Jordan University of Science and Technology
Faculty of Engineering
Chemical Engineering Department

Preliminary Design of Hydrogen Peroxide Production Plant

Graduation Project

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We dedicate this project first and foremost to Allah the Almighty who has seen us throughout the work to its success. Our project supportive supervisor; Eng. Rowaida Al Zoumot., deserves a very special acknowledgement for her guidance during this project. Completion of this project would not have been possible without her encouragement and moral support. In addition to all the educational staff who contributed to our reaching this stage armed with all the information they provided us. Special and deepest gratitude to our parents for their love, care, support and continued prayers.

ABSTRACT

The objective of this project is to design a hydrogen peroxide plant with a production capacity of 50,000 tons per year using the enzymatic conversion process in Aqaba, Jordan. This plant requires set of equipments to achieve the production goal, 6 reactors, 3 distillation columns, 1 flash vessel, ultrafiltration, 1 shell and tube heat exchanger, 4 storage tanks (1 for formaldehyde, and 3 for hydrogen peroxide), compressor, mixer, and 2 pumps. The volume of one reactor, one hydrogen peroxide storage tank and mixer, is 128.8, 498.2, 571.16 m^3 respectively. The distillation columns of this process are mainly differed in their heights and diameters due to different reflux ratios between them, it was found that the first column that separates almost pure hydrogen peroxide from the rest of the materials, is the longest among all distillation columns used. The precooling of the reactor feed stream is achieved by high heat exchanger transfer rate with 113 m^2 heat transfer area. As the enzyme must be recovered from the reactor outlet stream, a 79 m^2 ultrafiltration unit is used and placed directly after the reactor. Any chemical plant requires pumps and compressors to transfer fluids between locations of different pressures. For this plant, the electrical power that is supplied to the 2 pumps and the compressor is 460 hp, 5.36 hp respectively, to accomplish their task. Choosing the right material of construction for the plant equipments is an important issue to ensure the highest possible standards of the finished product and reduce maintenance costs as possible, by considering the corrosion properties of the materials handled in the process, the MOC for this plant equipments is selected to be stainless steel and carbon steel. Regarding the profitability study of this plant, it is considered profitable with a rate of return of 76.9 %.

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CHAPTER ONE: INTRODUCTION

Hydrogen peroxide is a chemical compound with a combination of two hydrogen atoms and two oxygen atoms that give a formula H_2O_2. It is a very pale blue liquid when it is in pure form, its slightly more viscous than water and its boiling point 150 ℃ which consider greater than boiling point of water by 50 degree. It is mainly used as an oxidizer agent. It was discovered by Louis Jacque Thenard, professor at the Collège de France in 1818. While, he was testing the effect of different acids on barium peroxide (BaO_2), he thought that forms of acids were modified. Eventually, Louis noticed that reacting barium peroxide with sulfuric acid(H_2SO_4) leads to produce a new compound that contained more oxygen than water called hydrogen peroxide and insoluble barium sulfate. It was the most common method of producing hydrogen peroxide until the mid-twentieth century. On the other hand, this compound is naturally produced by photo chemical reactions in the atmosphere, also it can be found in water in small quantities and in very low concentrations in the environment. In 1888, the American Medical Association journal comprised the approval of the presence of hydrogen peroxide in rainwater by Messner in1863. Hydrogen peroxide was commercially used for the first time in the 1800s. Since the 19th century, the demand for this compound has increased dramatically. By the end of the nineteenth century, several formulas were presented of hydrogen peroxide. Nevertheless, Petre Melinkishvili was the first who proves that the correct formula of HOOH (H_2O_2). Nowadays, the production of this chemical is about half a billion kilograms annually. [1,2]

CHAPTER TWO : LITERATURE OF SURVEY

2.1CHEMICAL IDENTITY

There are various IUPAC names of this chemical other than its common name (hydrogen peroxide), which are dihydrogen peroxide, dihydrogen dioxide, bis(hydridooxygen), and dioxidane. This chemical compound with H_2O_2 chemical formula and H-OO-H structural formula has a very light blue color, colorless in solution, and slightly sharp odor. It is classified as microbial, strong oxidizer with active oxygen content of about 47%, bleaching agent, and antiseptic. Its chemistry is dominated by the nature of its unstable peroxide bond because of its peroxide ion(O^{-2}),consisting of two hydroxyl groups joined by a covalent oxygen-oxygen $(O-O)^{-2}$

single bond. Hydrogen peroxide generally has physical properties like water, except for its higher density, and much lower vapor pressure at room temperature compared to water. Besides, due to its inflammability property, it can be easily to handle with in comparison with other oxidizing agents like white fuming nitric acid, red fuming nitric acid, nitrogen tetroxide, and liquid oxygen, it is completely miscible with water, transforms to solid at low temperatures, and has unusual exothermic decomposition into oxygen and water. As for its corrosiveness, it depends on the quantity of the dissolved oxygen produced, and therefore the impaction of the amount of iron and the PH on corrosiveness is greater than of the concentration of hydrogen peroxide. At the last, the amount of this chemical is represented by wight percentage. [3,4]

Table 1 :Physical and chemical properties of H_2O_2

Property	**Description**
Melting point (ºC)	-0.43
Boiling point (ºC)	150.2
Density at25ºC/gmL $^{-1}$	1.4425
Viscosity at 20ºC (mPa s)	1.249
pK_a at 20ºC	11.75
Specific gravity (50%)	1.17 to 1.21
Water solubility	Completely soluble
Solubility in other solvents	Soluble in polar organic solvents

2.2PROCESS PRODUCTION

The oldest way of the production of hydrogen peroxide was from the hydrolysis of barium peroxide with sulfuric acid. Recently, hydrogen peroxide may be produced by the following processes, the most common, auto-oxidation process which is considered as an indirect synthesis, direct synthesis that is still not developed enough, Anthraquinone process by the hydrogenation and oxidation of an organic working solution, electrochemical process, wet chemical process, enzymatic production methods and the most recent, MIT developed process, that produces this chemical continuously from just air, water, and electricity, by using simple, inexpensive, and portable device. The electrochemical processes occur through two electrodes, the cathode that reduces the oxygen (O_2), and the anode that oxidizes the hydrogen or water with requisite protons and electrons.[5]

2.3USES AND APPLICATIONS

Hydrogen peroxide is known by its versatility, it has greater than 100 years of usage. Its first modern use was to bleach straw hats at the beginning of the 20th century. In the recent years, what increased the interest of this chemical is that it can mostly be used in many fields and for many applications, it is used as a strong oxidizing agent than chlorine (Cl_2), chlorine dioxide (ClO_2), and potassium permanganate ($KMnO_4$). Through the addition of oxygen, hydrogen peroxide removes the pollutants from air as well as from wastewater, and it treats both pollutants which are difficult to oxidize such as (dissolved solids and pesticides) and those that are easy to oxidize such as (iron and sulphides). In addition to the wastewater treatment application, hydrogen peroxide is applied in the electronics industry as a cleaning and oxidizing agent, in mining industry as an aid in gold and uranium production processes, in pulp and paper industry as a bleaching and de-inking agent in 1994, and it accounts for 80% of its use. Moreover, food processing industry, some compounds such as starches, natural oils and natural sugars are bleached by this chemical, an antiseptic packaging industry in the manufacturing of the packaging material surface that is in contact with food, and textile manufacturing industry, are other applications of hydrogen peroxide in industry. Hydrogen peroxide has a wide disinfection uses as a popular contact lenses disinfection mechanism, and to disinfect drinking water that was first applied in the 1950s in Eastern Europe because of its biocidal and oxidative efficiency. In medicine, hydrogen peroxide is used as a teeth whitener by oxygenating the stains, while in cosmetics, it is considered as the main component of hair dyes. H_2O_2 should be diluted down to 3% as a base for most applications which is safer way to handle with either for transporting or storage.[6]

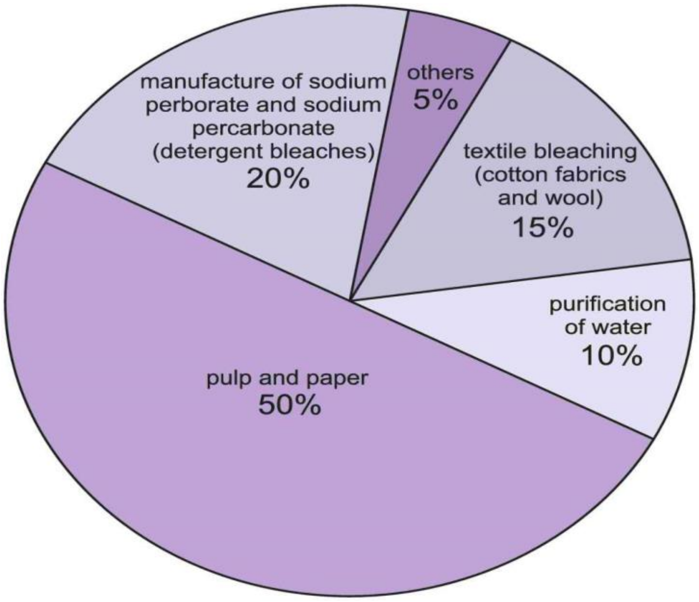
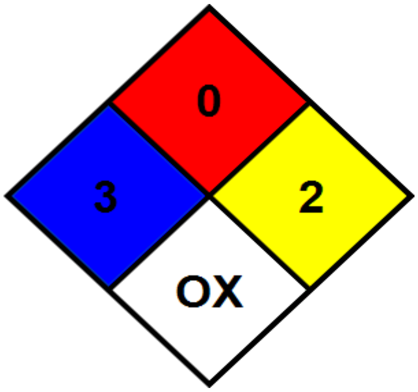


Figure 1:Uses of hydrogen peroxide. (The Essential Chemical Industry (online), 2021)

2.4 HEALTH AND SAFETY HAZARD

Hydrogen peroxide can pose serious health and safety hazards when it is found in high concentration. Hydrogen peroxide is a strong oxidizer that can be corrosive to the eyes, skin and respiratory system so you should wear eyes, face and skin protectors up to 5 ppm wear respiratory equipment. It is considered dangerous material when explosive vapor form comes with contact with certain organic compound. Hydrogen peroxide itself doesn't burn, but its decomposition liberates oxygen which support's combustion, to control the fire involving H_2O_2 use large quantities of water.[7]

Dimond	Hazard	Value	Description
	◆ HEALTH	3	Short exposure could cause serious temporary or residual injury.
	◆ Flammability	0	Not burn under typical fire condition
	◆ Reactivity	2	Normally unstably and may form potentially explosive mixture with water.
	◇ Personal productivity	---	This detonates an oxidizer which can greatly increase the rate of combustion

2.5 STORAGE OF HYDROGEN PEROXIDE

Keep H_2O_2 in organic container in cold area out of direct sunlight and away from combustibles. To prevent the release of vapor into work environment, provides local exhaust ventilation. Storeroom should be made of non-combustible material with impermeable flower. [8]

CHAPTER THREE: MARKET SURVEY

3.1MARKET OVERVIEW

Globally, in 2019 the global hydrogen peroxide demand was nearly 4.1 million metric tons, the global market size of hydrogen peroxide was reached a value of 2.94 billion US$ by the end of 2020, during the period between 2021-2030 the market is further expected to grow at compound annual growth rate (CARG) of 5.1%.

The market was negatively impacted by COVID 19 in early 2020 due to supply of chain distractions and trade restrictions.

Based on the application, the global market of hydrogen peroxide differs from field to another, relating to the demand of the market on certain application of this chemical as shown in the figure below. Recently the chemical synthesis dominates the market since it has the largest share.

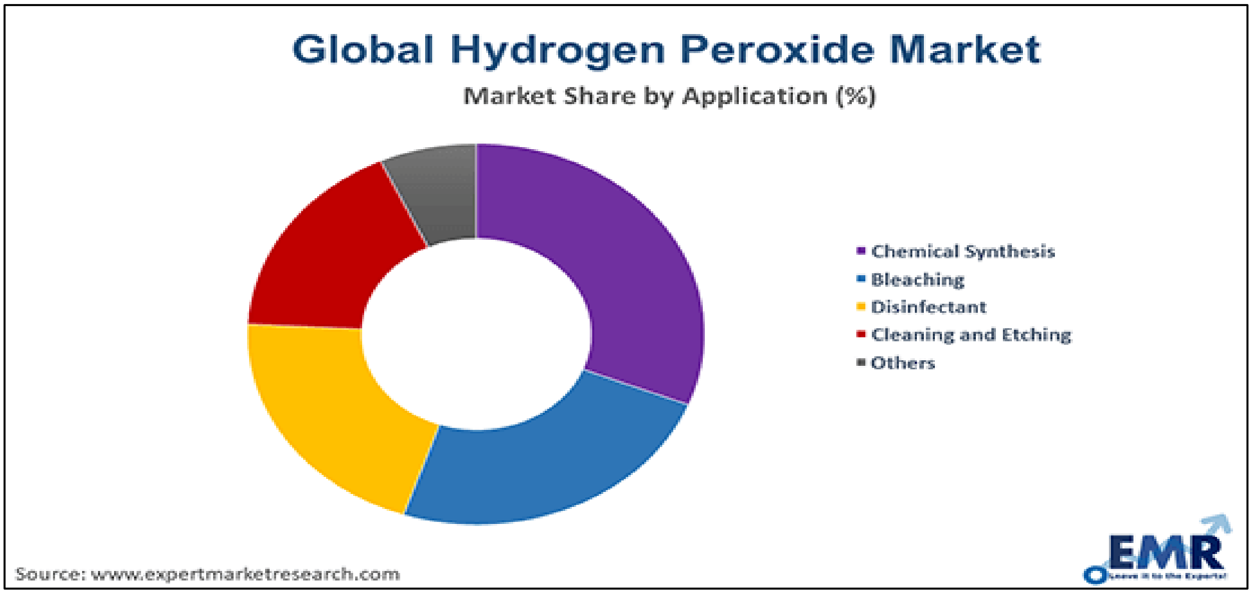


Figure 2:Global Hydrogen Peroxide Market (Global Hydrogen Peroxide Market Report and Forecast 2021-2026, 2021)

It is very important to consider top five producing countries of H_2O_2 which are China, USA, Belgium, Thailand, and Canada. [12]

Table 2:Annual Production of Hydrogen Peroxide In 2015(expressed as H2O2 100%) [12]

World	4.3 million tonnes[1]
China	1.5 million tonnes[2]
US	400 000 tonnes[2]
Belgium	300 000 tonnes[2]
Thailand	300 000 tonnes[2]
Canada	200 000 tonnes[2]

Table (2) shows the production of H_2O_2 for several countries as well as the world production in 2015, it shows that China is the most H_2O_2 producing country in the world due to high demand and presence of hydrogen peroxide manufactures in the region along with a low-cost production factor in China.

Finally, the world consumption of hydrogen peroxide plays a major rule to make fully analysis for market. Figure (3) represents world consumption of hydrogen peroxide.

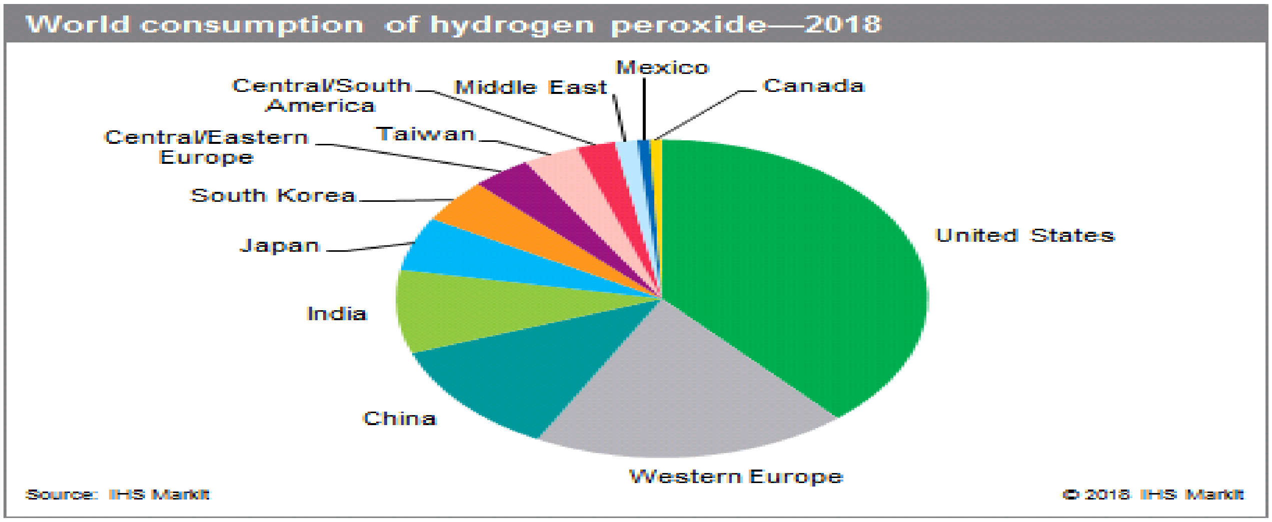


Figure 3:World Consumption of Hydrogen Peroxide (To make better decisions, you need to see the big picture., 2021)

It was noted from figure 3 that United states was the most country that consumes hydrogen peroxide in 2018 followed by Western Europe. China was ranked as the third country when the consumption of hydrogen peroxide was the spot of the talking due to the increase of the usage of H_2O_2 as a bleaching agent in paper industry in China since annual production of paper in China was 130 million ton which made the consumption of hydrogen peroxide there so high. [11]

3.2GLOBAL MARKET (IMPORTS & EXPORT)

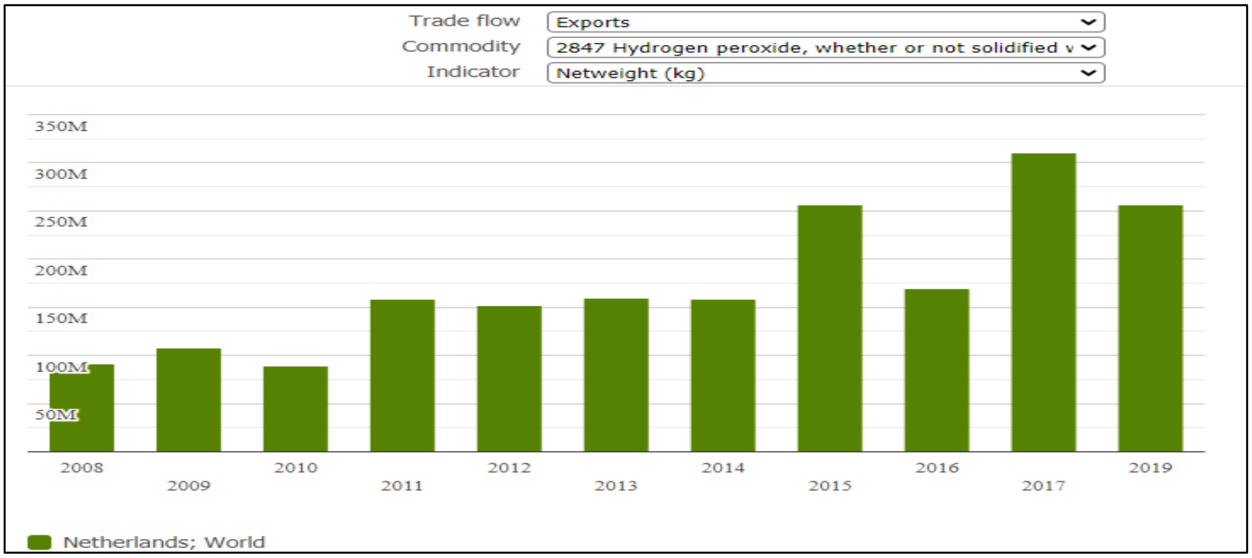
3.2.1 MARKET OF EXPORT

Netherland's market was considered the largest markets of export with 256 M Kg of H_2O_2 in 2019, as its export market value in 2019 was 94 million US$ followed by Belgium that exports 215 M Kg of H_2O_2 with market value 88.6 million US$.

Some of export counties in different locations over the world will be presented with respect to both amount of export in Kg of H_2O_2 and the value in US$ dollars in 2019 as follows:

- United States export 90.8 million Kg (51 million US$)
- China export 32.9 million Kg (32.9 million US$)
- Saudi Arabia export 39.6 thousand Kg (19.2 thousand US$)
- Palestine export 9.8 million Kg (5.3 million US$)
- South Africa export 4.46 million Kg (1.96 million US$)
- India export 674.6 thousand Kg (600 thousand US$) [13]

The figure below shows the market export of H_2O_2 in Netherland's in the Forecast period (2008-2019) as it was noted that the export amount was increased by the years due to the needed of H_2O_2 in the industrial factor.



Trade flow Exports

Commodity 2847 Hydrogen peroxide, whether or not solidified v

Indicator Netweight (kg)

Figure 4:Market Export of H2O2 in Netherland (China | CN: Market Price: Monthly Avg: Inorganic Chemical Material: Hydrogen Peroxide 27.5% | Economic Indicators, 2021)

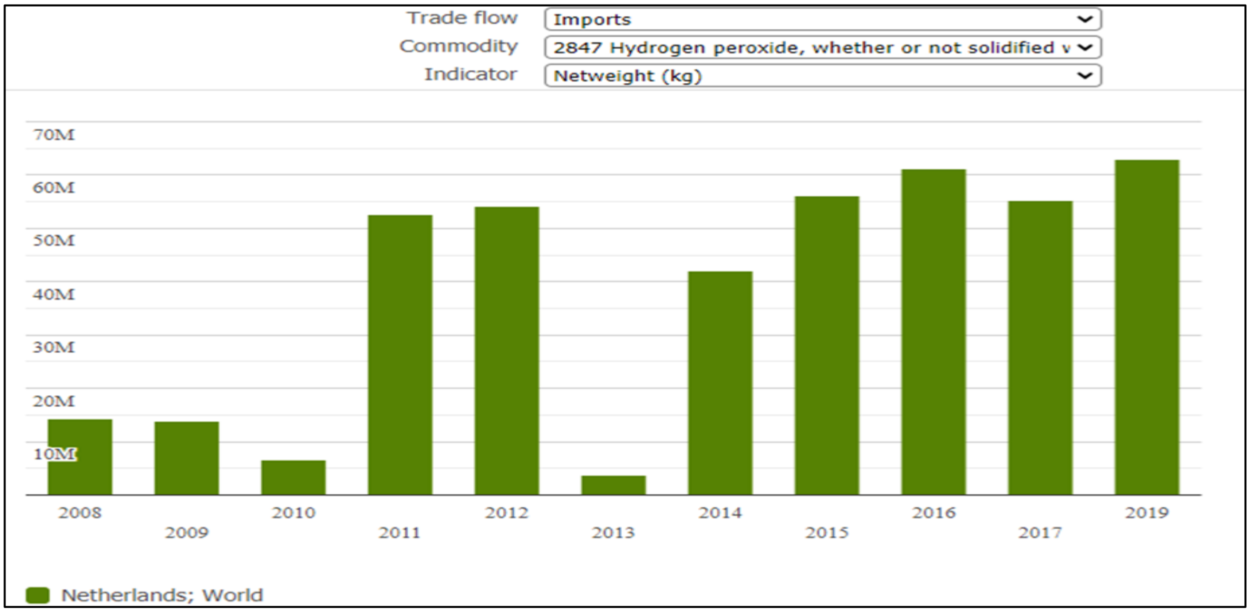
3.2.2 MARKET OF IMPORT

Germany market was considered the largest markets of import to this chemical with 126 M Kg of H_2O_2 in 2019, as its import market value in 2019 was 56 million US$ followed by Italy that imports 96.1 M Kg of H_2O_2 with market value 39.6 million US$.

The main importers of H_2O_2 in the world in 2019 are:

- Netherland's imports 63 million Kg (28 million US$)
- France imports 61.6 million Kg (28 million US$)
- Palestine import 42.2million Kg (19 million US$)
- Korea import 36.7 million Kg (18.8 million US$)
- Yemen import 272 thousand Kg (115 thousand US$) [13]

The figure below shows the market import of H_2O_2 in Netherland's in the Forecast period (2008-2019).



Trade flow: Imports

Commodity: 2847 Hydrogen peroxide, whether or not solidified w

Indicator: Netweight (kg)

Figure 5:: Market import of H2O2 in Netherlands from (2008-2019) (Netherlands | Imports and Exports | World | Hydrogen peroxide, | Value (US$) | 2008 - 2019, 2021)

By compering figure (4) with figure (5), it was noticed that Netherland export of hydrogen by the was more than its import.

3.3LOCAL MARKET (IMPORTS & EXPORT)

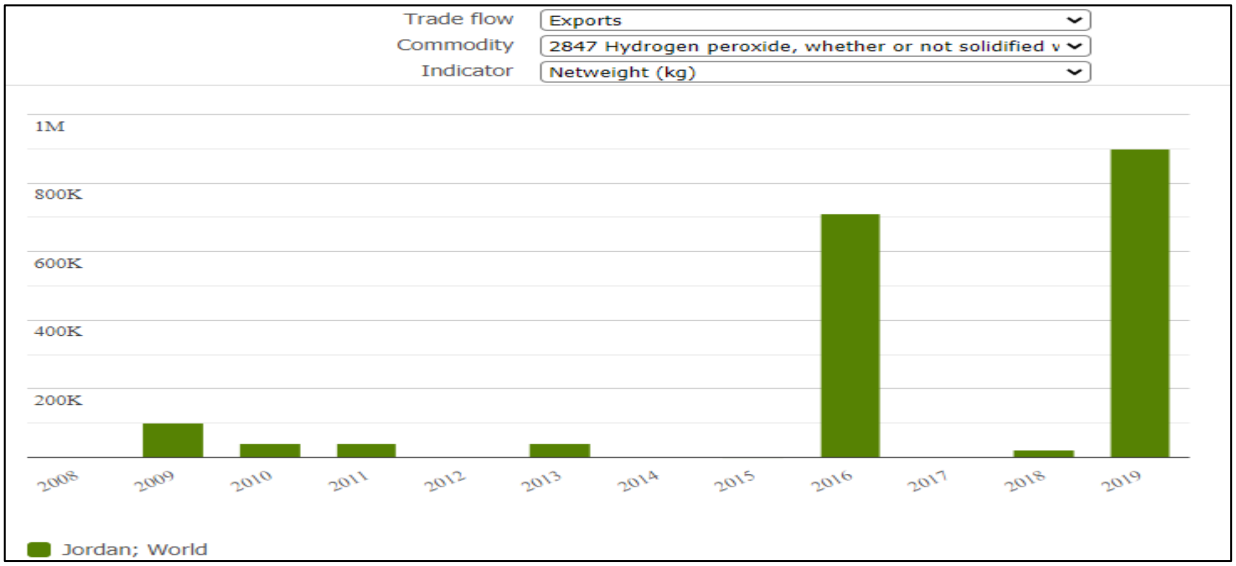
3.3.1 MARKET OF EXPORT (JORDAN)

The total amount of H_2O_2 exported by Jordan to the world in 2019 was 900 thousand Kg with 593 thousand US$.

Jordan exports hydrogen peroxide to the following countries:

- Belgium
- Germany
- United Kingdom
- Sweden

The figure below shows the market export of H_2O_2 in Jordan in the Forecast period (2008-2019).



Trade flow: Exports

Commodity: 2847 Hydrogen peroxide, whether or not solidified v

Indicator: Netweight (kg)

Figure 6:: Export of H_2O_2 from Jordan to World from 2009 to 2019 (Jordan | Imports and Exports | World | Hydrogen peroxide, | Value (US$) | 2008 - 2019, 2021)

As shown in the figure (6), the export from Jordan to world started in 2009 to reach its maximum amount of export in 2019.

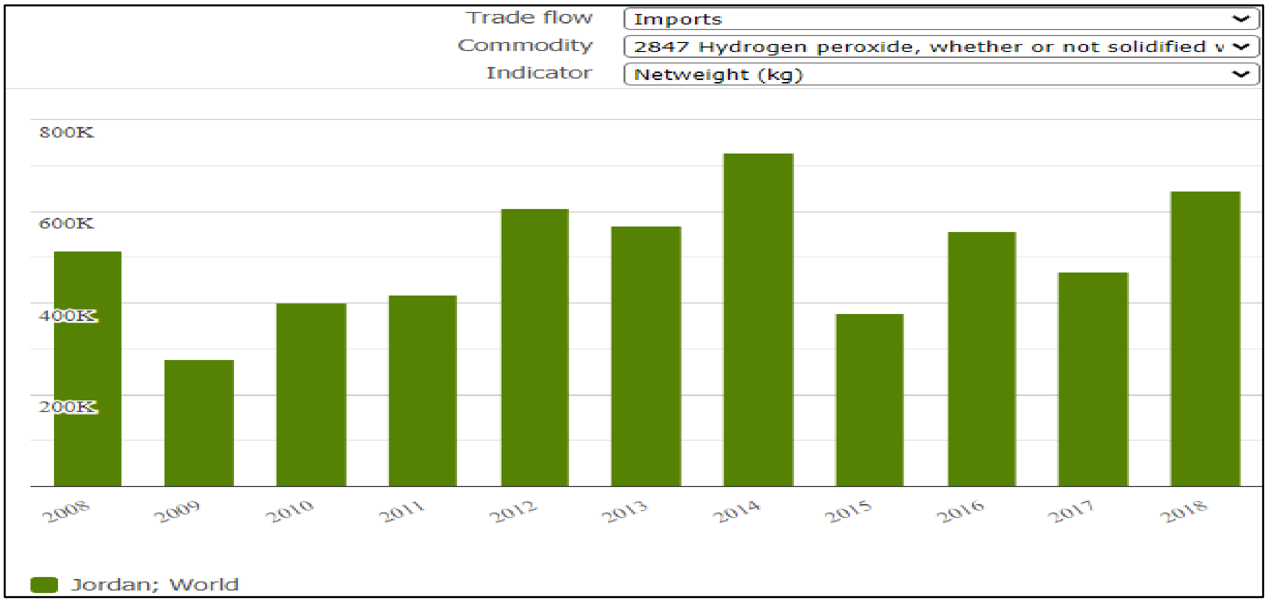
3.3.2 MARKET OF IMPORT (JORDAN)

The total amount of H_2O_2 imported by Jordan in 2019 was 644.6 thousand Kg with 386.8 thousand US$, which indicated that hydrogen peroxide industrial sector in Jordan is so small.

The direction of import to the Jordan market of H_2O_2 in the year 2019 was as follow:

- Thailand
- Belgium
- Palestine
- Spain
- Turkey [13]

The figure below shows the market import of H_2O_2 in Jordan in the Forecast period (2008-2018), and the imported amount reached the highest value in 2014 with 772 thousand Kg of H_2O_2.



Trade flow: Imports

Commodity: 2847 Hydrogen peroxide, whether or not solidified v

Indicator: Netweight (kg)

Figure 7:: Market Import of H_2O_2 in Jordan (Jordan | Imports and Exports | World | Hydrogen peroxide, | Value (US$) | 2008 - 2019, 2021)

3.4PRICE TRENDS

3.4.1 GLOBAL HYDROGEN PEROXIDE PRICES

As it was mentioned earlier that China is the biggest production country of hydrogen peroxide in the world, so that the global price of H_2O_2 will mainly be governed by China market price, and so on to show the hydrogen peroxide price fluctuation during the period of (2006-2019), the changing of its price in China at that period was taken as an example which is represented by figure 9.Price on that figure represented on RMP/ton

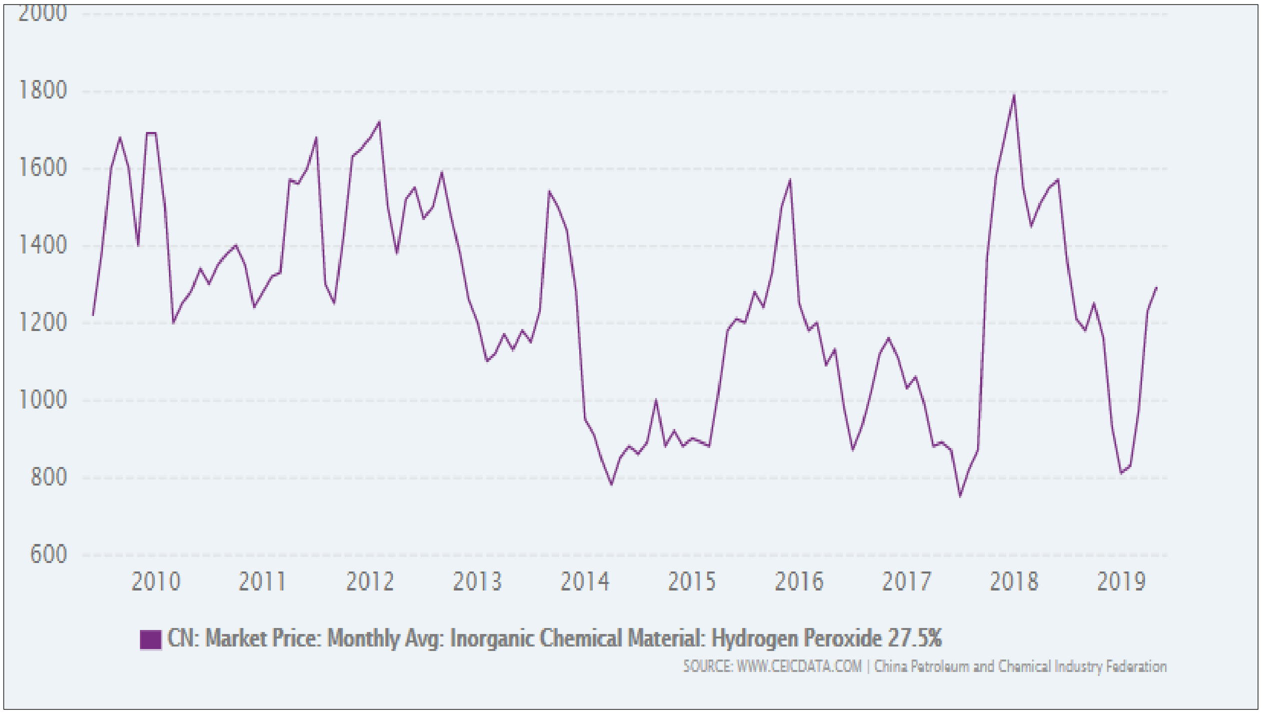


Figure 8:Changing in the price of Hydrogen Peroxide during the period between 2009 and 2019. (China | CN: Market Price: Monthly Avg: Inorganic Chemical Material: Hydrogen Peroxide 27.5% | Economic Indicators, 2021)

From figure (8), it was noticed that maximum price for H_2O_2 in 2018 during this periods, the demand on the hydrogen peroxide product such as paper and pulp was totally increase.

3.4.2 LOCAL HYDROGEN PEROXIDE PRICES

According to Eng.Abdulhamid, who worked in Jordan Engineer Association,[20] the average local price of hydrogen peroxide in Jordan 1000 JD / ton. In metric tons unite price of 50% H_2O_2 is 327.558 JD for 10 metric tons.[16]

3.4.3 RAW MATERIAL PRICES

1- Methanol price

A- Price of methanol in Middle East in 2020

MONTH / WEEK	WEEK 1	WEEK 2	WEEK 3	WEEK 4
APRIL	220 USD	210 USD	210 USD	185 USD
MARCH	260 USD	265 USD	265 USD	260 USD
FEBRUARY	280 USD	285 USD	280 USD	280 USD
JANUARY	290 USD	290 USD	285 USD	290 USD

Table 3:Prices on Middle East USD/MT – Packing in bulk(https://ishtarcompany.com/knowledge-base/methanol-prices/)

Price of the methanol decrease in 2020 that due to drop in oil prices as well as corona virus spread over the world.[17]

Table (3) shows the methanol price from April to January in the middle east in 2020, it was noticed that price over the months increase that due to middle east market return to be activated after its make new rule to confrontation corona virus.

B- Price of methanol in Jordan

Price of industrial methanol with purity equal 99% is $1000/ton when its export from China[18].

2- Oxygen Price

Industrial oxygen price with cylindrical size 50 L is equal $100 per pieces this price when O_2 comes from China [19]

3.5 HYDROGEN PEROXIDE MARKET SIZE 2016 TO 2025

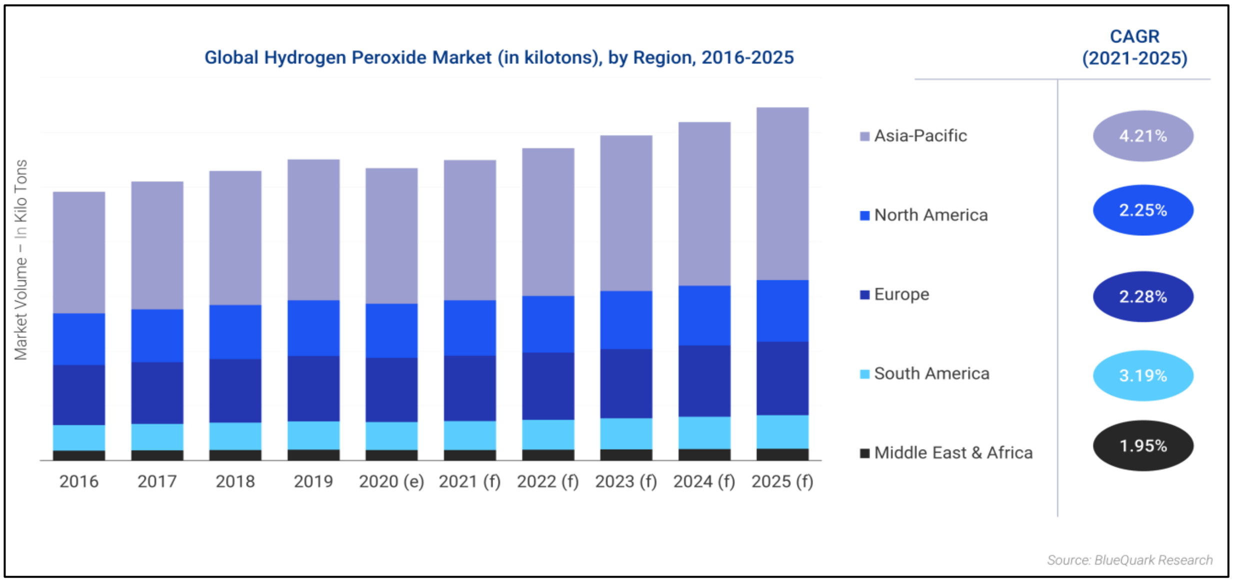


Figure 9:Future Hydrogen peroxide market volume & CAGR(/https://www.bluequarkresearch.com/reports/global-hydrogen-peroxide-market)

Figure (9) illustrate future outlook in hydrogen peroxide market by major player region. It was noted that demand of H_2O_2 will increase with time. North America is expected to have fast-growing due to the continuous increase in the environmental degradation and technological advancement.[21]

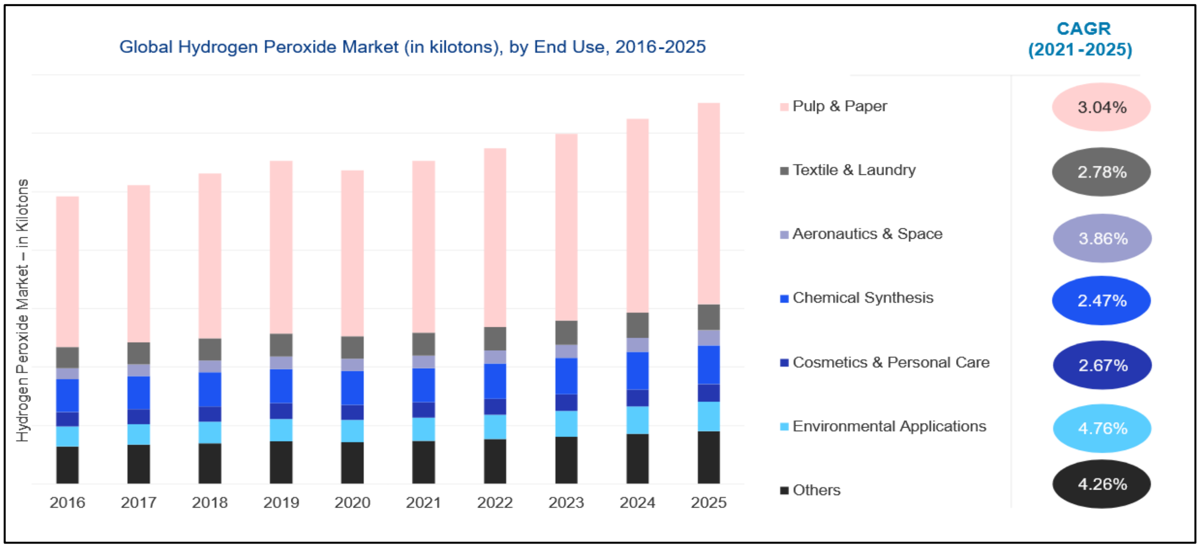


Figure 10:Future Hydrogen peroxide market by end uses

(https://www.bluequarkresearch.com/reports/global-hydrogen-peroxide-market)

Figure (10) shows that the most growth sector from H_2O_2 product is Pulp and paper since hydrogen peroxide improves paper quality as well as environmentally friendly than chlorine compound when its use as bleaching.[21]

3.6 PRODUCTION CAPACITY

There is no plant in Jordan producing hydrogen peroxide.[20]

Since H_2O_2 has a lot of application so it is very important to establish a factory to produce H_2O_2. Production capacity of hydrogen peroxide is mainly impacted by the export and import balance. According to section 2.3 which was talk about Jordan import and export amount of H_2O_2, the suggested capacity of the hydrogen peroxide plant in Jordan would be 1.5 M Kg/year since average import which give indication about Jordan demand of hydrogen peroxide almost equals 700 thousand Kg/year remained 800 thousand Kg/year from suggested capacity will be exported to near countries such as Saudi Arabia and U

CHAPTER FOUR

4.1PROCESS SELECTION

The figure below shows the processes from 1800 to 2000 of produce hydrogen peroxide.

