y= 0.0336

F=A*Ppr+(0.0336+0.03362+0.03363-0.03364)/(1-0.0336)3-B*0.03362+C*D*0.0336D-1

F= -2.069 * 10-5 (close to zero)

Z = A*Ppr/Y

= 0.03423*0.9/ 0.0336

Z = 0.92

54//×10/

1.38×3.04

VL=7.6

$$\rho = \frac{2.7\gamma_g p}{zT}$$

ρL= 56.18 lb/ft^3

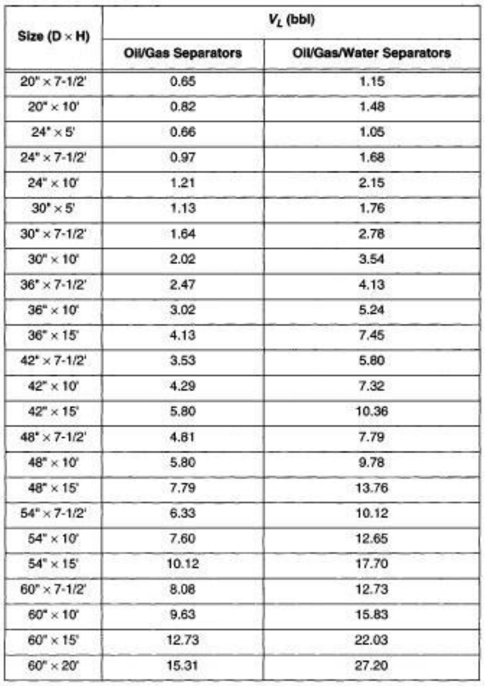
ρg= (2.7*0.68*604)/(0.9*130)

ρg = 2.044 lb/ft^3

$$q_{st}=\frac{2.4D^2Kp}{z(T+460)}\sqrt{\frac{\rho L-\rho g}{\rho L}}$$

$$q_{st}=\frac{2.4(20/12)^2*0.35}{0.92(130+460)}\sqrt{\frac{56.18-2.044}{56.18}}$$

q_{st}= 4.2*10^{-3}



Size (D × H)	V_L (bbl)	
	Oil/Gas Separators	Oil/Gas/Water Separators
20" × 7-1/2'	0.65	1.15
20" × 10'	0.82	1.48
24" × 5'	0.66	1.05
24" × 7-1/2'	0.97	1.68
24" × 10'	1.21	2.15
30" × 5'	1.13	1.76
30" × 7-1/2'	1.64	2.78
30" × 10'	2.02	3.54
36" × 7-1/2'	2.47	4.13
36" × 10'	3.02	5.24
36" × 15'	4.13	7.45
42" × 7-1/2'	3.53	5.80
42" × 10'	4.29	7.32
42" × 15'	5.80	10.36
48" × 7-1/2'	4.81	7.79
48" × 10'	5.80	9.78
48" × 15'	7.79	13.76
54" × 7-1/2'	6.33	10.12
54" × 10'	7.60	12.65
54" × 15'	10.12	17.70
60" × 7-1/2'	8.08	12.73
60" × 10'	9.63	15.83
60" × 15'	12.73	22.03
60" × 20'	15.31	27.20

Table 4 sizing of vessel w.r.t V $_l$

$54^{//}\times10^{/}$

V_L=7.6

$$qL=\frac{1440*7.6}{1}$$

qL =10944 bbl/day

$e = P_iD_i / 2f - P_i$

Operating pressure= $P_i = 41.69(1.10)$

$P_i = 45.86$ bar

For carbon steel at 60 degree C.

Design stress=f= $125 * 10^6$ Pa

Diameter of cylinder= 1.38 m

e= $(45.86*10^5*1.38)/2(125*10^6)-(45.86*10^5)$

e= 0.0254 m

Hemispherical domed head will have thickness half of that of the wall thickness

Fully radiographed, J=1

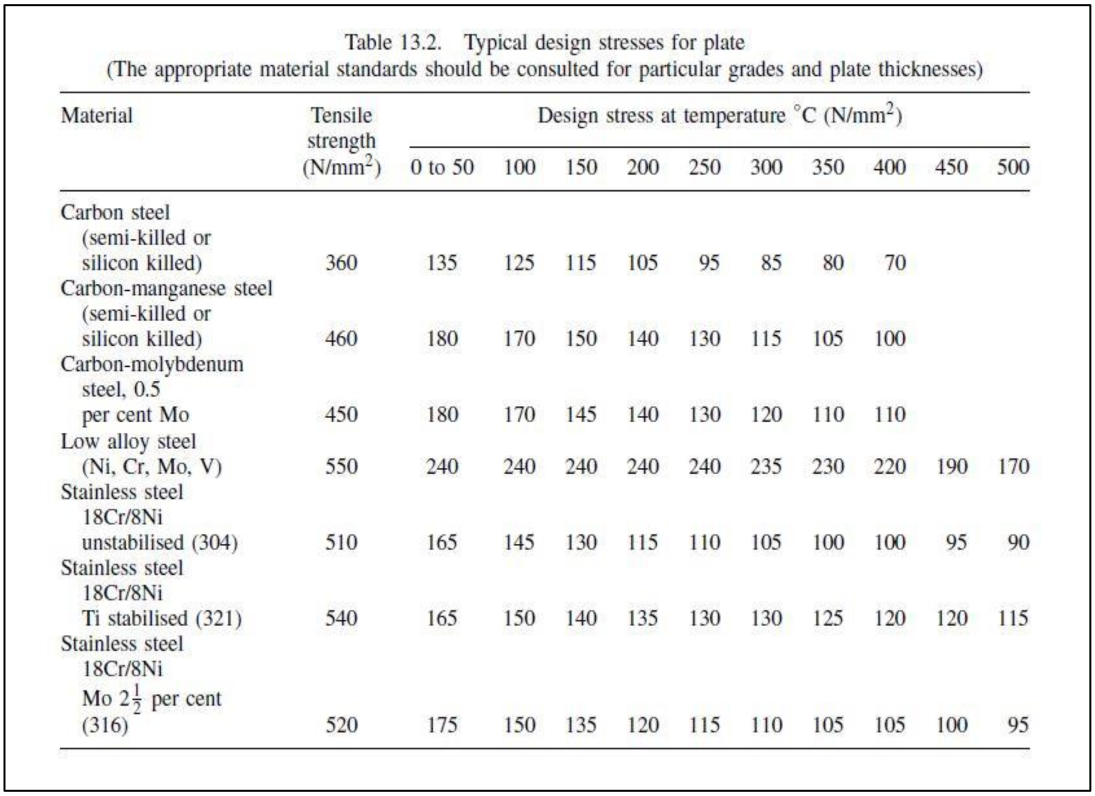


Table 13.2. Typical design stresses for plate
(The appropriate material standards should be consulted for particular grades and plate thicknesses)

Material	Tensile strength (N/mm^2)	Design stress at temperature °C (N/mm^2)									
		0 to 50	100	150	200	250	300	350	400	450	500
Carbon steel (semi-killed or silicon killed)	360	135	125	115	105	95	85	80	70		
Carbon-manganese steel (semi-killed or silicon killed)	460	180	170	150	140	130	115	105	100		
Carbon-molybdenum steel, 0.5 per cent Mo	450	180	170	145	140	130	120	110	110		
Low alloy steel (Ni, Cr, Mo, V)	550	240	240	240	240	240	235	230	220	190	170
Stainless steel 18Cr/8Ni unstabilised (304)	510	165	145	130	115	110	105	100	100	95	90
Stainless steel 18Cr/8Ni Ti stabilised (321)	540	165	150	140	135	130	130	125	120	120	115
Stainless steel 18Cr/8Ni Mo $2\frac{1}{2}$ per cent (316)	520	175	150	135	120	115	110	105	105	100	95

Table 5 Design stress factor of vessel

3.2 Reactor Design (CSTR):

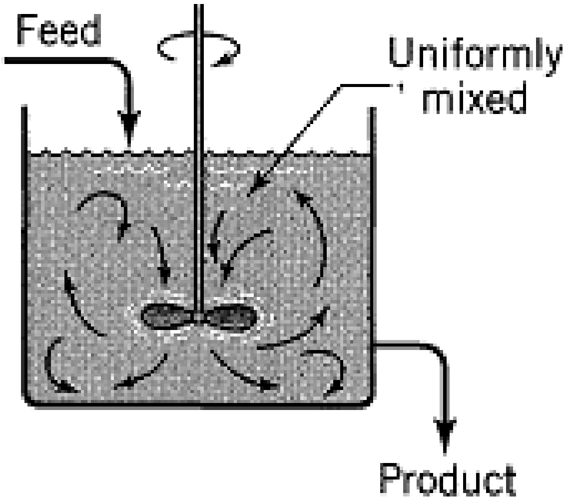


Figure 11 CSTR Reactor

3.2.1 Design of Demulsification unit:

Reaction equation

$C_{25}H_{54}COO^{-}Na^{+} + C_{25}H_{54}Cl^{-}N^{+} \rightarrow (C_{25}H_{54})_2COO^{-}N^{+} + NaCl$

$F_{Products}$= 9.4 +1.04+9.4

$F_{Products}$ = 19.84 moles/hr product

Concentration of Aliquat-336= $C_{0\text{-}A\text{-}336}$= 1.55×10^{-3} M

Concentration of Resins= $C_{0\text{-}Resins}$= 1.73×10^{-3} M

3.2.2 Kinetics of reaction:

$k=k_0e^{Ea/RT}$

k= $1.976 * 10^2$ 1/Ms

The reaction is first order with respect to both the reactants so it will be 2nd order reaction overall.

$-r_A = kC_A C_R$

$-r_A = k C_{A0} (1 - X_A) (M - X_A)$

Where, $M = C_{R0} / C_{A0}$

$X_a = 90\%$

$-r_A = 1.976 * 10^{-2} * (1-0.9)((1.73/1.55)-0.9)$

$r_A = 5.01*10^{-4}$ M/s

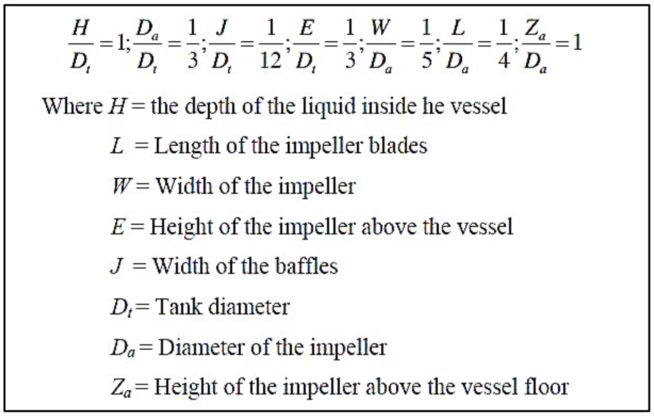
Using equation for CSTR:

$\tau/C_{A0} = V/F_{A0} = X_A/r_A$

$\tau = X_A * C_{A0}/r_A$

$= 0.9*1.55*10^{-3}/5.01*10^{-4}$

= 2.78 hrs



$$\frac{H}{D_t} = 1; \frac{D_a}{D_t} = \frac{1}{3}; \frac{J}{D_t} = \frac{1}{12}; \frac{E}{D_t} = \frac{1}{3}; \frac{W}{D_a} = \frac{1}{5}; \frac{L}{D_a} = \frac{1}{4}; \frac{Z_a}{D_a} = 1$$

Where H = the depth of the liquid inside he vessel

L = Length of the impeller blades

W = Width of the impeller

E = Height of the impeller above the vessel

J = Width of the baffles

D_t = Tank diameter

D_a = Diameter of the impeller

Z_a = Height of the impeller above the vessel floor

Figure 12 Impeller dimensions

3.2.3 Total Volume of Reactor:

Molar flow rate of the product

$\tau = V / v0$

$V = \tau * v0$

$V= 2.67*1249$

$V= 3334.83$ L

$V= 3.334$ m3

Providing 25% allowance

$=3.334 +0.8335$

$V= 4.1675\ m^3$

Height and Diameter:

Ideal height to dia ratio is in range of 1-2.

$h/D = 1.5$

$V= \pi/4$ D2h

$h= 1.5D$

$V= 3.14/4 *1.5(D)3$

$D= 1.52$ m

$h = 2.28$ m

Impeller Calculations:

$Da = 0.51$ m

$E= 0.51$ m

$L = 0.1275$ m

$W= 0.102$ m

$Za= Da = 0.51$ m

Blade speed is 1.5-3 m/s

$n=NDa$

where,

n= tip speed

N= rotating speed

NR= D2Nρ/μ

N= n/D

N= 2/0.51

N= 3.92 rps

Turbulance in whole tank is inevitable so Nr should be above 10000.

μ = 0.0917 pa.s

Baffle Calculations:

J= 0.1275 m

Baffle height form bottom = Da/2 = 0.225 m

Baffle Spacing= Da/4= 0.1275 m

3.3 Design of Heat Exchanger (E4001A):

3.3.1 Definition:

A heat exchanger is a heat transfer device that is used for transfer of internal thermal energy between two or more fluids available at different temperatures.