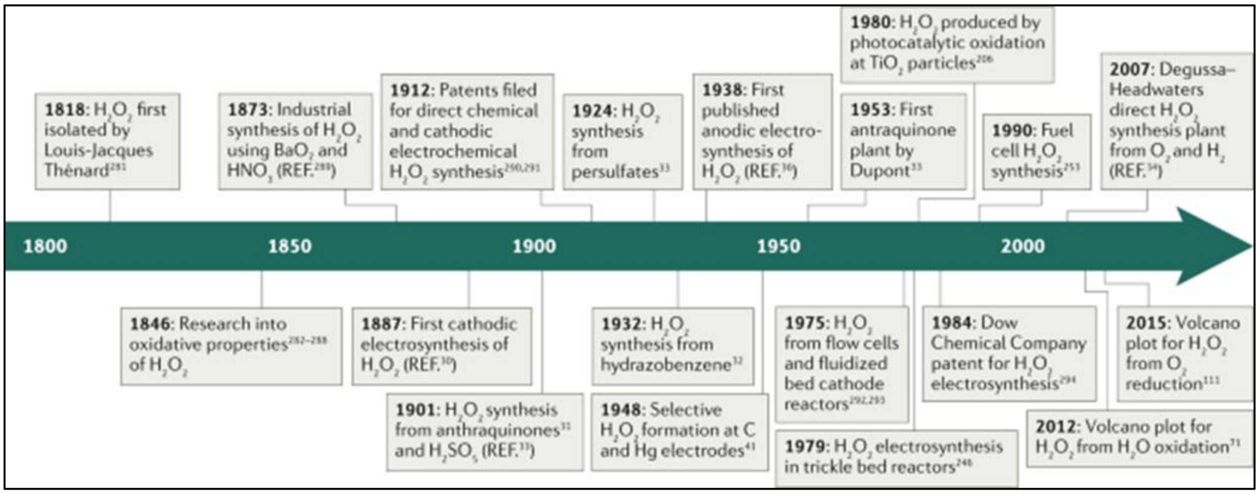


Figure 11:processes of produce hydrogen peroxide. (https://www.nature.com/articles/s41570-019-0110-6)

As mentioned before the hydrogen peroxide manufacturing processes, five processes are chosen for discussion and comparison, which are the Enzymatic process, anthraquinone autooxidation process, direct synthesis process and electrochemical process.

4.1.1 ANTHRAQUINONE AUTOOXIDATION PROCESS

This process is the most common method of manufacturing hydrogen peroxide, as more than 99% of this chemical is synthesized with this process. Hydrogen peroxide is often supplied from this process as 40%w/w, but it can be also concentrated to 50-70%w/w solution. 2-ethyl anthraquinone, hydrogen, 2-ethyl anthrahydroquione, water, and air are the raw chemicals that used for this process. And because of the low-cost essay availability of its raw chemicals, this process is preferred than another. it is four main stages as a form of sequential loop where the 2-ethyl anthraquinone, that is used as H_2 carrier, is reused. In hydrogenation stage catalytic reaction between H_2 and the 2-Ethyl anthraquinone at 320 K to form anthrahydroquinone (H_2Q) is taking place, preceding this main process is the preparation of working solution that contains dissolved anthraquinone and organic solvents, then filtration stage that is filtered out the catalyst from the solution. After that, oxidation stage takes place where the solution is oxidized by blowing air to form H_2O_2, thereafter extraction stage in which a liquid -liquid extraction column used to remove the hydrogen peroxide.[22]

The reactions take place during this process are as follows:

1- Hydrogenation of 2-Ethyl anthraquinone by hydrogen gas

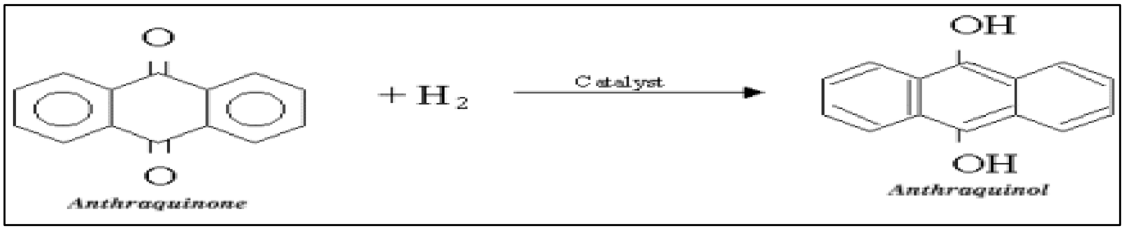


Figure 12:Hydrogenation of anthraquinone ((https://www.essentialchemicalindustry.org/images/stories/350_hydrogenperoxide/35-H2O2_04_(2).jpg)

2- Oxidation of anthraquinol

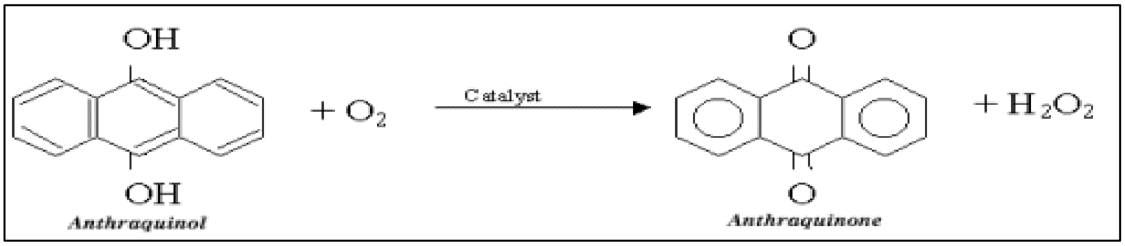


Figure 13:Oxidation of anthraquinol (https://www.cheresources.com/h2o22.gif)

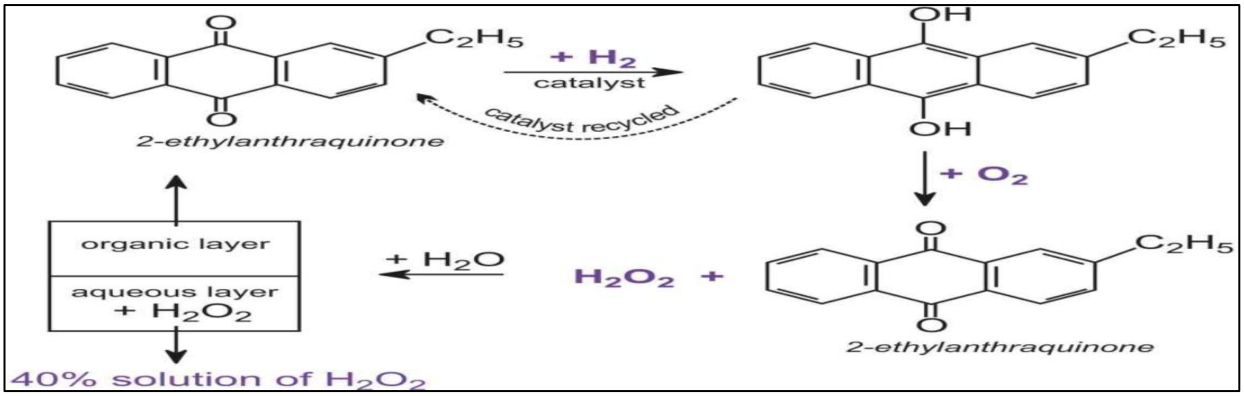


Figure 14:the synthesis loop for hydrogen peroxide manufacturing

(https://www.essentialchemicalindustry.org/chemicals/hydrogen-peroxide.html)

The synthesis loop and peroxide processing depends on many utilities, one of them is the hydrogen plant that is used as a supplying source for the hydrogen gas used at the first stage of the process, where the raw chemical 2-ethyl anthraquinone is hydrogenated by this gas. The hydrogen gas produced from this plant comes out of it with high purity because of subsequent high temperature shift catalytic stage and reforming stage.

Another utility used is the demineralized water, which is added as a feed to the extractor, and the concentrator with high purity. It is purified by conventional degassing followed by cation and anion exchange.

Due to the probability of the presence of trace levels of the solvents in the gas stream produced from the oxidizer, another utility requires is the waste gas system where the gas contained solvent is fed to the absorber. The solvent is absorbed on the activated carbon surface that is used as an absorbent in the column, this agent is then purged with nitrogen gas and the solvent is recovered for further synthesis loops.

Package boiler that generates steam in the stage of concentration the hydrogen peroxide from 40% to 50-70% solution, and calciner system, which washes the catalyst to remove the organics that might block the catalytic sites and affects its activity, are other required utilities for this process.

As a safety procedure, fire system utility is provided the process with foam injection, since all the materials used in this process are highly flammable, such as the natural gas, hydrogen, and organic solvents.[23]

This process is considered an eco-friendly process, liquid wastes that are resulted from the cooling tower, and demineralization plant wash water, both are the major sources of waste. While the gaseous emissions of solvents are reduced to the acceptable concentrations by the EPA.

This process is mainly used to produce hydrogen peroxide that is used in specific applications that require 40% w/w, such these applications are for disinfecting tools, cleaning surfaces, oral clean, and gardening. However, the concentrated product with 50-70% w/w of this compound produced after the purification and concentration stage of this process is used to prevent infection of minor cuts and burns, and to relieve minor mouth irritation. Also, it helps to remove mucus as a mouth rinse.[23]

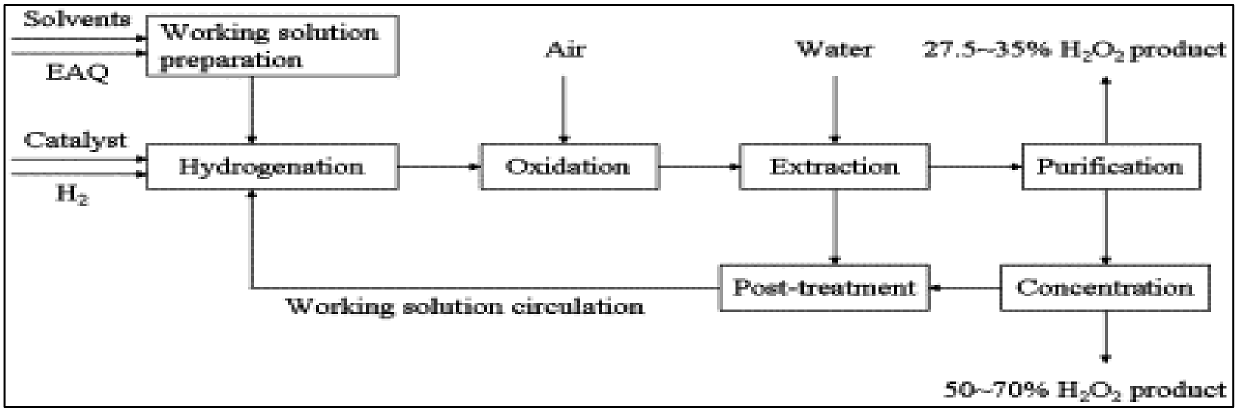


Figure 15:anthraquinone block flow diagram (https://nzic.org.nz/app/uploads/2017/10/1E.pdf)

4.1.2 DIRECT SYNTHESIS OF H_2O_2 WITH CATALYST FROM H_2 AND O_2

The first process that was capable to produce H_2O_2 from hydrogen (H_2) and oxygen (O_2) was done by H. Henkel and W. Weber in 1914.[24]

This process also called green hydrogen peroxide production process since it does not have negative environmental impact owing to the production of unwanted wastes.

The process mainly occurs by direct reaction of two gaseous mixture: an oxygen spices (free and bound) with hydrogen in pressurized vessel along of water, Palladium (Pd) in its pure form or alloyed with certain metals is use as a catalyst, the process is stabilized by adding mineral acids or halides to prevent the decomposition of hydrogen peroxide to water. [25]

The process was illustrated on Figure (17):

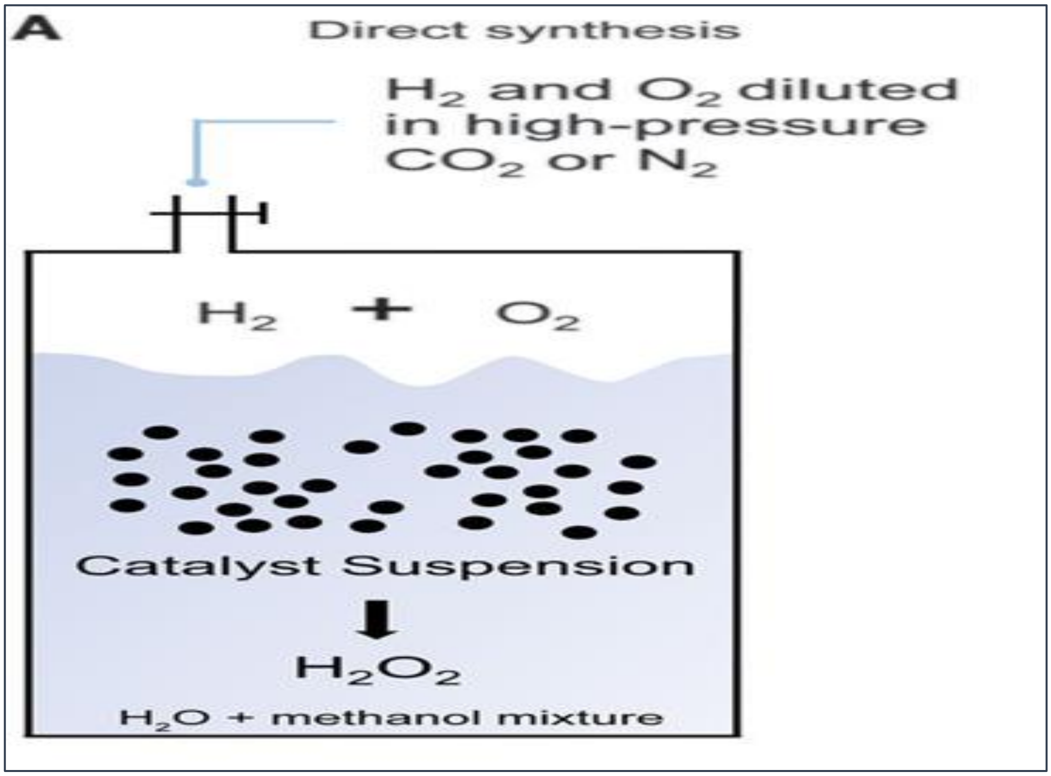


Figure 16:Direct synthesis of H_2O_2 with catalyst from H_2 and O2/(https://science.sciencemag.org/content/366/6462/226)

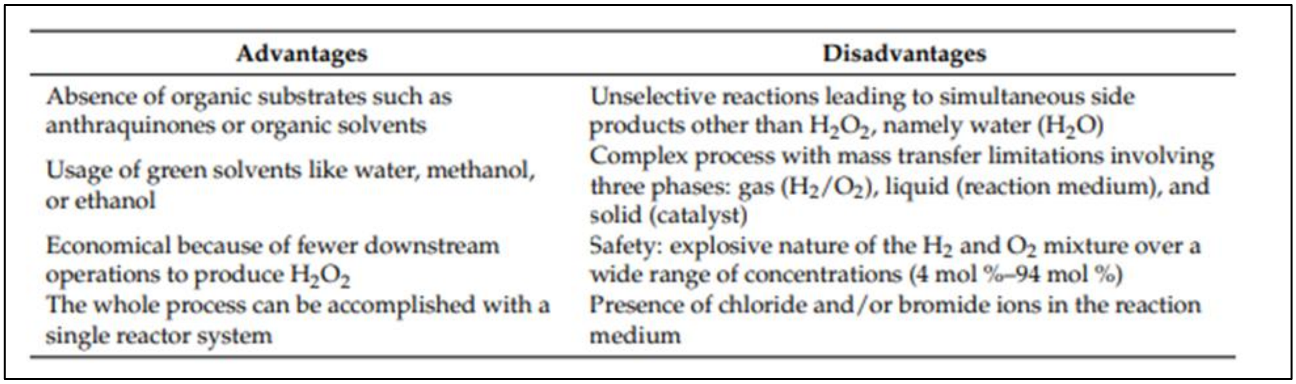
Equation (1) shows the direct synthesis of H_2O_2

$$\begin{tikzcd}
 & & & H_2O + 0.5O_2 \\
H_2O+0.5\,O_2 & H_2 + O_2 \arrow[l] \arrow[r] & H_2O_2 \arrow[ur] \arrow[dr] & \\
 & & & 2\,H_2O
\end{tikzcd}$$

As seen direct synthesis consist of two parallel reactions (reduction and oxidation), the reaction either produce water or hydrogen peroxide depending on the reaction condition. [26]

If the reaction take place in batch reactor at 20 atm and 30 ℃ the reaction produces 6.6 mol% of aqueous H_2O_2 after 20 h. [27]

Table 4:Advantage and disadvantage of direct synthesis process [28]



Advantages	Disadvantages
Absence of organic substrates such as anthraquinones or organic solvents	Unselective reactions leading to simultaneous side products other than H_2O_2, namely water (H_2O)
Usage of green solvents like water, methanol, or ethanol	Complex process with mass transfer limitations involving three phases: gas (H_2/O_2), liquid (reaction medium), and solid (catalyst)
Economical because of fewer downstream operations to produce H_2O_2	Safety: explosive nature of the H_2 and O_2 mixture over a wide range of concentrations (4 mol %–94 mol %)
The whole process can be accomplished with a single reactor system	Presence of chloride and/or bromide ions in the reaction medium

Table (4) list the advantage and disadvantage for direct synthesis process. The mainly observation on this table that direct synthesis process can be done by using single reactor system which indicate that this process is sample.

Direct synthesis process is just for small scale industry (pilot plant) because of three major drawbacks. Firstly, for safety issue H_2/O_2 mixture should be outside the flammability limited of H_2 in O_2 which typical between (4- 94% v) in 1 atm & 25 ℃ .Secondary, Pd catalyst is active for both the production and decomposition of H_2O_2. Thirdly, excessive use of solvent for the process.[29]

4.1.3 ELECTROCHEMICAL PROCESS

Electrochemical production of H_2O_2 is well suited to distributed manufacturing. This process is developed enough to play a major part of hydrogen peroxide production with AO process as they required capital investment and can be operated at ambient pressure and temperature in small plant.

Electrochemical process involves two main process which is Degussa-weissenstein process and Muncher process. These processes will be disused in detail below.[30]

A) EGUSSA-WEISSENSTEIN PROCESS

The basis of Degussa-Weissenstein process is an oxidation of sulfuric acid then hydrolyzed it to produce hydrogen peroxide. The process will describe in detail by using block diagram below.

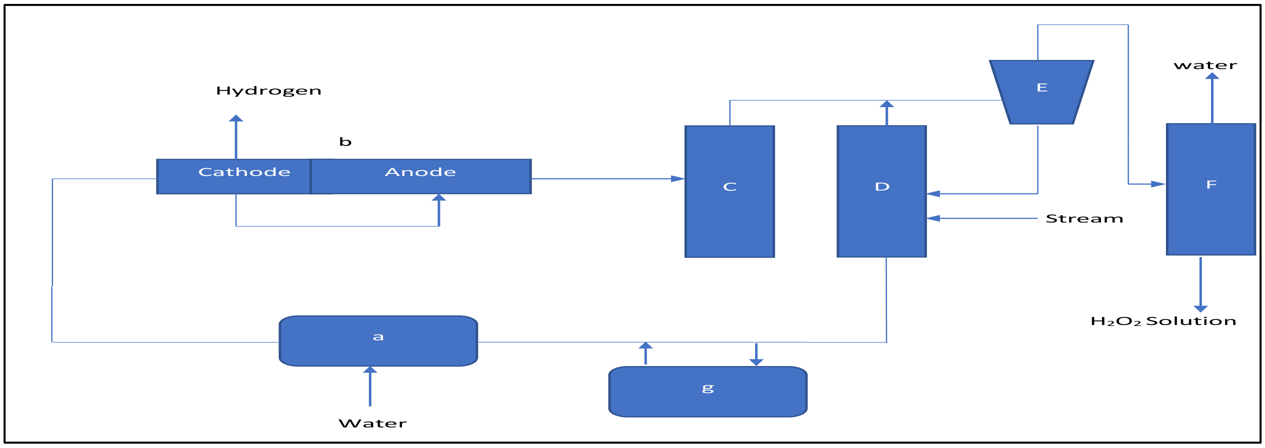


Figure 17:Degussa-Weissenstein Process (Ullmann's Encyclopedia of Industrial Chemistry, 2021)

a) Electrolyzer feed tank; **b)** Electrolyzer; **c)** Evaporation and hydrolysis; **d)** Hydrogen peroxide expulsion and hydrolysis; **e)** acid separator; **f)** fractional column; **g)** acid purification.

Sulfuric acid in electrolyzer feed tank, block (a), is adjected to concentration equal 550-570 g/l bypass a stream of water in the feed tank, then the solution enters the electrolyzer, block (b), by electrolytic process 50% of aqueous solution of sulfuric acid on electrolysis gives peroxodisulfric acid ($H_2S_2O_8$) which obtained by electrolytic oxidation of sulfuric acid by the following equation:

$$\left.\begin{aligned}&2H_2SO_4 \rightarrow 2HSO_4^- + 2H^+\\&2HSO_4^- \rightarrow H_2S_2O_8 + 2e^-\end{aligned}\right\}\quad\text{(At the anode)}$$

$$\left.\begin{aligned}&H_2S_2O_8 + 2H_2O \rightarrow 2H_2SO_4 + H_2O_2\\&2H^+ + 2e^- \rightarrow H_2\end{aligned}\right\}\quad\text{(At the cathode)}$$

To raise the chemical potential between cathode and anode some chemical such as hydrochloric acid are used.

On the cathode section heavy metal ion are precipitated so that residual hydrogen peroxide is destroyed as a consequence hydrogen peroxide do not form in the cathode section, but it is formed by hydrolysis of peroxodisulfate ion to peroxomonosulfate anion as the following equation:

$$S_2O_2^{-8} + H_2O \rightarrow HSO^{-4} + SO_3(OOH)^-$$

$$SO_3(OOH)^- + H_2O \rightarrow HSO^{-4} + HOOH$$

This hydrolysis process is carried out in two steps 80-90% complete in unite (c), the remine parentage complete in unite (d). also, in hydrolysis unite (c) hydrogen peroxide is distillated from sulfuric on the percentage 50%, other 50% was done in unite (d).

Then to separate sulfuric acid from hydrogen peroxide and water the stream enters acid separate unite (e) so that two streams are exits from unite (e) which is steam that contained and acid steam.

First stream contains (H_2O&H_2O_2) which will then pass to flash separate unite (f) that separate H_2O from H_2O_2. Second stream contain H_2O_2 with sulfuric acid which retains to unite (d) where hydrogen peroxide is expelled from the concentrated sulfuric acid solution by counter current steam. To reuse sulfuric acid that produce as a distillated from unit (d) this stream enters in acid purification unite(g) to remove all traces before reuse in new turn. By using Degussa-Weissenstein process an aqueous solution with maximum content of 45% H_2O_2 is obtained.[31]

B) MUNCHER PROCESS

Developed by Pietzsch and Adolph at the Elektrochemische Werke, Munich. Used potassium eroxodisulphate instead of sulfuric acid to produce H_2O_2.

In the Muncher process, ammonium peroxydisulfate is converted to potassium peroxydisulfate, which is then hydrolyzed.

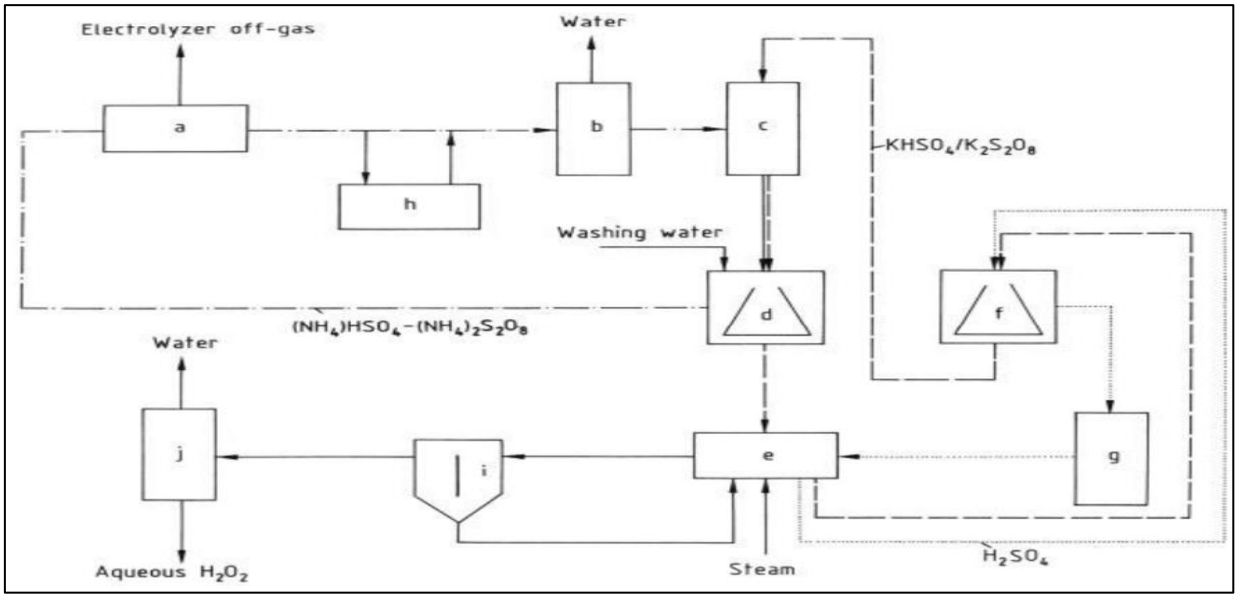


Figure 18 :Muncher Process to produce H_2O_2

a) Electrolyzer; **b)** Vacuum cooler; **c)** Conversion (crystallization); **d)** Centrifuge,

e) Hydrolysis–distillation; **f)** Centrifuge; **g)** Sulfuric acid purification; **h)** Purification of cycle solution; **i)** Acid separator; **j)** Fractionating column.

The circulating solution contains 300 g of $(NH_4)_2SO_4$(Ammonium sulfate), 40 g of K_2SO_4, 100 g of H_2SO_4(Sulfuric acid), and 57.6 g of $S_2O_2^{-8}$ per liter before entering the electrolyze. Its peroxydisulfate concentration is increased to 115.2 g/L by the electrolytic process. After electrolysis, the circulating solution is cooled and concentrated in a vacuum cooler (b) by evaporation of water and then fed into a crystallizer (c), then half of the peroxydisulfate is precipitated as potassium peroxydisulfate by addition of potassium hydrogen sulfate and separated in a centrifuge (d). Next the liquor diluted with washing water is returned to the electrolyze. Solid potassium peroxydisulfate is hydrolyzed(e), and hydrogen peroxide is steam distilled. The remaining potassium hydrogen sulfate– sulfuric acid residue is separated in a centrifuge(f), then Sulfuric acid is recycled to the hydrolysis and distillation stage (e), via a purification stage (g) and Potassium hydrogen sulfate is returned to the crystallizer(c) for precipitation of peroxydisulfate. To purify the circulating solution, a side stream is mixed with ammonia in the purification step (h) to precipitate iron. The hydrogen peroxide–water vapors formed during hydrolysis and distillation are led through an acid separator(i) and separated in the rectification step(j). Because hydrolysis is carried out batchwise, the initial vapor mixture is rich in hydrogen per oxide, giving a distillate of 60 wt % which then drops to 20 wt %. [31]

The following equations describe the reactions that take place in Muncher Process:

$$2(NH_4)HSO_4 \longrightarrow (NH_4)_2HSO_4+H_2$$

$$S2O_8^{-2}+H_2O \longrightarrow HSO_4^{-}+HSO_5^{-}$$

$$HSO_5^{-}+H_2O \longrightarrow HSO_4^{-}+H_2O_2$$

4.1.4 THENARD PROCESS

Thenard process involves reacting barium peroxide with nitric acid to produce a low concentration of aqueous hydrogen peroxide; and it can be greatly improved by using hydrochloric acid. Hydrogen peroxide is formed in conjunction with barium chloride, and both are soluble in water. The barium chloride is then removed by precipitation with sulfuric acid. Vacuum distillation is used in this process to obtain pure hydrogen peroxide.[32]

These equations explain how the process occurs:

$$BaO_2 + 2HCL \rightarrow BaCL_2 + H_2O_2$$

$$BaCL_2 + H_2SO_4 \rightarrow BaSO_4 + 2HCL$$

$$BaO_2 + H_2SO_4 \rightarrow BaSO_4 + H_2O_2$$

Associated Problem

- Low concentration of the product
- Low reaction rate
- This process required time and vacuum distillation to produce higher concentration of hydrogen peroxide which will increase the cost of the process

4.1.5 ENZYMATIC PROCESS

It is considered one of the other processes used to produce and supply the market demand for hydrogen peroxide. This process has recently discovered and developed as an alternative for the most common process used for H_2O_2 production, the anthraquinone process, due to its ability to overcome the limitations of anthraquinone. The major comparison points are that it requires less capital-intensive equipment in comparison with the requirement of large intensive facilities throughout the anthraquinone process, as well as the use of inexpensive feed stocks to manufacture important chemicals like hydrogen peroxide, which could be economically predominant to the processes which has been mentioned before. The process basically depends on the principle of enzymatic conversion of lower alkyl alcohols, in particular methanol to corresponding aldehydes (formaldehyde), and the targeted component, hydrogen peroxide in presence of oxygen at operating conditions which involve low temperatures. The process may be employed in batch or continuous operation. It is important to identify a suitable organism and alcohol oxidase enzyme, and it has focused on yeasts which are common to fermentation. This conversion process may be used to convert ethanol to acetaldehyde and hydrogen peroxide, allyl alcohol to acrolein and hydrogen peroxide, or methanol to formaldehyde and hydrogen peroxide which has been adopted here. The methanol oxidase enzyme is derived from a strain of yeast cell, either Hansenula, Pischia, Candida, or the most common Pischia pastoris and Hansenula polymorpha. While for any conversion process, regarding to the type of alcohol used, to both aldehyde and hydrogen peroxide implicates the addition of oxygen or oxygen containing gas to the reaction mixture, this is called aeration step. This process is preferable to operate at lower temperatures about 10^oC to -22^oC, since as the enzymatic reactions of methanol oxidase carried out at temperatures below 10^oC and at substrate concentrations up to 30% or above, higher catalytic capacity and production yield could be achieved with H_2O_2 concentrations up to 50% and conversion of about 90%. The following diagram represents the block flow diagram for the enzymatic conversion process of hydrogen peroxide production.[33]

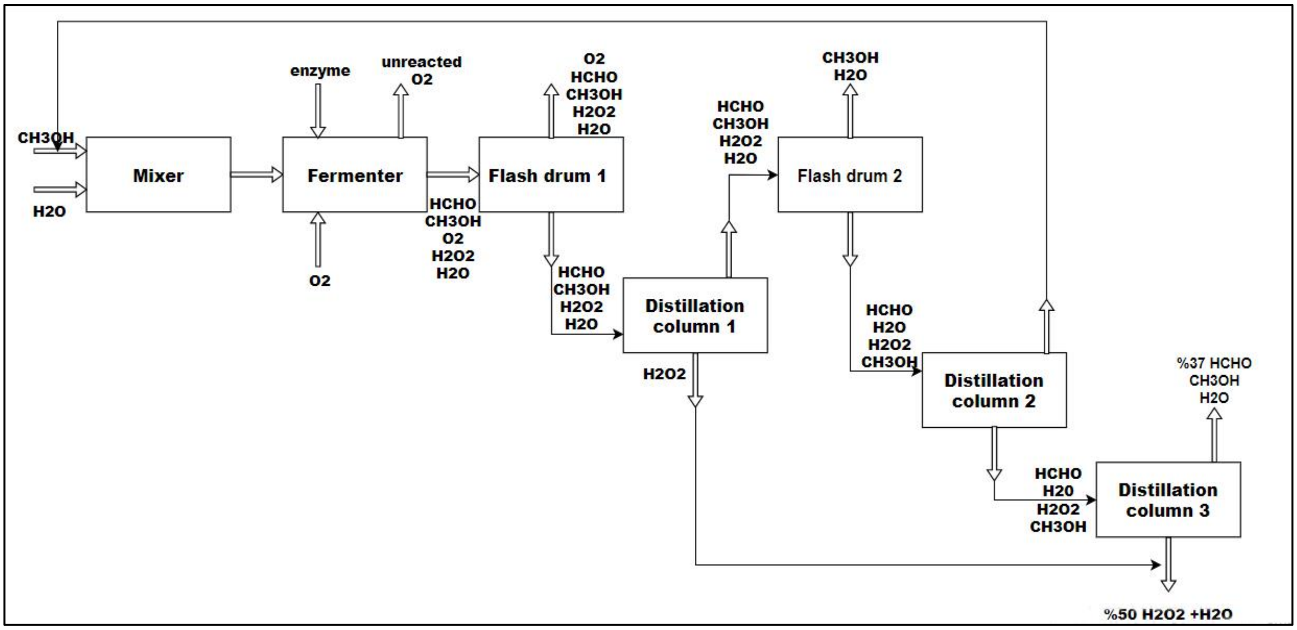


Figure 19:block flow diagram for the enzymatic conversion process of hydrogen peroxide production

4.2 COMPARISON BETWEEN H_2O_2 PRODUCTION METHODS

Table 5:Comparison between H_2O_2 synthesis processes

Process	**Indirect Synthesis Auto-Oxidation Process**	**Direct Synthesis**	**Degussa - Welssenstein Process**	**Munchner Process**	**Enzymatic conversion process**
Principle	Sequential hydrogenation and oxidation of organic molecules. Release of H_2O_2 and recycle of the organic molecules.	$H_2+O_2 \rightarrow H_2O_2$	Electrolyte oxidation of sulfuric acid then hydrolysis the product to form H_2O_2.	Ammonium peroxydisulfate is converted to potassium peroxydisulfate, which is then hydrolyzed.	Enzymatic conversion of methanol in presence of oxygen inside CSTR fermenter to produce formaldehyde and hydrogen peroxide
General Feature	A very famous, well-known, and complex process developed during a very long time.	A very simple process but still lack of maturity	Complex process which describes as a green hydrogen peroxide production.	Complex process	Simple process, and proven to be better than the others
Catalyst	PD in hydrogenation step	PD and Au based catalysis.	Electrochemical catalyst.	–	Methanol oxidase enzyme
Safety	Safe	Can be made safe	Safe	Safe	Safe
On Site Production	Impossible	Possible	Impossible	Impossible	Impossible
Purity	50%-70%	6.6%	45%	20%	50%
Raw Material	H_2, ethyl anthraquinone	H_2, O_2	Sulfuric acid, water	Sulfuric acid, water, ammonium sulfate, potassium sulfate	Methanol oxidase, Oxygen, water
Scale Production	Large	Small	Large	large	Large
Utility	Steam, water	High pressure	Steam	Steam	Steam& R134a and cooling water

There are many processes to produce hydrogen peroxide, these processes were display in table (5) .The most important aspect included in the comparison between those process is the purity of hydrogen peroxide that produce ,scale production as well as catalyst used within production process .As table 5 shows the most process that gives hydrogen peroxide with high purity is auto- oxidation beside enzymatic conversion process. On the other hand, the comparison takes place based on the scale production, it was found that all included process in table 3 used for large production except direct Synthesis.

The most used process for hydrogen peroxide is anthraquinone auto oxidation process but this process required large capital intensive to produce H_2O_2, beside that this process is not safe until some safety procedure provided within the process this will make the process too expensive. To addressing this problem in the anthraquinone process, enzymatic conversion process will use to produce H_2O_2 since this process is much safer comparing with AO process, in addition it contains the following advantages:

1- Less expensive raw material &catalyst.

2- No harmful emission.

3- No side reaction.

4- It has enzyme recovery.

5- High selectivity.

CHAPTER FIVE : PROCESS DESCRIPTION

Enzymatic process is chosen to be the selected hydrogen peroxide synthesis process due to certain comparison points that was display in chapter 4 section 2. The detailed description for its process flow diagram helps to understand the principle of the process in deep.

5.1 PROCESS FLOW DIAGRAM

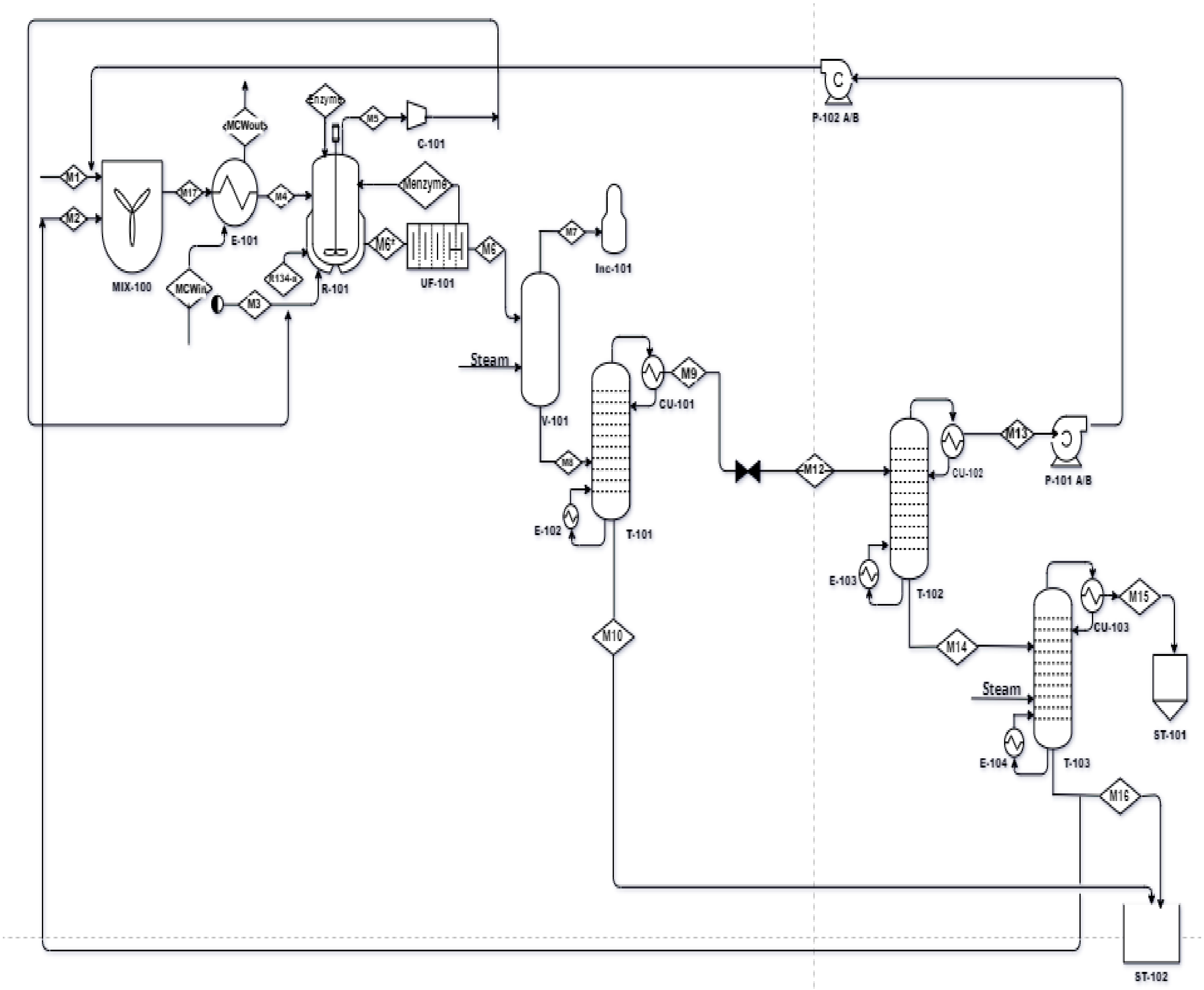


Figure 20:Enzymatic process flow diagram

E-101: Shell and tube heat exchanger	T-101: Multi component distillation column
R-101: Continuous stirred fermenter	V-103: Flash column
C-101: Compressor	T-102: Multi component distillation column
UF-101: Ultrafiltration membrane	T-103: Multi component distillation column
V-101: Flash column	P-101/P-102/P-103/P-104/P105/P-106: Pump
Mix-100: Mixing tank	

Table 6:PFD equipments and their uses

Symbol	Equipment	Use
E-101	Shell and tube heat exchanger	To cool stream M_{17} before entering the reactor
R-101	CSTR Fermenter	To mix M_3 and M_4 to produce formaldehyde and hydrogen peroxide
UF-101	Ultrafiltration membrane	To recover excess outlet enzyme
V-101	Flash separator column	To separate large amounts of oxygen from M_6
P-101,102, …	Pump	To increase the pressure of the streams
V-103	Flash separator column	To separate further negligible amounts of oxygen from M_9
T-101	Distillation column	To separate pure hydrogen peroxide from other components in M_8
T-102	Distillation column	To separate and remove methanol from water
C-101	compressor	To increase (compress) the reactor outlet stream M_5 pressure
-	Incinerator	Furnace, to remove solid wastes from M_7
T-103	Distillation column	To separate formaldehyde from water
-	Storage tanks	To store the remaining chemicals in the streams
E-102, 103, 104	Steam boilers	To provide heat that is needed for the column to perform its function
CU-101, 102, 103	condensers	To cool the vapors of the components of the mixture, allowing them to condense to liquid form to be collected
Mix-101	Mixer	To mix methanol and water streams before entering them to the heat exchanger

Table 7: Temperatures and pressure for each stream in PFD

Stream name	Temperature (ºC)	Pressure (bar)
M_1	49.5	11.9
M_2	203.4	11.49
M_{17}	139	11.15
$Mcw_{(in)}$	20	1
$Mcw_{(out)}$	52	1
M_4	50	11.15
M_3	38.9	11.15
M_5	-22 and goes out at 90.2	8 and compressed to 11.15
R-134a	-28	0.75
M_{6*}	-22	11.15
M_6	-22	11.15
M_7	68	0.35
M_8	68	0.35
M_9	66.6	0.29
M_{10}	84.9	0.35
M_{11}	68.1	0.29
M_{12}	68.1	0.29
M_{13}	64.8	1.22
M_{14}	112.9	1.53
M_{15}	207.4	19.97
M_{16}	237.9	20.36

5.2 FULLY DETAILED DESCRIPTION OF THE PROCESS FLOW DIAGRAM

The use of the enzyme, methanol oxidase, to convert alcohol to formaldehyde and hydrogen peroxide in the presence of oxygen is simply occurred through the following reaction:

$$CH_3OH + O_2 \xrightarrow{Methanol\ Oxidase} HCHO + H_2O_2$$

The preparation of the enzyme, that is used to catalyze the oxidation reaction of methanol, is done by translation and transcription of an appropriate gene, which is Hansenula polymorpha gene, since the preferred alcohol oxidizes enzymes are functionally identical to the alcohol oxidizes derived from the strain of this gene. After the preparation process of the required enzyme, it becomes ready to be introduced into the reaction zone besides oxygen to start the enzymatic conversion reaction of alcohol to aldehyde and hydrogen peroxide.

Referring to the enzymatic reaction formula, it can be observed that the consumption of the dissolved oxygen molecules in the conversion process is stoichiometrically, since as each mole of methanol converted into product, one mole of oxygen consumed from the reaction mixture. During the aeration step of the reaction mixture, it should be avoided to loss the volatile aldehyde product into the environment because of its harmful effects to all living things.

The process flow diagram, figure1 is a representation for the process sequence starting with a feed stock of methanol, water, and oxygen and ending with the storage of the produced formaldehyde and the targeted chemical, hydrogen peroxide with concentration of 50%.

Methanol (fresh and recycled) in stream M_1 at 49.5^oC and 11.9 bar, and water (fresh and recycled) in stream M_2 at 203.4$^oC$and 11.49 bar, are introduced to the Mixer (Mix-101) to get a methanol solution in stream M_{17} at 139^oC and 11.15 bar, then it is required to cool this stream before entering it to the reactor to avoid the thermal shock that may occur since the reaction takes place at -22°C by entering it to the heat exchanger (E-101) and it exits at 50°C and 11.15 bar as stream M_4, this cooling operation is achieved by entering cooling water to the tube side of E-101 at 20°C and leaving it at temperature equal 52°C . After that, both streams M_4 of methanol and M_5 of oxygen (fresh and recycled) are entered a batch stirred tank reactor (R-101) in which the enzyme, methanol oxidase catalysis the oxidation reaction. The obtainable reactor conversion of oxygen approaches 90 %. Methanol oxidase enzyme is used in this reaction at high concentration, 110 ug/ml in the reaction vessel, so that oxygen became the rate limiting of the reaction.

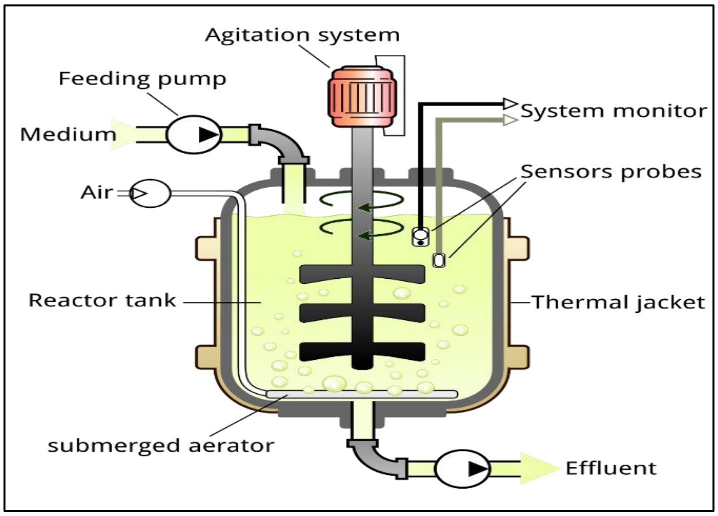


Figure 21: Stirred Bioreactor

The oxidation reaction of methanol to produce formaldehyde and hydrogen peroxide inside the reactor takes place at lower temperatures between 10 to -50°C, depending on the freezing point of alcohol water mixture.

The reaction is designed to be occurred at temperature -22°C, since it is considered as the best temperature for this type of reactions which is because a fact that the enzyme activation is the highest at that temperature. Reaction temperature must be remained at -22°C; this can be achieved by refrigeration loop as shown in the figure below. This loop can be illustrated simply as follows, a cooling jacket is added to the reactor, and a refrigerant R-134a that flows through it is then sent through two compressors in series to reach a pressure of 8 bar. Then through a heat exchanger (E-201), the refrigerant condensed and subcooled to 35°C, and then flashed to pressure of 0.75 bars and temperature of -28°C. Thereafter, the refrigerant is sent back to the cooling jacket and exceeded the same process.

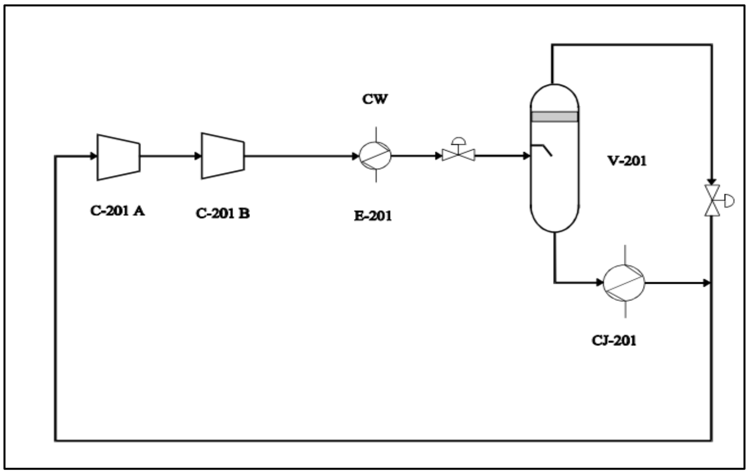


Figure 22:Refrigeration

From the top of the reactor, stream M_5 of unreacted oxygen at pressure 8 bars is introduced to compressor (C-101) to be compressed to 11.15 bars, then it is recycled back and mixed with stream M_3.

The outlet stream of the reactor, Stream M_6^*, is sent to ultrafiltration unit (UF-101), which is composed of hollow fiber membranes that allow for more cost-effective separations and reactions, this unit is purposed to recover the excess outlet enzyme from M_6^* and to get stream M_6 free from enzyme that is negatively affect the desired product. The representation diagram for this membrane is shown in the figure below.

Figure 23:hollow fiber ultrafiltration membrane

Flash separator (V-101) receives the permeate of the ultrafiltration unit, stream M_6, which is composed of formaldehyde, hydrogen peroxide, methanol, oxygen, and water at -22°C and 11.15 bars, V-101 is needed to eliminate most of the oxygen as well as traces of other organic products in stream M_6 by flashing it to separate the stream into two phases, liquid, and vapor phase. The liquid phase stream which is named as M_8 is sent to sieve tray distillation column (T-101). However, from the top of the flash separator, stream M_7 is delivered to unit called incinerator, which is defined as burning furnace that is commonly used to get rid of wastes.

.

Figure 24:Proburn incinerator

Stream M_{10} is the bottom outlet of (T-101), it involves almost pure separated hydrogen peroxide (97%) at 85^{o}C and 0.35 bars. Whilst stream M_9, the top stream of (T-101) is introduced to the second sieve tray distillation column (T-102) which is mainly required to separate methanol from water, the top stream M_{13} of T-102 is mostly contained methanol, and small amounts of formaldehyde is pumped, using two pumps in series (P-101, P-102), up to 11.15 bar to be recycled back to stream M_1. Bottom stream of (T-102) which is M_{14} at 112.9^{o}C and 1.53 bar, that is composed mainly of formaldehyde and water, is fed to third sieve tray distillation column (T-103) at 20 bars using a side stream of steam to ease the separation process because of the similar boiling points between the compounds to be separated. At 207.4^{o}C and 19.97 bars, the top stream M_{15} left the column with solution of about 27% formaldehyde in water and stored inside atmospheric storage tank, ST-101. While the bottom M_{16} is mostly water at 239.7^{o}C and 20.36 bars. Some of this stream (M_{16}) is mixed with the bottom outlet stream M_{10} of (T-101) to get the desired product with its suitable concentration, 50% hydrogen peroxide, and the remain of M_{16} is recycled back to M_2.

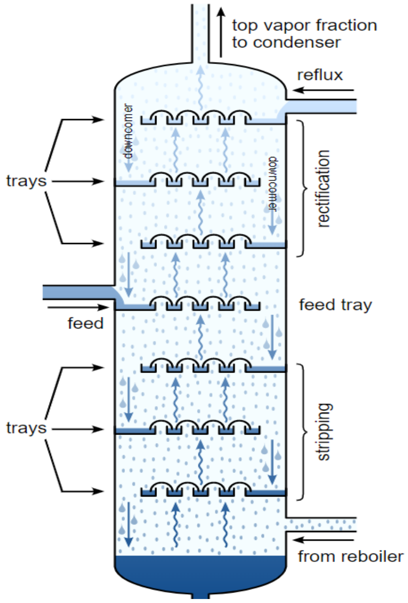


Figure 25: Tray distillation column

CHAPTER SIX

6.1 PLANT LOCATION

In this section, we will choose many locations in Jordan that fit the plant requirement, discussing these locations and make a comparison between them, the choose of location, comparing and selecting one will be in terms of an important factors to be discussed in this chapter.

- **Raw material**

There are many important aspects to such an important matter and those might include the following: material availability and distance from source of supply. In addition, Transportation and storage are also considered as crucial factors.

- First of many steps must be the research for a nearby supplier of the raw material needed for the process in order to save transportation costs as well as saving time and effort.

- Considering the material availability includes raw material prices, distance from source of supply, transportation expenses and storage requirements and therefore are the most important step.[37]

- **Climate**

Caution to weather conditions is required because of humidity, hot, or cold extremes that may cause the initiation of trigger effect on plant equipment.[37]

- **Water supply**

For the purposes of cooling, washing, and generating steam during manufacturing processes in industrial facilities, there must be a source for supplying these facilities with water through rivers, lakes, artesian wells, and sea water.[38]

- **Waste Disposal**

It includes all activities and procedures required to control wastes from its inception until its final disposal, by collecting, transporting, and treating it until disposal, and at the same time monitoring and organizing the waste management process, laws, technologies, and economic mechanisms related to waste.[38]

- **Labor supply**

Since the labor is one of the most important factors needed to operate the plant, its necessary to locate the plant where both skilled and unskilled labors are available.

Mostly, local labor will be chosen to decrease unemployment ratio in Jordan as well as importing labor is usually costly and involves administrative problem [37].

This factor also deals with the following sections:

1. Productivity of labor
2. Pay scale.
3. Restrictions on hours of work per week

- **Utilities**

When selecting a site, make sure that an adequate need of the plant is available such as energy source (steam and electricity) and water supply, with respect to the ease of access to these utilities as well as their cost. [37]

- **Transportation**

There are five basic models in physical transportation which are water (cheapest), railway, highways (the most expensive), pipeline and air transportation.

Keep transportation cost in mind, try to minimize the total transportation cost by decreasing the distance between raw material, plant and customers.[37]

- **Location of consumption market**

It is one of the priorities to be considered in choosing the plant location. Market trend and competition, industrial market, and the scope of export to neighboring countries are things should be kept in mind in market analysis. There are other factors involved such as cost of the product distribution, time required for shipping, and taxes. As a result, the closer the market to the plant, the more effective the shipping costs is.[37]

- **Taxation and legal instructions**

Taxes analysis is critical before deciding where to locate the plant such as the income tax, which is a tax by the federal government, sales tax which is charged at the point of purchase for certain goods, and property tax that is assessed on the property owned. While it could be the way to encourage new projects and investments is to reduce the taxes at the beginning of the project lifetime. Legal restrictions are applied on both exporting and importing.[37]

Three main industrial cities were compared, all these cities are located in Jordan. The criterion that was used in the selection of plant location were shown in the table below:

Table 8 :Site selection criteria

Criterion	Weighing	Compared industrial zones in Jordan					
		(PBI Aqaba industrial area) Aqaba		(Sahab) Amman		(Al-Hassan industrial area) Irbid	
Raw material	10	8	80	7	70	6	60
Transportation	7	6	42	5	35	4	28
Water supply	7	5	35	6	42	6	42
Utilities	8	5	40	6	48	4	32
Climate	5	3	15	4	20	5	25
Labor Supply	7	5	35	6	42	6	42
Taxation and legal Restrictions	5	5	25	4	20	4	20
Waste disposal	6	5	30	4	24	4	24
Market	8	7	56	6	48	5	40
Sum	63	358		349		313	

According to table (7) PBI Aqaba industrial area which locates 15 Km from Aqaba's seaports [39] it was ranked as the best location to start up a hydrogen peroxide plant.

The most important factor to be considered in plant location is the availability of raw material as it illustrates from table 5 Aqaba industrial area and it was ranked to be the most locations that makes raw material available any time since it is mainly methanol and other raw material import from China through Aqaba’s port , as it was mentioned PBI just locates 15 Km from Aqaba's sea as a consequence the distance between raw material and the plant is small which make also the transportation between them so easy and take short time also Aqaba industrial area makes transportation of hydrogen peroxide to possibly middle east market easy and not too costly comparing to other location that was considered. The table shows also that Aqaba industrial area was consider to have the largest point when the market is considered as a part of comparison that’s due to all local company that produce paper as a main product don't manufacture paper from its raw material at which H_2O_2 is used since local company bring paper as a role from China and Pakistan as a consequence middle east market is the main target market of our plants without neglect local market demand. [40]

Figure 26:Site location that chosen in Aqaba, Governorate in Jordan

6.2 PLANT LAYOUT

Plant layout, which is prepared after the process flow diagram and before the piping structural details, is a mechanism involves knowledge of the space requirements for the facilities and their proper arrangement so that continuous and steady movement of the production cycle takes place. Also, it can be an important part in determining construction and manufacturing costs because the aim is safe and profitable design.

There are many factors to choose the distribution of equipment inside the plant, the most important of which is the direction of the wind, the location of the equipment that produces the flammable material must be opposite to the direction of the wind to prevent the occurrence of fires and thus harm workers. In other words, the equipment should be distributed in a way that ensures that it is not damaged, and workers are not harmed.[41]

Wind direction is shown in the following figure which are presents the wind rose chart for Aqaba City.

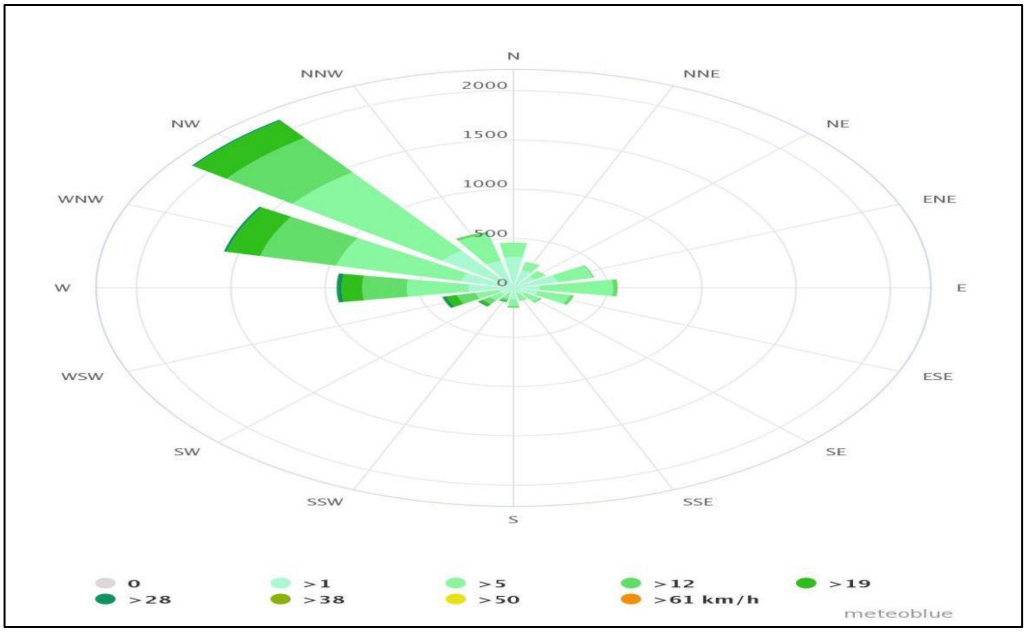


Figure 27:Prevailing wind in Aqapa region

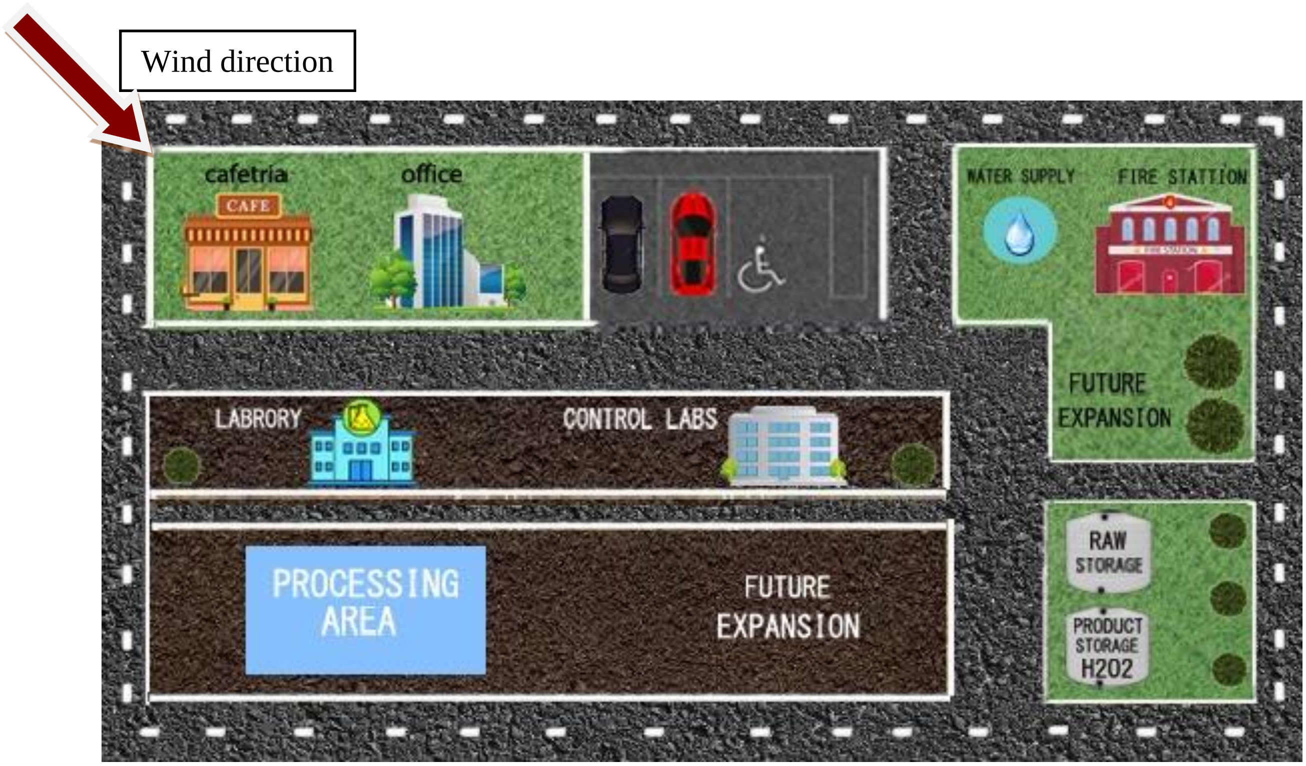


Figure 28:Predicted plant layout for Hydrogen peroxide production plant

CHAPTER SEVEN : MATERIAL BALANCE

7.1 MATERIAL BALANCE ON DISTILLATION COLUMN (T-101):

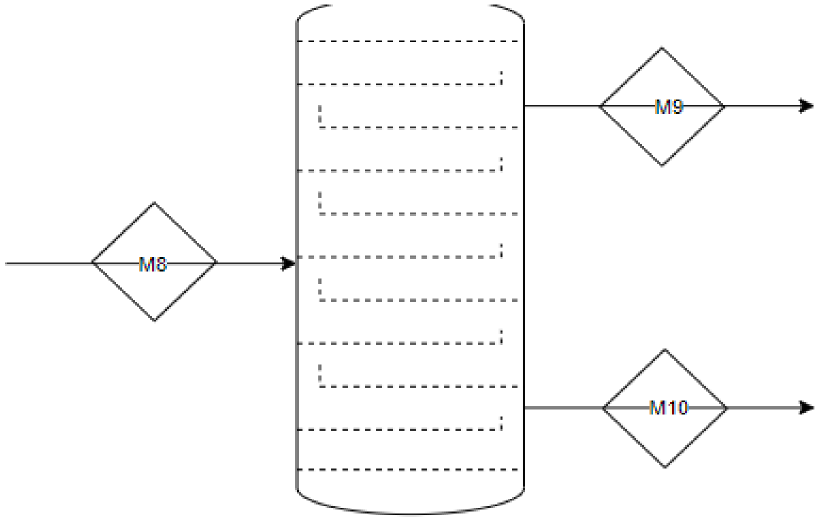


Table 9:Overall and Component Material Balance on Multi-component Distillation Column (T-101)

	Input	Output	
Streams	**M_8**	**M_9**	**M_{10}**
Units	**Kgmole/hr**	**Kgmole/hr**	**Kgmole/hr**
Components	**------**	**------**	**------**
HCHO	**1011.78894**	**997.5257721**	**0**
CH_3OH	**1530.174631**	**1511.402685**	**0**
H_2O	**27668.05558**	**27658.66914**	**0**
H_2O_2	**1011.78894**	**11.78894094**	**1000**
O_2	**0**	**------**	**------**
Total	**31221.80809**	**30179.38653**	**1000**
Total(in/out)	**31221.80809**		**31221.80809**

7.2 MATERIAL BALANCE ON FLASH COLUMN (V-103):

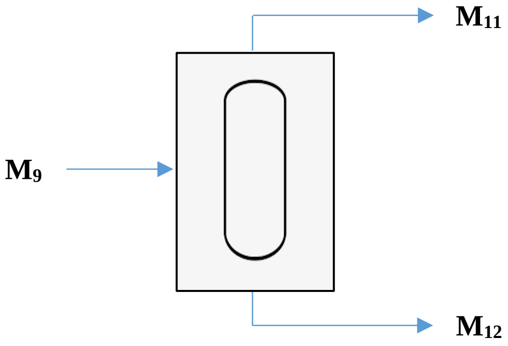


Table 10:Overall and Component Material Balance on Flash Column (V-103)

	Input	**Output**	
Streams	**M_9**	**M_{11}**	**M_{12}**
Units	**Kgmole/hr**	**Kgmole/hr**	**Kgmole/hr**
Components	-------	-------	------
HCHO	**997.5257721**	-------	**997.5257721**
CH_3OH	**1511.402685**	**0.00512**	**1511.402685**
H_2O	**27658.66914**	**0.01536**	**27658.66914**
H_2O_2	**11.78894094**	-------	**117.8894094**
O_2	-------	-------	------
Total	**30179.38653**	**0.02048**	**30285.487**
Total(in/out)	**30179.40701**		**30179.38653**

7.3 MATERIAL BALANCE ON DISTILLATION COLUMN (T-102):

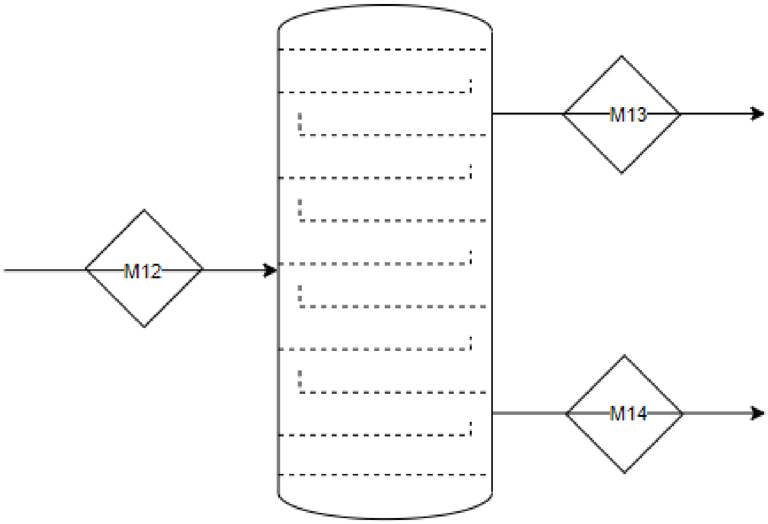


Table 11:Overall and Component Material Balance on Multi-component Distillation Column(T-102)

	Input	Output	
Streams	**M_{12}**	**M_{13}**	**M_{14}**
Units	**Kgmole/hr**	**Kgmole/hr**	**Kgmole/hr**
Components	------	------	------
HCHO	1009.616994	0.041153854	1009.575838
CH_3OH	1511.402685	1325.869818	127.974402
H_2O	27658.66914	458.060288	27272.767
H_2O_2	117.8894094	------	11.3755024
O_2	------	------	------
Total	30297.57822	1783.97126	28421.69275
Total(in/out)	30297.57822	------	30297.57822

7.4 MATERIAL BALANCE ON DISTILLATION COLUMN (T-103):

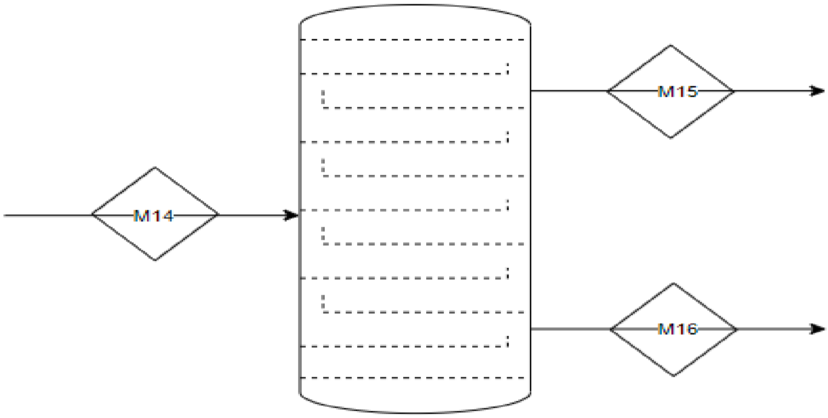


Table 12:Overall and Component Material Balance on Multi-component Distillation Column(T-103)

	Input	Output	
Streams	**M_{14}**	**M_{15}**	**M_{16}**
Units	**Kgmole/hr**	**Kgmole/hr**	**Kgmole/hr**
Components	------	------	------
HCHO	1009.575838	1011.242742	------
CH_3OH	127.974402	128.772484	------
H_2O	27284.14251	2645.138318	24639.00434
H_2O_2	11.3755024	------	11.3396118
O_2	------	------	------
Total	28433.06825	3785.153544	24650.34395
Total(in/out)	28433.06825		28433.06825

7.5 MATERIAL BALANCE ON FLASH COLUMN (V-101):

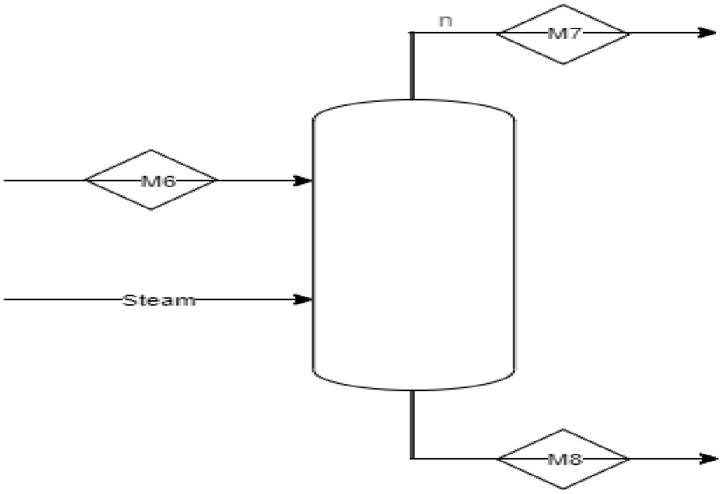


Table 13:Overall and Component Material Balance on Flash Column (V-101)

	Input	Output	
Streams	**M_6**	**M_7**	**M_8**
Units	**Kgmole/hr**	**Kgmole/hr**	**Kgmole/hr**
Components	**------**	**------**	**------**
HCHO	**1000.11744**	**1.02472**	**999.2977184**
CH_3OH	**1531.42983**	**4.09888**	**1530.174631**
H_2O	**27681.37552**	**13.32136**	**27668.05558**
H_2O_2	**1000.11744**	**1.02472**	**999.2977184**
O_2	**6.250734**	**6.14832**	**0**
Total	**31219.29096**	**25.618**	**31196.82565**
Total(in/out)	**31219.29096**		**31219.29096**

7.6 MATERIAL BALANCE ON FERMENTER (R-101):

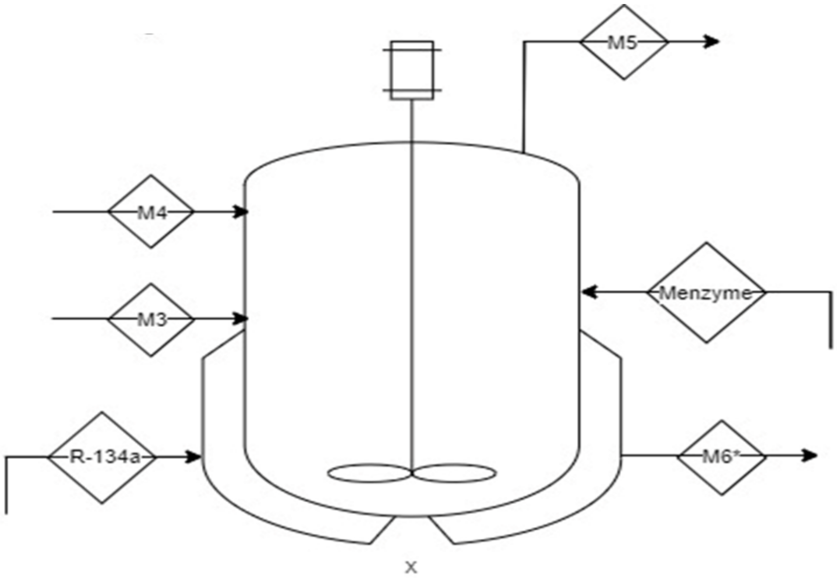


Table 14:Overall and Component Material Balance on Fermenter (R-101)

	Input		**Output**	
Streams	**M_3**	**M_4**	**M_5**	**M_6**
Units	**Kgmole/hr**	**Kgmole/hr**	**Kgmole/hr**	**Kgmole/hr**
Components	**-------**	**-------**	**-------**	**-------**
HCHO	**0**	**-------**	**-------**	**1012.618908**
CH_3OH	**0**	**1774.707732**	**3.4879123**	**1531.42983**
H_2O	**0**	**19331.6378**	**-------**	**27659.49795**
H_2O_2	**0**	**7.60589028**	**-------**	**1012.618908**
O_2	**11251.33**	**-------**	**1120.632468**	**6.250734**
Total	**11251.33**	**21113.95142**	**1124.12038**	**31222.41633**
Total(in/out)	**21113.95142**		**21113.95142**	

7.7 MATERIAL BALANCE ON HEAT EXCHANGER:

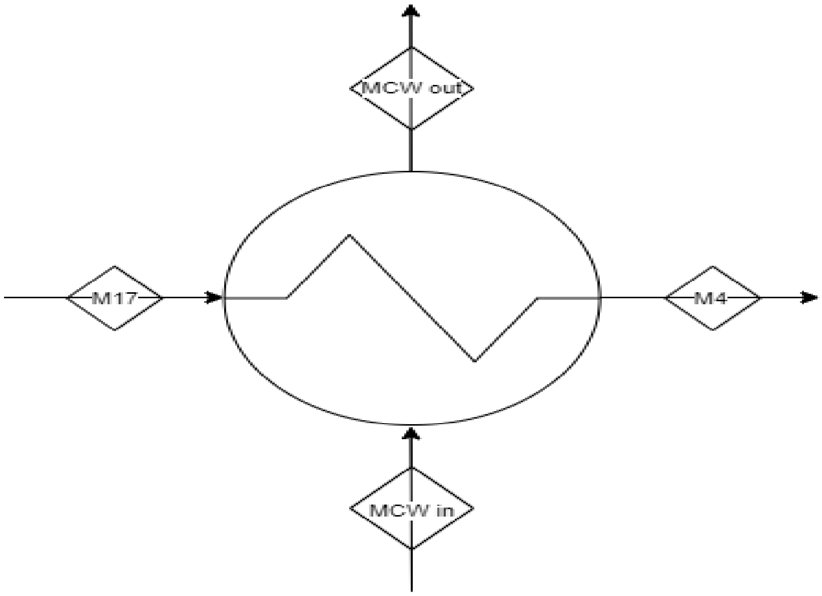


Table 15:Overall and Component Material Balance on heat exchanger

	Input	Output
Streams	**M_{17}**	**M_4**
Units	**Kgmole/hr**	**Kgmole/hr**
Components	**------**	**------**
CH_3OH	**1859.76**	**1859.76**
H_2O	**20258.1823**	**20258.1823**
H_2O_2	**7.970432**	**7.970432**