In most exchangers the fluids are separated by heat transfer surface and ideally, they don't mix. The word exchanger really applies to all types of equipment in which heat is exchange but is often used specifically to denote equipment in which heat is exchanged between two process fluids such as;

Heaters & coolers:

Exchangers in which process fluid is heated or cooled by a plant service stream are called heaters or coolers.

Vaporizer:

If the process stream is vaporized the exchanger is termed as vaporizer.

Reboiler:

If the stream is essentially completely vaporized, then the exchanger is Reboiler. It is associated with the distillation column.

Evaporator

For the purpose of concentration of solution, the exchanger is called an evaporator.

Fixed exchanger:

It is used for exchangers heated by combustion gases, such as boilers.

Chiller:

One stream is a process fluid being condensed at sub atmospheric temperature and the other boiling refrigerant or process stream.

3.3.2 Application:

Heat exchangers has application in industries, as chillers, in power sector, transportation, heat recovery units and temperature control units in chemical and process industries.

3.3.3 Classification of heat exchangers:

In general, industries heat exchanger classified according to them

- Construction
- Process transfer
- Degree of surface compactness
- Flow arrangement

- Pass arrangement
- Phase of process fluid
- Heat transfer mechanism
- Flow arrangement

Classification in accordance with construction:

In accordance with construction heat exchangers are:

- Plate heat exchanger (gas kited, spiral, plate coil)
- Tubular heat exchanger (double pipe, shell and tube, coiled tube)
- Extended surface exchangers (tube fin, plate fin)
- Regenerator

Classification according to transfer process:

- Direct contact (cooling towers)
- Indirect contact (double pipe, shell and tube, coiled tube)

Classifying in accordance with surface compactness:

A compact heat exchanger has a high density surface that is the ratio of heat transfer area (A) to its volume (V). The heat transfer volume is the same. It's about 700m2 / square meter. They also have higher thermal efficiency (95% vs. 60% to 80% for STHE), which makes them useful especially for energy-intensive industries.

Classifying in accordance with flow arrangement:

The basic flow arrangements in heat exchanger are:

- Parallel flow
- Cross flow
- Counterflow

Depends on the required efficiency of the heat exchanger, fluid flow pathways and the packaging envelope, the permissible thermal stress, the temperature, etc. All these factors decide the type of flow arrangement.

Parallel flow heat exchangers:

In parallel flow heat exchangers, two fluids enter the exchanger at the same end, and travel in parallel to one another to other side.

Counter flow heat exchangers:

In counter flow heat exchangers, which are often more efficient, the fluids enter the exchanger from opposite ends. Countercurrent exchange is a mechanism used to transfer some component of fluid from one flowing current of fluid to another across a permeable barrier between them. It is extensively used in biological systems for a wide variety of purposes.

Cross flow heat exchanger:

In a crossflow heat exchanger, the fluids travel perpendicular to one another.

Classifying in accordance with pass arrangement:

A fluid is thought to have passed if it flows once in its entire length through a section of the heat exchanger. Whether a single pass or multi pass is given, the fluid is reversed in a multi pass system and flows two or more times over the flow volume. Compact shell and tube and plate exchangers are necessary for multi-pass arrangements.

Classifying in accordance with phase of fluid:

Classification is based on the fluid phase of gas + gas, liquid + fluid, gas + liquid, etc.

Classification according to heat transfer mechanism:

The basic heat transfer mechanism employed for heat transfer from one fluid to another is:

- Single phase convection, forced or free
- Two phase convection (condensation or evaporation)
- Combined convection or radiation

3.3.4 Types of heat exchangers:

The principal types of heat exchangers used in chemical process and allied industries are as follows:

- Double pipe heat exchangers
- Shell & tube heat exchangers
- Plate & frame heat exchangers
- Plate fin exchangers
- Spiral heat exchangers
- Air cooled: coolers & condensers
- Direct contact: cooling & quenching
- Agitated vessels
- Fired heaters

3.3.5 Selection criteria of heat exchangers:

- Cost: want to have the least expensive exchanger
- Efficiency: want a most efficient, minimally energy-loss transmission exchanger and minimum fluid pressure drop
- Space: Want to have a small exchanger
- Material: want a material-compatible exchanger that doesn't take much to build with process streams
- Sustainability: you want an exchanger that is easy to clean
- Easy to construct
- Thermal and hydraulic requirements
- Environmental, health and safety consideration and regulations

3.3 Shell & Tube Heat Exchanger

Main Parts:

- The stationary header, where the fluid enters through the tubes of heat exchanger
- Rear header, it's the side of the exchanger from where the fluid leaves through tubes.
- Tube bundle, tube bundle is combination of tubes, baffle plates, and tie-rods.
- Shell, a shell is the outer layer of exchanger which holds the tube bundle.

Scope & limitation of shell & tube heat exchanger:

Less effective if in comparison to plate heat exchanger.

Cleaning of tubes is troublesome.
Capacity cannot be varied. Have to redesign for change in capacity.
Consume a lot of area space as compared to plate type.

Brief Description:

The primary objective is to increase the temperature of de-emulsified crude so it can be sent to LLE column for de-asphalting process. as a certain temperature is required for efficient extraction of asphalts.

Method of Design:

Starting with heat balance we determine flow of water required to achieve desired temperature increase. Followed by Tube and shell side dimension calculations for no. of tube, S.A of tubes, pitch, baffle dia, no. of baffles. Calculating LMTD, UA value, mass velocity, renolds no. Jh factor, coefficient and corrected coefficient to find actual value of UA, followed by pressure drop which should be in range.

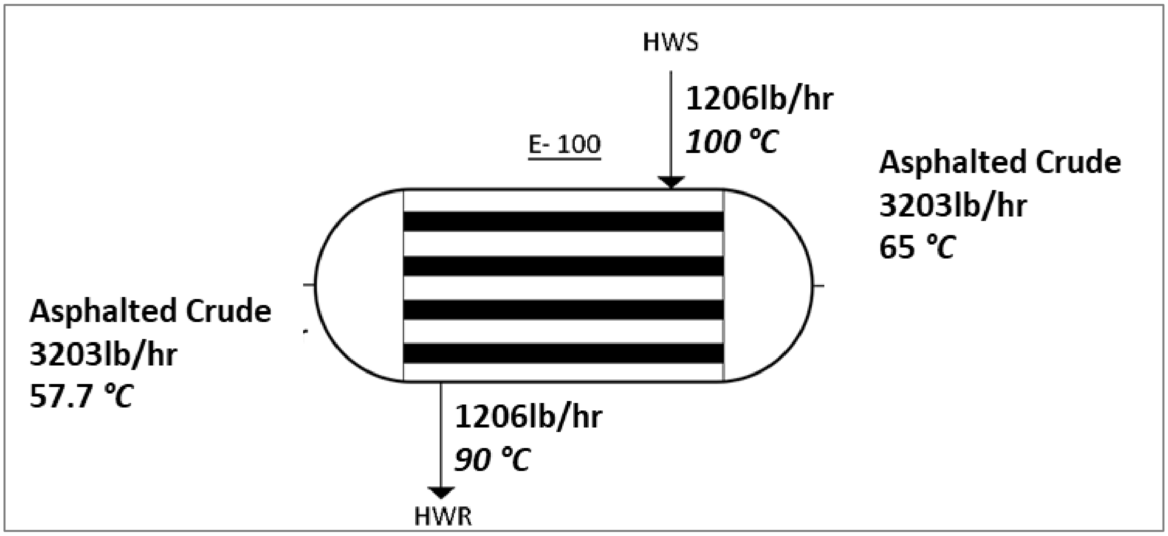


Figure 13 Shell & tube Heat Exchanger

DESIGN CALCULATIONS:

Heat Balance:

$Q = \dot{m} \times Cp \times \Delta T$

$Q = 1453 \times 2.275 \times 7.3$

Q = 24130 kJ/h = 22870 BTU/h

$22870 = m_w \times 4.19 \times 10$

Water flow= 547.12kg/h

Water flow= 1206lb/h

Q= 24130 kj/hr

Δt= 91.3F

UA =Q/T = 732W/F

Tube dimensions:

Tube outer diameter= do = ¾ in. = 0.016

d_{in} = 0.560 in = 0.0142

Type of tubes = 13 BWG

Length of a tube = 5 m.

No. of tubes passes = 2

For no. of tubes:

Assuming value of U= 300W/m^{2}F

Area = 732/300= 2.44m^2

S.A per tube = πDL= 0.25m^2

No. of tubes= 2.44/0.25

No. of tubes= 10

Pt = 1.25 *0.016= 0.02m

Clearance = Pt – do= 0.004m

Shell dimension:

No. of shell passes: 1

Baffle spacing= 0.51m.

Square pitch = 1.25*do =0.9 in

Bundle diameter:

Db = do (Nt/Kt)^1/n1 = 0.438m

Db= 0.438m

Clearance: 0.033

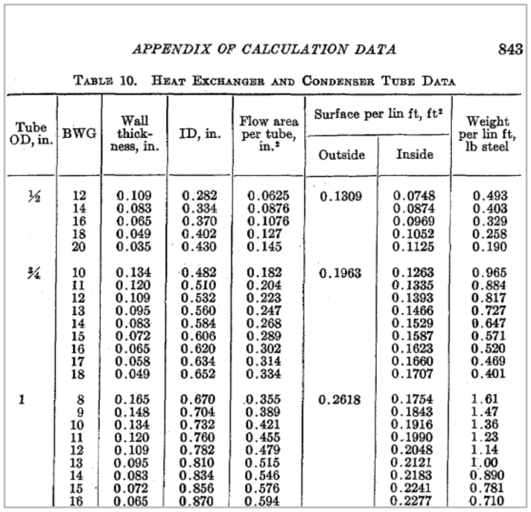
Ds= 0.438+0.033 = 0.472 m

No. of baffles:

Baffle spacing =Ls= 0.7*Ds=0.329m

No. of baffles **=** (L/Ls)-1 = (5/0.33) -1

No. of baffles= 14



APPENDIX OF CALCULATION DATA 843

TABLE 10. HEAT EXCHANGER AND CONDENSER TUBE DATA

Tube OD, in.	BWG	Wall thickness, in.	ID, in.	Flow area per tube, in.^{2}	Surface per lin ft, ft^2		Weight per lin ft, lb steel
					Outside	Inside	
½	12	0.109	0.282	0.0625	0.1309	0.0748	0.493
	14	0.083	0.334	0.0876		0.0874	0.403
	16	0.065	0.370	0.1076		0.0969	0.329
	18	0.049	0.402	0.127		0.1052	0.258
	20	0.035	0.430	0.145		0.1125	0.190
¾	10	0.134	0.482	0.182	0.1963	0.1263	0.965
	11	0.120	0.510	0.204		0.1335	0.884
	12	0.109	0.532	0.223		0.1393	0.817
	13	0.095	0.560	0.247		0.1466	0.727
	14	0.083	0.584	0.268		0.1529	0.647
	15	0.072	0.606	0.289		0.1587	0.571
	16	0.065	0.620	0.302		0.1623	0.520
	17	0.058	0.634	0.314		0.1660	0.469
	18	0.049	0.652	0.334		0.1707	0.401
1	8	0.165	0.670	0.355	0.2618	0.1754	1.61
	9	0.148	0.704	0.389		0.1843	1.47
	10	0.134	0.732	0.421		0.1916	1.36
	11	0.120	0.760	0.455		0.1990	1.23
	12	0.109	0.782	0.479		0.2048	1.14
	13	0.095	0.810	0.515		0.2121	1.00
	14	0.083	0.834	0.546		0.2183	0.890
	15	0.072	0.856	0.576		0.2241	0.781
	16	0.065	0.870	0.594		0.2277	0.710

Table 6 ID calculation

LMTD Calculation:

Table 16. Hot and Cold Fluid Temperatures and their Differences

Hot fluid		**Cold fluid**	**Difference**
100°C	Higher temp	65 °C	35°C
90°C	Lower Temp	57.7**°C**	32.3°C
10°C	Differences	7.3°C	2.7°C

Table 7 LMTD calculations

LMTD = $\Delta T_2 - \Delta T_1 / \ln(\Delta T_2/\Delta T_1)$

ΔT_2 = Temperature difference of hot fluids

ΔT_1 = Temperature difference of cold fluids

LMTD = 33.63 °C

R = $T_1 - T_2 / t_2 - t_1$ = 10/7.3 = 1.369

S = $t_2 - t_1 / T_1 - t_1$ = 7.3/100-57.7 = 0.172

F_t = 0.98

$\Delta t = F_t * \text{LMTD} = 0.98*33.63$

Δt= 32.95 C = 91.31 °F

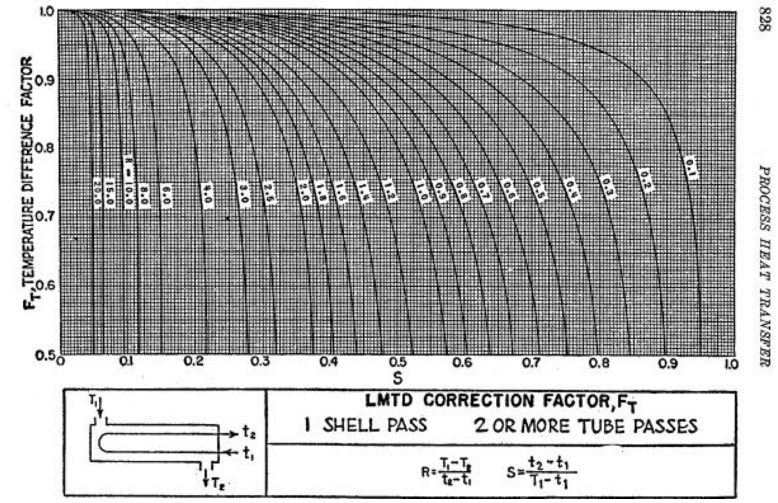


Figure 14 R and S calculation

Calorific temperatures:

$\Delta t_c / \Delta t_h$ = 15/42.3

$\Delta t_c / \Delta t_h$ = 0.355

Kc=0.84 Fc= 0.36

Tc = T2+Fc(T1-T2)

Tc = 90 + 0.36(100-90)

Tc = 93.6C = 200.48 °F

tc = t_1+Fc(t2-t1)

tc = 57.7 + 0.36(7.3)

tc = 60.33C = 140.59 F

SHELL SIDE	**TUBE SIDE**
de=(1.1/ do)(pt^2 – 0.917 do^2)= 0.011m **Flow Area:** a_s = I.D * C^n * B/144P_T a_s = 0.02 m^2 **Mass velocity:** G_s= W/a_s G_s = 5603.7697 lb/h ft^2 **Reynolds Number:** At T_c=188.96 F μ = 2.275*2.42= 5.505 lb/h ft D_e= 0.024 m= 0.078 ft Re_s = De * Gs/μ Re_s = 79 **J_H:** J_H = 4.8	**Total Flow Area:** a_t‘ = 0.247 in^2 a_t = N* a_t‘ /144 * n a_t = 0.008 ft^2 **Mass velocity:** G_t= W/a_t G_t = 140618lb/h ft^2 **Reynolds Number:** At tc=140.59 F μ = 4.19*2.42= 10.14 lb/h ft D= 0.0467 ft Re_t = D * Gt/μ Re_t = 647 **J_H:** J_H = 5.2

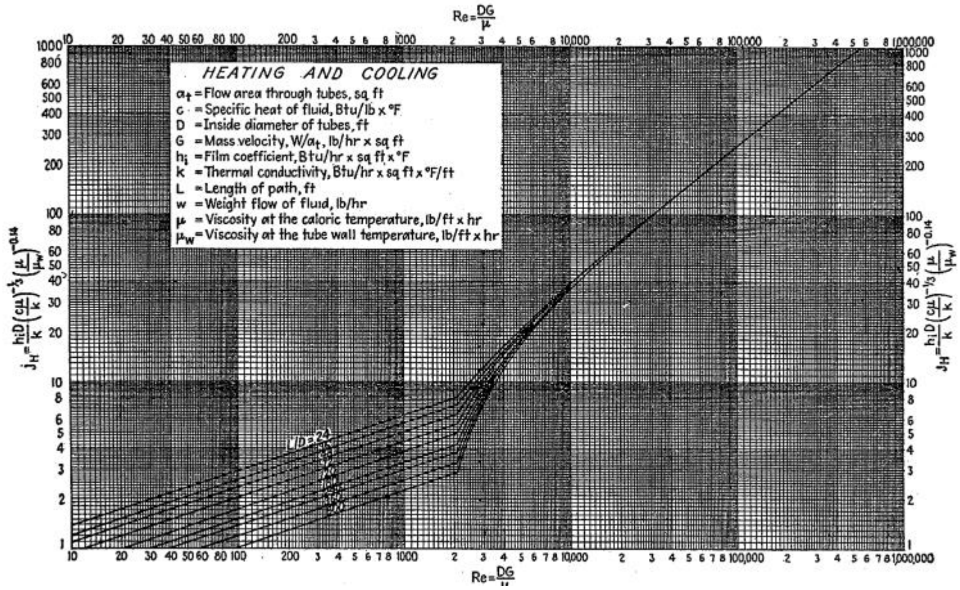


Figure 15 J_h factor for tube side

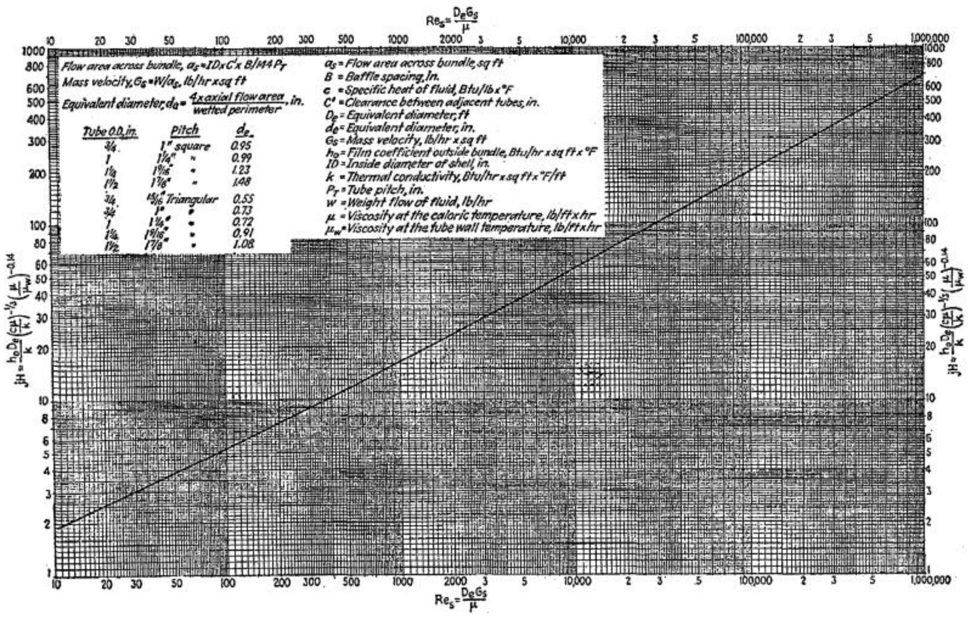


Figure 16 J_h factor for Shell side

Now, at Tc, k(c μ/k)$^{1/3}$ = 0.235 **Heat transfer coefficient:** h0/Øs = (J$_h$ /De)* k(c μ/k)$^{1/3}$ h0/Øs = 14.46 **Tube wall temp:** Tw= tc + (h0/Øs)(Tc-tc)/(h0/Øs)+($h$io/Øt) Tw = 164 F At tw, μ$_w$ = 0.19* 2.42= 0.459 lb/h ft Øs = (μ/μ$_w$)$^{0.14}$ Øs = 1.4157 **Corrected Coefficient:** h0 = 10.21BTU/h. ft^2 F	Now, at Tc, k(c μ/k)$^{1/3}$ = 0.223 **Heat transfer coefficient:** hi/Øt = (J$_h$ /De)* k(c μ/k)$^{1/3}$ hi/Øt= 24.83 $h$io/Øt= hi/Øt * I.D/O.D= 16.71 * 0.0467/0.0524 = 22.19 At tw, μ$_w$ = 0.155* 2.42 = 0.375 lb/h ft Øt = (μ/μ$_w$)$^{0.14}$ Øt = 1.5866 **Corrected Coefficient:** hi0= 13.98BTU/h. ft^2 F

Clean overall coefficient:

Uc= 6.9 BTU/h. ft^2 F

Design overall coefficient:

a” = 0.1963 ft^2 /lin.ft

A = 10.76 * 10

A= 107.6 ft^2

UA= 742.4 BTU/h. ft^2 F

Hence UA proposed satisfy the assumed 732 BTU/hr ft^2 F

R$_D$ = 742.4-732/(742.4*732)

R$_D$ = 1.84*10^{-5}

Dirt factor is also less than 0.003

SHELL SIDE	**TUBE SIDE**
Pressure Drop	**Pressure Drop**
For Re_s = 307 f = 0.004 ft2/in2 De= 0.95/12= 0.079ft No. of crosses: N+1 = 12L/B N+1 = 12 * 7/15 N+1 =5.6 ΔPs: $\Delta Ps = f*Gs^2*Ds*(N+1)/$ $(5.22*10^{10}*De\text{Ø}s)$ ΔPs = 0.0031psi Less than 10psi, hence acceptable.	For Re_t = 39 f = 0.012 ft2/in2 , s= 0.81 ΔPt: $\Delta Pt= f*Gt^2*L*n/5.22*10^{10}*Ds\text{Ø}t$ $\Delta Pt= 1.095*10^{-4}$ psi Gt = 8809lb/h ft^2 $V^2/2g' = 0.97$ $\Delta Pr = 4n V^2 /s 2g= 4*2*0.97/0.89$ ΔPr = 8.5 $\Delta P_T = \Delta Pr + \Delta Pt$ $\Delta P_T = 8.5+0.1*10^{-4}$ ΔP_T = 8.77 psi Less than 10psi, which is in acceptable range

Sine the pressure drop on both ends is less than 10psi, our second condition satisfies for the design.

Unit	**HEAT EXCHANGER**
Type	Shell & Tube heat exchanger
Function	Heating Crude oil
Operation	Continuous
Heat Duty	22870Btu/h
Heat Transfer Area	107.6ft^2
Overall Heat Transfer Coefficient	U_d A= 742.4 Btu/h. ^{o}F

	Shell side	**Tube Side**
Fluid Circulated	Cold Fluid	Hot Fluid
Flow Rates	3203lb/hr	1206lb/hr
Temperatures	Inlet= 57.7°C Outlet= 65 °C	Inlet= 100 °C Outlet= 90 °C
Pressure Drop	0.003psia	8.77psia
Specifications	Diameter D_s=1.548ft J_h=4.8	Diameter D_o=0.052ft J_h=5.2

Table 8 Design Description for Heat Exchanger E-

3.4 Design of LLE column:

3.4.1 Process Description:

Extraction column is used to perform liquid-liquid extraction here for removal asphalting from crude oil with a technique known as solvent de-asphalting. Propane is used as a solvent. The crude is charged from the top and solvent from bottoms. The counter flow interaction ensures maximum contact area for both fluids. Propane along with crude leaves through extract and remained asphalts and heavy compounds are obtained from raffinate.

3.4.2 Choice of solvent:

In LLE process solvent is the fluid which dissolve the solute from the old solvent. The new solvent solubility should be greater than previous one to successfully perform extraction. In case of de-asphalting, solvent selection is based on the feed of crude. For heavy crude solvent of C-5 or C-7 is used. For medium and light crude C-3 & C-4. As mentioned previously, our feed is feed is wet gas, hence propane is the best option to proceed with.

3.4.3 Design Calculations:

A ternary diagram for Propane de-asphalting of crude is available and theoretical no. of stages calculated using this are eight.

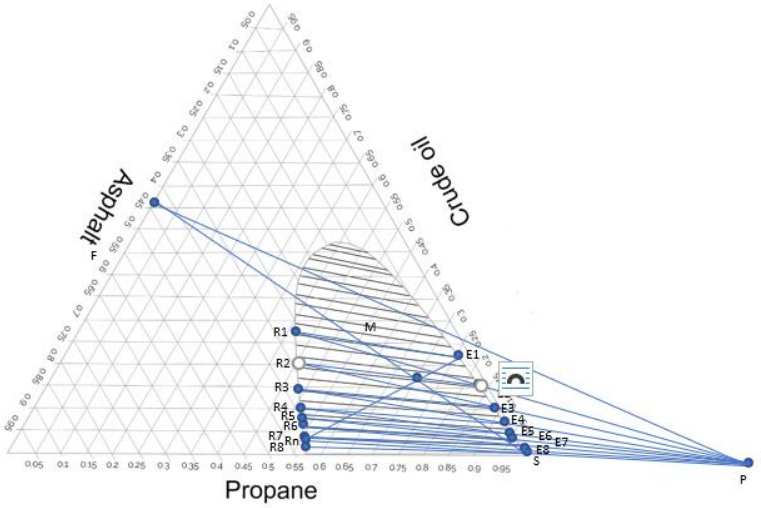


Figure 17 LLE ternary Diagram

Total no. of stages calculated are, 8 stages

Conc. asphalt = XA , Conc. Crude oil= XC , Conc. Solvent= Xs

Feed:

XA= 0.4318 , Xc=0.5682 , Xs=0

Solvent:

XA=0 , Xc=0 , Xs=1.0

propane to the extraction column must be 4 times the feed.

Solvent (C-5)= 3000 kg

Feed = 1453 kg

Overall Balance

C-5 + crude = Raffinate+ Extract

$3000 + 1453 = Rn + E1$

$Rn + E1 = 4453$ kg

$E1 = 4453 - Rn$

At E1 we get,

2.5 percent asphalt, 75 percent solvent & 22.5 percent crude.

Asphaltene Balance

$N_{CO}\,(x_A) + n_{C5}(x_A) = R_n(x_A) + E_1(x_A)$

$1453(0.4318) + 3000(0) = E_1\,(0.00125) + R_n\,(0.43)$

$627.40 + 0 = 0.00125(4453 - R_n) + 0.43R_n$

$326.6 = 0.4287R_n + 5.566$

$R_n = 321.0/0.4287$

$R_n = 748.68$ kg

$R_n + E_1 = 4453$ kg

$E_1 = 4453 - R_n$

$E_1 = 4453 - 748.68$

$E_1 = 3704.31$ kg

From literature study, following properties table is obtained to be used for design calculations.

Parameters	***Values***
Flow rate of crude oil:	C = 1453 kg/h
Density of crude oil:	ρ_c = 216 kg/m^3
Volumetric flowrate of crude oil:	q_c =0.001867 m^3/s
Viscosity of crude oil:	μ_c = 0.0013 kg/m.s
Flow rate of propane:	P = 3000 kg/h
Density of propane:	ρ_P = 486.5 kg/m^3
Volumetric flowrate of propane:	q_P =0.001713 m^3/s
Viscosity of propane ether:	μ_P =0.00008859 kg/m.s
Interfacial tension :	σ = 0.019 N/m
Conversion factor:	g_c = 1 kg.m/N.s^2
Acceleration of gravity:	g =9.80 M/s^2

Table 9 Parameters of feed and solvent

Columns Perforations:

Orifice diameter = do = 0.005m

Jet dia = dj

Triangular pitch on 15mm centers = 0.015m

$\Delta\rho = \rho_c - \rho_P$ = 216-486 = 270 kg/m3

$do/(\sigma.g_c/\Delta\rho g)^{0.5}$ = 1.865

1.865> 0.785

So,

$do/dj = 0.12 + 1.51\, do/(\sigma.g_c/\Delta\rho g)^{0.5}$ = 2.937

do/dj= 2.937

dj= do/2.937 = 0.005/2.937

dj= $1.702 * 10^{-3}$ m

The velocity through perforations (orifice):

$Vo = 2.69\, (dj/do)^2\ [\sigma/ dj\, (0.5137p_p + 0.4719p_c)]$

$V_o = 2.69\ (1.702*10^{-3}/0.005)^2\ [0.019/1.702*10^{-3}\ (0.5137*486+0.4719*216)]^{0.5}$

$V_o = 0.0555$ m/s

If Velocity is less than 0.1, it should be set as $V_o = 0.1$m/s

Perforated area:

$A_{per} = q_p / V_o = 0.001713/0.1$

$A_{per} = 0.01713\ m^2$

Number of perforations:

$N_o = A_{per} / (\pi d_o^2/4) = 0.01713/\ \pi 0.005^2/4$

$N_o = 872$

Plate area for perforations:

$A_p = A_{per} /\ 0.907\ (d_o/p)^2 = 0.1699\ m^2$

Downspouts:

Set,

Continuous phase velocity = terminal velocity of dispersed phase drop

$V_d = V_t$

$d_p = 0.0007$m

$V_t = 0.864(\Delta\rho)^{0.5742} * (dp)^{0.7037} * (g)^{0.5742} / (p_c)^{0.4446} * (\sigma.g_c)^{0.01873} * (\mu_c)^{0.1108}$

$V_t = V_d = 0.0905$m/s

Downspout area:

$A_d = q_c/V_d$

$A_d = 0.001867/0.0905$

$A_d = 0.0206 m^2$

Total Plate Area:

$At = (Ap + 2Ad)/0.8$

$At = 0.264 m^2$

Tower Diameter:

$D_T = (4 At/\pi)^{1/2}$

$D_T = 0.57 m$

Stage efficiency:

Assuming tray efficiency to be 0.7

Eo = 0.7

Number of actual stages :

N_T = Number of theoretical stages= 8

$N_a = N_T/Eo = 8/0.7 = 11.4$

Actual stages = 12

Tower Height:

$H_T = (Na - 1)Ct + (Na*Ct/10) + 0.1 H_T$

Set :

Tray spacing = $C_T = 0.45 m$

$H_T = (12-1)0.35 + (12*0.35/10) + 0.1 H_T$

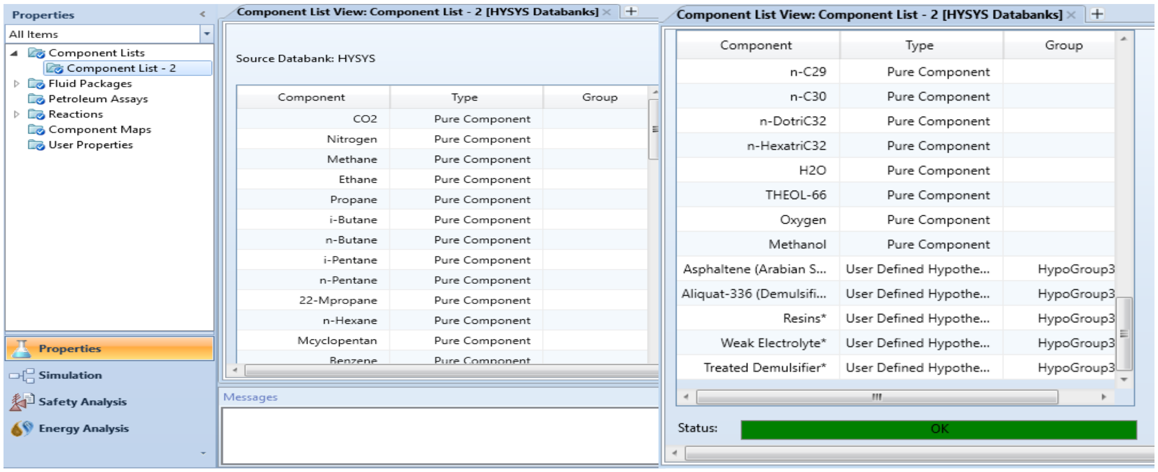
$H_T = 4.744 m$

CHAPTER 4

PROCESS MODELLING & SIMULATION

4.1 Selection of Components

The components were selected from the HYSYS library and some of them were added manually as hypothetical components using their boiling point ranges and densities.



Component List View: Component List - 2 [HYSYS Databanks]

Source Databank: HYSYS

Component	Type	Group
CO2	Pure Component	
Nitrogen	Pure Component	
Methane	Pure Component	
Ethane	Pure Component	
Propane	Pure Component	
i-Butane	Pure Component	
n-Butane	Pure Component	
i-Pentane	Pure Component	
n-Pentane	Pure Component	
22-Mpropane	Pure Component	
n-Hexane	Pure Component	
Mcyclopentan	Pure Component	
Benzene	Pure Component	

Component List View: Component List - 2 [HYSYS Databanks]

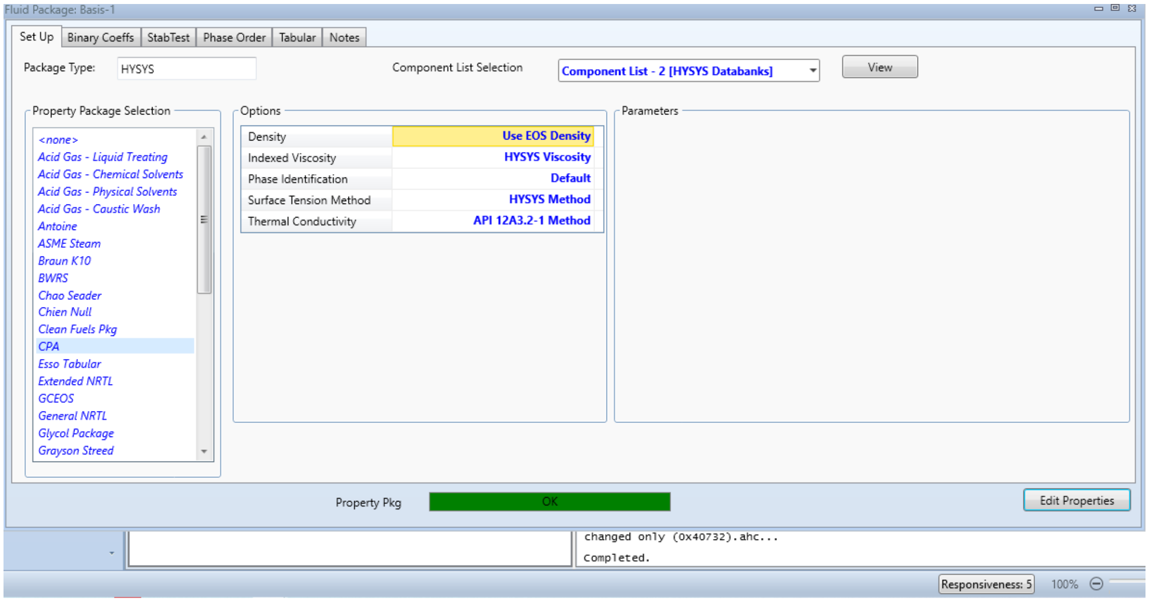
Component	Type	Group
n-C29	Pure Component	
n-C30	Pure Component	
n-DotriC32	Pure Component	
n-HexatriC32	Pure Component	
H2O	Pure Component	
THEOL-66	Pure Component	
Oxygen	Pure Component	
Methanol	Pure Component	
Asphaltene (Arabian S...	User Defined Hypothe...	HypoGroup3
Aliquat-336 (Demulsifi...	User Defined Hypothe...	HypoGroup3
Resins*	User Defined Hypothe...	HypoGroup3
Weak Electrolyte*	User Defined Hypothe...	HypoGroup3
Treated Demulsifier*	User Defined Hypothe...	HypoGroup3

Status: OK

Figure 18 component list

4.2 Selection of Fluid Package

The fluid package was selected to be CPA, both Peng Robinson and CPA are used for petroleum components but CPA has an advantage over here, that in addition to petroleum compounds it can also tackle polar components in it. As we have strong electrolyte in our feed which causes emulsion we preferred CPA over Peng Robinson to pursue our simulation with.



Package Type: HYSYS

Component List Selection: Component List - 2 [HYSYS Databanks]

Options

Density	Use EOS Density
Indexed Viscosity	HYSYS Viscosity
Phase Identification	Default
Surface Tension Method	HYSYS Method
Thermal Conductivity	API 12A3.2-1 Method

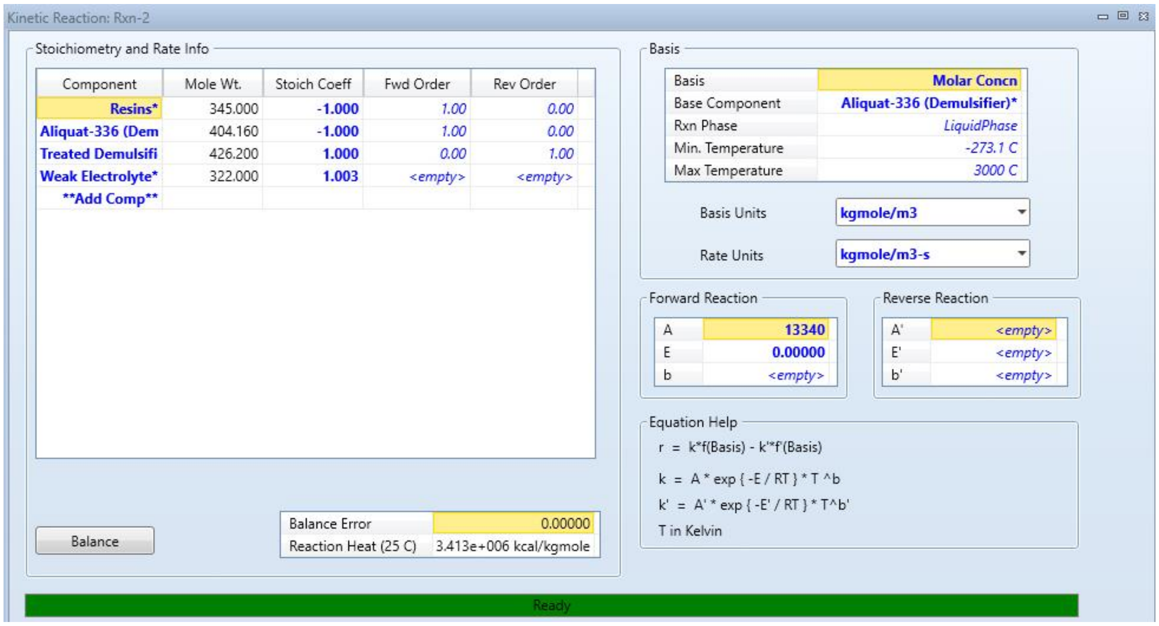
Property Pkg: OK

```
changed only (0x40732).ahc...
Completed.
```

Figure 18 component list

4.3 Defining the Kinetic Reaction Set

A kinetic reaction set was added on the basis of the reaction data obtained from the literature.



Component	Mole Wt.	Stoich Coeff	Fwd Order	Rev Order
Resins*	345.000	-1.000	1.00	0.00
Aliquat-336 (Dem	404.160	-1.000	1.00	0.00
Treated Demulsifi	426.200	1.000	0.00	1.00
Weak Electrolyte*	322.000	1.003	<empty>	<empty>
Add Comp				

Balance Error	0.00000
Reaction Heat (25 C)	3.413e+006 kcal/kgmole

Basis

Basis	Molar Concn
Base Component	Aliquat-336 (Demulsifier)*
Rxn Phase	LiquidPhase
Min. Temperature	-273.1 C
Max Temperature	3000 C

Basis Units: kgmole/m3

Rate Units: kgmole/m3-s

Forward Reaction

A	13340
E	0.00000
b	<empty>

Reverse Reaction

A'	<empty>
E'	<empty>
b'	<empty>

Equation Help

r = k*f(Basis) - k'*f'(Basis)

k = A * exp { -E / RT } * T ^b

k' = A' * exp { -E' / RT } * T^b'

T in Kelvin

Figure 19 Reaction set

4.4 Process Flowsheet Modelling and Simulation

The process flow diagram was replicated as the flowsheet on Aspen HYSYS as shown by the snippet. The molar flows were the same as that of our manual calculations. The temperature and pressure parameters were also defined according to the requirements and then the simulation was run to obtain the desired de-emulsified & de-asphalted crude oil.