M_{cw} in = M_{cw} out = 30472.0113 Kmol/hr

7.8 MATERIAL BALANCE ON MIXER (M-101)

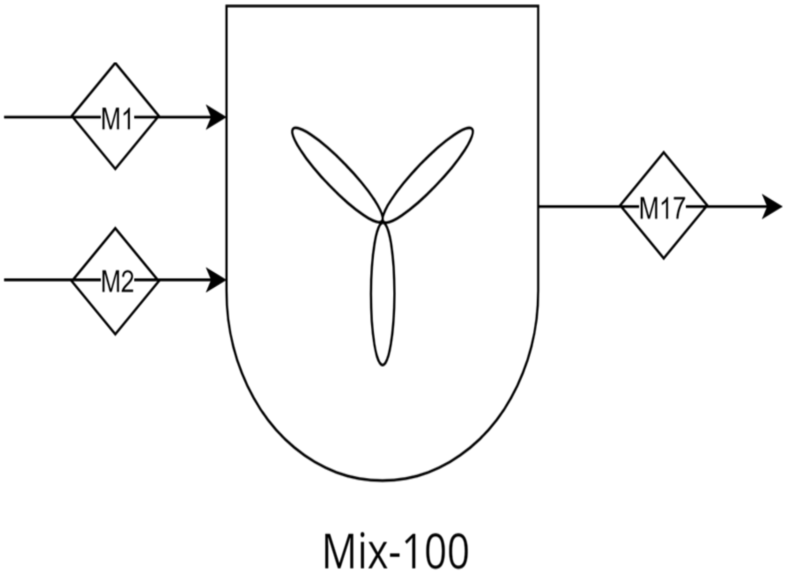


Table 16 : Overall and Component Material Balance on mixer

Streams	**M_1**	**M_2**	**M_{17}**
Units	**Kg mol/hr**	**Kg mol/hr**	**Kg mol/hr**
Components	**-**	**-**	**-**
CH_3OH	**1859.76**	**------**	**1868.6235**
H_2O	**354.241**	**19916.117**	**20263.2745**
H_2O_2	**------**	**7.9704324**	**7.9704324**

7.9MATERIAL BALANCE ON ULTRAFILTRATION MEMBRANE (UF-101)

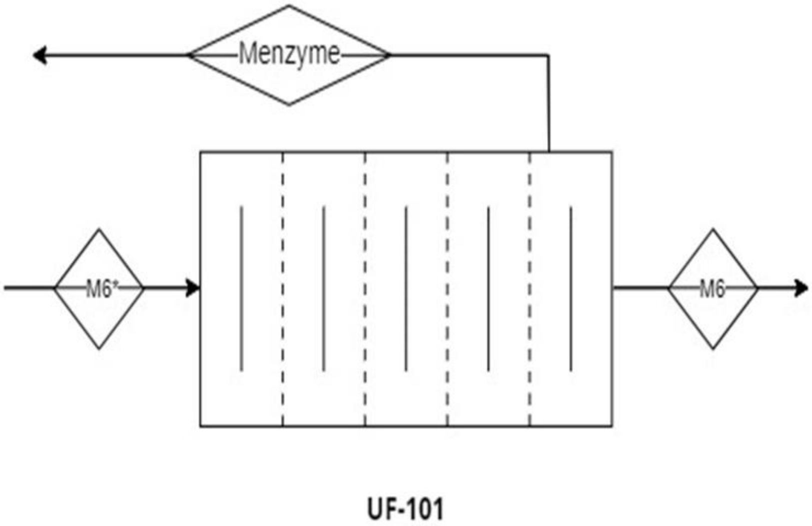


Table 17:Overall and Component Material Balance on ultrafiltration membrane

Streams	**M_6***	**M_6**
Units	**Kg mol/hr**	**Kg mol/hr**
Components	**----**	**-----**
HCHO	**1013.734016**	**1014.319**
CH_3OH	**1533.116258**	**1544.963**
H_2O	**27689.95691**	**27673.25**
H_2O_2	**1013.734016**	**1014.057**
O_2	**6.25761738**	**6.397626**
Enzyme	**34.41689559**	**------**

7.10 OVERALL BALANCE ON PLANT:

7.10.1 TOTAL MASS FLOW INPUT:

Table 18:Total mass flow input

Streams	Total Mole Flow (Kg mole/hr.)
M_3	10126.20
M_4	22140.09
Total	32266.3

7.10.2 TOTAL MASS FLOW OUTPUT:

Table 19:Total mass flow output

Streams	Total Mole Flow (Kg mole/hr.)
M_5	1012.62
M_7	25.618
M_{10}	1000
M_{11}	0.02048
M_{13}	1789.298
M_{15}	3787.426
M_{16}	24651.33
Total	32266.3

CHAPTER EIGHT : ERNERGY BALANCE

8.1 ENERGY BALANCE ON MIXER (M-101)

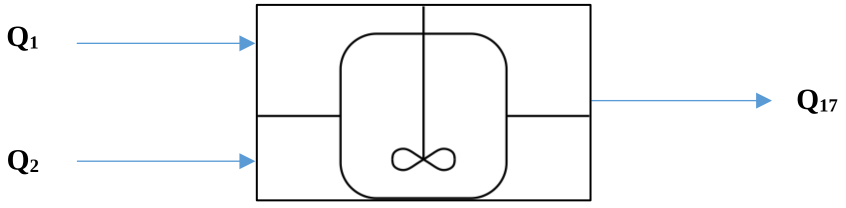


Table 20:Calculated values after applying energy balance on mixer(M-101)

Stream	Value (KJ/hr)
Q_1	5094554.209
Q_2	422771691.6
Q_{17}	215622473.4

8.2 ENERGY BALANCE ON HEAT EXCHANGER (E-101):

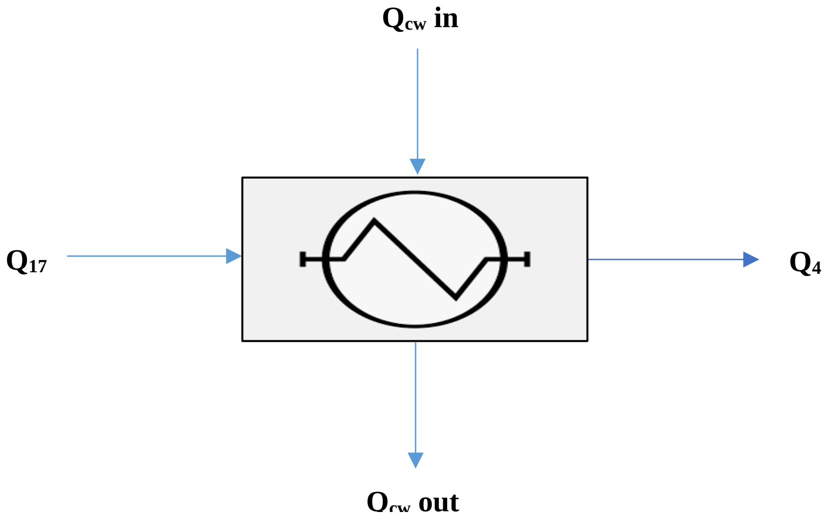


Table 21:Calculated values after applying energy balance on heat exchanger

Stream	Value (KJ/hr)
Q_4	42241299.85
Q_{cw}	173381173.6

8.3 ENERGY BALANCE IN THE FERMENTER (R-101)

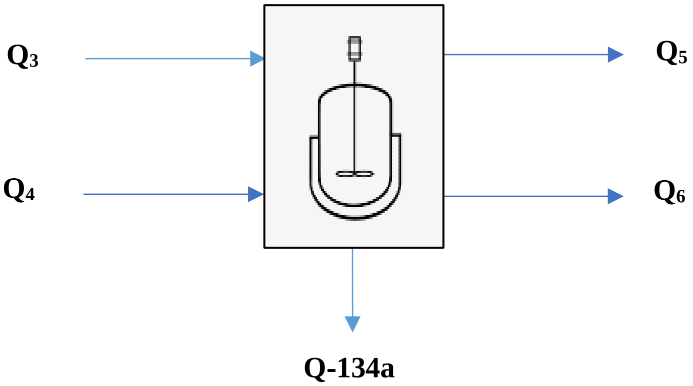


Table 22::Calculated values after applying energy balance on fermenter

Stream	Value (KJ/hr)
Q_3	38.366
Q_5	3.81116
Q_6	- 8704449.818
QR-134a	50945729.15

8.4 ENERGY BALANCE AROUND FLASH COLUMN (V-101)

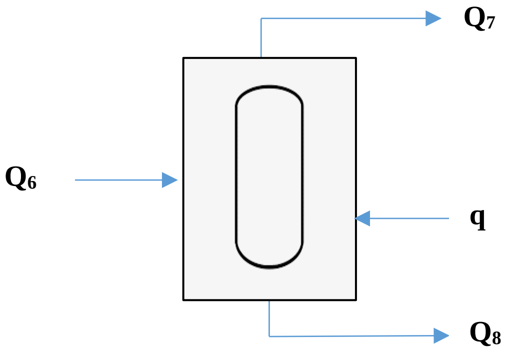


Table 23:Calculated values after applying energy balance on flash column(V-101)

Stream	Value (KJ/hr)
Q_7	71253.66756
Q_8	437800075.3
q	-446575778.8

8.5 ENERGY BALANCE AROUND DISTILLATION COLUMN (T-101)

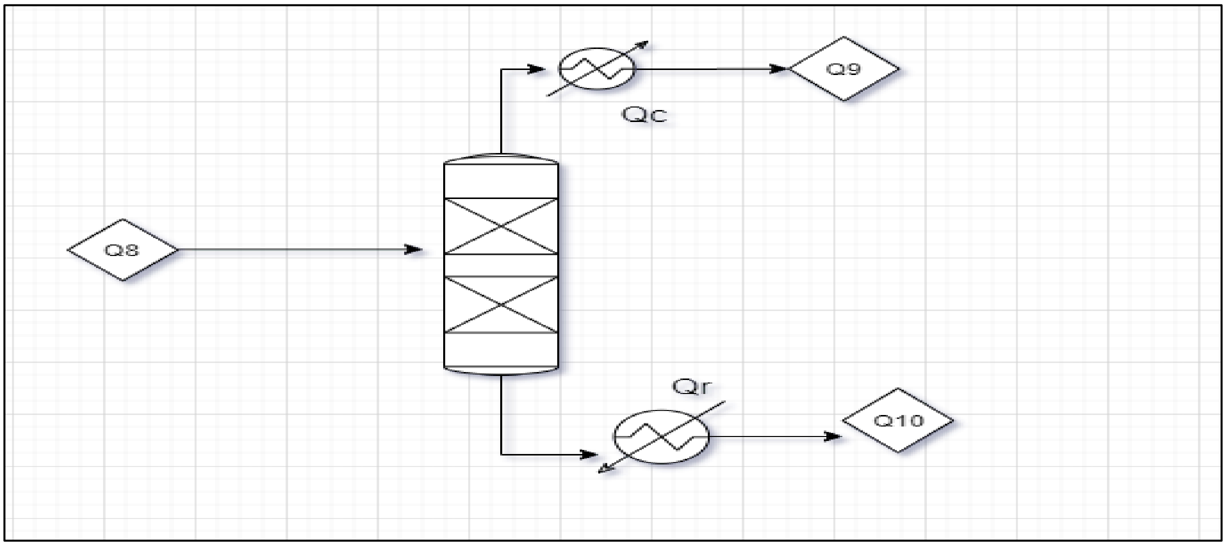


Table 24:Calculated values after applying energy balance on distillation column(T-101)

Stream	Value (KJ/hr)
Q_9	**92615918.17 kJ/hr**
Q_{10}	**2965649 kJ/hr**
Q_C	**1446.440 KJ/hr**
Qr	**- 342219954.6 KJ/hr**

8.6 ENERGY BALANCE ON FLASH COLUMN (V-103) :

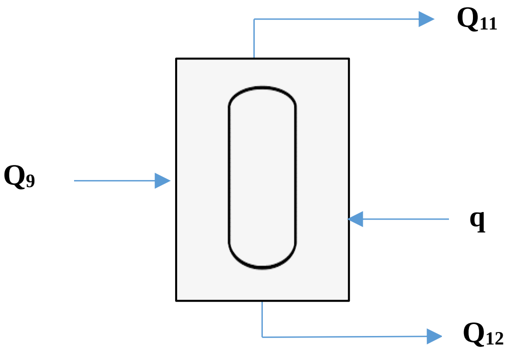


Table 25:Calculated values after applying energy balance on flash column(V-103)

Stream	Value (KJ/hr)
Q_{11}	70.28484987
Q_{12}	97612406.88 kJ/hr
q	-4996558.995 kJ/hr

8.7 ENERGY BALANCE AROUND DISTILLATION COLUMN (T-102)

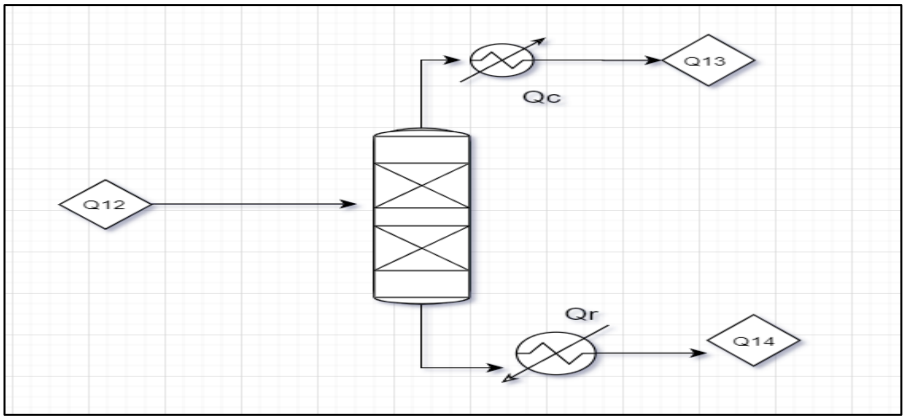


Table 26::Calculated values after applying energy balance on distillation column(T-102)

Stream	Value (KJ/hr)
Q_{13}	6003539.443
Q_{14}	185113253.8
Q_C	149.5
Qr	93504386.32KJ/hr

8.8 ENERGY BALANCE AROUND DISTILLATION COLUMN (T-103)

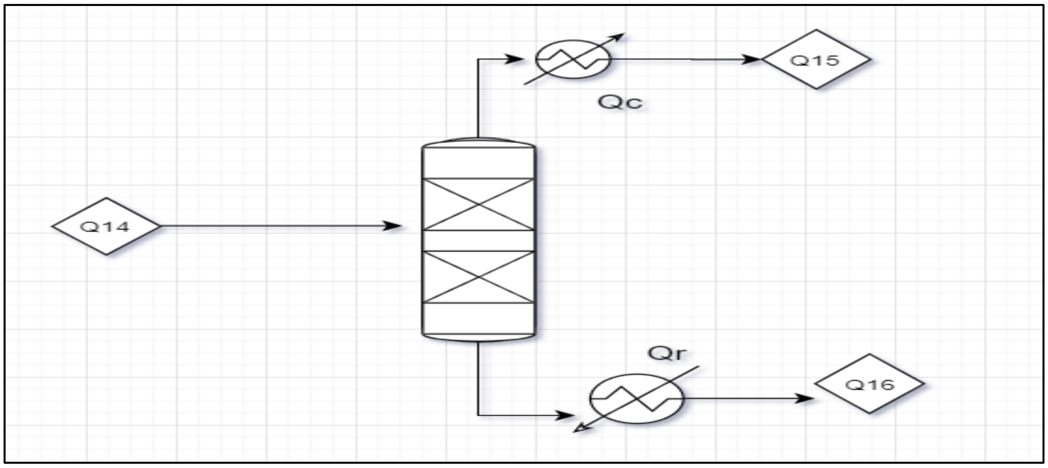
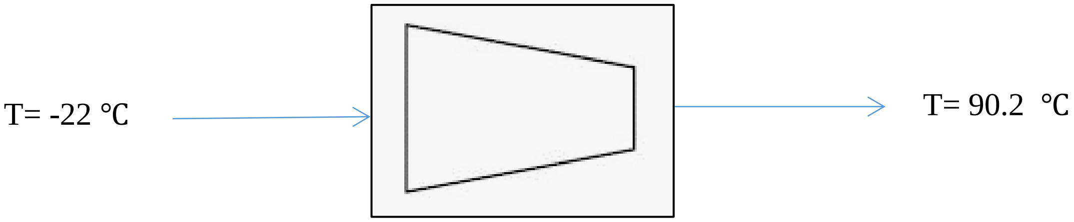


Table 27:Calculated values after applying energy balance on distillation column(T-103)

Stream	**Value (KJ/hr)**
Q_{15}	**51075888.55**
Q_{16}	**173588708.1 kJ/hr**
Q_C	**299.5 KJ/hr**
Qr	**90626931.9 KJ/hr**

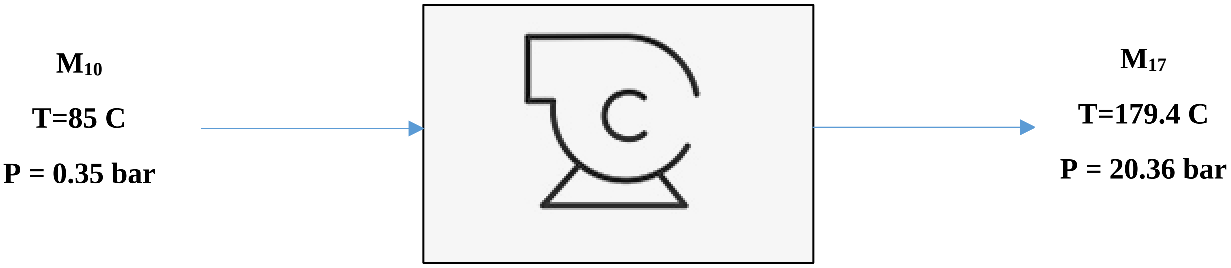
8.9 ENERGY BALANCE ON THE COMPRESSOR (C-101):



By applying energy balance around compressor C-101,it is found that :

Ws = - 3870841.081 KJ/hr

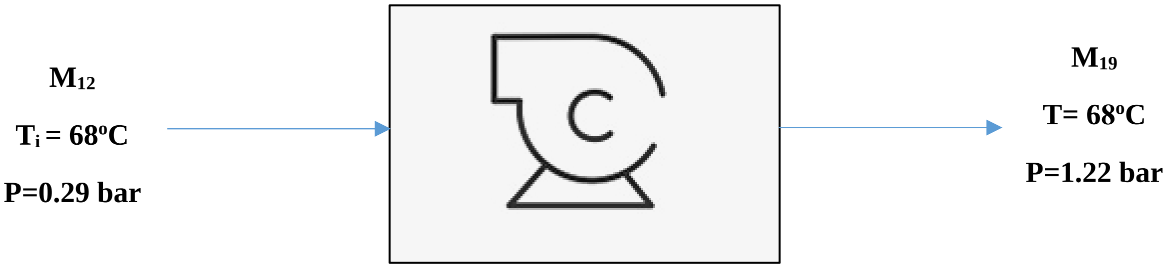
8.10 ENERGY BALANCE ON PUMP (P-101):



By applying energy balance around pump(P-101),it is found that :

W_s = 4580288 KJ/hr = 1272.3kw

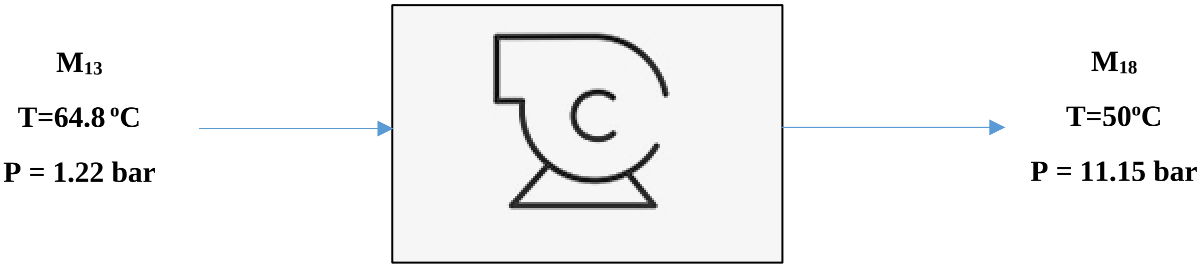
8.11 ENERGY BALANCE ON PUMP (P-102):



By applying energy balance around pump(P-102),it is found that :

Ws =3206.5 kJ/s =3206 kw

8.12 ENERGY BALANCE ON PUMP (P-103)



By applying energy balance around pump(P-103),it is found that :

W_s = -2349846.944 KJ/hr = -652.7kw

CHAPTER NINE: EQUIPMENT'S DESIGN

9.1 MIXER 101 DESIGN

9.1.1 INTRODUCTION

Mixing tanks are containers used to blend different substances together to get a uniform mixture of them, in this process stirred mixer with agitator is used to get a complete mixed liquid (methanol -water solution). According to the fact that methanol is considered low corrosive to carbon steel below 50% solution, the chosen material of construction is carbon steel. The preferred shape of the mixing tank in this process is cylindrical since it tends to have stronger design but will typically require baffles while the head shape is tori spherical head (dished ends) since the pressure in the tank is lower than 15 bar, (Peters, 2004). Several orientations are existed for mixing tank but it's selected to be vertical mounted on concrete foundations , as the Liquids in quantities > 10,000 gal (38 m^3) (Peters, 2004). Impeller within mixing tank can be classified as: Radial impellers, Axial impellers and Close-clearance impeller since viscosity of liquid equal 0.511 centipoise, which consider very low viscosity, the type of impeller chosen is a radial flat blade turbine which is used with low and moderate viscosity liquids in baffled tanks.

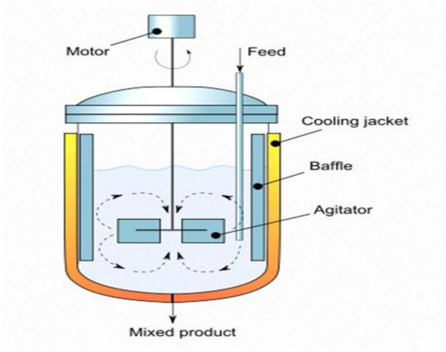


Figure 29 Figure 2:Mixing tank [62]

9.1.2 SAMPLE CALCULATON FOR MIXER DESIGN

1- Volume of tank:

The design is based on 3 mixing tanks needed for 24 hours:

$$\text{Amount of liquid to be mixed} = \frac{59549.9\ Kg}{hr} * \frac{24\ hr}{3\ mixer}$$

$$\text{Amount of liquid to be mixed} = 476399.2\ \text{Kg/mixer}$$

Average Density of mixture to be mixed at 25 ºC temperature and 1 atm = 926.766 Kg/m^3

$$\text{Volume of liquid to be mixed } = \frac{\text{Mass}}{\text{Avg density}} = \frac{476399.2\ kg}{926.766\ Kg/m^3} = 514.045\ m^3$$

Vessels above 500 gallons (1.9 m^3) are never filled more than 90% of their volume (Peters, 2004)

$$\text{Tank Volume} = \frac{514.045}{0.9} = 571.16\ m^3$$

Tank dimensions:

$$Assume\ Length = 1.5 * Diameter$$

$$Volume = \frac{\pi * D^2 * L}{4} = \frac{\pi * D^2 * (1.5 * D)}{4} = \frac{1.5\pi}{4} * D^3$$

$$D = \sqrt[3]{\frac{4 * Volume}{1.5\pi}} = \sqrt[3]{\frac{4 * 571.16}{1.5\pi}} = 7.86\ m$$

$$L = 1.5 * D = 1.5 * 7.86$$

$$L = 11.78\ m$$

2- Thickness calculation:

Cylindrical shell thickness can be found using this equation below (Peters, 2004)

$$t = \frac{P\,ri}{S\,Ej - 0.6\,P} + Cc \ldots (1)$$

t: shell thickness (m)

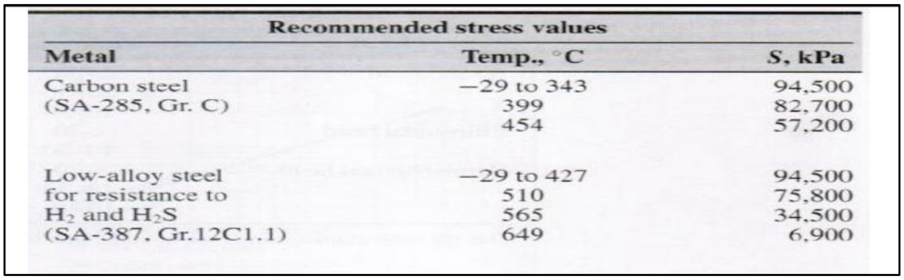
r_i: Inside radius of shell, before corrosion allowance is added (m) = 3.93m

P: Maximum allowable internal pressure (KPa), (gauge) = 101.325 KPa

E j: Efficiency of joints expressed as a fraction = 0.85 for spot examined weld (From table 12- 10) (Peters, 2004). [63]

Cc: Allowance for corrosion (m) = 0.003 m [63], a corrosion allowance of about 3 mm for a 10-year life is a reasonable value. (Peters, 2004)

S: Maximum allowable working stress (KPa) = 94500 KPa for carbon steel (From table 12-10) (Peters, 2004). [63]



Recommended stress values

Metal	Temp., °C	S, kPa
Carbon steel (SA-285, Gr. C)	−29 to 343 399 454	94,500 82,700 57,200
Low-alloy steel for resistance to H_2 and H_2S (SA-387, Gr.12C1.1)	−29 to 427 510 565 649	94,500 75,800 34,500 6,900

Figure 30: Recommended stress values for some metals

$$t = \frac{101.325 * 3.93}{94500 * 0.85 - 0.6 * 101.325} + 0.003 \ldots (2)$$

$$t = 0.0079\,m = 0.79\,cm$$

3- Head thickness calculation:

- Thickness of the head can be found using the equation below (Peters, 2004):

$$t = \frac{0.885\ P\ La}{S\ Ej - 0.1\ P} + Cc \ldots (3) \quad , La = ri = 3.93$$

$$t = \frac{0.885 * 101.325 * 3.93}{94500 * 0.85 - 0.1 * 101.325} + 0.003$$

$$t = 0.0074\ m = 0.74\ cm$$

4- Agitation

The design is based on 3 agitations needed for 3 tanks.

Total density inside the tank = 926.766 kg/m^3

The viscosity is 0. 000511 Pa.s = 0.511 centipoise

5- Ratio constant for flat blade turbine design

- **Assumptions:**

- The tank is fully baffled to prevent vortex formation. Baffles are always used in turbulent flow systems (low viscosity fluids).
- Number of baffles: 4 (fully baffled)
- Number of impeller blades: 6
- Speed of flat blade turbine: 10-150 rpm, assume 90 rpm

Table 28:Results of mixing tank agitation

Da	**E**	**W**	**J**	**L**	**Kt**
Dt/3	Dt/3	Da/5	Dt/12	Da/4	1.6
2.62	2.62	0.524	0.655	0.655	1.6

D_t: Tank diameter (m)

D_a: Impeller diameter (m)

E: Height of impeller above vessel Flore

J: Baffled width

L: Length of impeller blades

W: Width of impeller blades

K_t: flat blade constant, for flat blade turbine with 6 blades, the value of K_t is 1.6

6- Impeller tip speed

$$\boldsymbol{V_{tip} = \pi n D_a} \ldots (4)$$

V_{tip} : Impeller tip speed (m/hr)

n: Speed of flat blade turbine = 90 rpm = 90/60 rps = 1.5 rps

Da: Impeller diameter (m)

$$V_{tip} = \pi * 1.5 * 2.62$$

$$\boldsymbol{V_{tip} = 12.34\ m/s}$$

7- Mixing and circulation time [91]

Circulation time: the average time required for a fluid element to complete one circulation around the tank, in other words , it is a measure of how long it takes the impeller to pump the same volume (V) of liquid as that contained in the tank. And it was measured by:

$$t_{circ} = \frac{V}{q_T} \ldots (5)$$, where:

V is the volume of liquid in the tank

q_T is the total circulation rate (m^3/s) , and for a standard radial flat blade (six-blade disk turbine), it is found from the below equation:

$$q_T = 0.92\, nD_a^3 \frac{D_t}{D_a} \ldots .(6)$$

$$q_T = 0.92 * 1.5\, rps * 2.62^3 m^3 * \frac{D_t}{D_t/3}$$

$$q_T = 74.4\, m^3/s$$

$$t_{circ} = \frac{514.045\, m^3}{74.4\, m^3/s} = 6.9\, s$$

Mixing time: It is the time to reach complete mixing (99%) , and it is achieved if the contents of the tank are circulated about 5 times , so:

$$\text{Mixing Time } (t_T) = 5 * t_{circ} \ldots (7) \Rightarrow t_T = 5 * 6.9\, s$$

$$\text{Mixing Time } (t_T) = \mathbf{34.55\ s = 0.57\ min}$$

8- Power dissipation by an impeller

$$P = K_t * n^3 * D_a^5 * \rho \ldots (8)$$

P: Power dissipation by an impeller (w)

$$\text{P} = 1.6 * 1.5^3 * 2.62^5 * 926.766$$

$$\mathbf{P = 616.72\ KW}$$

Efficiency = 0.8(the actual power should not be over 85% of the calculated power requirement) [64]

Actual power = P / 0.8 = 714.6/0.8 = 893.25 KW

9- Gradient velocity:

$$G = \sqrt{\frac{P}{V * \mu}} \ldots . (9)$$

G: Gradient velocity (s^{-1})

P: Power dissipation by an impeller (kW)

V: volume (m^3)

μ: viscosity (Pa.s)

$$G = \sqrt{\frac{893.25 * 1000}{571.16 * 0.000511}}$$

$$G = 1749.43 \text{ s}^{-1}$$

9.1.3 DETAILED DESIGN SUMMARY

Table 29 mixing tank specification sheet

Identification	
Name of Equipment	Mixer
Item No.	Mix-100
Equipment Type	Cylindrical mixing tank
Number of units required	3 units
Function	To mix methanol and water streams before entering them to the heat exchanger
Material property	
Materials handled	Water , Methanol , traces of hydrogen peroxide
Flow rate (Kg/hr)	59549.9
Operating temperature (℃)	25
Operating pressure (atm)	1
Density (kg/m^3)	926.766
Design specifications	
Material of construction	Carbon Steel
Diameter (m)	7.86
Length (m)	11.79
Shell Type	Cylindrical shell
Orientation	Vertical
Shell thickness (cm)	0.79
Head Type	Tori spherical head
head thickness (cm)	0.74
Cost $	1,559,376

Table 30: Mixing tank agitation specification sheet

Agitation	
Number of units required	3
Type of impeller	Flat blade turbine
Number of impellers	1
Number of baffles	4
Number of blades	6
Impeller diameter (D_a) (m)	2.62
Baffled width(J) (m)	0.65
Height of impeller above vessel Flore(E) (m)	2.62
Width of impeller blades (W)(m)	0.52
Length of impeller blades (L)(m)	0.65
Speed of impeller (rpm)	90
Tip velocity (m/s)	12.34
Mixing time (min)	0.57
Power (KW)	893.25
Gradient velocity (s^{-1})	1749.43

9.2 HEAT EXCHANGER DESIGN (E-101)

9.2.1 INTRODUCTION

Heat exchangers are devices that used to transfer heat from one medium to another, this medium is either gas, liquid or a mixture of them, the medium may be directly connected to each other or separated by a solid wall that prevents their mixing. Heat exchangers improve the efficiency of the system by transferring heat from systems that do not need, to systems that can be used in a useful and good way.

There are several applications of heat exchangers, which are: heating a coolant using heat from a hotter liquid and vise verse, boiling a liquid using hotter fluid heat, boiling a liquid while condensing a hotter gaseous liquid and condensation of a gaseous liquid with a coolant. [65]

9.2.2 HEAT EXCHANGER TYPE SELECTION: [66]

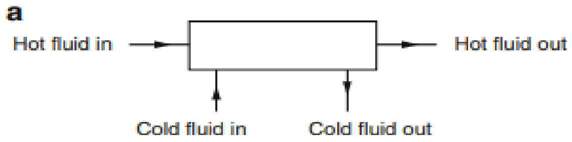
There are four types of heat exchangers, which are:

1. The double tube heat exchanger or what is known as the tube inside the tube structure, where the first fluid flows through the inner tube while the second fluid flows around the first fluid in the outer tube, this type is characterized by its reasonable price in addition to its size that makes it ideal for narrow spaces as it increases flexibility manufacturing process planning.
2. Shell and tube heat exchangers which is considered the most versatile among the rest of the types, as it is designed with a number of tubes placed inside a cylindrical shell, this type is used when there is a need to cool or heat a large amount of liquids or gases, and its design is characterized by allowing a wide range of temperatures, in addition to that, it is easy to be cracked, which makes cleaning and repair also easy.
3. Tube in tube heat exchanger which consists of two tubes, one for each liquid, where these tubes are coiled together to form an outer and an inner pattern. The applications of this type revolve around high temperatures and high pressures. In addition to that, its production is higher with greater efficiency.
4. Plate heat exchangers: Metal plates are used in this type to transfer heat between two liquids. The plate is a metal shell with spaces inside each plate that act as hallways for liquids to flow through. This type is characterized by a high rate of heat transfer compared to others, due to the presence of a larger surface area in contact with liquids, the cost of this type is large while the efficiency is high, so it is preferred to be used in places such as power plants because of its durability and low repair rates.

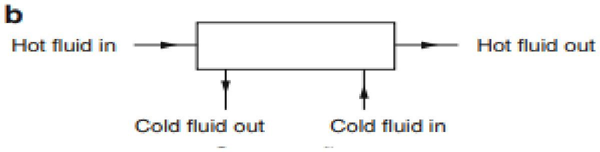
Because of the **high flow rate and temperature of the streams fed to the heat exchanger, and as it is discussed previously,** the most suitable type for this operation is **shell and tube heat exchanger** that designed in accordance with TEMA (Tubular Exchanger Manufacturing Association) specifications.

9.2.3. FLOW PATH CONFIGURATION TYPE SELECTION:[67]

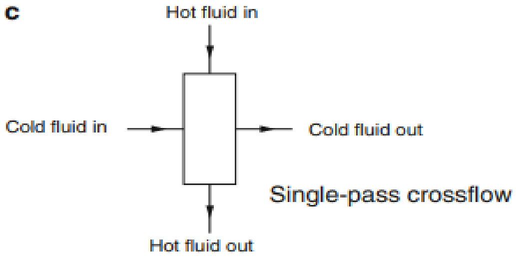
a. Concurrent, or parallel flow: In this type two fluid streams enter together at one end, flow in the same direction, and leave together at the other end.



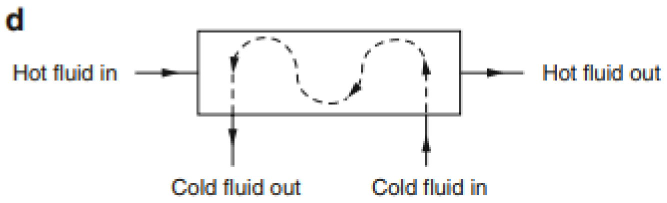
b. Countercurrent, or counter-flow: In this type, the two working fluids flow in opposite directions.



c. Single-pass crossflow: One fluid moves through the heat transfer matrix at right angles to the flow path of the other fluid.



d. Multipass crossflow: One fluid stream is transmitted back and forth through the flow path of the other fluid stream, and this gives a crossflow approximation.



Multipass crossflow

The best design for a shell-and-tube heat exchanger is a counter current configuration in which the heat transfer between the fluid is the maximum, and the efficiency is higher than others, since this type creates a more uniform temperature difference between the fluids over the entire length of the fluid path. [68,69]

9.2.4 TUBE AND SHELL SIDE FLUID SELECTION [70]

There are several factors that must be taken into account when deciding where to place the fluid in shell and tube heat exchanger design, the first of which is the fouling factor, where the fluid that more likely to foul must be placed on the tube side, since the cleaning of tubes is easier than the shell. Another factor is corrosion where the most corrosive fluid must be placed inside the tube side, which is easy to replace at a low price in the case of corrosion. In addition to that, the temperature change in the shell side is greater than in tubes. So, for these reasons, **water** in our process was placed in the **tube side** while the **reactor feed mixture** was in the **shell side.**

9.2.5 SHELL AND TUBE PASSES ARRANGEMENT [71,72]

The tubes are arranged within the shell and tube heat exchanger in passes. These passes can be one, two, four, six or eight. The number of passes can be defined as the number of times the fluid passes through the fluid in the shell. In a single pass heat exchanger, the fluid goes through the shell once. As the number of passes increases, the heat transfer coefficient increases. The figure below shows the passes for the tube side.