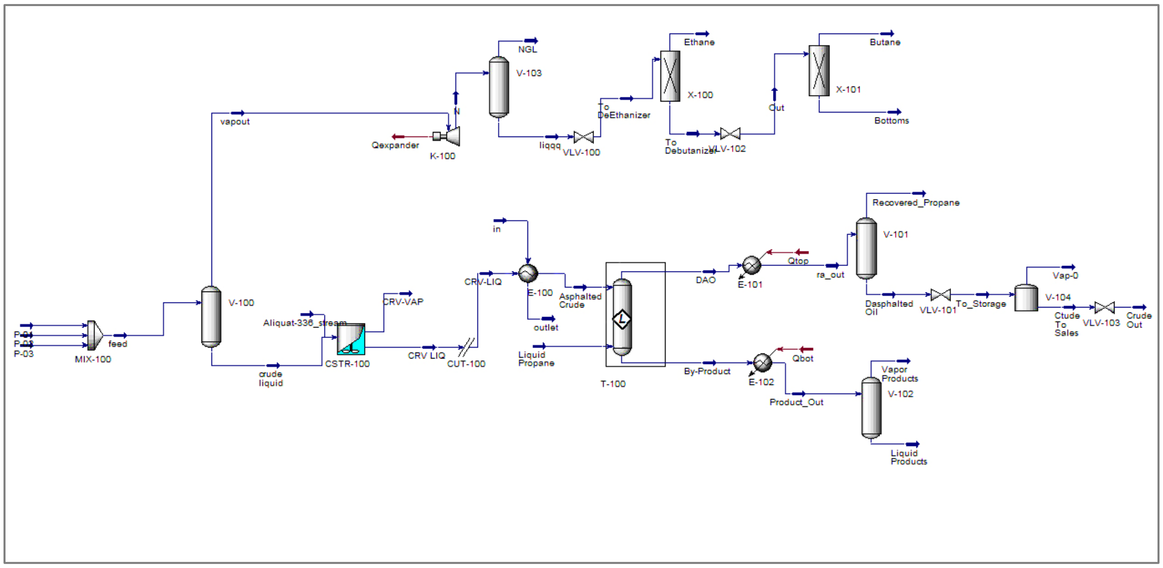
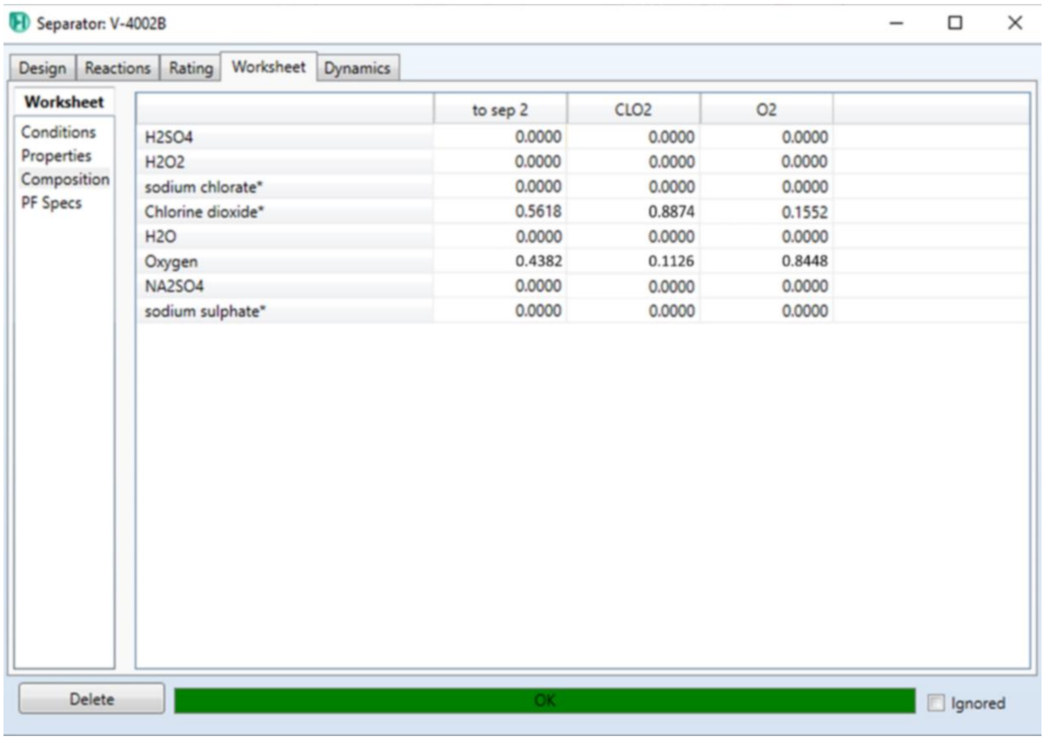


Figure 20 PFD Simulation

4.5 Separator

After the successful simulation of the model, the results were obtained that were close to our theoretical calculations. The final separator shows a recovery of chlorine dioxide up to a percentage mole fraction of 88.7%

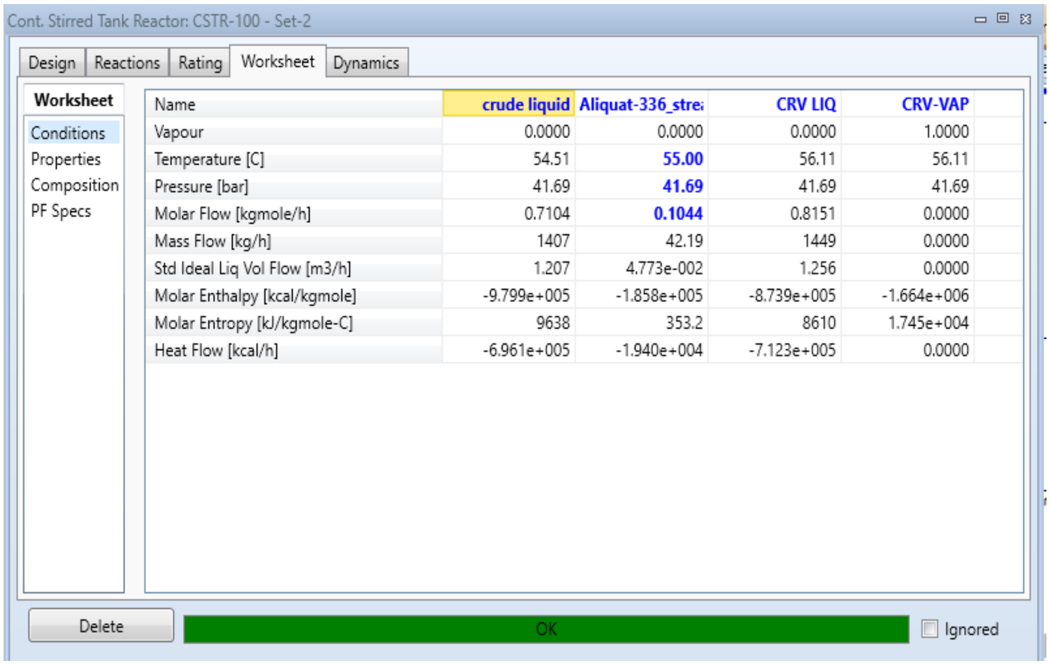


	to sep 2	CLO2	O2
H2SO4	0.0000	0.0000	0.0000
H2O2	0.0000	0.0000	0.0000
sodium chlorate*	0.0000	0.0000	0.0000
Chlorine dioxide*	0.5618	0.8874	0.1552
H2O	0.0000	0.0000	0.0000
Oxygen	0.4382	0.1126	0.8448
NA2SO4	0.0000	0.0000	0.0000
sodium sulphate*	0.0000	0.0000	0.0000

Figure 21 separator composition

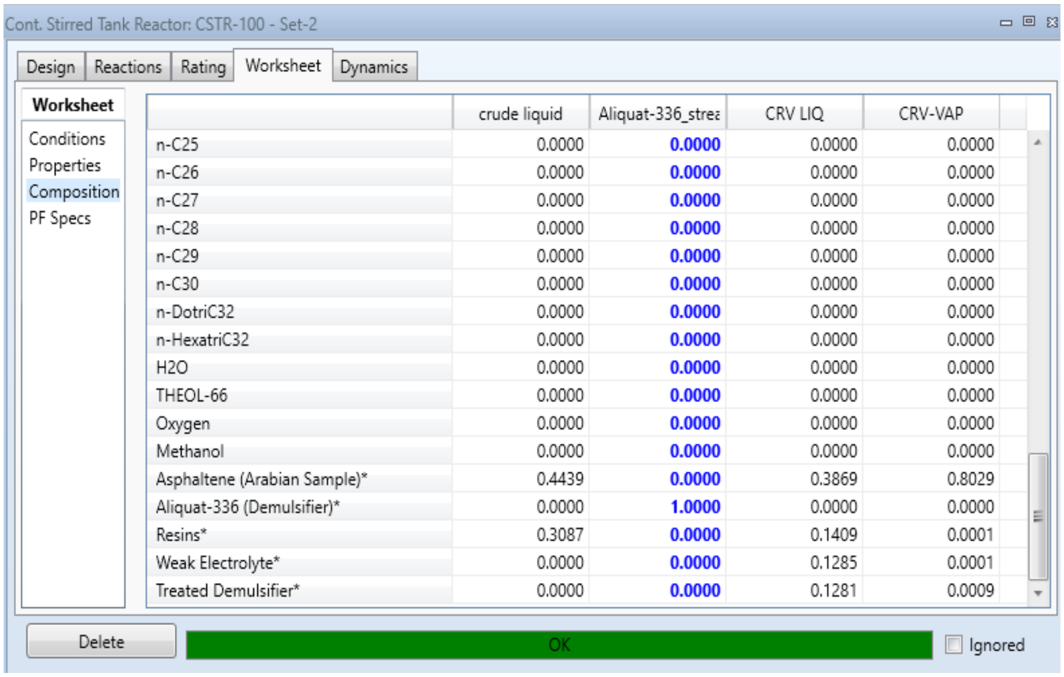
4.6 CSTR

From simulation results we can see the resins are converted to week electrolyte and the results are close to theoretical calculations.



Name	crude liquid	Aliquat-336_stre	CRV LIQ	CRV-VAP
Vapour	0.0000	0.0000	0.0000	1.0000
Temperature [C]	54.51	55.00	56.11	56.11
Pressure [bar]	41.69	41.69	41.69	41.69
Molar Flow [kgmole/h]	0.7104	0.1044	0.8151	0.0000
Mass Flow [kg/h]	1407	42.19	1449	0.0000
Std Ideal Liq Vol Flow [m3/h]	1.207	4.773e-002	1.256	0.0000
Molar Enthalpy [kcal/kgmole]	-9.799e+005	-1.858e+005	-8.739e+005	-1.664e+006
Molar Entropy [kJ/kgmole-C]	9638	353.2	8610	1.745e+004
Heat Flow [kcal/h]	-6.961e+005	-1.940e+004	-7.123e+005	0.0000