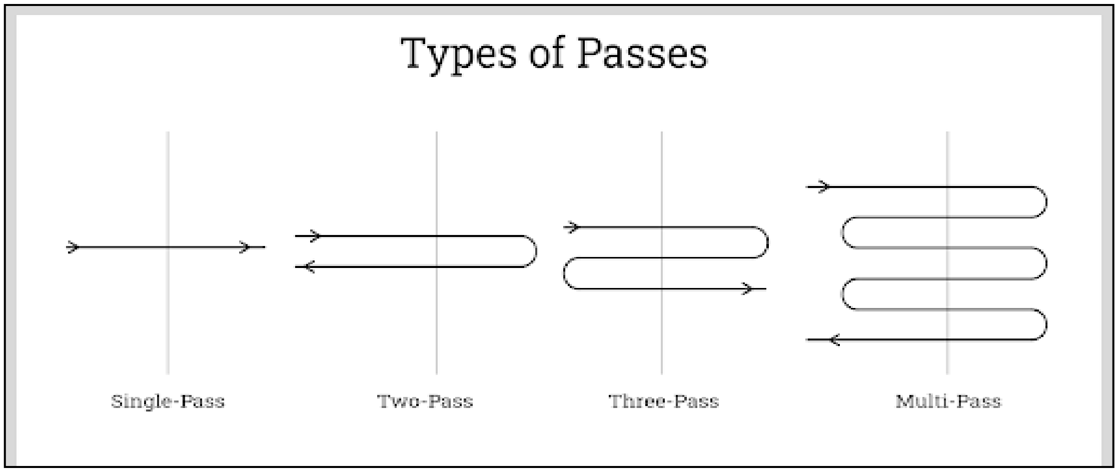


Figure 31: Types of passes in the tube side of heat exchanger

For shell side, the passes of it are one, two and three shell passes as shown in the figure below:

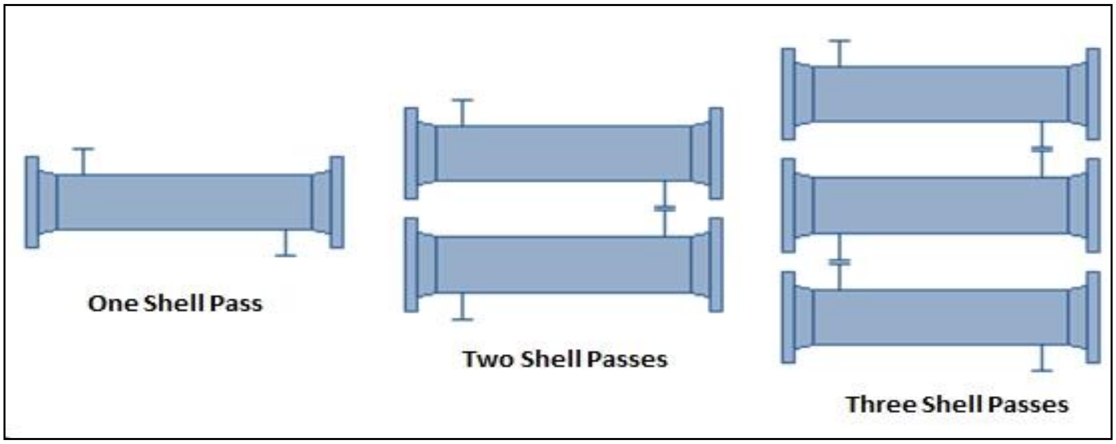


Figure 32:Types of passes in the shell side of heat exchanger

The best type of passes in a shell-and-tube heat exchanger design is one shell and two tube passes, the 2-tube passes type is considered the most frequently used and in which the flow travels 2 times the plate length before exiting on the other side, the 2-tube passes type increases the velocity of the fluid in tube side and this leads to provide higher film heat transfer coefficient in it.

9.2.6 TUBE SHEET TYPE SELECTION:

The tube sheet is made from a flat round piece of plate, a sheet with holes drilled to accept tubes or pipes in accurate location and pattern relative to each other. It is used to support and insulate tubes in heat exchangers. The tubes are connected to the tube sheets by hydraulic pressure or by roller expansion [73]. There are three types of tube sheets in shell and tube heat exchanger; the first one is the fixed tube sheet in which the tubes are straight through the shell, and at both ends they are fixed. The second one is the U-tube at which the tubes start in the top part of the tube sheet, make a U, and come back to the bottom part of the tube sheet. The last type is the floating head in which one end of the tube is fixed to the tube sheet, and the other end is kept free for expansion. [74] Among these types, fixed tube sheet was chosen because of: [75]

1. The simplicity of its construction in addition to being economical.
2. Minimal gasket joints.
3. Reduced possibility of fluid leakage due to shell welding on the tubular plate on both sides.
4. No intermixing of shell side and tube side fluid.
5. No restrictions on the number of side passes of the tube.
6. Easy tube replacement due to the use of straight tubes
7. Tubes can be cleaned mechanically or chemically.
8. Minimum by passing of shell side fluid between tube bundle and shell.

9.2.7 MATERIAL OF CONSTRUCTION TYPE SELECTION:

Carbon steel was chosen as a material of construction since it is one of the cheapest alloy that can resist corrosion.

9.2.8 TUBE PITCH TYPE SELECTION

Tube pitch in heat exchangers is the shortest distance (tube center to tube center) between two adjacent tubes. There are different types of tube arrangement such as triangular pitch, square pitch, diagonal square pitch, etc.

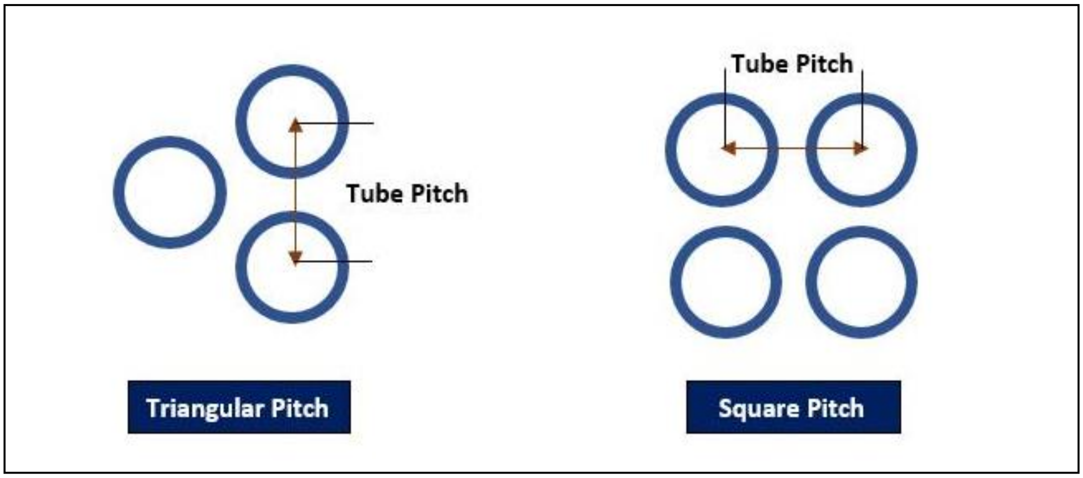


Figure 33:Types of tube pitch [76]

The shell side heat transfers co-efficient in case of triangular pitch as compared to the square pitch under similar condition, fluid flow and tube size, is large. The triangular pitch provides a more compact arrangement, usually resulting in smaller shell, and stronger header sheet for a specified shell-side flow area. It is preferred when the operating pressure difference between the two fluids is large (high turbulence) as in our process. So, the triangular pitch arrangement was used. [77]

9.2.9 BAFFLE TYPE SELECTION:

Baffles are the inter connectors of the tubes and shell of the heat exchangers. The shell side fluid will pass through the direction depending on the baffle’s arrangements. The crossflow pattern on the shell side caused by presence of baffles, enables better contact of the shell side fluid with the tube surface. This enhances the heat transfer efficiency in a shell & tube exchanger. [77]

- **Baffle type**

There are different types of baffles such as conventional segmental baffles, deflecting baffles, overlap helical baffles, and the rod baffles. The most commonly used is segmental baffles since it makes the fluid flow in a tortuous manner across the tube bundle in the shell side and it improves the heat transfer by enhancing turbulence and local mixing on the shell side of heat exchanger. [78]

- **Baffle cut**

It is the percentage of the shell inside diameter that is not covered by the baffle and also it is defined as the height of segment that is cut in each baffle to allow the shell side fluid to flow across the baffle. The normal recommended value of baffle cut is 20% to 25% of shell diameter that used in shell and tube heat exchanger design. Horizontal and vertical cuts are the two types of it, and the vertical cut was chosen since it can simplify the fabrication of two-pass exchangers. [78]

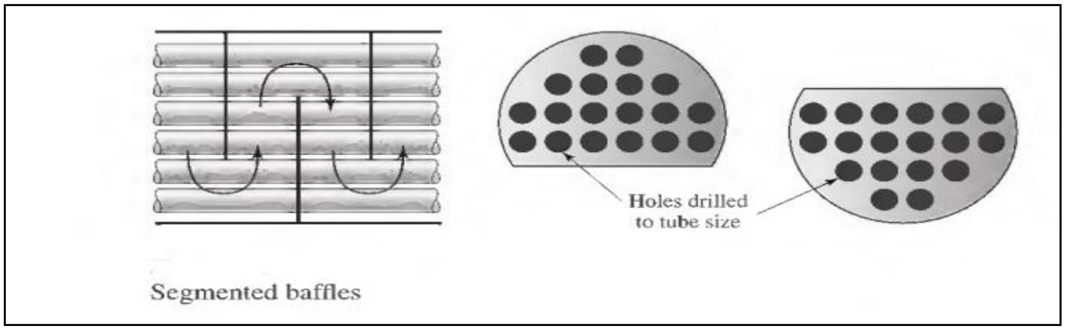


Figure 34:segmental baffles with vertical cut [80]

9.2.10 TUBES ARRANGEMENT TYPE SELECTION:

There are two arrangements of tubes that flowing through tube banks, which are the in-line arrangement and staggered arrangement. The staggered arrangement was chosen in our process since it provides larger pressure drop and heat transfer than the in-line arrangement.

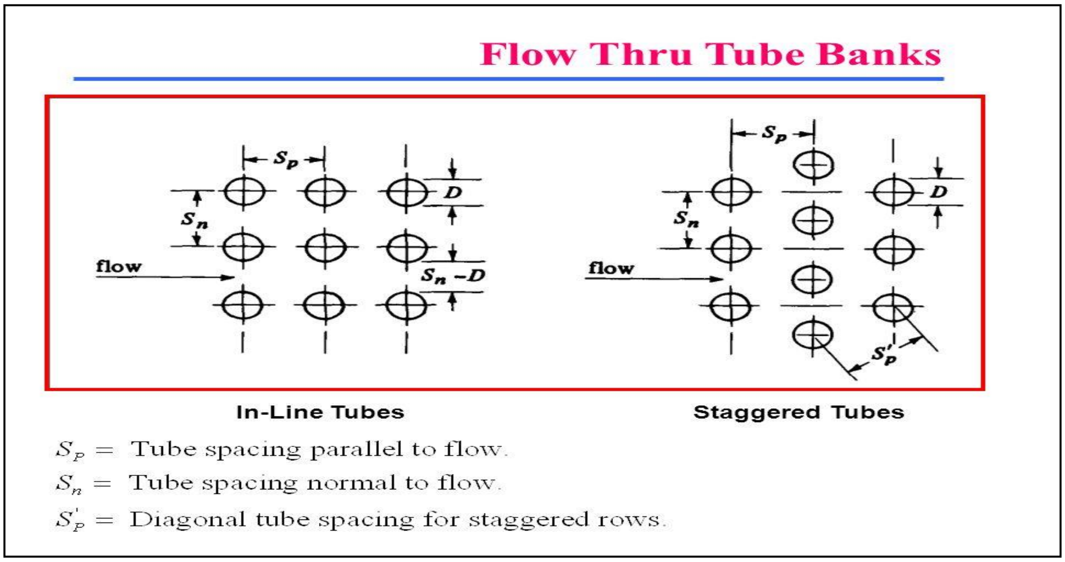


Figure 35:Flow through tube banks

9.2.11 FOULING RESISTANCE

The fouling of heat exchangers is the accumulation of unwanted deposits on heat transfer surfaces. The fouling layer imposes an additional resistance to heat transfer and the narrowing of the flow area due to the presence of deposit, results in an increased velocity for a given volumetric flow rate. Furthermore, the deposit is usually hydro-dynamically rough so that there is an increased resistance to the flow of the fluid across the deposit surface. Therefore, the consequences of fouling are, a reduction in exchanger efficiency and other associated operating problems including excessive pressure drop across the exchanger. [81]

9.2.12 SAMPLE CALCULATION FOR HEAT EXCHANGER DESIGN

A 1- shell- 2- tube passes with fixed tube sheet heat exchanger is designed to cool the fermenter feed mixture using liquid water. The feed mixture at a flow rate of 59481.45 Kg/hr enters the shell side at 168.49 °C and leaves at 50°C. It is assumed that the water at a flow rate of 70932.95 Kg/hr enters the tube side at 20°C and leaves at 52°C.

Table 31:Shell and Tube side streams information for (E-101)

Properties	**Tube Side**	**Shell Side**
	Cold (water)	Hot (Fermenter feed mixture)
Tin	20	168.49
Tout	52	50
ΔT	32	-118.49
T_{avg}	36	109.2
Cp (KJ/Kmol.K)	75.3 [82]	82.066
Density (kg/m^3)	993.72 [83]	929.017
Viscosity (Pa. s)	0.000705 [84]	0.000254
Thermal conductivity (W/m. K)	0.62744 [85]	0.6374

1- Heat transfer rate:

$$Q = Q_{mixture} = Q_{cooling\ water} = Q_{17} - Q_4 \dots (10)$$

$$Q_{cooling\ water} = 28319561.84 - 5898638.875$$

$$Q_{cooling\ water} = 22420922.97\ KJ/hr$$

$$Q_{cooling\ water} = nC_p(T_{out} - T_{in}) + n\lambda \dots (11)$$

$$22420922.97 = n * 75.9(367.65 - 293.15) + n * 35$$

$$n = 3940.719\ kmol/hr$$

$$Mass\ flow\ rate\ of\ water\ in\ tubes\ (M) = n * MW \dots (12)$$

$$M = \frac{3940.719\ kmol}{hr} * \frac{18\ Kg}{Kmol}$$

$$M = 70932.958\ Kg/hr$$

2- Mean Temperature Difference (∆Tm)

$$\boldsymbol{\Delta T_m = F_T * \Delta T_{LM}} \dots \dots (13)$$

The flow is assumed to be counter current flow:

$$\Delta T_{LM} = \frac{(T_{\ h,in} - T_{c,out}) - (T_{h,out} - T_{c,in})}{ln \frac{T_{\ h,in} - T_{c,out}}{T_{h,out} - T_{c,in}}} = \frac{(168.49 - 52) - (50 - 20)}{ln \frac{(168.49 - 52)}{(50 - 20)}} \dots \dots \dots (14)$$

$$\boldsymbol{\Delta T_{LM} = 63.75}$$

$$R = \frac{(T_{h,in} - T_{h,out})}{(T_{c,out} - T_{c,in})} \dots\dots(15) \Rightarrow R = \frac{168.49 - 50}{52 - 20}$$

$$\boldsymbol{R = 3.7}$$

$$\text{Thermal effectiveness factor } (P) = \frac{(T_{c,out} - T_{c,in})}{(T_{h,in} - T_{c,in})} \dots\dots(16) \Rightarrow P = \frac{52 - 20}{168.49 - 20}$$

$$\boldsymbol{P = 0.215}$$

From figure (14-4) “plant design and economics for chemical engineering 5th”, at P= 0.215 and R = 3.7, the value of correction factor can be found, and it was equal to:

$$\boldsymbol{F_T = 0.96}$$

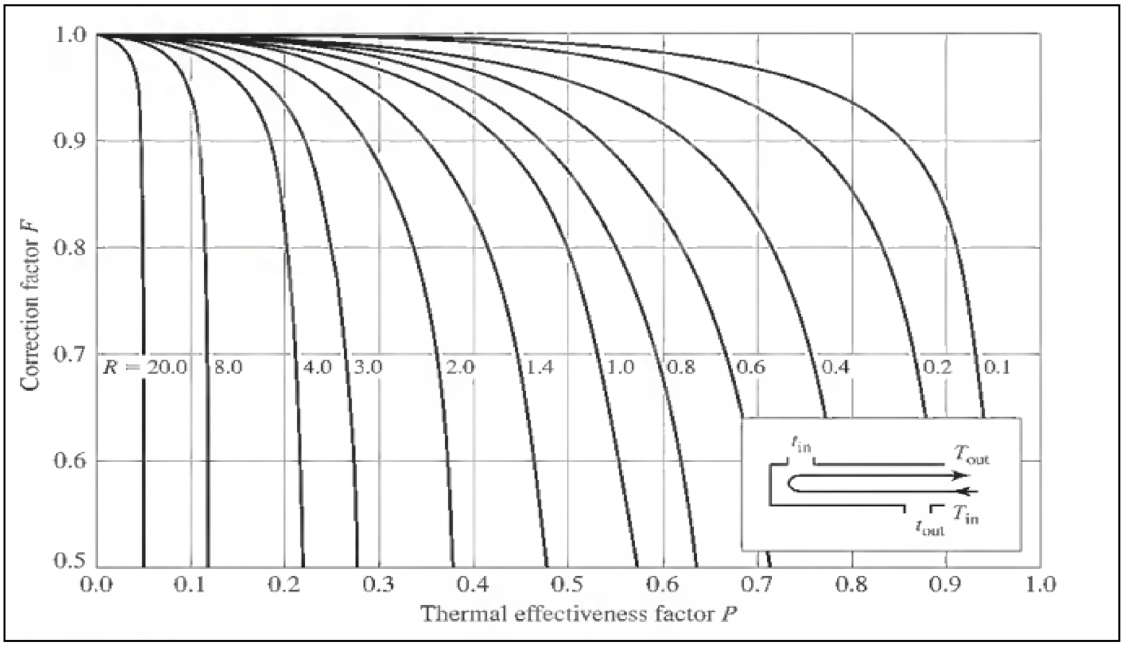


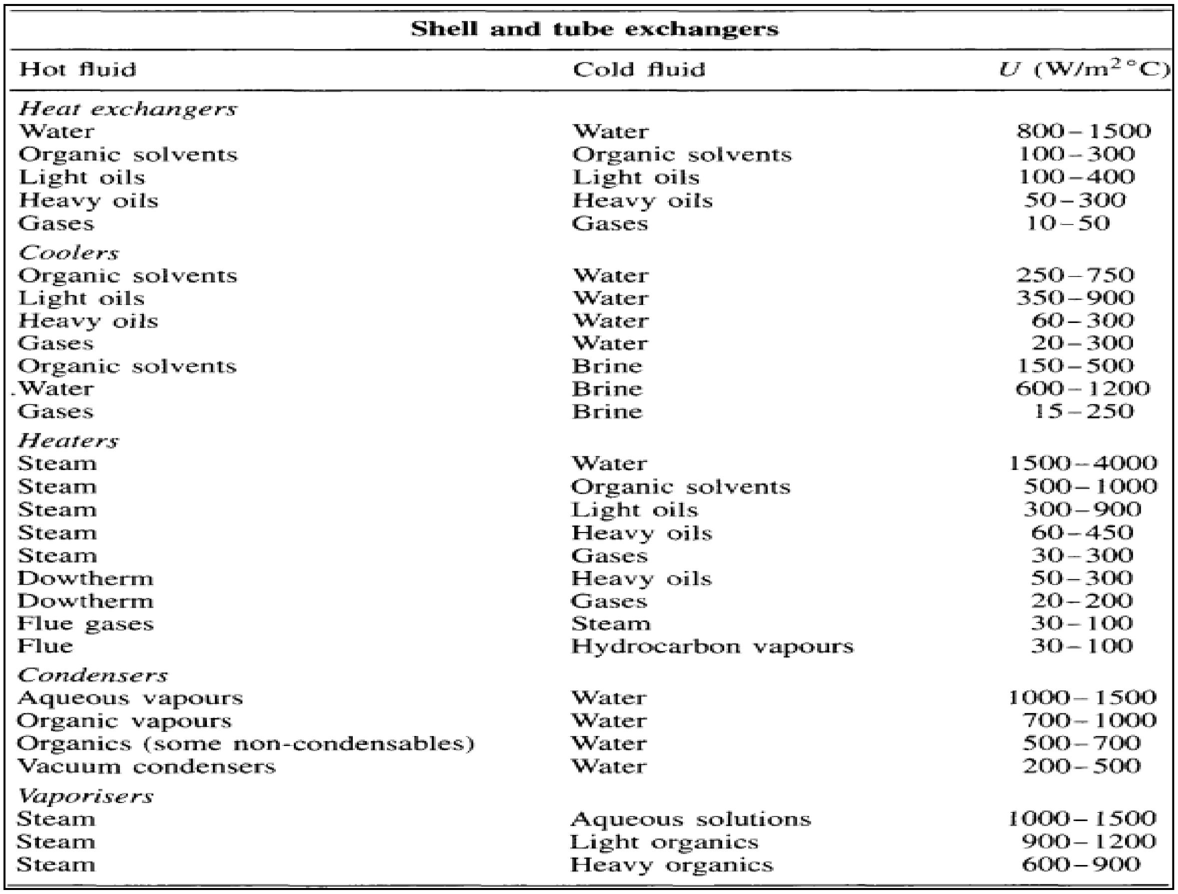
Figure 36:Chart for determining correct mean temperature-difference driving force for an exchanger with one shell pass and two or more even-numbered tube passes [63]

$$\boldsymbol{\Delta T_m = F_T * \Delta T_{LM}}$$

$$\boldsymbol{\Delta T_m = 0.96 * 63.75 = 61.2}$$

3- Overall heat transfer coefficient (U):

From table (12-1) “Coulson Richardson s Chemical Engineering vol_6”, the range of assumed value of overall heat transfer coefficient (U) for water-to-water heat exchanger is (800-1500) W/m^2.K , trial value is assumed to be 900 W/m^2.K



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

Figure 37:Typical values of the overall heat-transfer coefficient for various types of heat exchangers [80]

$$\mathrm{Q} = \mathrm{U} * \mathrm{A} * \Delta T_m \ldots \ldots (17)$$

$$A = \frac{22420922.97\ \mathrm{KJ}}{\mathrm{hr}} * \frac{m^2.K}{900\ W} * \frac{1}{61.2\ K} * \frac{hr}{3600\ s} * \frac{1000\ J}{KJ} * \frac{W.s}{J}$$

$$Heat\ transfer\ area\ (A) = 113.06\ m^2$$

4- Tube Layout and Sizing:

The design is done under these assumptions:

- Nominal outside tube diameter (Do) = (3/4)” = 0.01905m.
- Birmingham wire gauge (BWG) = 12 (Carbon steel tubes will normally be 12 or 14 gauge which includes an allowance for corrosion)
- Tube length (Lt) = 16 ft = 4.8768 m
- From table (D-12) “plant design and economics for chemical engineering 5th” ,

At Do = 0.75” and BWG = 12 , the inside tube diameter (Di) can be determined, and it is equaled to 0.532” = 0.013513 m

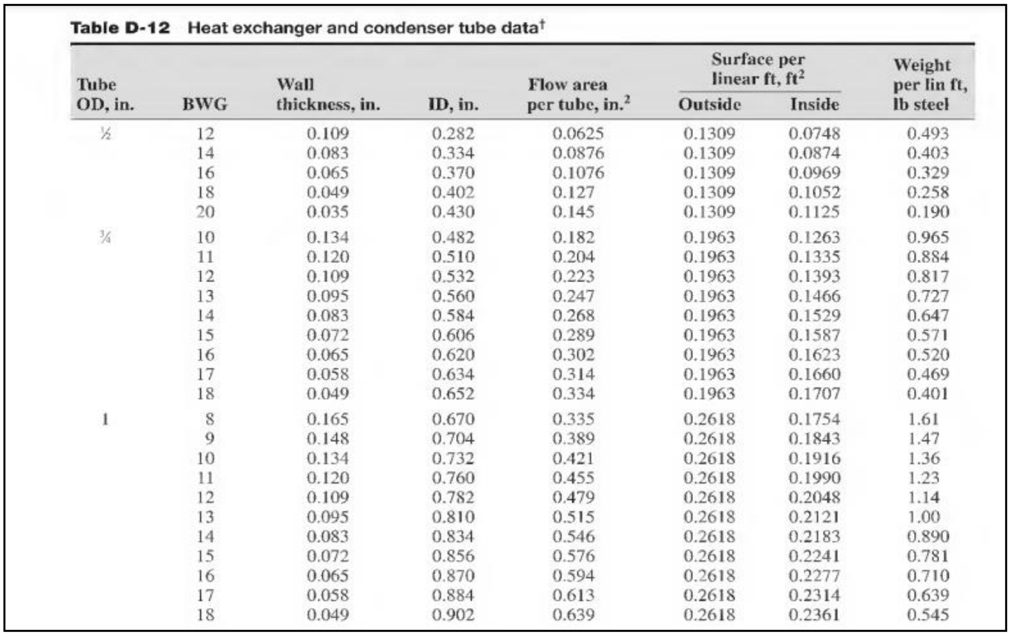


Table D-12 Heat exchanger and condenser tube data†

Tube OD, in.	BWG	Wall thickness, in.	ID, in.	Flow area per tube, in.^{2}	Surface per linear 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	0.1309	0.0874	0.403
	16	0.065	0.370	0.1076	0.1309	0.0969	0.329
	18	0.049	0.402	0.127	0.1309	0.1052	0.258
	20	0.035	0.430	0.145	0.1309	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	0.1963	0.1335	0.884
	12	0.109	0.532	0.223	0.1963	0.1393	0.817
	13	0.095	0.560	0.247	0.1963	0.1466	0.727
	14	0.083	0.584	0.268	0.1963	0.1529	0.647
	15	0.072	0.606	0.289	0.1963	0.1587	0.571
	16	0.065	0.620	0.302	0.1963	0.1623	0.520
	17	0.058	0.634	0.314	0.1963	0.1660	0.469
	18	0.049	0.652	0.334	0.1963	0.1707	0.401
1	8	0.165	0.670	0.335	0.2618	0.1754	1.61
	9	0.148	0.704	0.389	0.2618	0.1843	1.47
	10	0.134	0.732	0.421	0.2618	0.1916	1.36
	11	0.120	0.760	0.455	0.2618	0.1990	1.23
	12	0.109	0.782	0.479	0.2618	0.2048	1.14
	13	0.095	0.810	0.515	0.2618	0.2121	1.00
	14	0.083	0.834	0.546	0.2618	0.2183	0.890
	15	0.072	0.856	0.576	0.2618	0.2241	0.781
	16	0.065	0.870	0.594	0.2618	0.2277	0.710
	17	0.058	0.884	0.613	0.2618	0.2314	0.639
	18	0.049	0.902	0.639	0.2618	0.2361	0.545

Figure 38:Heat exchanger and condenser tube data [63]

- The thickness of the tube $(tw) = D_o - D_i = \ 0.01905 \text{ m} - 0.013513 \text{ m} = \ \ 0.005537$

$\text{A} = \ \text{n}\,\pi\, D_o L_t \ldots\ldots (18)$

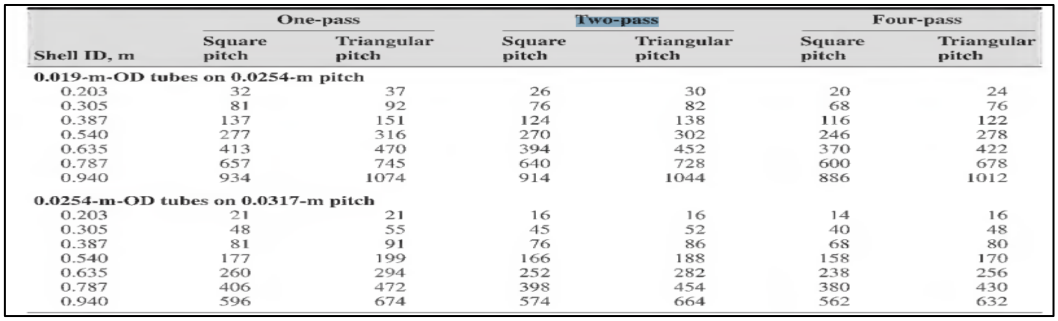
- $113.06 = n * 3.14 * 0.01905\ m * 4.8768\ \text{m}$
- $number\ of\ tubes = \boldsymbol{n} = \mathbf{388}$

a- Tube pitch

- Minimum tube pitch recommended for shell and tube heat exchangers is about 1.25 times the outside diameter of the tube.
- $\text{Pitch} = 1.25 * D_o = 1.25 * 0.01905 = 0.0238125$
- $\text{Clearance} = \text{Pitch} - D_o = 0.0238125 - 0.01905 = 0.0047625$

b- Baffle spacing:

- Minimum baffle spacing recommended in a shell and tube heat exchanger is 0.2 times of shell diameter.
- From table (14-9) “plant design and economics for chemical engineering” , for two passes triangular pitch with (OD= ¾’’=0.019 m) and 388 tubes, the inside diameter of the shell is:



	One-pass		Two-pass		Four-pass	
Shell ID, m	Square pitch	Triangular pitch	Square pitch	Triangular pitch	Square pitch	Triangular pitch
0.019-m-OD tubes on 0.0254-m pitch						
0.203	32	37	26	30	20	24
0.305	81	92	76	82	68	76
0.387	137	151	124	138	116	122
0.540	277	316	270	302	246	278
0.635	413	470	394	452	370	422
0.787	657	745	640	728	600	678
0.940	934	1074	914	1044	886	1012
0.0254-m-OD tubes on 0.0317-m pitch						
0.203	21	21	16	16	14	16
0.305	48	55	45	52	40	48
0.387	81	91	76	86	68	80
0.540	177	199	166	188	158	170
0.635	260	294	252	282	238	256
0.787	406	472	398	454	380	430
0.940	596	674	574	664	562	632

Figure 39:Number of tubes in conventional tube sheet layouts[63]

- ID shell = 0.59 m

- $\frac{1}{5}\text{ID shell} < \textbf{Baffle spacing} < \text{ID shell}$

- $0.11\text{ m} < \text{ B } < \text{ 0.55 m}$

- $\text{B} = 0.334$

- **Baffle type:**

Use segmental baffles with vertical cut, then find number of baffles

- $\text{Number of baffles } (N_b) = \frac{L_t}{B} - 1 = \frac{4.8768}{0.334} - 1 = 13.57 = 14$

5- Heat transfer coefficients (h_i , h_o)

a. Heat transfer coefficient for tube side (h_i Calculation):

$$\text{The crossflow area (A flow, in)} = \frac{\text{\# of tubes}}{\text{\# of passes}} * \frac{\pi}{4} * D_i^2 \dots\dots (19)$$

$$A(\text{flow, in}) = \frac{388}{2} * \frac{\pi}{4} * 0.013513^{\ 2}$$

$$\boldsymbol{A = 0.02777\ m^2}$$

$$Velocity\ inside\ tube = \frac{\frac{\dot{m}}{\rho}}{\text{A flow, in}} \dots\dots (20)$$

$$= \frac{51783.05636\text{ Kg}}{hr} \frac{1}{0.020261\ m^2} * \frac{m^3}{993.72\text{Kg}} * \frac{hr}{3600s}$$

$$\boldsymbol{Vi = 0.71379\ m/s}$$

$$Re = \frac{\rho in * Vi * Di}{\mu} \dots\dots\dots (21)$$

$$Re = \frac{993.72 * 0.71379 * 0.013513}{0.000705}$$

$\boldsymbol{Re = 13595.615} > 2100$ (Turbulent)

$$Prandtl\ number(Pr) = \frac{C_p * \mu}{k} \ldots\ldots\ldots (22)$$

$$Pr = \frac{\frac{4.18\ \mathrm{KJ}}{\mathrm{Kg.K}} * 0.000705\ \mathrm{Pa.s}}{\frac{0.62744W}{m.K}} = \mathbf{4.69}$$

$$h_i = 0.023 * Re^{0.8} * P_r^{\frac{1}{3}} * \left(\frac{\mu}{\mu_m}\right)^{0.14} * \frac{K_i}{D_i} \ldots\ldots\ldots (23)$$

$\frac{\mu}{\mu_m} \approx 1$, since the viscosity of fluid at the bulk fluid temperature is very closed to the viscosity of fluid at tube wall temperature.

$$\boldsymbol{h_i} = 0.023 * 13607.15608\ ^{0.8} * 4.69^{\frac{1}{3}} * 1 * \frac{0.62744}{0.013513} = \frac{\mathbf{3624.085297W}}{\boldsymbol{m^2.K}}$$

b. **Heat transfer coefficient for shell side (h_o Calculation):**

$$\text{The cross} - \text{flow area (A flow, out)} = \frac{\text{(Shell ID)} * \text{(Clerance)} * \text{(Baffle spacing)}}{\text{Tube pitc}h} \ldots\ldots (24)$$

$$\text{A flow, out} = \frac{0.59\text{m} * 0.00476 * 0.33457\text{m}}{0.0238125} = 0.0395\ m^2$$

$$Velocity\ inside\ shell = \frac{\frac{\dot{m}}{\rho}}{\text{A flow, out}} \ldots\ldots\ldots (25)$$

$$= \frac{59306.82/929.0175}{0.0395} = 0.4491\ m/s$$

Shell side equivalent diameter (de) for triangular pitch arrangement:

$$d_e = \frac{1.1 * (P_t^2 - 0.907 * D_o^2)}{D_o} \ldots\ldots\ldots (26)$$

$$d_e = \frac{1.1 * (0.0238125^2 - 0.907 * 0.01905^2)}{0.01905} = 0.0137\ m$$

$$Re = \frac{\rho * V_{insinde\ shell} * d_e}{\mu} \dots\dots\dots (27)$$

$$Re = \frac{929.0175 * 0.4491 * 0.0137}{0.000254} = 22559.5227\ (turbulent)$$

$$P_r = \frac{C_{P(avg,mixture)} * \mu}{k} \dots\dots\dots (28)$$

$$\text{Pr} = \frac{4.146 * 0.000254}{0.63740} = 1.65$$

$$h_o = \frac{a_o}{F} * Re^{0.6} * P_r^{\frac{1}{3}} * \frac{k}{D_o} \dots\dots\dots (29)$$

$a_o = 0.33\ (tubes\ are\ staggered)$

$F = 1.6\ (safety\ factor)$

$$h_o = \frac{0.33}{1.6} * 17656.22^{\ 0.6} * 1.65^{\frac{1}{3}} * \frac{0.63740}{0.01905}$$

$$\boldsymbol{h_o = 3339.1755 \frac{W}{m^2.k}}$$

6- The overall heat transfer coefficient (calculated):

Using **Carbon steel** as material of construction for tube and shell side

$$\frac{1}{U} = \frac{1}{h_o} + \frac{1}{h_{od}} + (\frac{D_o}{2k_w} * \ln\left(\frac{D_o}{D_i}\right)) + \frac{D_o}{D_i h_{id}} + \frac{D_o}{D_i h_i} \dots\dots\dots\dots (30)$$

The thermal conductivity for tube wall material (Carbon steel) is Kw = 43 (W/m.K)

- **Fouling resistances**

From table (14-3) “Coulson Richardson s Chemical Engineering vol_6”, the values of shell and tube fouling resistances are:

$(h_{id})^{-1}=1*10^{-4}$ m^2. K/W → h_{id} = 10000 W/m^2 . K

$(h_{od})^{-1}=2*10^{-4}$ m^2. K/W → h_{od} = 5000 W/m^2 . K

$$\frac{1}{U}=\frac{1}{3339.175}+\frac{1}{5000}+\left(\frac{0.01905}{2*43}*ln\frac{0.01905}{0.013513}\right)+\frac{0.01905}{0.013513*10000}+\frac{0.01905}{0.013513*3626.54}$$

$$\frac{1}{U}=0.001105$$

$$U=\frac{904.77\ W}{m^2}.K \ \rightarrow \text{(Very close to our assumption)}$$

7- Pressure drop calculation:

a. Pressure drop in tube side:

$$\Delta P=N_P*\left[8J_f*\frac{L}{D_i}*\frac{\mu}{\mu_w}^{-M}+2.5\right]*\frac{\rho_t U_t^2}{2}\ldots\ldots\ldots(31)$$

Where:

J_f: Tube side friction factor

N_p: Number of passes

M=0.14 for turbulent flow at Re > 2100

ρ_t: Density of fluid in tube side (Kg/m^3)

L: Length per tube (m)

μ: Viscosity of fluid at the bulk fluid temperature.

μ_w: Viscosity of fluid at tube wall temperature

U_t = tube-side velocity =0.7143 m/s

From figure (12.24) "Coulson Richardson s Chemical Engineering vol_6",

At $\text{Re} = 13595.61491, J_f$ can be determined and it was equaled to 0.0045

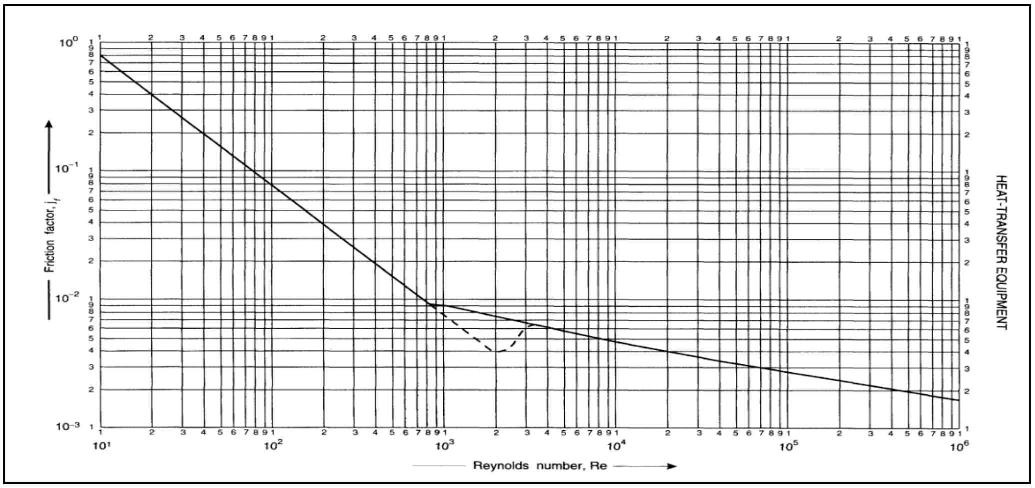


Figure 40:Tube-side friction factors[80]

$$\Delta P = 2 * \left[8 * 0.0045 * \frac{4.8768}{0.013513} * (1)^{(-0.14)} + 2.5\right] * \frac{993.72 * 0.7143^2}{2}$$

$\frac{\mu}{\mu_w} \approx 1$ since the viscosity of fluid at the bulk fluid temperature is very closed to the viscosity of fluid at tube wall temperature.

$$\Delta P = 7857.076\ Pa = 7.857\ KPa$$

The maximum allowable pressure drop = 35 Kpa, and (7.857 < 35)

So, the pressure drop in tube side is acceptable.

b. Pressure drop in shell side:

$$\Delta P_s = 8 * J_f * \frac{D_s}{d_e} * \frac{L}{B_s} * \frac{\rho_s U_s^2}{2} * \frac{\mu}{\mu_w}^{-0.14} \quad \dots\dots\dots(32)$$

Where:

B_s: baffle spacing =0.334

J_f: Shell side friction factor

Ds: Shell diameter (m) = 0.59

de: Shell side equivalent diameter (m) = 0.0137

L: Length per tube (m) = 4.87

μ: Viscosity of fluid at the bulk fluid temperature =0.000254

μ_w: Viscosity of fluid at tube wall temperature.

U_s =0.4491 , Re =22559.522

From figure (12-30) “Coulson Richardson s Chemical Engineering vol_6”, the value of J_f can be determined and it was equal to 0.046.

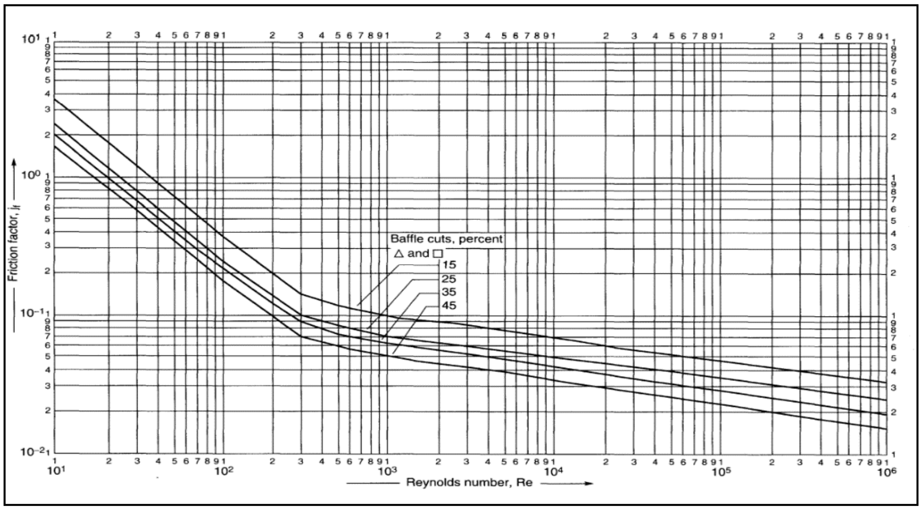


Figure 41:Shell-side friction factors, segmental baffles [80]

$$\Delta P_s = 8 * 0.046 * \frac{0.59}{0.0137} * \frac{4.87}{0.334} * \frac{929.017 * 0.4505^2}{2} * (1)^{-0.14}$$

$$\Delta P_s = 21592.349\ Pa = 21.59\ KPa$$

The maximum allowable pressure drop = 35 Kpa , and (21.59 < 35) ; so, the pressure drops in shell side acceptable.

9.2.13 DETAILED DESIGN SUMMARY

Table 32: Summery of E-101 Design

Specification Sheet for (E-101)		
Item number	E-101	
Number required	1	
Function	Pre-Cooling the reactor feed mixture	
Process information		
	Tube side	**Shell side**
Fluid Circulated	Water	Fermenter feed mixture
Material of construction	Carbon steel	Carbon steel
Mass flow rate (Kg/hr)	70932.95	59481.45
Heat Duty (KJ/hr)	22420922.97	22420922.97
Temperature in (C)	20	168.49
Temperature out(C)	52	50
h_o (W/m^2 . K)	-	3339.17
h_i (W/m^2 . K)	3626.54	-
h_{id} , h_{od} (W/m^2 . K)	10000	5000
Re	13607.16	22559.52
Pr	4.696	1.652
A_{flow} (m^2)	0.02777	0.0395
Velocity (m/s)	0.714	0.449
Pressure drop (KPa)	7.85	21.59
Dimensions		
Heat transfer area (m^2)	113.06	
Tube inside diameter (m)	0.013513	
Tube outside diameter (m)	0.01905	
Number of tubes	388	
Length per one tube (m)	4.876	
Shell inside diameter (m)	0.59	
Number of passes	2	
Pitch (Triangular)	0.02381	
Clearance	0.00476	
Baffle spacing	0.334	
Number of baffles	14	
Overall heat transfer coefficient U (W/m^2 . K)	904.77	
Cost $	**24,843**	

9.3 BIOREACTOR R-101 DESIGN

9.3.1 INTRODUCTION

Reactor is the heart of chemical engineering industry; it is an enclosed volume (process vessel) in which the chemical reaction takes place. Bioreactor, which is used in this process for hydrogen peroxide and Formaldehyde production, is the vessel in which a suitable biochemical environment is achieved by regulating the transfer of oxygen, pH and temperature. The desired chemical transformation inside the bioreactor is attained by utilizing the activity of the catalyst.

There are several types of bioreactors and here are some of them:

1. Stirred tank bioreactor, it is known for its flexibility, and ability to provide tremendous mass transfer and advantageous mixing, so it is one of the most frequently used fermenter.
2. Airlift bioreactor, this type does not have agitating system, but contents in the vessel are agitated with the air introduced from the bottom.
3. Packed bed bioreactor, it is normally used in wastewater engineering with biofilms attached.

The bioreactor is carried out in small or large size reactor depending on the product quantity, and it is accomplished in several configurations either batch, fed batch, or continuous. However, most bioreactors can be modeled as batch reactors, because of the following reasons:

- They provide sufficient time of contact between reactants and enzymes.
- They maintain the uniform concentration and temperature within the reaction mixture.

As a result, the process mode is chosen to be batch configuration.

Reactor (R-101) is selected to be stirred tank reactor with a batch operation and a cylindrical shape, and it consists of different parts:

- Baffles (vortex breaker), to provide better mixing.
- Sparger (aerator), to introduce the sterile air or oxygen to the media.
- Cooling jacket, to maintain the temperature of the process.
- Impeller (agitator), to maintain a continuous stirring, distribution of the oxygen throughout the system, enhance heat and mass transfer, and prevention cells from settling down.

Two types of impellers are widely used in the industry such as : axial and radial flow impellers. [86] the choosing agitator for the process is marine propeller for the following reasons:

1. It increases the homogeneity of the mixing
2. There are flexible in operation
3. Mostly used in the mechanical mixing of low to medium viscosity fluid where the fluid that inter the reactor is consider low viscous.
4. Low cost of propeller comparing with turbine agitator.

9.3.2 SAMPLE CALCULATION FOR BIOREACTOR: R-101 DESIGN [87,88]

According to "Elementary of chemical engineer edition 5 chapter 9"

The reaction time for biocatalyst batch reactor is:

$$t = \frac{Km}{Vmax} ln \frac{1}{1-X} + \frac{CA*X}{Vmax} \dots \dots \dots (33)$$

Where:

V_{max}: maximum reaction rate of the enzyme, it is equals to 0.3137 mM/min

K_m: Michaelis -Menten constant, it depends on the enzyme concentration, and in our case, it equals to 0.4 mM since the oxygen is the limiting reactant.

[**CA**]: concentration of substrate (methanol)

X: conversion of oxygen, it is selected to be 90% conversion

According to patents ([90] , page 8 last line) that the reaction is preferred when feed stock of methanol is at 1% ,2%,3% w/v. In this project the concentration of methanol was taken to be 1% w/v which is equal 156.2 mM, substitute value of Vmax, K_m, and C_{A0} in the previous equation then:

$$t = \frac{0.4}{0.3137} ln \frac{1}{1-0.9} + \frac{156.2*0.9}{0.3137} = 448.14 \text{ min } = 7.46 \text{ hr}$$

Then volume of batch reactor is

$$V = \frac{\dot{m}*tc}{\rho L} \dots \dots \dots (34)$$ [98]

$\dot{m}$:mass flowrate of the inlet mixture**, 62490 kg/hr**

ρl: density of liquid for the inlet mixture, **995.76 kg/m^3**

The mass flowrate of the mixture is calculated by taking the summation of the multiplication of the molar flowrate of each component with its molecular weight, while the liquid density is considered by taking the summation of the multiplication of the density of each component at the inlet temperatures with its composition in the inlet streams.

t_c: total batch time which is assumed according to “chemical process engineering design and economy chapter 6” is,

$$t_b = t_R + t_c + t_e + t_f \ldots\ldots\ldots (35)$$

t_c: Cooling time, and it is assumed to be **1.5hr**

t_f: Feed charging time, and it is assumed to be **1 hr**

t_e: Empty time (discharging) and assumed to be **0.5hr**

t_r = Reaction time = 7.46 hr

Go back to the relation.

$$t_b = 7.46 + 1.5 + 1 + 0.5 = 10.46\ hr$$

$$V = \frac{61329*10.46}{995.76} = 644.24\ m^3$$

with a volume allowance of 20% of free space in the reactor, the total volume will be

V_R = 644.24*1.2 = 773.09 m^3

Since the reactor volume is considering high comparing to the large scale industrial fermenter (2-500 m^3) [89]; then it is decided to use 6 reactors in parallel with V_R (for one reactor) = 773.09/6 = 128.84 m^3

1. Height and Diameter of reactor: H, D

Referring to most design books, it is shown that the height of the column or reactor is to be 1.5 times the diameter (H=1.5D)

As we know that the volume of the cylindrical reactor is

$V_R = Area * Hieght \ldots\ldots\ldots (36)$

$128.84 = (3.14/4) * D^2 * 1.5D$

$D_i = 4.77\ m , H = 7.17\ m$

Mechanical design: [88]

It is selected to be a stainless-steel reactor because formaldehyde and hydrogen peroxide corrode carbon steel.

1- Wall thickness: t_m

First, it is required to calculate the total pressure, which is the summation of the static to the operating pressure of the reactor as follows,

$\text{Static pressure} = \rho * g * h \ldots\ldots\ldots (37)$

$$\text{Static pressure} = \frac{995.7 * 9.81 * 7.17}{1000} = 70\ \text{Kpa}$$

$\text{Operating pressure} = 11.15\ \text{bar} = 1115\ \text{Kpa}$

$\text{Total pressure} = \text{Static} + \text{operation} = 70 + 1115 = 1185\ \text{KPa}$, (At the base of the reactor)

For safety purposes, 10% extra pressure is added to the total pressure to be

Total pressure = 1185*1.1 = 1303.5 KPa

$$t_m = \frac{P * Di}{2(S * E_j - 0.6\,P)} \ldots\ldots\ldots (38)$$

Di: inside diameter of the reactor, 4.77 m

S: maximum allowable stress, it equals 165000 KN/m^2 for the selected material of construction [80 page 826]

Ej: welded joint efficiency, the recommended efficiency is 0.85 [80 page 827], given that a double welded butt is used

Back to the relation,

$$t_m = \frac{1303.5 * 4.77}{2 * (165000 * 0.85 - 0.6 * 1303.5)} = 0.022 \text{ m}$$

t_m = 0.022 m, for SS reactors the wall thickness must be less than 25 mm, and this condition is achieved here so the design is good. "Coulson and Richardsons chemical engineering volume6, page827" [80]

To take the corrosion in consideration, and because most of design codes and standards specify a minimum allowance of 1mm, the corrosion allowance is added to the wall thickness by 1mm (0.001m), "Coulson and Richardsons chemical engineering volume6, page 827" [80]

so, the total wall thickness will be:

t_m = 0.022 + 0.001 = 0.023m

2- Outside diameter: D_o

It is found as follows,

$D_o = D_i + 2\ t_m D_o \ldots \ldots \ldots (39)$

$= 4.77 + 2 * 0.022$

$D_o = 4.829m$

3- Minimum thickness for reactor head: t_H

To select the shape of the reactor head, the following condition should be used

For pressure $> 150Psi\ , Ellipsoidal\ head\ is\ used$

For pressure $< 150\ Psi\ , Torispherical\ head\ is\ used$

Our total pressure at the base of the reactor is 1303.5 KPa = 189 Psi,

Hence 189 Psi >150 Psi, so the reactor head shape is selected to be ellipsoidal, and the minimum thickness for ellipsoidal heads is calculated as follows,

$$t_H = \frac{1.104 * \mathrm{P} * \mathrm{D}i}{2 * \mathrm{Ej} * \mathrm{S} - 0.2 * P} \ldots \ldots \ldots (40)$$

$t_H = 1.104 * (1303.5 * 4.7) / ((2 * 0.85 * 165000) - (0.2 * 1303.5)$

$t_H =$ 0.024m, after adding the corrosion allowance

Agitator Design: "Unit Operations of Chemical Engineering 5 th , page264" [91]

Agitator dimensions: [92/ for marine propeller]

Da: impeller diameter, it is calculated as Da = 0.15*Dt = 0.15*4.77 = 0.72

E: impeller height above the vessel floor, and it is equal to the impeller diameter, E = 0.71m

L: length of impeller blade, it is calculated as L = Da/4 = 0.71/4 = 0.175 m

W: width of impeller blade, it is calculated as W = Da/5 = 0.71/5 = 0.142m

J: width of baffle, it is calculated as J = Da/12 = 4.77/12 = 0.4 m

H: height of baffle = height of reactor = 7.17 m

Number of impellers = 3

Number of blades = 3 since for marine propeller, number of blades is 3 or 4. [92]

1- Shape factors:

$S_1 = D_a/D_t = 0.71/4.77 = 0.15$

$S_2 = E/D_t = 0.71/4.77 = 0.15$

$S_3 = L/D_a = 0.175/0.71= 0.25$

$S_4 = W/D_a = 0.142/0.71= 0.2$

$S_5 = J/D_t = 0.4/4.77 = 0.08$

$S_6 = H/D_t = 7.17/4.77 = 1.5$

Baffle design:

Number of baffles = 4

Width of one baffle = D_t /10 = 0.477m

Height of baffle =height of reactor= 7.17 m

2- Power consumption:

Np: power number = 0.3 for 3 blades [92]

$$\text{Nspeed: impeller speed} = \frac{5}{\pi * D_a} \ldots\ldots\ldots.(41)$$

$$\text{Nspeed} = \frac{5}{3.14 * 0.71} = 2.24\ \text{RPS} = 2.24 * 60 = 134.5\ \text{RPM}$$

Tip velocity is ranged from 3-6 m/s, and it is selected to be 4 m/s

$$\text{Power consumption} = \frac{N_p * N^3 * {D_a}^5 * \rho}{gc} \ldots(42)$$

$$= \frac{(6 * (2.24^3) * (0.71^5) * 995.7)}{32.17}$$

$$\text{Power consumption} = 19.3\ \text{hp}$$

Assume the impeller is 85% efficient, then

$$\text{Actual power} = 19.3 * 0.85 = 16.4\ \text{hp}$$

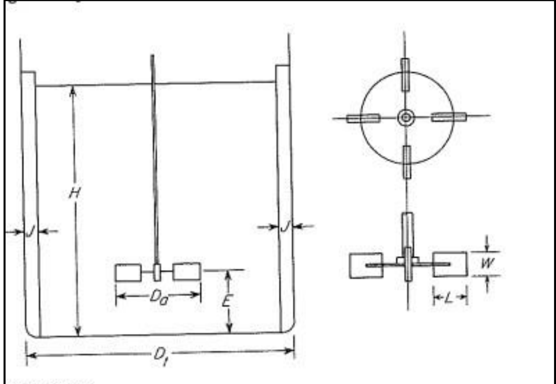


FIGURE 9.7
Measurements of turbine. (*After Rushton et al.*[42])

- Cooling jacket Design

From Energy balance calculation for plant capacity 50,000 ton /year of H_2O_2 the amount of heat removed by using R-134 A is 5012043.98 KJ/hr

- Inlet temperature of R-134a = -28 C
- Outlet temperature of R-134a = 35 C

$$\boldsymbol{Q_{R-134a} = \dot{m}_{mR-134a} * C_P \Delta T + \dot{m}\lambda} \ldots\ldots\ldots.(43)$$

- Cp R-134a at average temperature and pressure 8 bar = 1.34849 KJ/Kg.K

λ = 216.97 KJ/Kg

$5012043.98 = \dot{m} * (1.34849 * (35 - -28) + 216.97)$

By solving the previous equation, the amount of refrigerate R-134 a = 16598.881 Kg/hr .

9.3.3 DETAILED DESIGN SUMMARY

Table 33:Specification Data For R-101

Equipment	Bioreactor
Function	To convert methanol into hydrogen peroxide and formaldehyde
Number of Reactors	6 in parallel
Operation	Batch mode
Type	Cooling Jacketed Stirred Tank Reactor
Material of Construction	Stainless steel 304
Residence Time (hr)	7.46
Total Volume of Reactor(m^3)	773.09
Diameter of one Reactor(m)	4.77
Height of Reactor(m)	7.17
Static Pressure (Kpa)	70.8
Operating Pressure (Kpa)	1115
Total Pressure (Kpa)	1303.6
Outside Diameter of Reactor(m)	4.82
Wall thickness(m)	0.022
Agitator Design	
Impeller Diameter(m)	0.72
Impeller Height above the floor(m)	0.72
Length of Impeller Blade(m)	0.175
Width of Impeller Blade(m)	0.142
Width of Baffle(m)	0.4
Number of Impellers	3
Number of Impellers Blades	3
Speed of Impellers (RPM)	134.5
Actual power (hp)	16.4
Cost $	2,791,061

9.4 ULTRAFILTRATION DESIGN

9.4.1 INTRODUCTION

Ultrafiltration unit is one of the separation processes that use either a hollow membrane or a sheet membrane to mechanically filter the inlet fluid that contains very small particulates by sieving and diffusion mechanics. UF works on a micro level which means that it can almost take all the suspended materials out of the fluid.

For the enzymatic process of methanol conversion into formaldehyde and hydrogen peroxide, the Ultrafiltration is required to recover the methanol oxidase enzyme that appears in the product stream using a hollow fiber membrane. This unit is widely used to purify and concentrate the enzyme, and it is also used to prevent the loss of enzyme activity.

The feed stream (M_6*) to this unit is mixture of methanol, oxygen, formaldehyde, hydrogen peroxide, and enzyme, this stream flows either inside the shell or in the fibers. The enzyme is retained back to the reactor through stream (M_{enzyme}), while the purified product stream pass through the membrane and exits it through stream (M_6) as a permeate.

9.4.2 SAMPLE CALCULATION FOR ULTRAFILTRATION DESIGN

1. **Flux of permeate: [93,94,95,96]**

The flow of the solvent through ultrafiltration membrane is described with an assumption of ideal cylindrical pores aligned normal to the membrane surface as follows:

$J_v = \frac{\in m * d_P^2 * \Delta P}{32 * \mu * L_p} \ldots \ldots \ldots . (44)$, where:

J_v: volumetric permeate flux, m^3/m^2

$\in_m$: membrane porosity, 0.056

$\mathbf{d_p}$: average pore diameter,$1.25*10^{-9}$ m

$\Delta \boldsymbol{P}$: transmembrane pressure, Kpa

μ: viscosity, Pa. s (Kg/m.s), its value is calculated in the appendix

L_p: average pore length, 2m

Transmembrane pressure can be found using the following relation,

$$\Delta P = \frac{P_i + P_o}{2} - P_p \ldots\ldots\ldots.(45)$$

P_i: inlet pressure (M_6*)

P_o: outlet pressure, retentate pressure (M_{enzyme})

P_p: permeate pressure ($_{M6}$)

$$\Delta P = \frac{1290000 + 1290000}{2} - 1115000 = 175000\, pa$$

By substitution the parameters,

$$J_v = \frac{0.056*\left(1.25*10^{-9}*2\right)^2*175000}{32*2.5*10^{-3}*2}$$

$\mathbf{J_v = 1.53*10^{-4}\ m^3/m^2}$

2. Filtering Area

The filtering area can be determined by:

$$A = \frac{F_P}{J_W} \ldots\ldots\ldots.(46)$$

$\mathbf{F_P}$**:** volumetric flow rate of permeate (M_6), the calculation of it is shown in the appendix.

A = 0.012182763/ $1.53*10^{\mathbf{-4}}$ **, A = 7.96E+01 m^2**

3. Number of pores per cm^2

It can be predicted using the following relation:

$$N = \frac{4}{\pi * d^2} * \varepsilon \ldots\ldots\ldots.(47)$$

$\boldsymbol{\varepsilon}$: porosity

d: pore diameter, m

And now,

N = (4/3.14*(1.25*10^{-9})^2) *0.056

N = 1.1414E+16 pores

9.4.3 DETAILED DESIGN SUMMARY

Table 34:Specification Sheet of ultrafiltration

Name of unit	**Ultrafiltration**
Function	Recovery for the methanol oxidase to maintain it in a reaction zone
Type of membrane	Hollow fiber membrane
Total Cost ($)	127,161
Permeate flux (m/s)	1.1414*10^{16}
Filtering area (m^2)	79.6
Number of holes	1.1414*10^{16}

9.5 FLASH DRUM (V-101) DESIGN

9.5.1 INTRODUCTION

Flash drum is considered as one type of gas-liquid separator which separates a gas from liquid stream under reduce pressure. In chemical industry, more than 70% of the total project cost is incurred on separation and purification equipment's, these give an indication about the importance of such process.

There are two types of Liquid – Gas separation equipment which are vertical and horizontal (knockout pot) [97] separators. Horizontal separator is used when three phase presents as well as when liquid content is small, in this project the separation is based on two phase and the liquid content is large as a result horizontal separator is not appropriate and the vertical separator is selected.

The main function of flash drum (V-101) is to remove gases that produced from reactor as O_2 from desire products by flashing stream M_6 where the pressure is reduced from 11.15 bar to 0.35 bar and produce vapor stream M_7 and Liquid stream M_8.

9.5.2 SAMPLE CALCULATION FOR FLASH DRUM (V-101) DESIGN

The design procedure was taken from "Chemical Process Engineering Design and Economic" [98]

Entrainment factor Kv = 0.03045 m/s (Assume no mist eliminator since mist is used when the efficiency of process needs to be 99 % which is not needed in this project)

Density of vapor = 0.3049 Kg/m^3

Density of Liquid = 932.17Kg/m^3

$$Gas\ velocity = K_v \sqrt{\frac{\rho_l - \rho_v}{\rho_v}} \ldots \ldots \ldots .(48)$$

$$= 0.03045 \ * \sqrt{\frac{932.17 - 0.3049}{0.3049}} = 1.683 \text{ m/s}$$

$$\text{Vapor volumetric flowrate } = \frac{n_7 * Mw}{\rho_v * 3600} \ldots \ldots \ldots (49)$$

$$\text{Vapor volumetric flowrate} = 7.122\ m^3/s$$

$$\text{Area of the Flash } = \frac{\text{Vapor volumetric flowrate}}{Gas\ velocity} \ldots \ldots \ldots (50)$$

$$= \frac{7.122}{1.683} = 4.231\ m^2$$

$$\text{Diameter } = \sqrt{\frac{4 * A}{3.14}} = 2.32 m^2$$

Liquid-phase surge time, t_s = 4 min =280 s (For Vertical Flash t_s between 3 to 5 min)

$$\text{Liquid flow rate } = \frac{n_6 * Mw}{\rho_l * 3600} = 0.0174\ m^3/s$$

$$\text{liquid} - \text{level height } = \frac{\text{Vapor volumetric flowrate} * \text{Liquid} - \text{phase surge time}}{\text{Area of the Flash}} \ldots \ldots \ldots (51)$$

$$= 0.985 \text{ m}$$

Total separator height = liquid-level height +(1.5 D + 0.457).……(52)

= 0.985+(1.5*2.32+0.457) = 4.9246 m

Total separator height / D =2.121

Since L/D ratio less than 5 so the design is **satisfactory**

9.5.3 DETAILED DESIGN SUMMARY

Table 35:Specification sheet for V-101

Number of equipment	1
Function	Remove gases that produce from reactor as O_2 from the desired products
Type	Vertical Flash drum
Area (m^2)	4.231
Diameter (m)	2.32
Liquid-phase surge time (min)	4

9.6 DISTILLATION COLUMN (T-101) DESIGN

Distillation is considered one of the most widely used mass transfer chemical processes in chemical industries in addition to absorption. In chemistry it is a particular method of purifying liquids, and in industry is used to purify chemicals. Generally, distillation is a process used to separate liquid or vapor mixture of two components (binary system) or more (multi component system) based on their required conditions to change their phase. For example, a mixture of liquids can be separated by heating the liquid to force components with different boiling points or relative volatility into the gas phase, the gas is then condensed back into liquid formed and collected, and then the process is repeated to purify the product. The operations of small scale differ enormously in size and complexity to the large scale of thousands of tons per year production operations.

9.6.1 DISTILLATION TYPE SELECTION:

There are several types of distillation columns that differ in design to carry out specific types of separation processes and in terms of complexity. For example, in a batch distillation, which is the most common type used in laboratory, the feed is introduced to the column in a batch wise and then the column starts to carry out the process, and a next batch of the feed is introduced after each achieving of the targeted task. However, the continuous distillation columns process a continuous stream of the feed with no interruption except for column problems. This type is more common in industry for its capability of handling high production. In this project, the aim of using distillation column (T-101) is to separate formaldehyde, methanol, and water from the targeted component which is the hydrogen peroxide to get it pure as possible. The type of distillation needed to process this separation is the continuous type with a production rate of 50000 tons/year of hydrogen peroxide which is equal to 91.8 Kmol/hr.

9.6.2 COLUMN TYPE SELECTION: PLATE OR PACKED

The selection criterion between the two types is done on the following basis:

- Plate columns can handle a wide range of gas and liquid flow rates.
- Packed columns are more suitable for low-capacity operations.
- Plate columns are normally easier to clean the plates when the liquid cause fouling or deposition of solids.
- Packed columns are more suitable for foaming and corrosive systems.
- It is easier to make the provision for installation of internal cooling coils or withdrawal of side streams from a plate column.
- In packed columns, there are always some uncertainly to maintain good liquid distribution, and so that it is difficult to estimate the efficiency of the packed columns.
- Plate columns efficiency can be predicted more accurately.

Upon the above points which illustrate that the advantages of the plate overpass the packed column advantages, the plate column was chosen as the type of the distillation columns in the process.

9.6.3 PLATE TRAY SELECTION CRITERIA: SIEVE, BUBBLE CAP, OR VALVE

The flow pattern of the gas and liquid across the trays can be either cross flow or counter flow, the cross flow is more widely practiced which is considered to be the flow pattern inside the column. There are three main types of trays which are: sieve, bubble cap, and valve trays with down comer. The comparison between the types of trays is based on the capacity, efficiency, entrainment, and pressure drop. For sieve and valve trays, all comparison factors between them are almost the same. While for bubble cap trays in comparison with sieve and valve, bubble cap has lower capacity and efficiency, but higher pressure drop and entrainment. An enhanced gas-liquid contact can be achieved having bubble caps on the tray at very low liquid flow rates, which are not achievable using sieve trays. However, sieve trays are the cheapest type, and suitable for almost all applications. As a result, it is chosen to design a plate distillation column with sieve trays.

9.6.4 DESIGN OF DISTILLATION COLUMN (T-101):

Assuming hydrogen peroxide: heavy key, HK

Assuming water: light key, LK

1- **Dew point calculations: Dew point**

   By trial and error using excel sheet, appendixes, the dew point (top) temperature value was found to be 96.3 °C, at the pressure 1 atm.

2- **Bubble point calculations: bubble point**

   By trial and error using excel sheet, it was found that the bubble point (bottom) temperature equals to 137.2 °C, at the pressure equals 1atm bar

3- Minimum reflux ratio: R_{min}

The value of minimum reflux ration is calculated by using the correlation from “Applied process design for chemical and petrochemical plant 3 edition “as:

$$R_{min} = \frac{1}{(XFLK)eff((\alpha LK)average-1)} \ldots\ldots\ldots (53)$$

$$(X_{FLK})_{eff} = \frac{XLKF}{XLKF+XHKF} \ldots\ldots\ldots (54) \Rightarrow (X_{FLK})_{eff} = \frac{0.886}{0.886+0.032} = 0.965$$

$\boldsymbol{\alpha}\mathbf{LK},\mathbf{avg} = \sqrt{\alpha\mathrm{LK,D} * \alpha\mathrm{LK,W}} \ldots\ldots (55)$, first relative volatility is at the top temperature, and the second is at the bottom temperature. αH_2O in the top= 2.68 ,in the bottom =5.11

then:$\boldsymbol{\alpha}\mathbf{LH},\mathbf{avg} = \sqrt{2.68 * 5.11} = 3.7$

$$R_{min} \frac{1}{0.965(3.7-1)} = 0.38$$

4- Actual reflux ratio: R

As a rule of thumb, the optimum reflux ratio (actual) for the distillation column is 1.5 of the minimum reflux ratios.

$$R_{Actual} = 1.5 * 0.38 = 0.5748$$

5- Minimum number of stages: N_{min}

It is estimated after applying certain values in Fenske equation [80].

$$N_{min} = \frac{\log(\frac{XLD * D}{XHD * D} * \frac{XHW * W}{XLW * W})}{log(\alpha LH, avg)} \ldots\ldots\ldots (56)$$

XLD: mole fraction of light key in distillate = 0.915

XHD: mole fraction of heavy key in distillate =0.00039

XHW: mole fraction of heavy key in withdrawn =0.970

XLW: mole fraction of light key in withdrawn = 0.0091

Back to the equation:

$$N_{min} = \frac{\log(249774)}{\log(3.7)}$$

N_{min}= 9.5 stages

6- Theoretical number of stages: $N_{theoretical}$

Gilliland method is used to obtain the theoretical number of stages needed to the process [80]

$$\frac{N - N_{min}}{N + 1} = 1 - \exp\left(\frac{1 \ + \ 54.4\varphi}{11 \ + \ 117.2\varphi} * \frac{\varphi - 1}{\varphi^{0.5}}\right) \dots \dots \dots (57)$$

$$\varphi = \frac{R - R_{min}}{R + 1} \dots \dots \dots (58)$$

$$\varphi = \frac{0.5748 - 0.38}{0.5748 + 1} = 0.1216$$

N (# of theoretical stages) after substitute the parameter in above equation = 18.84 stages

7- column efficiency: E_o

It is found by using the following relation [63]

$$E_o = 0.492 * \left(\mu_{avg} * \left(\alpha \frac{LK}{HK}\right)_{avg}\right)^{(-0.245)} \dots \dots \dots (59)$$

E_o: column efficiency

μ_{avg}: average viscosity of the feed mixture, and it is a summation of multiplying the mole fraction by the viscosity of each component in the feed stream, and equals to 0.1196, note that the values of the viscosity were found at 20^{o}C due to lack of sources for its value at required temperature (T=96^{o}C).

back to the efficiency relation,

$$E_o = 0.492 * \left((0.1196 * 3.7)\right)^{(-0.245)} * 100$$

Eo = 60 %

8- Actual number of stage: N_{real}

It can be found by dividing the theoretical number of plates over the efficiency of the column as follows:

N = $N_{theoritical}$ / Eo ……….(60)

Theoretical stages is assumed to be 18.84 -1 because of taking the reboiler as an equivalent stage

$$N = \frac{18.84 - 1}{0.60}$$

N_{real} = 29.7 real stages

9- Feed stage location:

Kirkbride method is used for the estimation of feed stage location [80]

$$\log\frac{Ne}{Ns} = 0.206 \log \left[\frac{\mathrm{XH}f}{\mathrm{XL}f} * \frac{W}{D} * \left(\frac{XLW}{XHD}\right)^2 \right] \ldots\ldots\ldots.(61)$$

Ne: number of stages in the enriching section (top section)

Ns: number of stages in the stripping section (bottom section)

XHF: mole fraction of heavy key in the feed , **XLF**: mole fraction of light key in the feed

By substituting the right hand of the equation,

$$\log\frac{Ne}{Ns} = 0.206\log(\frac{0.032}{0.886} * \frac{91.8}{2775.282} * [\frac{0.009}{0.00039}]^2) \ = \ -0.038$$

$$\frac{Ne}{Ns} = 10^{-0.038} = 0.916 \dots \dots \dots (62)$$

Ne + Ns = 29.7………(63)

After solving the two equations (62 & 63), we get:

Ne = 14.2 , Ns =15.5

The feed plate is determined by adding 1 to the number of enriching plates or subtracting 1 from the number of stripping plates.

So that, the feed is on stage 15.2

Tray Dynamics

a. **For top section:**

1- **Flow parameter: F_{LV}**

$$F_{LV} = \frac{LN}{VN} * \left(\frac{\rho V}{\rho L}\right)^{0.5} \dots \dots \dots \dots . (64)$$

LN: top section liquid flow rate, LN = R*D = 0.57 *2775.8 = 1595.39 Kmol/hr.

VN: top section vapor flow rate, VN= D*(1+R) =2775.8*(1+0.57) = 4370.68Kmol/hr

$\boldsymbol{\rho l}$: liquid density of the top stream, it is calculated by getting the summation of the multiplication of the density by the mole fraction of each component in the top stream, and it is found to be 923.757 Kg/m^3

$\boldsymbol{\rho v}$: vapor density of the top stream, it is calculated by getting the summation of the value of the ideal gas law substitution for each component in the top stream $\frac{P\ MW}{R\ T}$, where P is the total pressure in the top stream which is equal to 1 atm , Mw is the molecular weight of the component, R is the gas constant, and T=96 ^{o}C , is the dew point temperature, and it is found to be 0.95 Kg/m^3

Back to the relation, [63]

$$F_{LV} = \frac{1595.39}{4370.68} * (\frac{0.95}{923.757})^{0.5}$$

F_{LV} = 0.0117

2- Net vapor gas velocity at the flooding conditions: Vnf

It can be calculated as follows [80] ,

$$V_{nf} = K * (\frac{\rho_l - \rho_v}{\rho_v})^{0.5} \ \ldots \ldots \ldots (65)$$

Back to figure 11.27,

According to Chemical Process engineering design and economics, table 6.25 the tray spacing for distillation operating pressure (1atm) = 0.45 m

X-axis is the flow parameter, which is calculated before, FLV = 0.0117 ~0.012

Y-axis: is the intersection of the flow parameter value with the tray spacing, which gives the value of K = 0.08

Back to the equation,

$$V_{nf} = 0.08\ (\frac{923.757 - 0.95}{0.95})^{0.5}$$

$$V_{nf} = 12.4\ \ m/s$$

3- Actual vapor velocity: V_n

The actual vapor velocity is determined by assuming that it is 70-90% of the net vapor velocity at flooding conditions, and it is selected to be 75 % of the net vapor velocity at flood conditions [63].

$$V_n = 0.75 * 12.4$$

$$V_n = 9.3\ m/s$$

4- Net area: A_n

It is calculated as follows [63] ,

$$A_n = \frac{VN * Mw}{Vn * 3600 * \rho v} \dots\dots\dots (66)$$

VN: vapor flow rate at the top section, kmol/hr

Mw: molecular weight of the top stream mixture, g/mol

Vn: actual vapor velocity, m/s

$\boldsymbol{\rho v}$: $vapordensiy, kg/m^3$

Back to the relation,

$$A_n = \frac{4370.68 * 29.07}{9.3 * 3600 * 0.95}$$

$A_n = 3.98\ m^2$

5- Cross sectional area of the top of the column: A_c

There is a relation between the cross-sectional area, the net area of the column, and the down comer area, which is as follows,

$A_c = A_n + A_d$ ……..(67) but it is assumed that the down comer occupies 10% of the cross-sectional area of the column ($A_d = 0.1*A_c$) [63]

So that,

$$A_c = 3.98 + 0.1 * A_c$$

$$A_c = 4.42\ m^2$$

6- Diameter of the column:

D_c : The diameter of the column can be calculated as

$$D_c = \left(\frac{4 * A_c}{\pi}\right)^{0.5} \ldots\ldots\ldots (68)$$

$$D_c = \left(\frac{4 * 4.42}{3.14}\right)^{0.5}$$

$$D_c = 2.37\ m$$

7- Liquid flow arrangement:

To determine the plate type, and the flow pattern of the fluid inside the column, figure 11.28 [80] can be used.

The X-axis of the figure is the column diameter Dc the Y-axis of the figure is the liquid volumetric flow rate of the top stream L

$$L = \frac{LN * \text{Mw}}{\text{ρl} * 3600} \ldots\ldots\ldots (69)$$

$$L = \frac{1595.39 * 29.07}{923.75 * 3600}$$

$$L = 0.009\ m^3/s\ (\text{X} - \text{axis})$$

Back to figure 11.28,

The intersection point between Dc= 2.3, and L= 0.009, is inside the region of cross flow (single pass). So that, it has selected single pass plate.

8- Flooding check:

It can be calculated as follows,

Percent flooding = U_n/V_{nf} ……….(70)

U_n: actual velocity, m/s

V_{nf}: flooding velocity, m/s

The actual velocity is calculated as,

$$U_n = \frac{Q_{max}}{A_n} \ldots\ldots\ldots (71)$$

Q_{max}: maximum volumetric flow rate

$$Q_{max} = \frac{VN * Mw}{\rho * 3600} \ldots\ldots\ldots. (72)$$

$$Q_{max} = \frac{4370.68 * 29.07}{0.95 * 3600} = 24.37\ m^3/s$$

$$U_n = \frac{24.37}{3.98} = 6.12\ \mathrm{m/s}$$

$$\text{Percent flood} = (6.12/12.4) * 100\%$$

Percent flood = 49% and according to figure 11.29, it can be observed that we are at the satisfy region, since fractional Entrainment is less than 0.1 [80]

9- Estimation of weep point:

To be able to detect whether there is weeping problem in the column or not (or the process satisfied or not), the following condition must be taken into consideration.

Actual minimum vapor velocity > U_{min}

U_{min}: minimum design vapor velocity through the holes

Actual minimum vapor velocity = 0.7*Q_{max} /Ah…….(73) , [104]

Aa: active area = Ac – 2*Ad…….(74) , **Aa** = 4.4 -2*0.44= 3.54 m^2

Ah: hole area = 0.1*Aa ………(75) , **Ah** =0.1*3.54 =0.354 m^2 (total area of tray holes)

Back to the actual minimum vapor velocity

$$\text{Actual minimum vapor velocity} = \frac{0.7 * 24.37}{0.354} = 48.18\ \mathrm{m/s}$$

$$U_{min} = \frac{K_2 - 0.9(25.4 - dh)}{\rho v^{0.5}} \dots \dots \dots (76)$$

dh: the hole diameter, and the recommended value is 5 mm

K_2: constant, can be found from figure11.30 [80] where X-axis**:** $h_w + h_{ow}$

h_w: weir height, and it is selected to be 50 mm, since it is the recommended value for atmospheric operations [99]

h_{ow}: weir liquid crest, and it can be calculated as follows,

$$h_{ow} = 750 * (\frac{Lw}{\rho l * lw})^{\frac{2}{3}} \ldots\ldots\ldots\ldots (77)$$

Lw: liquid flowrate = 8.3 kg/s

lw: weir length, From figure 11.31 its equal 1.73 m

back to the relation,

$$h_{ow} = 750 * (\frac{8.3}{923 * 1.73})^{\frac{2}{3}} = 22.5 \text{ mm}$$

the point, where the X-axis ($h_w + h_{ow}$ = 72.7 mm) meets the curve of figure11.30, is K_2 value (K_2=30.6)

Back to U_{min} relation,

$$U_{min} = \frac{30.6 - 0.9(25.4 - 5)}{0.95^{0.5}} = 12.66\, m/s$$

Since U_{min} < actual minimum vapor velocity there is no weeping in the system, and the process is satisfied.