Figure 22 CSTR Worksheet

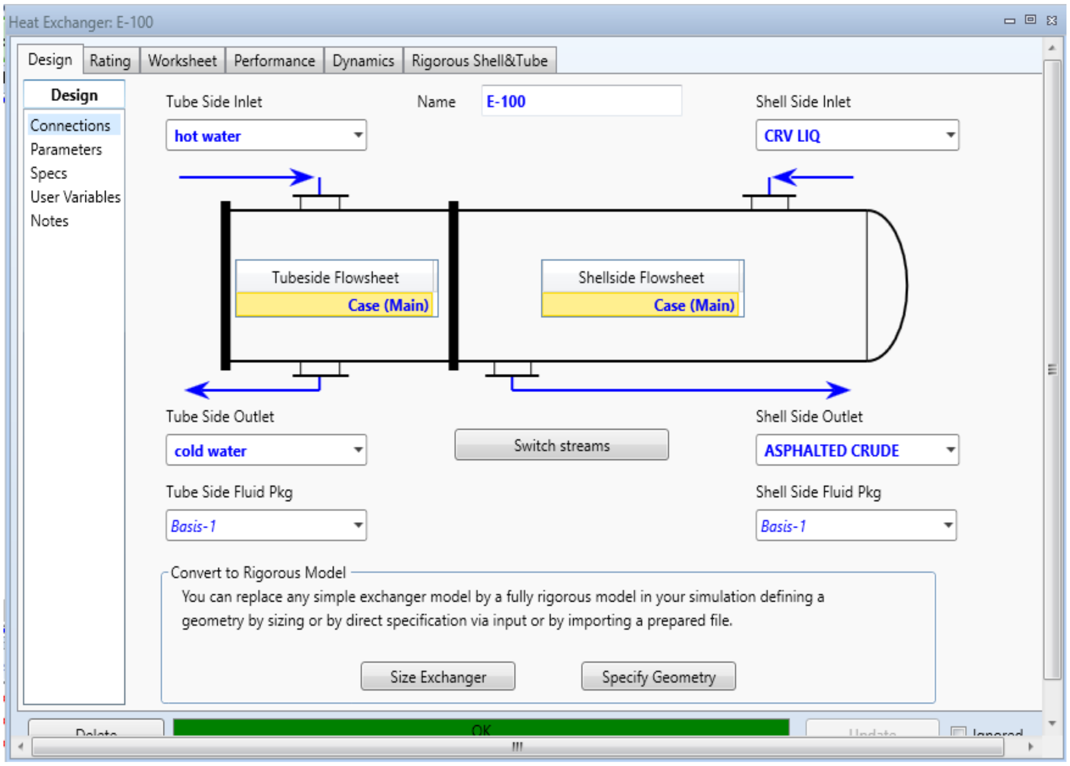


	crude liquid	Aliquat-336_strea	CRV LIQ	CRV-VAP
n-C25	0.0000	0.0000	0.0000	0.0000
n-C26	0.0000	0.0000	0.0000	0.0000
n-C27	0.0000	0.0000	0.0000	0.0000
n-C28	0.0000	0.0000	0.0000	0.0000
n-C29	0.0000	0.0000	0.0000	0.0000
n-C30	0.0000	0.0000	0.0000	0.0000
n-DotriC32	0.0000	0.0000	0.0000	0.0000
n-HexatriC32	0.0000	0.0000	0.0000	0.0000
H2O	0.0000	0.0000	0.0000	0.0000
THEOL-66	0.0000	0.0000	0.0000	0.0000
Oxygen	0.0000	0.0000	0.0000	0.0000
Methanol	0.0000	0.0000	0.0000	0.0000
Asphaltene (Arabian Sample)*	0.4439	0.0000	0.3869	0.8029
Aliquat-336 (Demulsifier)*	0.0000	1.0000	0.0000	0.0000
Resins*	0.3087	0.0000	0.1409	0.0001
Weak Electrolyte*	0.0000	0.0000	0.1285	0.0001
Treated Demulsifier*	0.0000	0.0000	0.1281	0.0009

Figure 23 CSTR composition

4.7 Shell & Tube heat Exchanger

Heat exchanger simulation design and worksheet is displayed here. The requirements are fulfilled and design calculations are analogous with manually determined.



Tube Side Inlet: hot water

Name: E-100

Shell Side Inlet: CRV LIQ

Tube Side Outlet: cold water

Shell Side Outlet: ASPHALTED CRUDE

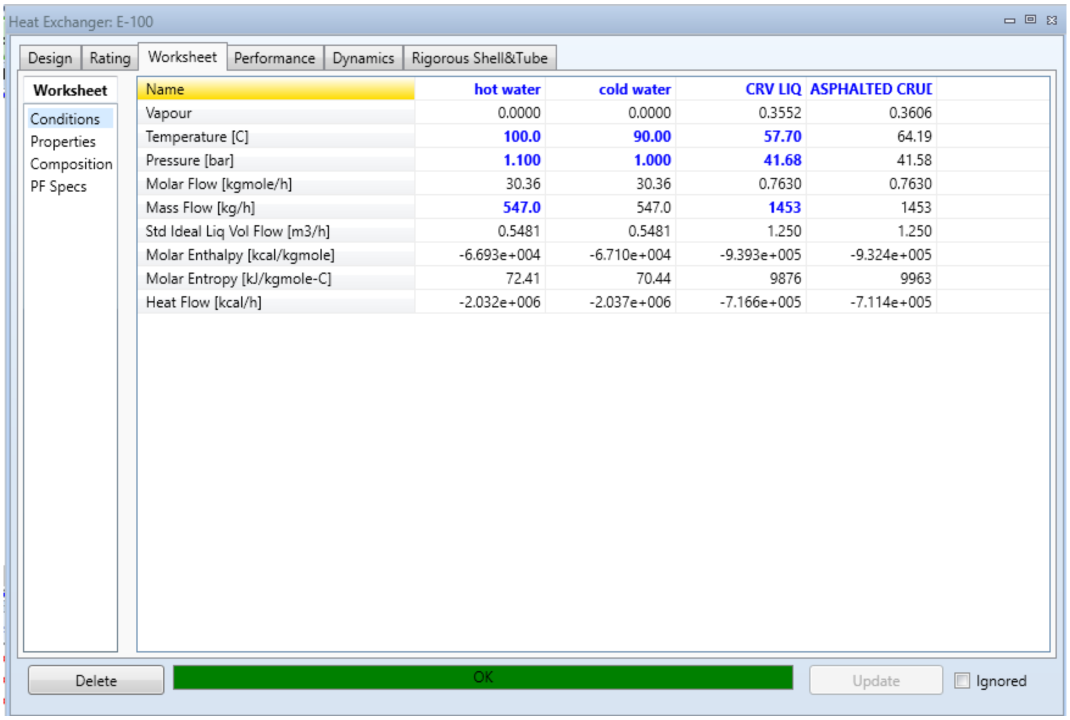
Tube Side Fluid Pkg: Basis-1

Shell Side Fluid Pkg: Basis-1

Convert to Rigorous Model

You can replace any simple exchanger model by a fully rigorous model in your simulation defining a geometry by sizing or by direct specification via input or by importing a prepared file.

Figure 24 Heat exchanger Design

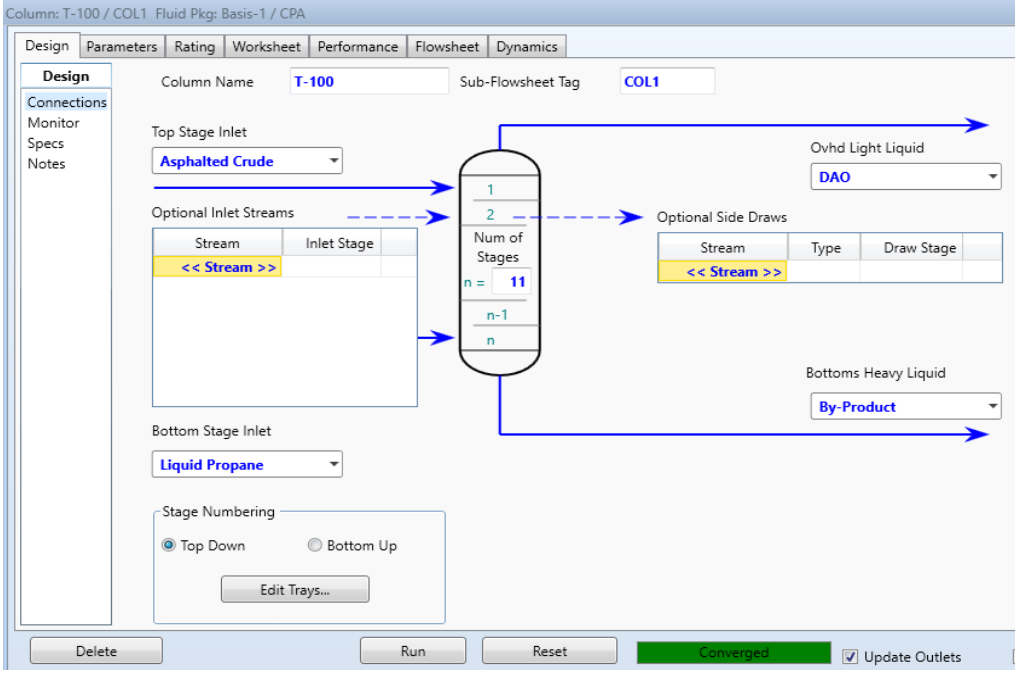


Name	hot water	cold water	CRV LIQ	ASPHALTED CRUI
Vapour	0.0000	0.0000	0.3552	0.3606
Temperature [C]	100.0	90.00	57.70	64.19
Pressure [bar]	1.100	1.000	41.68	41.58
Molar Flow [kgmole/h]	30.36	30.36	0.7630	0.7630
Mass Flow [kg/h]	547.0	547.0	1453	1453
Std Ideal Liq Vol Flow [m3/h]	0.5481	0.5481	1.250	1.250
Molar Enthalpy [kcal/kgmole]	-6.693e+004	-6.710e+004	-9.393e+005	-9.324e+005
Molar Entropy [kJ/kgmole-C]	72.41	70.44	9876	9963
Heat Flow [kcal/h]	-2.032e+006	-2.037e+006	-7.166e+005	-7.114e+005

Figure 25 Heat exchanger worksheet

4.8 Extraction Column

The simulation of LLE column gave us 11 stages which is close to the expected 12 stages calculated theoretically. The simulation result is very close to our study and verify our results.


Column: T-100 / COL1 Fluid Pkg: Basis-1 / CPA

Column Name: T-100

Sub-Flowsheet Tag: COL1

Top Stage Inlet: Asphalted Crude

Ovhd Light Liquid: DAO

Num of Stages: n = 11

Bottom Stage Inlet: Liquid Propane

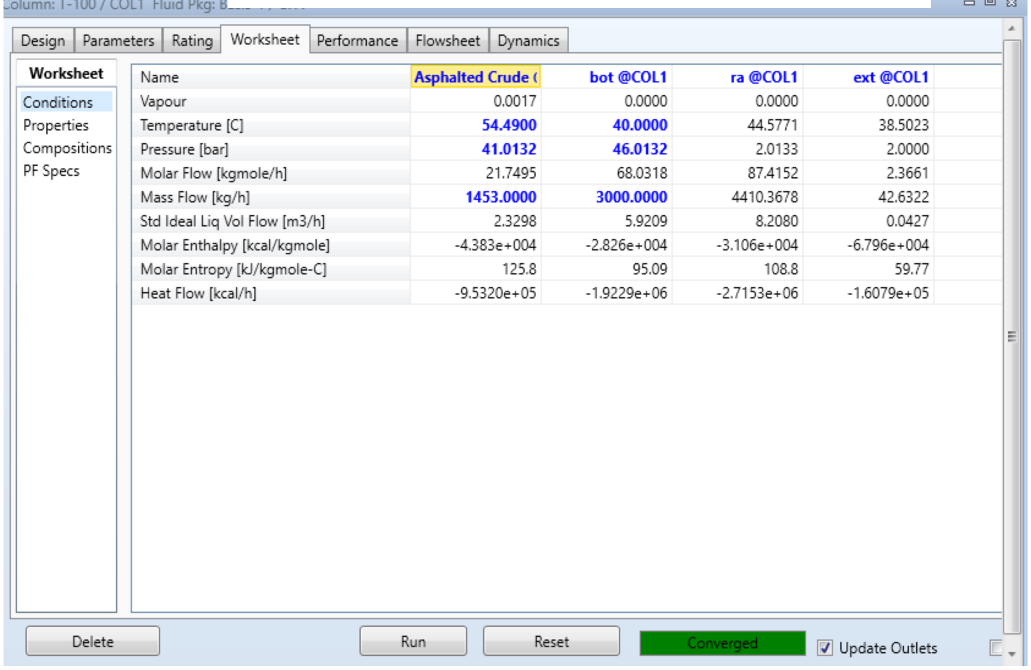
Bottoms Heavy Liquid: By-Product

Stage Numbering: ● Top Down ○ Bottom Up

Converged

☑ Update Outlets

Figure 26 Column Design


Column: T-100 / COL1 Fluid Pkg: B...

Name	Asphalted Crude (	bot @COL1	ra @COL1	ext @COL1
Vapour	0.0017	0.0000	0.0000	0.0000
Temperature [C]	54.4900	40.0000	44.5771	38.5023
Pressure [bar]	41.0132	46.0132	2.0133	2.0000
Molar Flow [kgmole/h]	21.7495	68.0318	87.4152	2.3661
Mass Flow [kg/h]	1453.0000	3000.0000	4410.3678	42.6322
Std Ideal Liq Vol Flow [m3/h]	2.3298	5.9209	8.2080	0.0427
Molar Enthalpy [kcal/kgmole]	-4.383e+004	-2.826e+004	-3.106e+004	-6.796e+004
Molar Entropy [kJ/kgmole-C]	125.8	95.09	108.8	59.77
Heat Flow [kcal/h]	-9.5320e+05	-1.9229e+06	-2.7153e+06	-1.6079e+05

Converged

☑ Update Outlets

Figure 27 Column worksheet

CHAPTER 5

INSTRUMENTATION

5.1 Introduction:

Instrumentation and Process Control play a key-role in the smooth functioning of a process plant. Instrumentation utilizes the hardware components that are also known as controllers to maintain and regulate the various variables that are essential for the process. To provide control and monitoring of our process, we have included the instrumentation and process control in our project. This is done by applying controls on major equipment, including the separator, reactor, heat exchanger and the extraction column.

5.2 Requirements of control:

Control system must meet the following requirements and objectives during operation of chemical plant:

Safety:

The safe operation is very crucial for the well-being of the personnel working at the processing plant. The operating factors should not breach the safety limits, i.e. the concentration, temperature and the pressure limits should not be breached.