10- Number of holes for one tray:

Number of holes = A_h/ a_h………(78)

A_h: total area of all holes in one tray, m^2

a_h: area of one hole, m^2 and it equals to $(\pi/4) * Dc^2$ = (3.14/4) *$(2.37)^2$ = 0.000019 m^2

Then, number of holes = 0.35 /0.000019 = 18041. 27 holes

11- Hole pitch:

Hole pitch area is calculated as follows [80] ,

Ap = $\boldsymbol{Ah}/(\ \mathbf{0.9}* (\boldsymbol{dh}*\boldsymbol{Ip})^{\mathbf{2}})$..........(79)

Ip = 3*dh …….(80) , it is assumed to be not less than 2.0 of hole diameter [80]

$A_p = 0.35 / (0.9 * (1/3))^2$

$A_p = 3.54\, m^2$

12- Dry tray pressure drop:

It can be calculated as follows,

$$H_d = 51 * \left(\frac{U_n}{C_o}\right)^2 * \frac{\rho_v}{\rho_l} \dots\dots\dots (81)$$

U_n: hole velocity = Q_{max}/A_h ……..(82), **U_n** = 24.37 /0.35 = 68.8 m/s

C_o: is a constant, estimated from figure 11.34 [80]

X-axis of figure 11.34: (Ah/Ap) *100% = 10%

For stainless steel the recommended plate thickness is 2.77 mm [100]

$$\mathrm{C_o} = 0.74 * \frac{\mathrm{Ah}}{\mathrm{Aa}} + \ \exp\ \left(0.99 \frac{\mathrm{t}}{\mathrm{dh}} - 0.56\right) \dots\dots..(83)\ \ [101]$$

Where t is the plate thickness , then $C_o = 0.74 * \frac{0.35}{3.54} + \exp\left(0.99 \frac{2.77}{5} - 0.56\right)$

$C_o = 0.74$

Back to the dry tray pressure drop relation:

$$H_d = 51 * \left(\left(\frac{68.8}{0.74}\right)^2\right) * \left(\frac{0.95}{923.757}\right) = 453.8 \text{ mm liquid}$$

13- Residual head [80]:

$$H_r = \frac{12500}{\rho_l} \dots \dots \dots (84)$$

$$H_r = \frac{12500}{923.757} = 13.53\, mm\ liquid$$

14- Plate pressure drop:

$$\boldsymbol{H_t = H_d + H_r + (h_w + h_{ow})} \dots \dots \dots (85)$$ **[80]**

$$\boldsymbol{H_t} = 453.8 + 13.53 + 72.7$$

$$\boldsymbol{H_t} = 539.827 \text{ mm liquid}$$

Plate pressure drop in Pa is calculated as :

$$\Delta Pt = 9.81 * 10^{-3} * H_t * \rho_l \dots \dots \dots (86)$$

$$\Delta Pt = 9.81 * 10^{-3} * 539.827 * 923.757$$

$$\Delta Pt = 2403.35\, Pa = 2.403 KPa$$

15- Column height:

The height of the column can be determined using the following equation [63] :

$$H_c = (N_{\text{act}} - 1) * \text{Hs} + \Delta\text{H} \dots \dots \dots (87)$$

N_{act}: actual number of plates

H_s: plate spacing = 0.45

ΔH:it is assumed to be 0.3 to 0.15 of H_c , and it is selected to be 0.15 of H_c

By solving the above equation,

H_c = 15.17 m

And in the same procedure the above calculations are done again for the bottom section of the column. However, when the pressure of the bottom section is calculated, the top section pressure is added to the multiplication of the number of plates by the pressure drop across the column.

9.6.5 DETAILED DESIGN SUMMARY

Table 36 Specification data for T-101

Column type	Plate Multicomponent Distillation Column
Tray type	Sieve tray
Operation	Continuous
Function	To separate H_2O_2 from other components
MOC	Stainless Steel 304
Light Key/Heavy Key	H_2O/ H_2O_2
Operation Pressure (atm)	1
Dew point°C	96
Bubble point°C	137
Actual number of trays	29.6
Feed stage location	15.2
Column efficiency (%)	60
Column height (m)	15
Column diameter(m)	2.37
Reflux ratio	3.57
Flooding percent (%)	49
Number of holes	18041.27
Hole area (m^2)	0.35
Hole diameter(mm)	5
Active area(m^2)	3.54
Downcomer area(m^2)	0.44
Cross sectional area(m^2)	4.47
Pressure drop(Kpa)	2.4
Cost $	252,660

Table 37 Specification data for T-102

Column type	Plate Multicomponent Distillation Column
Tray type	Sieve tray
Operation	Continuous
Function	Recovery of Methanol from H_2O
MOC	Stainless Steel 304
Light Key/ Heavy Key	CH_3OH/ H_2O
Actual number of trays	14.2
Column efficiency %	60
Column height (m)	6.9
Column diameter (m)	1.89
Flooding percent (%)	54
COST $	190,336

Table 38 Specification data for T-103

Column type	Plate Multicomponent Distillation Column
Tray type	Sieve tray
Operation	Continuous
Function	Separation of HCHO from H_2O
MOC	Stainless Steel 304
Light Key/ Heavy Key	HCHO/ H_2O
Actual number of trays	26
Column efficiency %	65
Column height (m)	13.2
Column diameter (m)	1.46
Flooding percent (%)	72.26
Cost $	194,787

9.7 STORAGE TANK DESIGN

9.7.1 INTRODUCTION

Storage tanks are containers that hold liquids, compressed gases or any medium used for the short or long-term storage. Industrially, storage tanks are come in different shapes and orientation, the most common shape is cylindrical. Consequently, it is chosen to be the shape for the storage tanks in the process. Storage tank can either be underground, horizontal or vertical, and since the liquid content stored in ST-101 and ST-102 are considered high then the tanks orientation is vertical mounted on a concrete foundation. According to the corrosivity of liquids inside them, the material of constriction is considered to be stainless steel 304.

Figure 42 storage tank

9.7.2 HYDROGEN PEROXIDE STORAGE TANK DESIGN

Tank Volume calculation: [105]

1 day storage capacity of H_2O_2 = $\frac{mass}{density} \ldots \ldots \ldots .(88)$

The density of the mixture at ambient temperature = 1077.43 kg/m^3

Mass flow rate (kg/day) =1463894.7

Volume of liquid need to be stored on one day = $\frac{1463894.75kg}{1077.43498\ \mathrm{kg/m3}} = 1358.685\ m^3$

The Volume of storage tank is =Volume of liquid * free space on tank

=1358.685 *1.1

= 1494.5 m^3

This volume of tank is considered high, so two storage tanks are needed, the volume of one tank is =1494.5/3= 498.2 m^3

Tank dimensions:

Assume:

L/D =3

$$v = \frac{\pi}{4} D^2 L \ldots \ldots \ldots (89)$$

$$D = \sqrt[3]{\frac{4\,Volume}{1.5\ D}}$$

D = 10.9 m

L = 3*10.9 = 32.7 m

Shell type and thickness calculation:

For vertical cylindrical shell.

$$t_{Shell} = \frac{Pr_i}{sE_j - 0 \cdot 6p} + CC \dots \dots \dots . (90)$$

r$_i$: 5.45m

P: 101.3 KPa

S: Maximum allowable working stress. KPa = 165000 KPa for stainless steel (From table 12-10)

E $_j$: Efficiency of joints expressed as a fraction = 0.85 for spot examined weld (From table 12-10) [105]

Cc: Allowance for corrosion = 0.002

$$t\ shell = \frac{101.3 \times 5.45}{(165000 \times 0.85) - (0.6 \times 01.3)} + 0.002 = 0.00594m = 0.0594\ cm$$

Head type and thickness calculation:

for hemispherical spherical head Shell thickness

$$t = \frac{0.885PLa}{SE_j - 0.1p} + cc \dots \dots \dots (91)$$

Where: La = ri (m)

$$t = \frac{0.885 \times 101.3 * 2.8}{(165000 \times 0.85) - (0.1 \times 101.3)} + 0.002 = 0.005486754m = 0.05486754cm$$

9.7.3 DETAILED DESIGN SUMMARY

Table 39 Specification Sheet of ST-102

Identification	
Name of Equipment	Storage Tank
Equipment shape	cylindrical
Number of units required	3
Function	To store hydrogen peroxide
Material property	
Volume of one tank m^3	498.18
Operating temperature (℃)	15-30 [106]
Operating pressure (atm)	1
Residence time (day)	1
Design specifications	
Material of construction	304 Stainless steel
Diameter (m)	10.9
Length (m)	32.7
Shell shape	Cylindrical shell
Orientation	Vertically mounted on a concrete foundation
Shell thickness (cm)	0.059412589
head Type	Hemispherical head
head thickness (cm)	0.05486754
Cost	455,000

Table 40 Specification Sheet of storage Tank ST-101

Identification	
Name of Equipment	Storage Tank
Equipment Shape	Cylindrical
Number of units required	1
Function	storage of 27% Formaldehyde
Material property	
Volume of one tank (m^3)	305.6
Operating temperature (℃)	25
Operating pressure (atm)	1
Residence time (day)	1
Design specifications	
Material of construction	304 Stainless steel
Diameter (m)	6.4
Length (m)	19.2
Shell shape	Cylindrical shell
Orientation	Vertically mounted on a concrete foundation
Shell thickness (cm)	0.43
head Type	Hemispherical head
head thickness (cm)	0.45
Cost $	182,000

9.8 INCINERATOR DESIGN

9.8.1 INTRODUCTION

The incineration is considered the best method of industrial waste disposal, there are two type of it which are Knock-out and Flare, the knockout is used to knock out the free liquid from the gaseous mixture while flare is used when the gaseous volatile organic compounds, which are considered dangerous to the environment, have to be removed. In this project the flare is selected as the used type of the incinerator since it removes gaseous mixture, that produce from the top of V-101, which are traces of oxygen, hydrogen peroxide, methanol as well as formaldehyde.

9.8.2 SAMPLE CALCULATION FOR INCINERATOR DESIGN

The input stream to the incinerator is stream M_7 which is composed of methanol and formaldehyde as volatile organic compounds.

$$B_v\text{: heating value of vent streams} = \sum B_{vi} * X_i = B_{v,CH_3OH} * X_{CH_3OH} + B_{v,HCHO} * X_{HCHO} \ldots (92)$$

$$B_v = 726\frac{KJ}{mol} * 0.16 + 563\frac{KJ}{mol} * 0.04, \text{the heat value of the components is found at T} = 25\text{C}$$

Components	Mole%	Mw	Average mw	Density	X*Density
HCHO	0.04	30	1.2	815	32.6
CH_3OH	0.16	32	5.12	1.2	0.192
			6.32		**32.792**

$$B_v = 138.68\frac{KJ}{mol} * \frac{mol}{6.32g} * \frac{1\,Btu}{1.05506KJ} * \frac{32.792Kg}{m^3} * \frac{1000g}{Kg} * \frac{2.063 * 10^{-3}m^3}{1scf}$$

$$B_v = 1407 \text{ Btu/Scf}$$

Since the heat value is greater than 1000Btu/Scf, then the maximum allowed velocity, V_{max} = 400 ft/s [104]

B_f: heat content of auxiliary fuel, and assumed to be 920 Btu/Scf [108]

1. Flare tip diameter:

$$\mathrm{D_{min}} = 1.95\sqrt{\frac{\mathrm{Q_{tot}}}{\mathrm{V_{max}}}} \dots \dots \dots . (93)$$

D_{min}: minimum flare diameter (inch)

Q_{tot}: total volumetric flowrate (acfm)

V_{max}: maximum allowable velocity (ft/s)

Q: flow rate of the waste gas stream (Scfm)

$$Q = \text{flowrate of stream } (M_7) = \frac{63.25Kmol}{hr} * \frac{6.32Kg}{Kmol} * \frac{m^3}{32.792Kg} * \frac{1Scf}{2.06 * 10^{-3}m^3} * \frac{1hr}{60min}$$

Q = 98.626 Scfm

$$\mathrm{D_{min}} = 1.95 * \sqrt{\frac{Qtot}{Vmax}} = 1.95 * \sqrt{\frac{98.626}{400}} = 0.968 \text{ inch} = 0.08 \text{ ft} = 0.0245\text{m}$$

1. Flare height: [48]

$$L^2 = \frac{\tau * f * R}{4 * K} \dots \dots \dots (94)$$

τ: fraction of heat intensity transmitted constant equals to 1

f: fraction of heat radiated, constant equals to 0.2

R: net heat release (Btu/hr)

K: maximum allowable radiation (500 Btu/hr-ft^2)

Back to the relation,

$$L^2 = \frac{1 * 0.2 * 833826.35}{4 * 500} = 83.38 ft^2$$

$\mathrm{L} = 9$ ft,

As the stack height less than the smallest commercially available height which is 30 ft**,** the height is taken as 30 ft.

9.8.3 DETAILED DESIGN SUMMARY

Table 41 Specification sheet for Flare

Name of unit	**Flare**
Function	Remove gaseous volatile organic compounds
Minimum diameter (ft)	0.08
Flare stack height (ft)	30
Cost $	13,906.5

9.9 COMPRESSOR DESIGN

9.9.1 INTRODUCTION:

Compressor is an equipment that is used in chemical plant to raise the pressure of gases. There are many types of compressors in industrial scale, the most common are: centrifuge, axial flow and reciprocating. The selection between them is mainly done either by the change of pressure between inlet and outlet stream or the pressure for the outlet stream. Referring to the plant design course slides for Dr. Mohammad AL Azzam, when the change in ΔP is less than 0.5 Psig then axial flow is chosen while for centrifugal compressor, P_{out} must be less than 50 Psig, the least type which is reciprocating can deliver pressure up to 4000 atm, Since the outlet pressure of the compressor(C-101) is 161 Psig then the chosen type is reciprocating. Reciprocating compressor can achieve its work either by single or multistage, in this process is chosen to operate reciprocating with multistage since it improves the volumetric efficiency as well as it reduces the leakage loss from the system. [116], According to "plant design and economy chapter 12" the actual process of compressor is followed polytropic system, work of compressor in real system does not occur at constant entropy, so compressor calculation will be based on polytropic equations.

The main function of C-101 is to increase the pressure of unreacted oxygen that leaves the reactor from 8 bar to 11.15 bar and recycle it back to R-101.

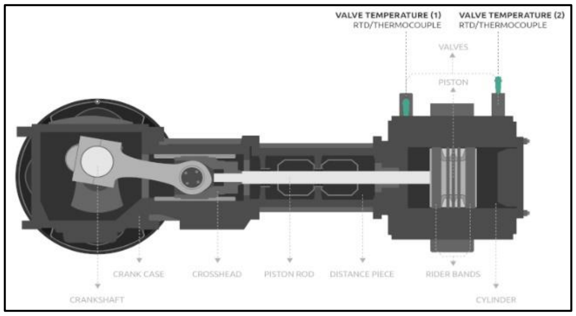


Figure 43: Reciprocating compressor

9.9.2 SAMPLE CALCULATION FOR COMPRESSOR DESIGN

Total mass flow rate =0.0028 Kmol/s

T_{inlet} = 90.2℃ , P_{in} = 8 atm , P_{out} = 11.15 atm

1. **Compressibility factor Z:**

Z(O_2) =0.993 [49]

Z(CH_3OH) =1.76*10^{-2} [111]

Z=Σ (Zi Xi) =9.95*10^{-2}

2. **Heat capacity ratio:**

K = Cp / Cv……..(95)

K(O_2)=1.3973 [112]

K(CH_3OH)=1.2188 [113]

$K_{mixture}$ = K =1.39548908

$$\eta = \frac{\frac{K-1}{k}}{\frac{n-1}{n}} \dots\dots\dots\dots (96)$$

η = 0.75, For reciprocating compressor, polytropic efficiency range is 70-85% [115]

then n after substitute K=1.4 and $\eta = 0.75$ is equal 1.6 .

3. Theoretical power

$$P = m\, R_1 T_1 Z\, [\left(\frac{P_2}{P_1}\right)^{\frac{n}{n-1}} - 1](\frac{n}{n-1}) \dots\dots\dots.(97)$$

Where:

P: Polytropic theoretical power Kw

m: total mass flowrate kg/s

R_1: gas constant =0.2596 kJ/kg. k

T_1: inlet temperature = 90.2℃

Z: compressibility factor =$9.95*10^{-2}$

P_2: outlet pressure =11.15 bar

P_1: inlet pressure =8 bar

n: polytropic exponent =1.6

Substitute all the parameters in the power equation then,

Theoretical power of is = 3 kw

Actual power = $\frac{\text{Theoretical power}}{\text{Effeciency}} \dots \dots \dots (98)$

$$Actual\ power = \frac{3}{0.75} = 4kw$$

4. Discharge temperature

T_2 : discharge temperature $= \mathrm{T_1} \left(\frac{\mathrm{P_2}}{\mathrm{P_1}}\right)^{(\mathrm{n}-1)/(\mathrm{Nst}*\mathrm{n})} \dots \dots \dots (99)$

Nst is the number of stages in compressor which is equal 2 then T_2= 96 ℃

9.9.3 DETAILED DESIGN SUMMARY

Table 42 Specification Sheet of compressor C-101

Identification	
Name of Equipment	Compressor
Function	To increase the pressure of unreacted oxygen .
Material property	
Inlet temperature (°C)	90
Inlet pressure (atm)	8
Outlet temperature (°C)	96
Outlet pressure (atm)	11.15
Design specifications	
Material of construction	Carbon steel
Multistage/Single stage	2 stage
Compressor efficiency	0.75
Actual power Kw	4

9.10 PUMP (P-101) DESIGN

9.10.1 INTRODUCTION

Pumps are used to transfer fluids from one location to another by increasing their pressure, they require power as an external source to accomplish their function such as electric energy.

There are many types of pumps where the selection between them requires fully information about the liquid to be handled such as fluid temperature, viscosity and density. As well as the total dynamic head, suction and discharge head. These type are:

1. Centrifugal pumps
2. Positive displacement pumps

3. Jet pump
4. Electromagnetic pumps

In this project the centrifugal pumps are used because the following reasons: [103]

1. Deal with large volume
2. Small, Space &less capital cost
3. Operate at high speed so that they can be driven directly by electrical motors. (Sinnott R. K., 1996)

Figure below shows the cross section of centrifugal pump.

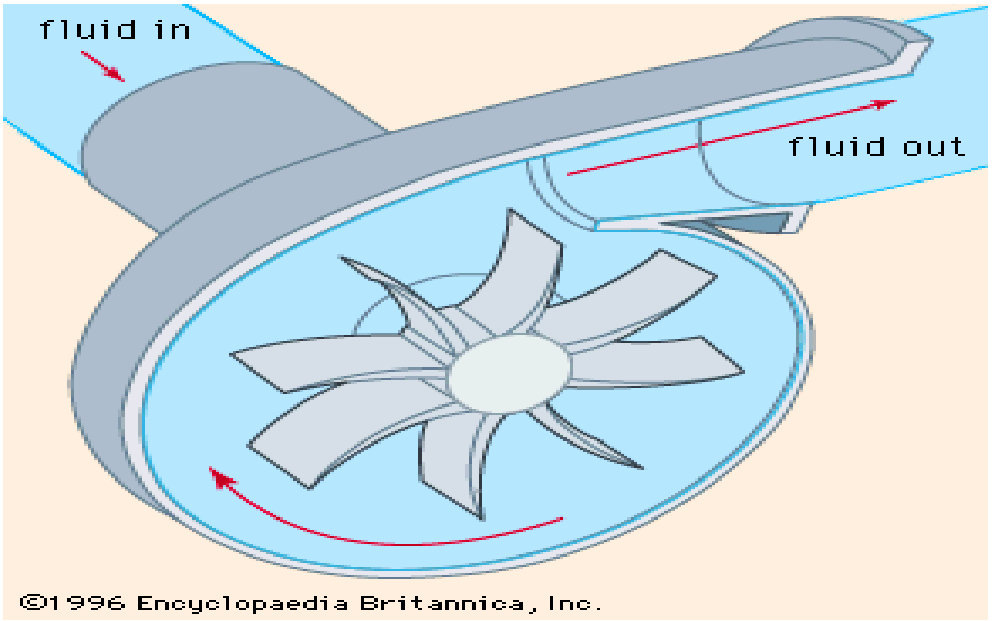


Figure 44 Centrifugal pumps[102]

9.10.2 SAMPLE CALCULATION FOR PUMP DESIGN P-101 A/B

P_1 = 1.22 atm = 123616.5 Pa

P_2 = 6 atm =607950 Pa

Average $MW_{13} = \sum Xi * Mwi = 28.42 \text{ Kg/Kmol}$

Average density $= \sum X_i * \rho = 255.26\text{Kg}/m^3$

Flow rate of M_{13} =164.4 Kmol/hr

Average viscosity of $M_{10} = \mu = \sum Xi * \mu i = 0.00122 \text{ Pa.s}$

$$\text{Volumetric flow rate of } M_{13} = \frac{Mole\ flow\ rate\ of\ stram\ M13 * Mw\ of\ M13}{density\ of\ M13} \ldots \ldots (100)$$

$$\text{Volumetric flow rate of } M_{13} = \frac{164.4 * 28.42}{255.26} = 18.30m^3/\text{hr} = 0.30m^3/s$$

Assume turbulent flow then:

$$D_{i,opt} = 0.363\ \rho^{0.13} * V^{0.45} \ldots \ldots (101)$$

$$= 0.363\ 1175.2^{0.13} * 0.044^{0.45} = 0.44\text{m}$$

Now check Re:

$$\text{Velocity} = \text{V} = \frac{Q}{A} \ldots \ldots \ldots (102)$$

$$= \frac{Q}{\left(\frac{\pi}{4}\right) * (Di\ out)^2} = \frac{0.044}{\left(\frac{\pi}{4}\right) * (0.44)^2} = 2.032\ m/s$$

$$Re = \frac{\rho\ V\ D}{\mu} = \frac{255.26 * 2.032 * 0.44}{0.00122} = 185826.42 > \ 2100\ , then\ the\ assumption\ is\ true.$$

Assume nominal pipe size (NPS) = 6 inch and Schedule # = 40, then from “plant design and economics for chemical engineering” appendix D-13:

DI = 6.065 in = 0.154051 m

OD = 6.625 in = 0.168275 m

Recalculate Re and Velocity for the true value of DI=0.1540 m

$$\text{Velocity} \ = \ \text{V} = \frac{Q}{A} = \frac{Q}{\left(\frac{\pi}{4}\right) * (DI)^2} = \frac{0.30}{\left(\frac{\pi}{4}\right) * (0.15405)^2} = 16.37\ \text{m/hr} \ = 0.019\ \text{m/s}$$

$$Re = \frac{\rho\ V\ D}{\mu} = \frac{255.6 * 16.37 * 0.1540}{0.00122} = 527534.65$$

Mechanical design equation:

$$W \ = \text{g}\Delta Z + \frac{\Delta P}{\rho} + \sum F + \frac{(\Delta V)^2}{2\alpha} \dots \dots \dots (103)$$

$\text{g}\ \Delta Z = \ 0$ (No change of high between two pumps)

$\alpha = 1$, (For turbulent flow)

$\frac{(\Delta V)^2}{2\alpha} = 0$ (assume no change in inlet and outlet velocities of the pump)

Mass flow rate of M_{10} =4672.6 Kg/hr =77.87 Kg/s

$$\frac{\Delta P}{\rho} = \frac{607950 - 123616.5}{255.26} = 1897.43\ \text{J/Kg} = \ 1897.43 * 77.87 = 1447766.47\ \text{W}$$

$$\sum F = F_{St.Pipe} + F_{\text{elbow + valves}} + F_{\text{contraction}} + F_{\text{expansion}} \ldots\ldots\ldots\ldots\ldots\ldots.(104)$$

$$F_{St.Pipe} = \frac{2f(V)^2L}{D} \ldots\ldots..(105)$$

Assume L=20 m

Roughness for Stainless steel ε = 0.000046

$$\frac{\varepsilon}{D} = \frac{0.000046}{0.154051} = 0.0003$$

Then from Moody chart f= 0.004

$$F_{St.Pipe} = \frac{2 * 0.004 * (16.3)^2 * 20}{0.1540} = \frac{278.56\text{J}}{\text{Kg}} = 278.561 * 77.87 \ = \ 21693.6 \text{ W}$$

$$F_{\text{elbow + valves}} = \frac{2fV^2Le}{D} \ldots\ldots\ldots(106)$$

Assume there is four elbows (90 std. radius) and two gate valves then Le/D =4*32+2*7=142

$$F_{\text{elbow + valves}} = \frac{2 * 0.004 * 16.3^2 * 142}{1} = 304.68\frac{\text{J}}{\text{Kg}} = 304.68 * 77.87 = \ 23727.6 \text{ W}$$

$$F_{\text{contraction}} = \frac{Kc\ v^2}{2\alpha} \ldots\ldots\ldots(107)$$

$\text{K}_\text{c} = 0.4 * \left(1.25 - \frac{\text{A}_1}{\text{A}_2}\right)$ ……..(108) , A_1/A_2 = 0 because A1 larger than A_2, the term A_2/A_1 almost = $zero$, then K_c = 0.5

$$F_{\text{contraction}} = \frac{0.5 * 16.3^2}{2} = 67 \text{ J/Kg} = 67 * 77.87 = 5221.74 \text{ W}$$

$$F_{\text{expansion}} = \frac{V1^2}{2\alpha} \ldots\ldots\ldots.(109)$$

$$= \frac{16.3^2}{2} = 134.1\,\frac{\text{J}}{\text{Kg}} = 134.1 * 77.87 \ = 10443.5\ \text{W}$$

$$\sum F = F_{St.Pipe} + F_{\text{elbow + valves}} + F_{\text{contraction}} + F_{\text{expansion}}$$

$$\sum F = 21693.6 + 23727.6 + 5221.74 + 10443.5 = 61086.5\ \text{W}$$

$$\text{Theoretical Power} = \sum F + \frac{\Delta P}{\rho} \ldots\ldots\ldots(110)$$

$$= 61086.5 + 147766.47 = \ 198006.1456\text{W} \ = \ 265.53\ \text{hp}$$

Efficiency for centrifugal pump; figure (12-17) “plant design and economics for chemical engineering” is between (55- 77%)

Assume efficiency of P-101 A/B = 60 %

Then, motor size $= \frac{\boldsymbol{Theoritical\ Power}}{\boldsymbol{Efficiency}} \ldots\ldots\ldots\ldots(111)$

$$= \frac{\mathbf{265.53}}{\mathbf{0.6}} \ = \ \mathbf{442.56\ hp}$$

9.10.3 DETAILED DESIGN SUMMARY

Table 43 :Specification Sheet of P-101 A/B

Identification	
Name of Equipment	Pump
Item Number	P-101 A/B
Equipment Type	Centrifugal Pump
Drive	Electrical
Material of Construction (MOC)	Stainless Steal
Function	Raise Pressure of Stream M_{13} that produce from T-103 to 6 atm
Description	
Flow Rate (m^3/sec)	0.31
Average Density (Kg/m^3)	255.257
Average Viscosity (Pa.sec)	0.0012
Schedule Number	40
Nominal Pipe Size (inch)	6
Optimum Diameter (m)	0.4373
Velocity (m/S)	16.376
Reynolds's Number	527534.6
Pump Efficiency	60%
Motor Size (hp)	442.56
Cost $	49,680

Table 44 Specification Sheet of P-102A/B

Identification	
Name of Equipment	Pump
Item Number	P-102 A/B
Equipment Type	Centrifugal Pump
Drive	Electrical
Material of Construction (MOC)	Stainless Steal
Function	Raise Pressure that produce from P-102 to 11.49
Description	
Flow Rate (m^3/sec)	0.31
Average Density (Kg/m^3)	255.257
Average Viscosity (Pa.sec)	0.0012
Schedule Number	40
Nominal Pipe Size (inch)	6
Optimum Diameter (m)	0.4373
Velocity (m/S)	16.376
Reynolds's Number	527534.6
Pump Efficiency	60%
Motor Size (hp)	491.6
Cost ($)	49,680

CHAPTER TEN: HAZOP STUDY

10.1 INTRODUCTION

A Hazard and Operability Study investigates each element in a process. The goal is to find potential situations that would cause that element to pose a hazard; it identifies hazardous scenarios that impact receptors, people, and environment, or limit the operability of the process as a whole. Different problems could arise through a hazard study such as lack of flow in transfer line or over-pressuring in a reactor or heat exchanger.

This is accomplished by fully defining the intent of each segment and then applying guide words to each segment called nodes as follows:

- No or nor – No part of the intent is achieved, and nothing else occurs (e.g., no flow).
- More – Quantitative increase (e.g., higher temperature).
- Less – Quantitative decrease (e.g., lower pressure).
- As well as – Qualitative increase (e.g., an impurity).
- Part of – Qualitative decrease (e.g., only one of two components in a mixture).
- Reverse – Opposite (e.g., backflow).
- Other than – No part of the intent is achieved, and something completely different occurs (e.g., the flow of the wrong material).

The first step in HAZOP analysis is to define the system and subsystems, and in each subsystem, it is necessary to define nodes. These nodes will limit the assessment of subsequent process deviations and include groups of equipment, alarms, valves, and so on. Depending on the HAZOP leader definition, it's been considered the causes and safeguards into the nodes assessed,out of node assessed, or both cases.

The guide words mentioned before are applied to an appropriate parameter. Afterward, the causes, consequences, and safeguards are assessed to complete HAZOP study, these steps are summarized in Figure (19).

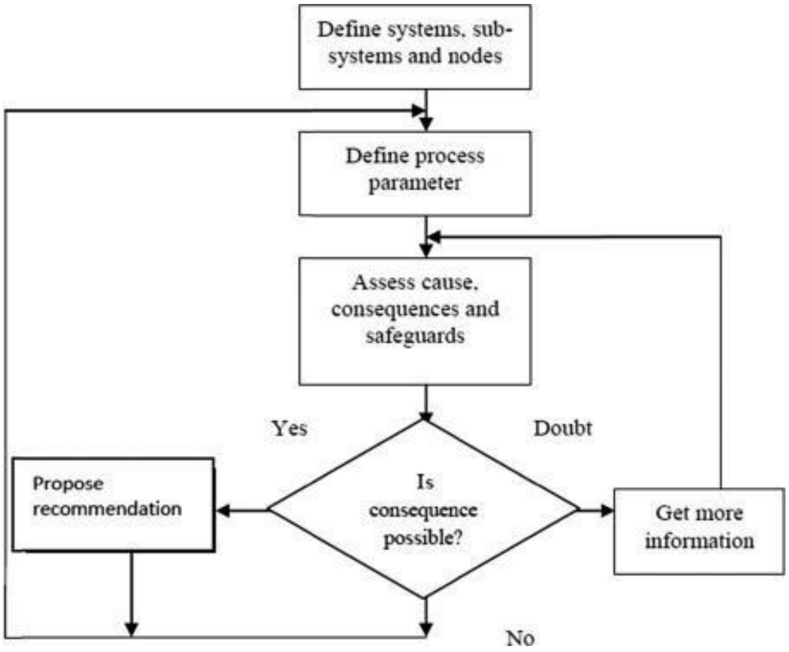


Figure 45 HAZOP steps

Hazard assessment is a vital tool in loss prevention throughout the life of the facility. Ideally, the assessment should be conducted during the conceptual design phase, final design stage, and pre-start-up period as well as when the plant is in full operation. In the conceptual design phase, many potential hazards can be identified, and significant changes or corrections made to minimize cost. Results of these assessments are key inputs to both site selection and plant layout decisions.The major hazards usually include toxicity, fire, and explosions; however, thermal radiation, noise, and various environmental concerns also need to be considered. A thorough hazard and risk assessment of a new facility is essential during the final design stage.

At this stage, the piping and instrument diagrams, equipment details, and maintenance procedures are finalized.However, since equipment often has not been ordered, it is still possible to make changes without incurring major penalties or delays. (Plant design and economy).

10.2 OBJECTIVE OF THE STUDY:

The objective of the HAZOP study was to identify the potential hazard and the operability problemsby reviewing reaction zone of the Hydrogen peroxide plant and to identify the action items to lessen such problems. The objective of the HAZOP study is:

- To provide process safety information.
- Find out of different HAZARD during plant operation.
- To establish safe operating procedures.
- To maintain safe work practice.

10.3 P&ID DIAGRAM

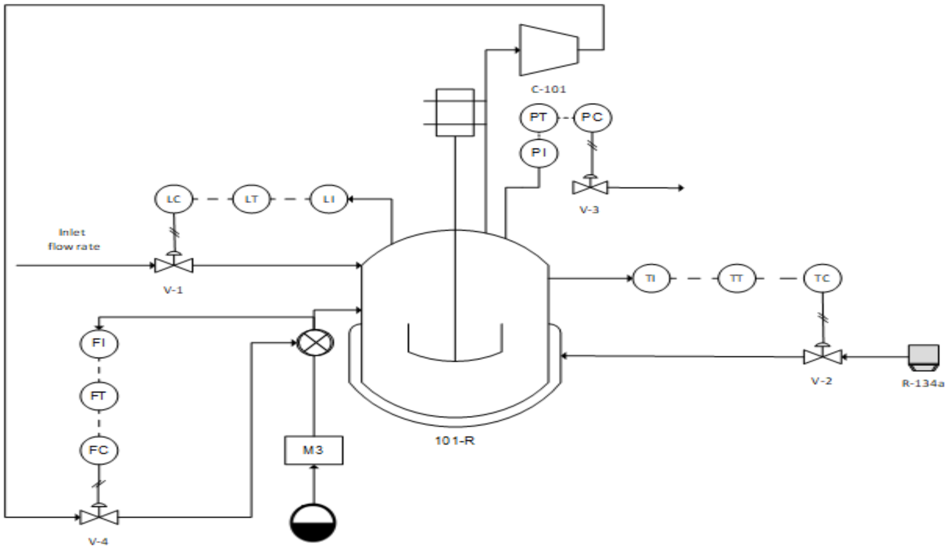


Figure 46:PID diagram for Reactor (R-101)

HAZOP worksheet					
Equipment	parameter	deviation	Potential causes	consequences	Action required
Reactor R-101	Flow rate	No	-Blockage or closing in M_3 feed pipes as well as control valve is closed.	-No reaction occurs. -Explosion at point of blockage in the pipes.	- Continuous check for inlet piping system. -Install flow control loop for continuous check of inlet flowrate. -Replace and clear blockage valves.
		less	-Leakage or partial blockage in piping system. -Controller fails and recycle valve is partially closed.	-Not enough oxygen for the reaction to take place which results in decreasing the product quantity.	-Continuous check for all inlet piping system. - Install flow control loop for continuous check of inlet flowrate. -install flow meter and low flow alarm to alert the operator.
		More	-High flow rate of M_3. -Controller fails, the control valve is fully opened or failed open.	-At high O_2 concentration, the enzyme activity is decreased so product quantity also decreased. -Reactor unit explosion. -Failing in the agitating system.	- Install flow meter and low flow alarm to alert the operator. - Install flow control loop
	Temperature	Lower	-High cooling R-134a flow rate since V-2 is fully opening.	-High reaction rate. -Production quality decreases.	- Check the reactor temperature by placing temperature indicator instrument and install control loop on the reactor.
		Higher	-Low R-134 a flow rate since V-2 is fully closed.	-Pressure build-up inside the reactor which will cause explosion.	-Use temperature indicator alarm. -Regular maintenance to the cooling jacket.

			-High reactor feed temperature due to failing in the heat exchanger.	- Rate of reaction is decreased.	
	Pressure	Lower	-Valve V-4 is failed open. -Problem in inlet and outlet gas flowrates.	-Product quality is decreased. -Chance of side reaction.	-Use pressure indicator combined with pressure controlling system and alarming system.
		Higher	-Increasing in inlet flow. -Problem in outlet gas. - Valve V-4 is failed open. -Increasing the reactor temperature	-Reactor explosion.	-Install relief valve -Use pressure indicator combined with pressure controlling system and alarming system.
	Level	High	Malfunction in inlet or outlet valve	-Change in product quality and quantity. -Flooding within the reactor.	-Continuous check for all inlet and outlet valves -Use level indicator with alarming system

CHAPTER ELEVEN: ECONOMIC EVALUATION

11.1 EQUIPMENT’S PURCHASED COST

Table 45:Equipment’s Purchased Cost

Equipment	Cost $ / year
Reactor R-101	**2,791,061**
Formaldehyde storage tank	182,000
Hydrogen peroxide storage tank	455,000
Pump 101A/B	49,680
Pump 102A/B	49,700
Distillation column T-101	252,660
Distillation column T-102	190,336
Distillation column T-103	194,787
Flash column V-101	45,856.8
Heat exchanger E-101	24,843
Mixture Mix-101	1,559,376
Incinerator	31,906.5
Compressor C-101	207,000
Ultrafiltration	127,161
TOTAL PURCHASED COST	6,161,367

11.2 COST OF RAW MATERIAL

a. Cost of methanol

- $\text{Cost} = \frac{348\ \$}{1.1\ ton} * \frac{ton}{1000\ kg} * \frac{32\ kg}{Kmol} * \frac{266\ kmol}{hr} * \frac{24\ \ hr}{1\ day} * \frac{350\ day}{1\ year} = 22.6\frac{\text{M\$}}{\text{year}} = 339{,}303.8\frac{\$}{\text{year}}$

b. Cost of oxygen [119]

- $\text{Cost} = \frac{5.5\ JD}{6\ m^3} * \frac{102\ Kmol}{hr} * \frac{32\ Kg}{Kmol} * \frac{m^3}{13.8\ Kg} * \frac{24\ hr}{day} * \frac{350\ day}{1\ year} = 1.82\text{M}\frac{\text{JD}}{\text{year}}$

$$= 2568745.05\frac{\$}{\text{year}} = \frac{2.56\ \text{M\$}}{\text{year}}$$

- **Cost of Catalyst**

The cost of methanol oxidase enzyme is 250 $/kg of catalyst [108]

Amount of catalyst needed $=$ concentration of catalyst within the reactor $*$ volume of liquid

$$= \frac{110\ ug}{ml}\frac{*644.24\ \text{m}^3*10^{-6}\ ug}{1g} * \frac{10^6\ ml}{1\ m^3} = 70.866\ \text{Kg of enzyme}$$

The annual cost of the catalyst is $= \frac{250\ \$}{Kg} * \frac{52{,}470\ Kg}{1} = 17{,}716.6\ \$$

Annual cost of raw material and catalyst = cost of O_2 + Cost of methanol + Cost of catalyst

$= 339303.8 + 2568745.05 + 17{,}716\ .6 = 2{,}581{,}862.55\ \$/\text{year}$

11.3 COST OF PRODUCT:

a. Cost of formaldehyde [120]

$$= \frac{1.4\ \$}{Kg} * \frac{92.8\ kmol}{hr} * \frac{30\ kg}{Kmol} * \frac{24\ hr}{day} * \frac{350\ day}{year} = 32.7\ \text{M \$/year}$$

b. Cost of hydrogen peroxide

$$= \frac{50000\ metric\ ton}{year} * \frac{327.5\ JD}{10\ metric\ tons} = 1.6\ \mathrm{MJD/year}\ = 2.25\ \mathrm{M\$/year}$$

11.4 OPERATING LABOR COST [121 ,122 ,123,63]

Equipment type	# of equipment	Operators/Unit/Shift	Total Number of Operators per shift
Mixer	1	0.3	0.3
Heat exchanger	1	0.1	0.1
Fermenter	6	1	6
Flash drum	1	0.2	0.2
Distillation column	3	0.35	1.05
Pumps	3	0	0
Storage tank	2	0.2	0.4
Compressor	1	0.15	0.15
Total Number of worker/ Shift			8.2

Average salary in Jordan = 400 $ /month if the worker works 6 days in week for 8 hr per day then they take 2$/hr.