Product specification:

Some appropriate instrumentation is being used to achieve desired product quality and quantity. This is essential to have a work system that adds stability to the process .

Environmental Regulation:

There are standards for the safe limits for factors like concentration, pressure and temperature of the exhaust and purge releases from a typical industrial plant.

The conditions for the work also need to lie within the less cost and more benefit to the society limit. These standards are consistently maintained by the plants and the law and regulatory authorities keep a regular check and balance on the industrial plants for it. These standards are met to fulfil the objectives of an organization related to the achievement of their operational and development goals.

For the most part, a control framework fulfills the accompanying:

Reducing the effect of external deviations

Keeping the plant in a stable working environment

Optimization of a chemical process

5.3 Control Parameters

There are variables that play a key role in controlling and regulating a process, these are defined as;

Manipulated variables:

The inputs given to the controller to control the desired part of the process are known as the manipulated variables. For example, in changing the opening of a valve to control the flow through a pipe, the valve opening is the manipulated variable.

Uncontrolled variables:

Some the variables are unable to be controlled by using the controllers and these variables are known as uncontrolled variables.

Controlled Variables:

These are the process conditions that the user decides to regulate using the controllers. For example; in the example of regulating the flow through a pipe by changing the valve opening, flow is the controlled variable.

5.4 Summary of the Defined Controllers

The controllers were defined on the basis of Proportional-Integral type, on the major equipment to monitor and control the fluid flows effectively. A number of level and flow controllers are required to be installed to maintain the liquid level in important equipment like the mixing vessels and the reactor.

The table below represents the summary of the controllers defined throughout our process flow diagram.

Equipment	Name of controller	Type of controller	Manipulated variable	Controlled variable
Separator	FC	PI	Valve opening	Product flowrate
	LC	PI	Flowrate	Product Level
Reactor	RC	PI	Valve opening	Product Output
	LC	PI	Flowrate	Product Level
Heat Exchanger	TC	PI	Steam/HW flowrate	Temperature
	FC	PI	Valve opening	Input Flowrate
Extraction Column	FC	PI	Valve opening	Product Output
	LC	PI	Flowrate	Product Level
	TC	PI	CW flowrate	Product Output

Table 20 Instrumentation Summary

5.5 Process Control

RC: Ratio controller

LC: Level controller

FC: Flow controller

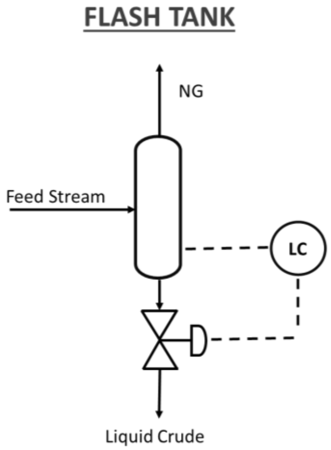


Figure 28 PID Flash tank

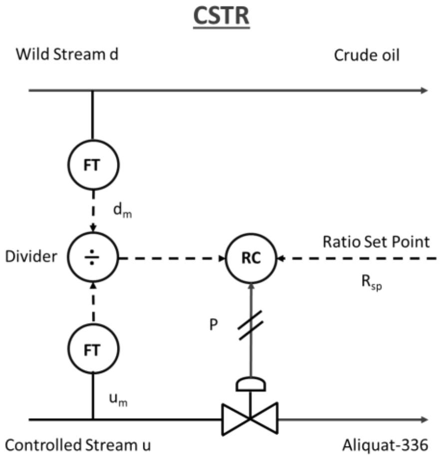


Figure 29 PID CSTR

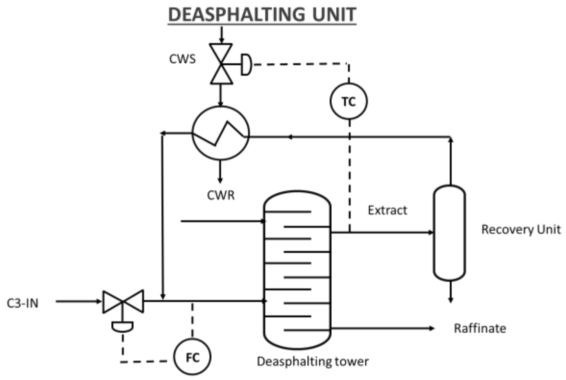


Figure 30 PID Extraction Column

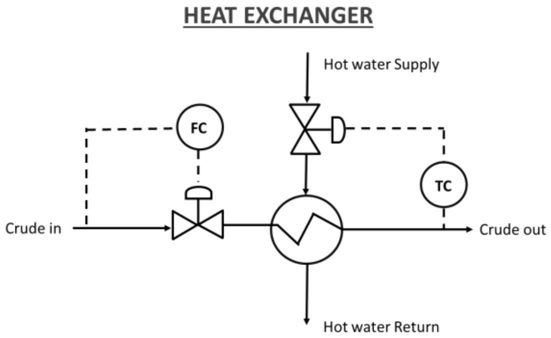


Figure 31 PID Heat Exchanger

CHAPTER 6

COST ESTIMATION AND ANALYSIS

6.1 Separator:

The cost for the agitated mixing vessel made up stainless-steel with a volume of 2.5 m3 is calculated by using Coulson and Richardson volume 6, and is then adjusted for inflation in 2021.

5700£ x (1)1.8 = 10300£

For cost in dollars;

Cost = 10300 x 1.8 = $18500

Cost = 18500 x 2.6 = $48100

6.2 Heat Exchangers:

For a shell and tube heat exchanger made up of carbon steel with

For four customized ordered Heat Exchangers;

Cost = 4500£ x 1.3 x 1.0 = 5850£

Cost = 5850£ x 1.8 = $10500

Cost = 10500 x 2.6 = $27300

6.3 Extraction Column:

The cost for extraction column made up of carbon steel with a volume of 2.4 m^3 is calculated by using Coulson and Richardson volume 6, and is then adjusted for inflation in 2021.

13000£ x (1.3)0.8 = $10800£

Cost in dollars,

Cost = 10800 x 1.8 = $19500

Cost = 19500 x 2.6 = $50700

6.4 Continuous Stirred Tank Reactor:

The cost is calculated by using Coulson and Richardson volume 6, and is then adjusted for inflation in 2021.

Cost = $54827 x 2.6 = $142550

Cost = $31000 x (4.16)0.4 = $54827

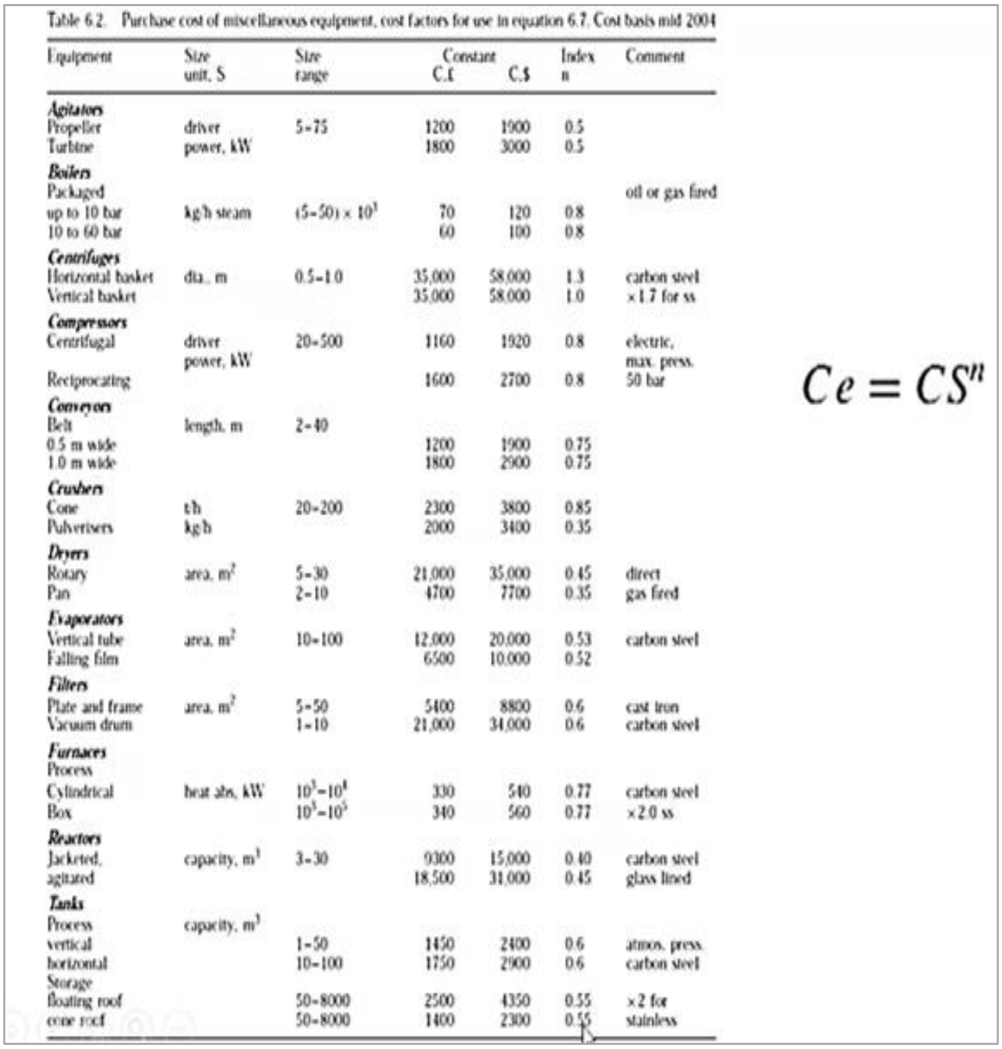
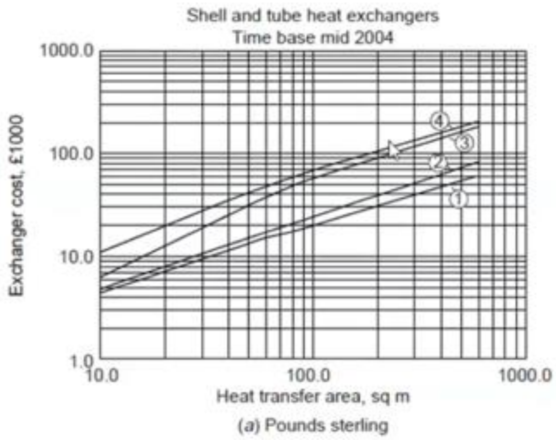


Table 6.2. Purchase cost of miscellaneous equipment, cost factors for use in equation 6.7. Cost basis mid 2004

Equipment	Size unit, S	Size range	Constant		Index	Comment
			C£	C$	n	
Agitators						
Propeller	driver	5–75	1200	1900	0.5	
Turbine	power, kW		1800	3000	0.5	
Boilers						
Packaged						oil or gas fired
up to 10 bar	kg/h steam	$(5–50) \times 10^3$	70	120	0.8	
10 to 60 bar			60	100	0.8	
Centrifuges						
Horizontal basket	dia., m	0.5–1.0	35,000	58,000	1.3	carbon steel
Vertical basket			35,000	58,000	1.0	×1.7 for ss
Compressors						
Centrifugal	driver power, kW	20–500	1160	1920	0.8	electric, max. press.
Reciprocating			1600	2700	0.8	50 bar
Conveyors						
Belt	length, m	2–40				
0.5 m wide			1200	1900	0.75	
1.0 m wide			1800	2900	0.75	
Crushers						
Cone	t/h	20–200	2300	3800	0.85	
Pulverisers	kg/h		2000	3400	0.35	
Dryers						
Rotary	area, m^2	5–30	21,000	35,000	0.45	direct
Pan		2–10	4700	7700	0.35	gas fired
Evaporators						
Vertical tube	area, m^2	10–100	12,000	20,000	0.53	carbon steel
Falling film			6500	10,000	0.52	
Filters						
Plate and frame	area, m^2	5–50	5400	8800	0.6	cast iron
Vacuum drum		1–10	21,000	34,000	0.6	carbon steel
Furnaces						
Process						
Cylindrical	heat abs, kW	$10^3–10^4$	330	540	0.77	carbon steel
Box		$10^3–10^5$	340	560	0.77	×2.0 ss
Reactors						
Jacketed,	capacity, m^3	3–30	9300	15,000	0.40	carbon steel
agitated			18,500	31,000	0.45	glass lined
Tanks						
Process	capacity, m^3					
vertical		1–50	1450	2400	0.6	atmos. press.
horizontal		10–100	1750	2900	0.6	carbon steel
Storage						
floating roof		50–8000	2500	4350	0.55	×2 for
cone roof		50–8000	1400	2300	0.55	stainless

$$Ce = CS^n$$

Figure 32 Cost CSTR



Materials	
Shell	Tubes
① Carbon steel	Carbon steel
② C.S.	Brass
③ C.S.	Stainless steel
④ S.S.	S.S.

Pressure factors	
1–10 bar	× 1.0
10–20	× 1.1
20–30	× 1.25
30–50	× 1.3
50–70	× 1.5

Type factors	
Floating head	× 1.0
Fixed tube sheet	× 0.8
U tube	× 0.85
Kettle	× 1.3

Figure 33 Cost vessel

6.5 Storage Tank:

A storage tank with of 0.42 m2 is attached at the end of the stream. The cost estimate of this tank is found to be $43,700.

6.6 Raw Material:

The raw material is crude oil which is directly extracted from well. The cost for the raw materials is composed of extraction process only.

6.7 Production Cost:

Purchase cost of equipment = PCE = $336800

PPC= 336800(1+0.4+0.7+0.2+0.1+0.15+0.5)

PPC= $1061000

Now for Fixed capital,

Fixed capital= 1061000 (0.3+0.05+0.1)

Fixed capital= $1538500

Working capital is 15% of fixed capital,

Working capital= 1538500*0.15= **$231000**

Total Investment= Fixed Capital + working Capital

Total Investment = 1538500 + 231000

Total Investment = $1769500

Annual Production cost= 419178$

Production cost per kg= 419178/1769500

Production cost per kg= USD 0.03/kg

6.8 Summary of the Expenses:

The cost of the equipment totals up to a value of $336800, that is the fixed capital for the project, the total investment is $1769500. As the project is a retrofit, the additional costing factors are omitted. The table below represents the cost breakdown.

Equipment	**Cost**
Separator	$48100
Reactor	$142550
Heat Exchanger	$27300
Extraction Column	$50700
Fixed Capital	$1538500
Working Capital	$231000
Total Investment	$1769500

Table 11 Cost Summery

6.9 Payout Period:

Our treated product is de-asphalted oil and revenue generation is calculated from current market value of DAO and multiply with production per day.

Revenue from DAO= $2874

Annual revenue = 2874 * 365

Revenue= $1,034795.32

With the total investment of almost 1.76 million and annual revenue of 1.03 million payout period is calculated to be 1.71 years.

Payback period= Total investment/Revenue

Payback period= 1769500/1034795.32

Payout Period= 1.71 Years

CHAPTER 7

HAZOP ANALYSIS

HAZOP is a risk management method that is used to foresee the potential threats and operational hazards that might be present in the system, to eliminate these threats and ensure the proper safety of working personnel present on the plant and the surroundings. Just like in every industry involving processes and safety concerns, we performed HAZOP analysis on a few of our key important equipment.

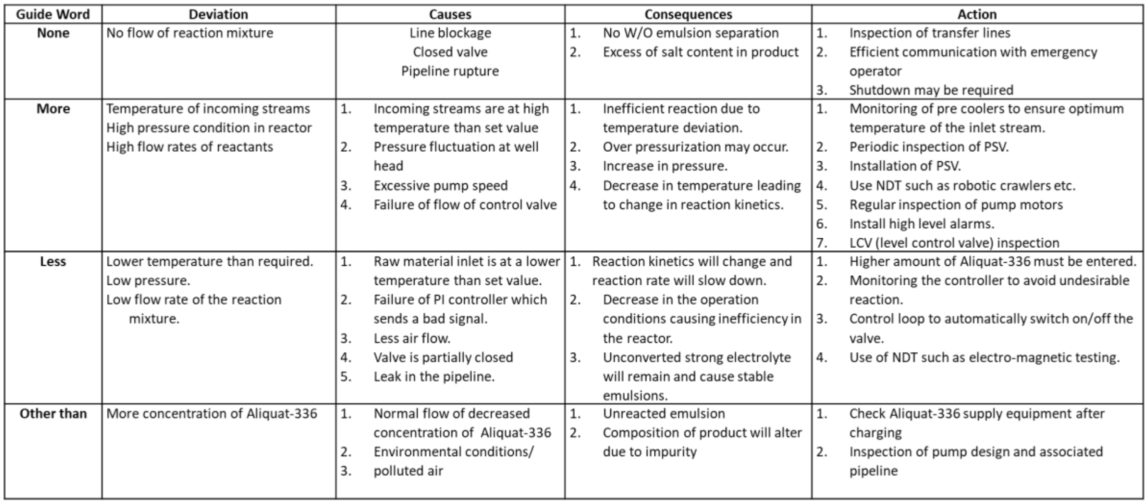
This type of analysis is basically performed in a typical two types of methods and these are Qualitative and Quantitative risk analysis. Qualitative is basically a prerequisite type of analysis whereas most industries usually adopt the Quantitative approach of analysis for better and thorough analysis of any situation. Both of these methods of analysis require different methods and require special types of people that are suited for this job to perform such tasks because it is a very risky analysis. There are a bunch of guide words for the type of deviations that can happen and as well as what are the causes, the direct consequences of the deviation and finally what action to take if a particular deviation occurs

HAZOP basically is finding out how a certain equipment or a process might deviate from what it was originally designed for, what are the possible scenarios, how to identify the problem even before there is a problem. It is basically finding out that everything is going just fine and probably identify certain issues that might cause some major problems in the future if the problem is not addressed at the right time. HAZOP can be summarized and understood by the following points:

- Try to identify possible issues in equipment/processes.
- Identify operation and design from an ideal working map.
- See how the problem can be prevented in order to reduce any type of risk that might involve it.
- Make the process more optimized to further.

7.1 HAZOP on the Reactor:

Intention: Main reaction occurs

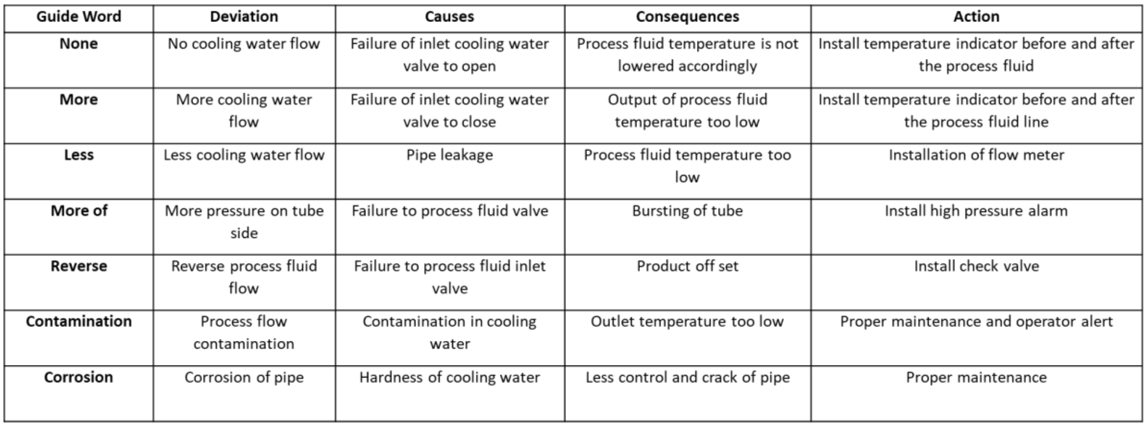


Guide Word	Deviation	Causes	Consequences	Action
None	No flow of reaction mixture	Line blockage Closed valve Pipeline rupture	1. No W/O emulsion separation 2. Excess of salt content in product	1. Inspection of transfer lines 2. Efficient communication with emergency operator 3. Shutdown may be required
More	Temperature of incoming streams High pressure condition in reactor High flow rates of reactants	1. Incoming streams are at high temperature than set value 2. Pressure fluctuation at well head 3. Excessive pump speed 4. Failure of flow of control valve	1. Inefficient reaction due to temperature deviation. 2. Over pressurization may occur. 3. Increase in pressure. 4. Decrease in temperature leading to change in reaction kinetics.	1. Monitoring of pre coolers to ensure optimum temperature of the inlet stream. 2. Periodic inspection of PSV. 3. Installation of PSV. 4. Use NDT such as robotic crawlers etc. 5. Regular inspection of pump motors 6. Install high level alarms. 7. LCV (level control valve) inspection
Less	Lower temperature than required. Low pressure. Low flow rate of the reaction mixture.	1. Raw material inlet is at a lower temperature than set value. 2. Failure of PI controller which sends a bad signal. 3. Less air flow. 4. Valve is partially closed 5. Leak in the pipeline.	1. Reaction kinetics will change and reaction rate will slow down. 2. Decrease in the operation conditions causing inefficiency in the reactor. 3. Unconverted strong electrolyte will remain and cause stable emulsions.	1. Higher amount of Aliquat-336 must be entered. 2. Monitoring the controller to avoid undesirable reaction. 3. Control loop to automatically switch on/off the valve. 4. Use of NDT such as electro-magnetic testing.
Other than	More concentration of Aliquat-336	1. Normal flow of decreased concentration of Aliquat-336 2. Environmental conditions/ 3. polluted air	1. Unreacted emulsion 2. Composition of product will alter due to impurity	1. Check Aliquat-336 supply equipment after charging 2. Inspection of pump design and associated pipeline

Table 12 HAZOP of reactor

7.2 HAZOP on the Heat Exchangers:

Intention: To bring about the required change in temperature for the respective process

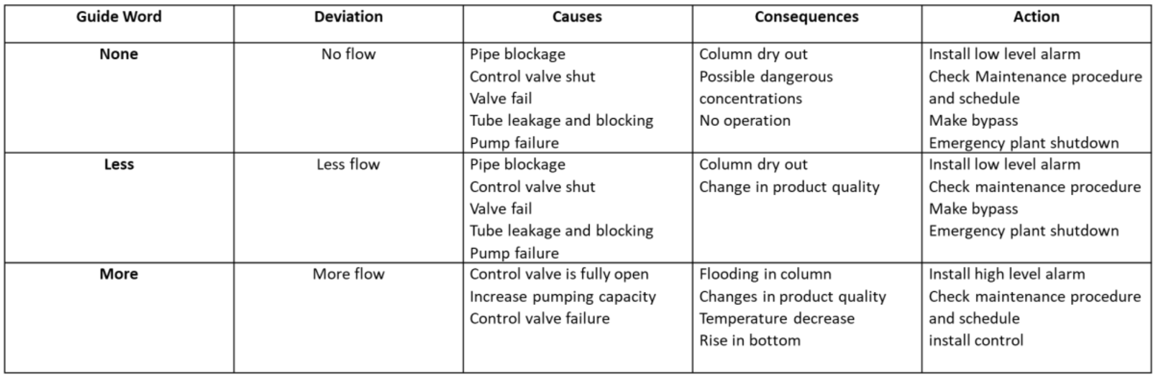


Guide Word	Deviation	Causes	Consequences	Action
None	No cooling water flow	Failure of inlet cooling water valve to open	Process fluid temperature is not lowered accordingly	Install temperature indicator before and after the process fluid
More	More cooling water flow	Failure of inlet cooling water valve to close	Output of process fluid temperature too low	Install temperature indicator before and after the process fluid line
Less	Less cooling water flow	Pipe leakage	Process fluid temperature too low	Installation of flow meter
More of	More pressure on tube side	Failure to process fluid valve	Bursting of tube	Install high pressure alarm
Reverse	Reverse process fluid flow	Failure to process fluid inlet valve	Product off set	Install check valve
Contamination	Process flow contamination	Contamination in cooling water	Outlet temperature too low	Proper maintenance and operator alert
Corrosion	Corrosion of pipe	Hardness of cooling water	Less control and crack of pipe	Proper maintenance

Table 13 HAZOP on Heat exchanger

7.3 HAZOP on Extraction Column:

Intension: the second phase of process where asphalts are treated with solvent de-asphalting.



Guide Word	Deviation	Causes	Consequences	Action
None	No flow	Pipe blockage Control valve shut Valve fail Tube leakage and blocking Pump failure	Column dry out Possible dangerous concentrations No operation	Install low level alarm Check Maintenance procedure and schedule Make bypass Emergency plant shutdown
Less	Less flow	Pipe blockage Control valve shut Valve fail Tube leakage and blocking Pump failure	Column dry out Change in product quality	Install low level alarm Check maintenance procedure Make bypass Emergency plant shutdown
More	More flow	Control valve is fully open Increase pumping capacity Control valve failure	Flooding in column Changes in product quality Temperature decrease Rise in bottom	Install high level alarm Check maintenance procedure and schedule install control

Table 14 HAZOP of Extraction Column

CONCLUSION

The results were considered after the installation of the control system for the de-emulsification chamber on Aliquat-336. And the results showed us that the concentration of the salt associated with the water was reduced from the 2.5 % to 0.1% by weight. The installation of the de-asphalting unit for the removal of the asphaltene from the crude was also advantageous as it reduced the overall asphalt concentration form 43 % in separated liquid crude to less than 2.5%. The production of the de-asphalted and de emulsified crude oil in the end added value to the sales and repute of the industry by helping them meet the contractual standards with the downstream refineries.

The economic perspective of the projected also proved to be the be profitable by successfully generating the annual revenue of 1.03 million dollars with the payback period of just 1.71 years.

After the completion of the project, we were able to achieve our desired objectives.
Following points briefly conclude our project.

- ✓ Design of a de-emulsification & de-asphalting column
- ✓ Simulation of the Process on Aspen HYSYS
- ✓ Decreased concentration of emulsions from 2.5 to 0.1%
- ✓ Asphalts reduction from 43 to less than 2.5%
- ✓ Controlled injection of aliquat-336 in reactor

Our project is in line with the Sustainable Development Goals by the United Nations which are:

- Work Growth & Decent Economics
- Innovation, Industry & Infrastructure
- Responsible Production & Consumption

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