Number of shifts in the plant at one day = 3

$$\text{Total number of Operates per shift } = \frac{8.2\ Operater}{1\ Shift} * \frac{3\ shift}{1} = 24.6\ \text{operator}$$

$$\text{Cost of operating labor } = \frac{2\ \$}{hr.operators} * \frac{8\ hr}{shift} * \frac{3\ shift}{day} * \frac{350\ day}{year} * \frac{25\ operator}{1}$$

$$\text{Cost of operating labor} = 420{,}000\ \$/\text{year}$$

According to “Ultrafiltration membranes and application by Alan S.Michael’s page 399 ”the labor cost =10 $/hr for UF then in one year the labor cost is:

$$= \frac{10\$}{hr} * \frac{24\ hr}{1\ day} * \frac{350\ day}{1\ year} = 84{,}000\$/\text{year}$$

Cost of UF operating labor in 2021 = 84,000 × 2.13 = 178,920 $

Then the total cost of operating labor is $= 178{,}920 + 420{,}000 = \mathbf{598{,}920}\ \$/\mathbf{year}$

11.5 COST OF UTILITY

Equipment type	Utility required	Annual utility required	Unit of Annual utility required	Cost of utility in 2021 $/year
Pump 101	Electricity	2,772,000	KW.h/Year	124,740
Pump 102	Electricity	2,931,600	KW.h/Year	131,922
R-101	R-134a	1754.21	M KJ/year	119,552
E-101	Cooling water	617.6	M Kg/year	105,136
UF	Cleaning fluid			2,982
T-103	Steam	130,121,445	Kg/year	1,219,498
V-101	Steam	1,498,191.42	Kg/year	14,041
Total cost of utility				1,717,871

The cost of utility based on the operating time of the plant = 350-day, annual utility required is calculated by multiply utility required that was calculated in equipment design or energy balance section, which is based on one day, by 350 days. Cost was calculated in 2021 after taking the cost of utility in 2002, from table 6.14 in “plant design and economy 5 edition “, and multiply it with the index ratio, while the cost of UF cleaning fluid is calculated according to “Ultrafiltration membranes and application by Alan S. Michaels page 399” which estimate that the cost of cleaning fluid / day for 100 m^2 is 5$.

The cost of 80 m^2 UF $= \frac{5\ \$}{day * 100m^2} * 80m^2 * \frac{350\ day}{1\ year} = 1400\$/\text{year}$

Cost of UF in 2021 $= 1400 * 2.13 = 2982\ \$/\text{year}$

Total utility cost = Cost of Electricity + Cost of R-134 a + Cost of cooling water + Cost of steam + Cost of cleaning fluid of UF = **1,717,871 $/year**

11.6 FIXED CAPITAL INVESTMENT

FCI is calculated according of "Plant Design and Economy 5th edition "Figure 6.6

Component	**Cost**
Purchased equipment (Delivered), E	6,161,367
Purchased equipment (Installation),0.39 E	2,402,933
Instrumentation (Installed), 0.43E	2,649,388
Piping (installed), 0.31 E	1,910,024
Electrical (Installed) , 0.10 E	616,137
Building (include scurvies) 0.15 E	924,205
Yard improvement 0.12 E	739,364
Service facility (instilled) 0.55 E	3,388,752
Total direct plant cost D	12,630,803
Engineering and supervisor, 0.32E	1,971,638
Construction expenses, 0.34 E	2,094,865
Legal expenses, 0.04 E	246,455
Constructor's fee, 0.19 E	1,170,660
Contingency ,0.37 E	2,279,706
Total indirect plant cost, I	7,763,323
Fixed –Capital Investment, D+I	20,394,126

11.7 TOTAL PRODUCT COST

Type	**Factor**	**Cost M$/year**
Raw material cost		2.58
Utility cost		1.717871
Operating labor cost		0.598
Operating supervision	0.15 of operating labor cost	0.0897
Maintenances and repaired	0.07 of FCI	1.427588803
Operating Supplies	0.15 of Maintenances and repaired	0.214138321
Laboratory Charge	0.15 of Operating labor cost	0.0897
Total Variable production cost		6.716998124

Total product cost (Without depreciation) = Total Variable production cost+ Fixed charge without depression + Plant overhead cost +Administrative cost+ Distribution cost + Research

Type	**Factor**	**Cost $/year**
Taxes (Property)	0.02 of FCI	407,883
Insurance	0.01 of FCI	203,941
Fixed charge without depression		611,824

Type	**Factor**	**Cost M$/year**
Plant overhead cost	0.5 (Operating labor cost +Operating supervision + Maintenances and repaired)	1.057644402
Administrative cost	0.20 Operating labor cost	0.1196
Total		1.177244402

Type	**Factor**
Distribution cost	0.2 of TPC
Research and development	0.05 of TPC
Royalties (if not on lump-Sum basis)	0.04 of TPC
Total	0.29 of TPC

Total product cost = (6.71+ 0.611) * 10^6 + 1.177+ 0.29 of TPC

By solving the above equation, the total product cost is 11.98 M$

11.8 COST OF MANFACTURE

$$\text{COM} = 1.23(C_{RM} + C_{WT} + C_{UT}) + 2.73C_{OL} + 0.304\ FCI$$

$$\text{COM} = 1.23 * (2.58 + 0 + 1.7) + 2.73 * 0.598 + 0.304 * 20.3 = 15\ M\$/\text{year}$$

CHAPTER TWELVE: PROFITABILITY PROJECT ANALYSIS

The followed method to determine if the project is profitable or not is the non-discounted method

The following data are required to draw cash flow diagram:

1. Land cost [63] =0.01 FCI = 203,941 $
2. FCI = 20,394,126 $
3. Working capital investment [63] = 0.175 FCI= 3,568,972
4. Plant start up at the end of second year
5. Cost of manufacture= 15 M$/year
6. Taxation in Jordan at 2021 = 8.7% [124]
7. Salvage value[63] = 0 $
8. Depreciation using double decline balance over 7 year
9. Project life 12 year
10. Total capital investment = fixed-capital investment + working capital = 23,630,719 $
11. Depreciation :(Peter,2006)

$$\text{Dk} = \frac{2}{n} * (\text{FCIL} - \sum dj)$$

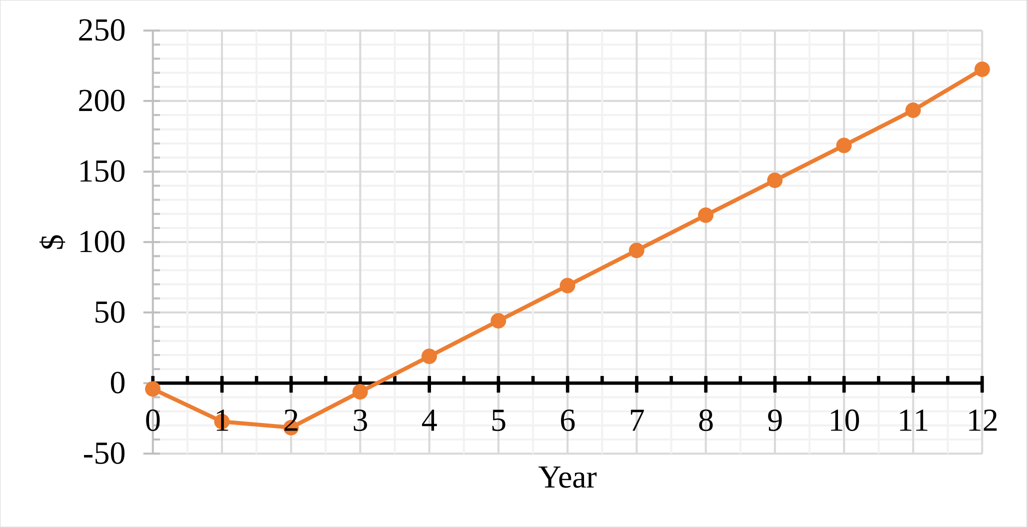


Figure 47 cash flow diagram for hydrogen peroxide plant

The pervious figure represents cash flow diagram for hydrogen peroxide plant. To determine if the project is profitable or not the following criterion should be analyzed:

1. Payback period (PBP), is the time required, after start-up, to recover the FCIL from figure (21) PBP equal 3.3 year
2. Cash Criterion

Cumulative cash position (CCP): The worth of the project at the end of its life = 153.5

Cumulative cash ratio (CCR): Sum of all positive cash / sum of all negative cash = 7.3

Rate of return on investment (ROROI), Rate at which we make money from our FCIL

$$ROROI = \frac{Total\ revenues - \mathrm{FCI_L}}{n * \mathrm{FCI_L}}$$

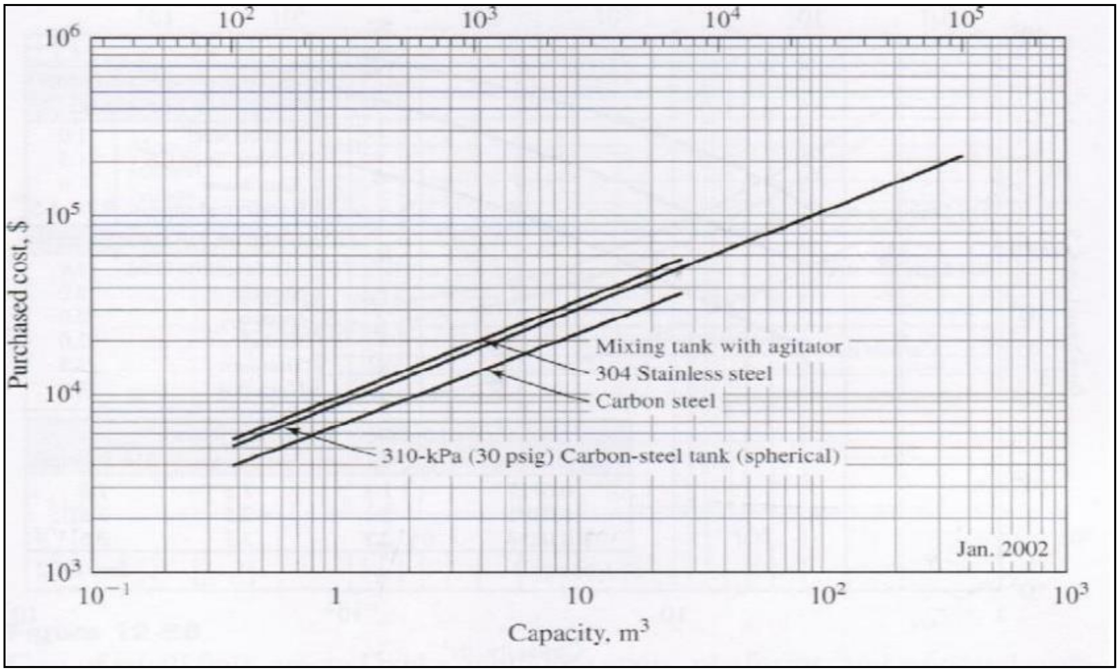
$$\mathrm{ROROI} = \frac{177.2 - 20.3}{10 * 20.3} * 100\% \ = 76.9\%$$

Total revenues =177.2 M$ (From the excel sheet on appendixes)

As a result of this analysis the project is **completely profitable.**

CHAPTER THERTEEN :DETAILED PURCHASE COST FOR EACH EQUIPMENT

13.1 MIXER PURCHASE COST:



The volume of each mixer = 190.4 m^3

The purchase cost for Carbon steel mixer at (V$_1$=20 m$^{3)}$ =30000$

So, the purchase cost at $(V_2 = 190.4)$ m^3 $= (\frac{V_2}{V_1} * \text{purchase cost for } V_1)$

The purchase cost at $(V_2 = \mathbf{190.4})$ m^3 $= \frac{190.4}{20} * 30000\ \$ = 285600$

For 3 mixers of capacity = 571.2 m^3, the purchase cost = 3*285600 $ = 856800 $

- CEPCI (2002) = 395.6
- CEPCI (2021) = 720.4

- Index ratio $= \frac{720.4}{395.6} = 1.82$
- $\boldsymbol{The\ total\ cost\ in\ 2021 = 856800\ \$ * 1.82 = 1,559,376\ \$}$

13.2 HEAT EXCHANGER PURCHASE COST:

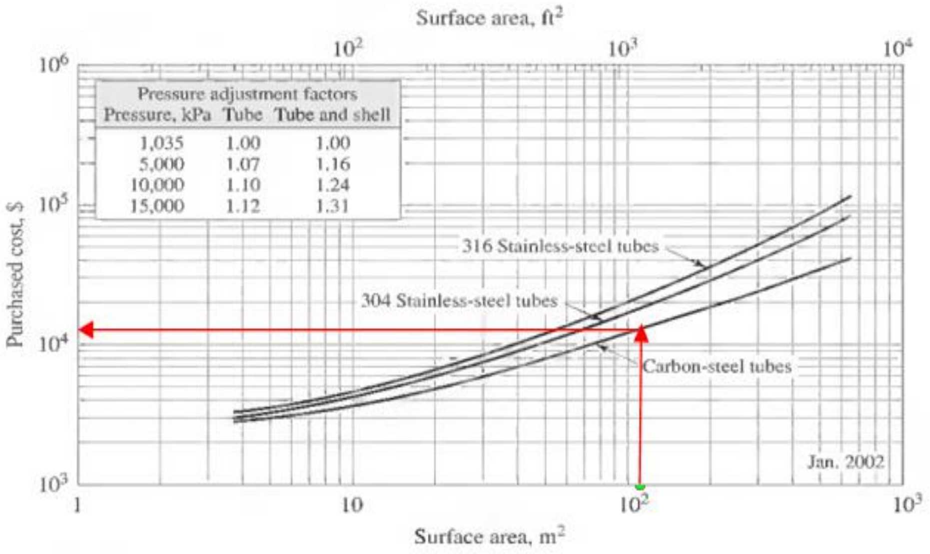
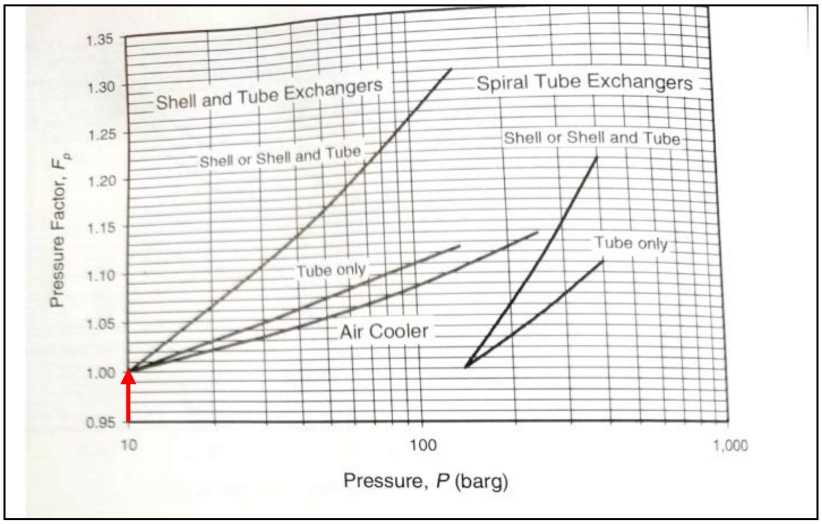


Figure 14-18
Purchased cost of fixed-tube-sheet heat exchangers with 0.019-m ($\frac{3}{4}$-in.) OD × 0.025-m (1-in.) square pitch and 4.88- or 6.10-m (16- or 20-ft) bundles and carbon-steel shell operating at 1035 kPa (150 psia)

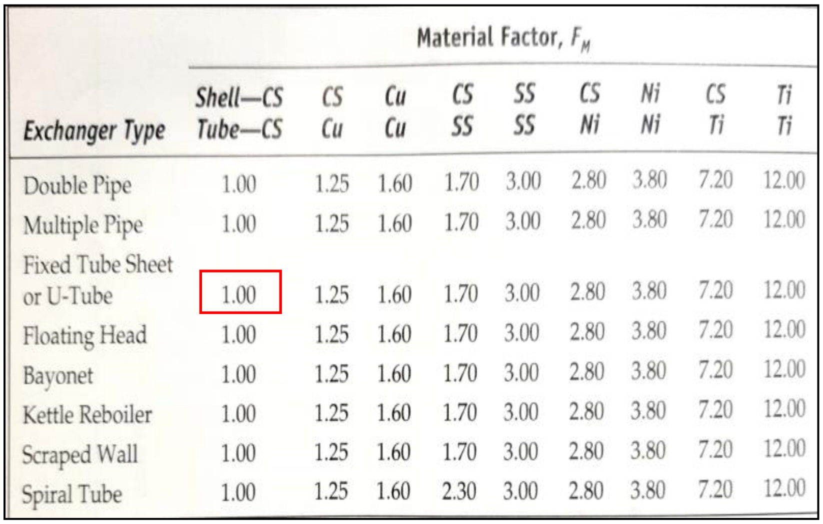
The purchase cost of heat exchanger with area =113.06 m^2 is =**13000$** if the MOC of the exchanger is Carbon steel.

Purchase cost Cp $= F_p * F_M * F_l * (13000\ \$)$, Where:

- F_p : Pressure factor at 150 psi (10 bar), from the figure below it is equal to (1)

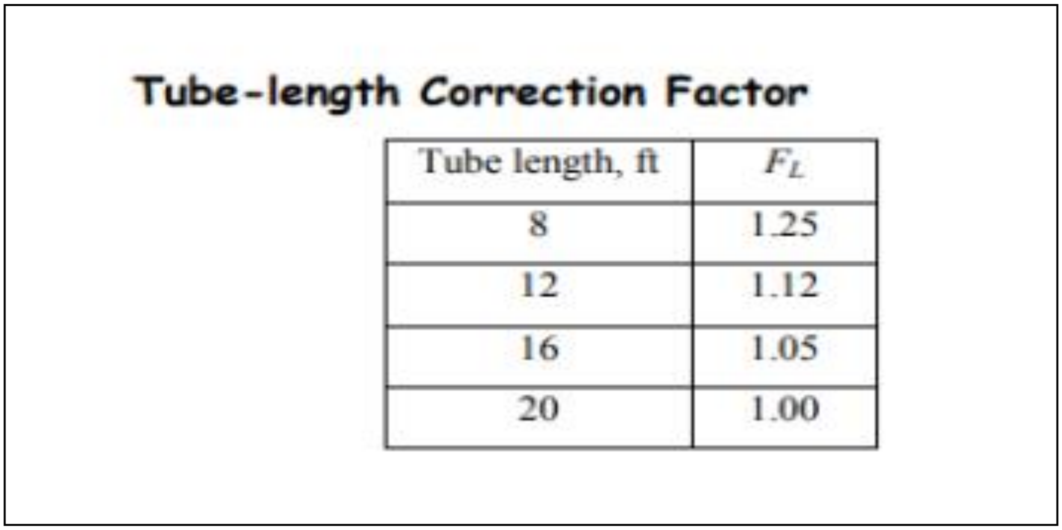


- F_M : Material factor for shell and tube heat exchangers with fixed tube sheet ,and from the figure below it is equal to (1)



Exchanger Type	Material Factor, F_M								
	Shell—CS Tube—CS	CS Cu	Cu Cu	CS SS	SS SS	CS Ni	Ni Ni	CS Ti	Ti Ti
Double Pipe	1.00	1.25	1.60	1.70	3.00	2.80	3.80	7.20	12.00
Multiple Pipe	1.00	1.25	1.60	1.70	3.00	2.80	3.80	7.20	12.00
Fixed Tube Sheet or U-Tube	1.00	1.25	1.60	1.70	3.00	2.80	3.80	7.20	12.00
Floating Head	1.00	1.25	1.60	1.70	3.00	2.80	3.80	7.20	12.00
Bayonet	1.00	1.25	1.60	1.70	3.00	2.80	3.80	7.20	12.00
Kettle Reboiler	1.00	1.25	1.60	1.70	3.00	2.80	3.80	7.20	12.00
Scraped Wall	1.00	1.25	1.60	1.70	3.00	2.80	3.80	7.20	12.00
Spiral Tube	1.00	1.25	1.60	2.30	3.00	2.80	3.80	7.20	12.00

F_L: Tube-length correction factor ,and it was equal to **1.05** at 16 ft tube length



Tube-length Correction Factor

Tube length, ft	F_L
8	1.25
12	1.12
16	1.05
20	1.00

So, the purchase cost of exchanger is:

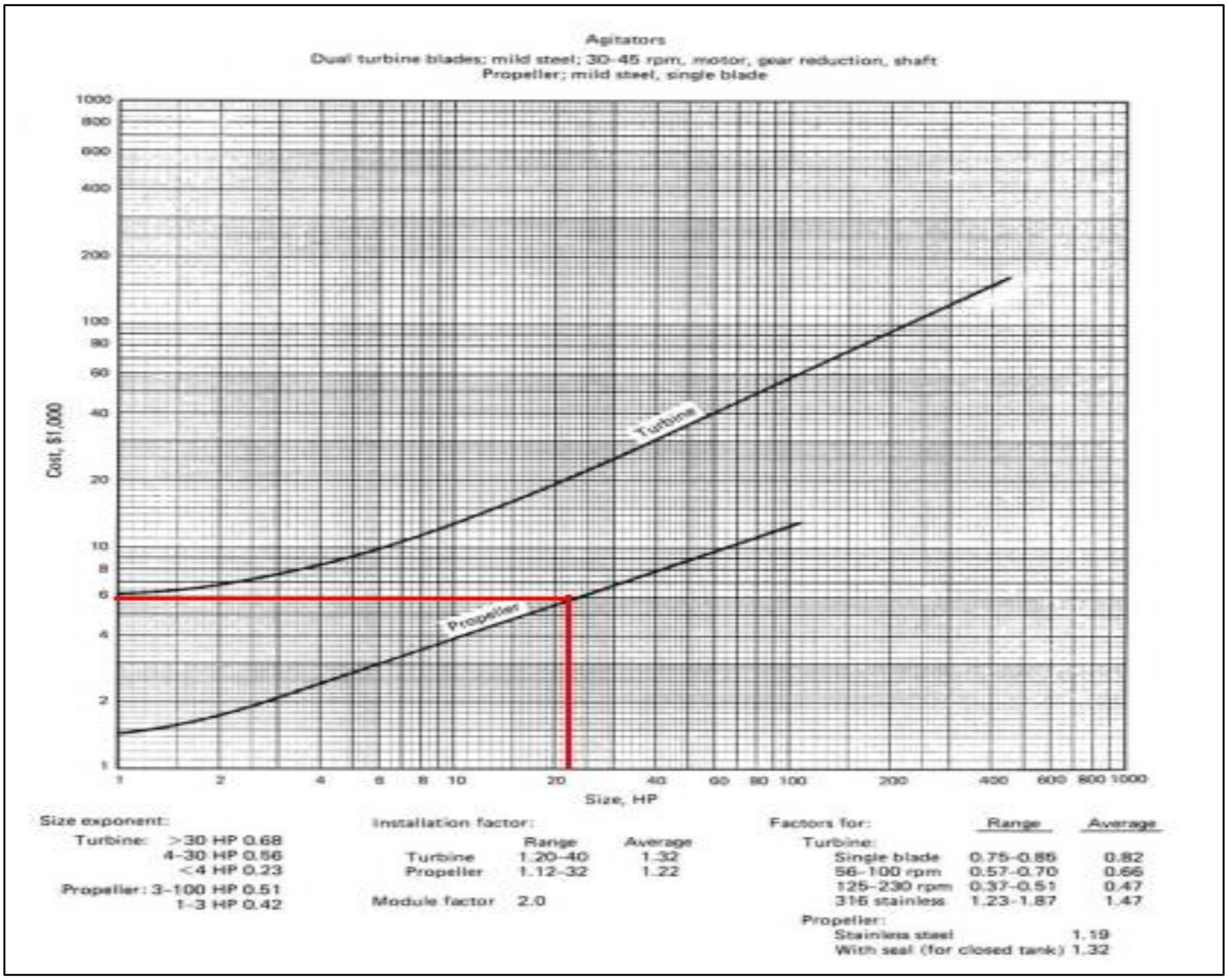
Purchase cost (Cp) $= F_p * F_M * F_l * (13000\ \$)$

(Cp) $= 1 * 1 * 1.05 * 13000\ \$$

$\mathbf{Cp = \ 13650}\ \$$

- CEPCI (2002) = 395.6
- CEPCI (2021) = 720.4
- Index ratio $= \frac{720.4}{395.6} = 1.82$
- $\boldsymbol{The\ cost\ in}\ \mathbf{2021 = 13650}\ \$ * \mathbf{1.82 = 24843}\ \$$

13.3 REACTOR PURCHASE COST



Size exponent:

Turbine: >30 HP 0.68
4-30 HP 0.56
<4 HP 0.23

Propeller: 3-100 HP 0.51
1-3 HP 0.42

Installation factor:

	Range	Average
Turbine	1.20-40	1.32
Propeller	1.12-32	1.22

Module factor 2.0

Factors for:	Range	Average
Turbine:		
Single blade	0.75-0.85	0.82
56-100 rpm	0.57-0.70	0.66
125-230 rpm	0.37-0.51	0.47
316 stainless	1.23-1.87	1.47
Propeller:		
Stainless steel		1.19
With seal (for closed tank)		1.32

First the cost of agitator is determining by the figure above where X axis's is the power requirement for the agitator which is found in design procedure of the reactor and y-axis is the cost in 1000$.

For agitator power = 22 hp the cost for propeller is 5000$

By multiply 5000$ with 1.19 to correct the cost for S.S as a material of construction the cost will equal 5950 $.

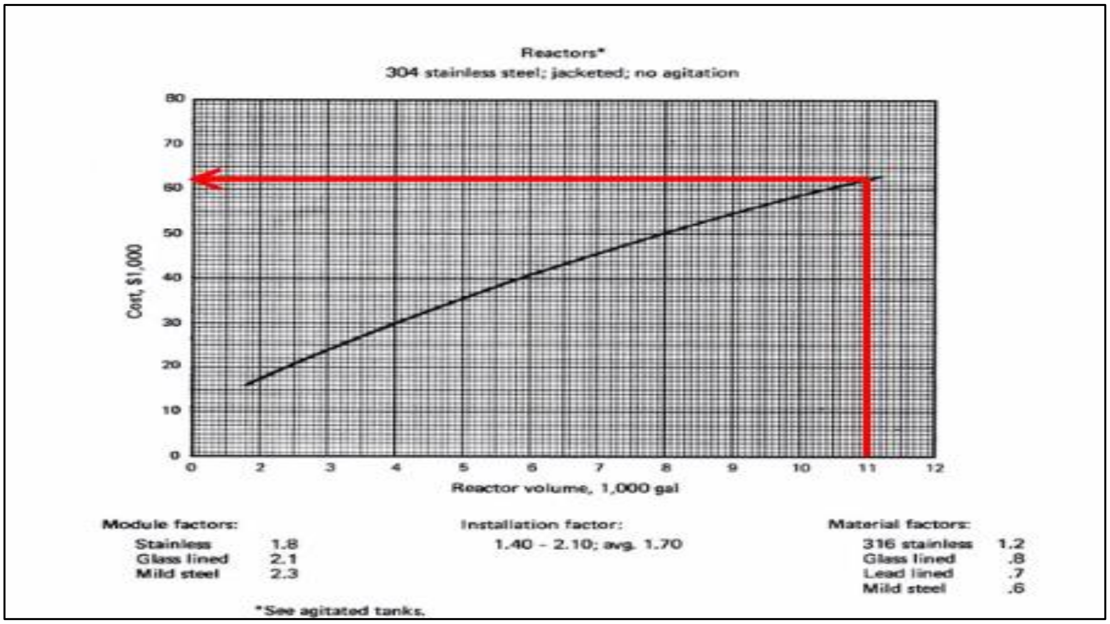
Since the numbers of the reactor needed in this process is 6 reactors then the number of agitator is 6 as a result the total cost of agitators in 1989 is =6*5950 =35,700 $

$\text{CEPCI in } (1989) = 355.4$

$\text{CEPCI in } (2021) = 720.4$

$\text{Index Ratio} = (720.4/355.4) = 2$

$\textbf{Cost of agitator in 2021} = \mathbf{2} * 35{,}700 = \mathbf{71{,}400}$ $



This figure illustrates the cost of reactor with jacketed and without agitator .

The volume of each reactor in parallel = 128.8 m^3

Cost for reactor volume 41.6 m^3 (11000 gal) = 61,000$

Cost of reactor with volume 128.8 m^3 is $= 61000 * \frac{128.8}{41.6} = 188{,}865$ $

For 6 reactors, the total cost in 1989 is = 6*188,865 =1,133,192 $

Correct the cost for material of construction to be stainless steel multiple the cost with 1.2

then cost = 1.2*1,133,192=1,359,830 $

$\text{CEPCI in }(1989)\ = 355.4$

$\text{CEPCI in }(2021)\ = 720.4$

$\text{Index Ratio } = \ (720.4/355.4)\ = 2$

$\textbf{Cost of reactor in 2021 } = \mathbf{2 * 1,359,830 = 2,719,661}\$$

The total cost of reactor = Cost of reactor with jacketed + cost of agitator

= $\mathbf{2,719,661}$+ $\mathbf{71,400}$ =2,791,061 $

13.4 ULTRAFILTRATION PURCHASE COST

According to "Ultrafiltration membranes and application by Alan Michaels page 399"

The cost of 1 m^2 of UF is 750 $, the UF area in this project is 79.6 m^2 then the UF cost in 1990 is

$= 750 * 79.6\ =\ 59,700\ \$$

$\text{Cost in 2021} = 59{,}700\ *\ 2.13 = 127{,}161\ \$$

13.5 DISTILLATION COLUMN T-101 PURCHASE COST

a. **Cost of vertical column:**

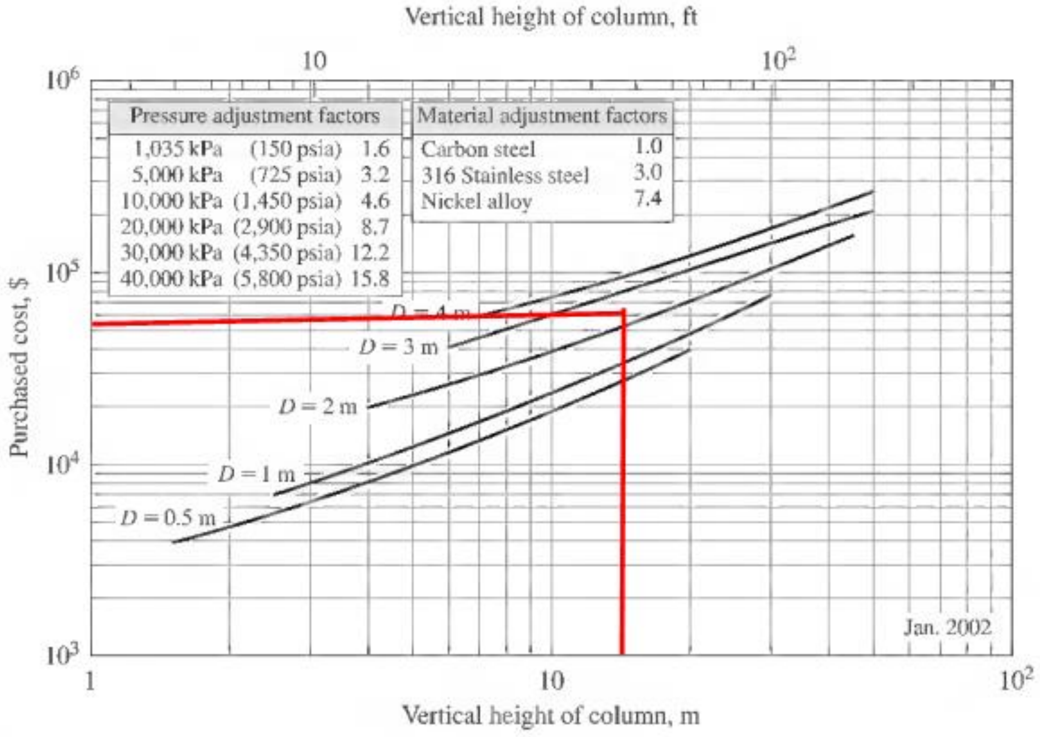


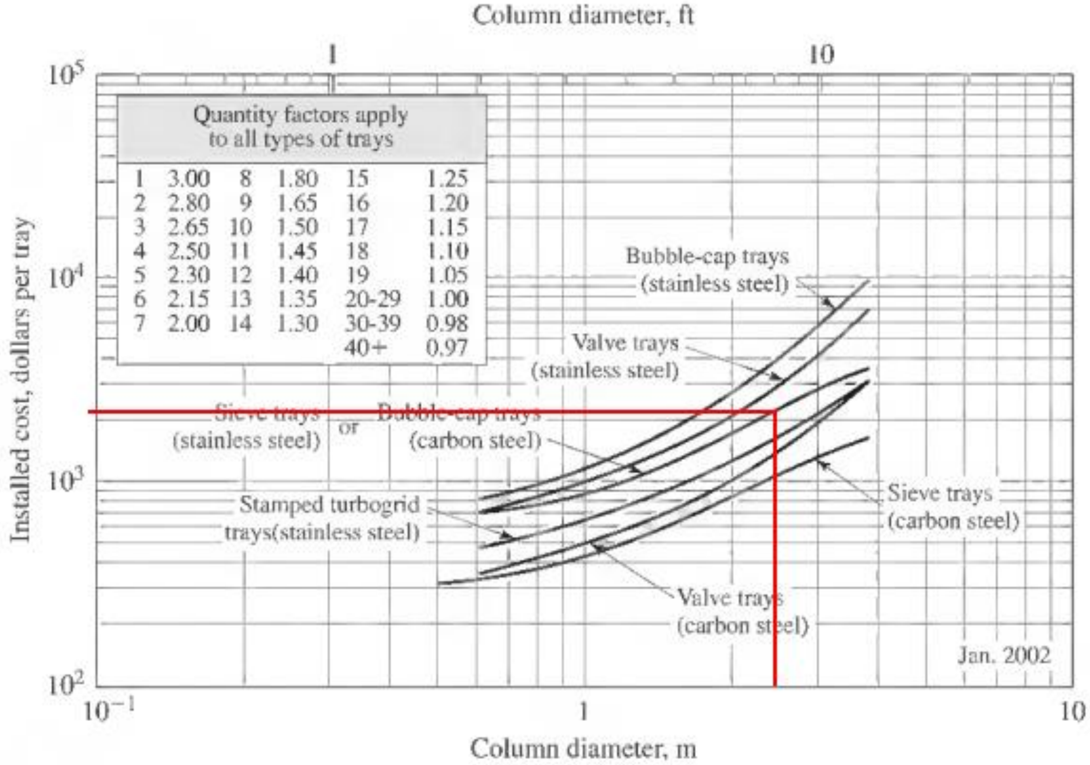
Figure 15-11
Purchased cost of vertical columns. Price does not include trays, packing, or connections.

The purchase cost of vertical column of T-101 figure above at dimeter = 2.37 m and height = 15 m is 40000 $ but this value must be multiplied with material adjustment factor of 316 stainless steel which is equal 3 then cost of the T-101 in 2002 is 120,000 $

$$\text{Cost at 2021} = \text{cost in 2002} * \text{Index ratio} = 120{,}000 * 2.07$$

$$\text{Cost at 2021} = 248{,}400\ \$$$

b. Cost of tray inside of column:



Quantity factors apply to all types of trays

1	3.00	8	1.80	15	1.25
2	2.80	9	1.65	16	1.20
3	2.65	10	1.50	17	1.15
4	2.50	11	1.45	18	1.10
5	2.30	12	1.40	19	1.05
6	2.15	13	1.35	20-29	1.00
7	2.00	14	1.30	30-39	0.98
				40+	0.97

Figure 15-13

Purchased cost of trays in tray columns. Price includes tray deck, bubble caps, risers, downcomers, and structural-steel parts.

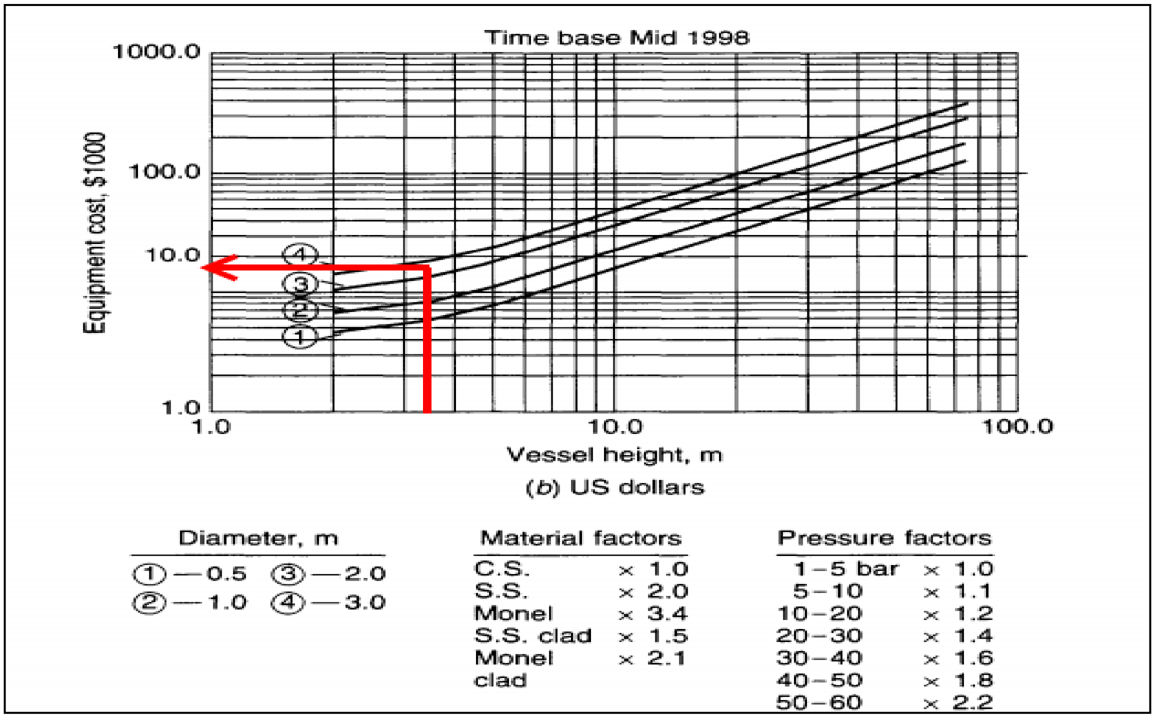
Purchase cost of one tray inside column from figure above at column dimeter = 2.37 m and tray type is sieve is 2100 $ but this value must be multiplied with quantity factor of column when number of real stage equal 29.6 =0.98, so the total cost of tray =0.98 * 2100 = 2058 $

$$\text{Cost of tray in } (2002) = 2058\ \$$$

$$\text{Cost of tray in}(2021) \ = \ 2058\ \$ * 2.07 = \ 4260.06\ \$$$

$$\text{Total cost of T} - 101 \text{ in } 2021 \ = \text{Cost of trays } + \text{column cost} = \ 4260.06 + 248{,}400 = 252{,}660\ \$$$

13.6 FLASH COLUMN V-101 PURCHASE COST



$\text{CEPCI (1998)} = 115.1$ [117]

$\text{CEPCI (2021 Oct.)} = 266.6$

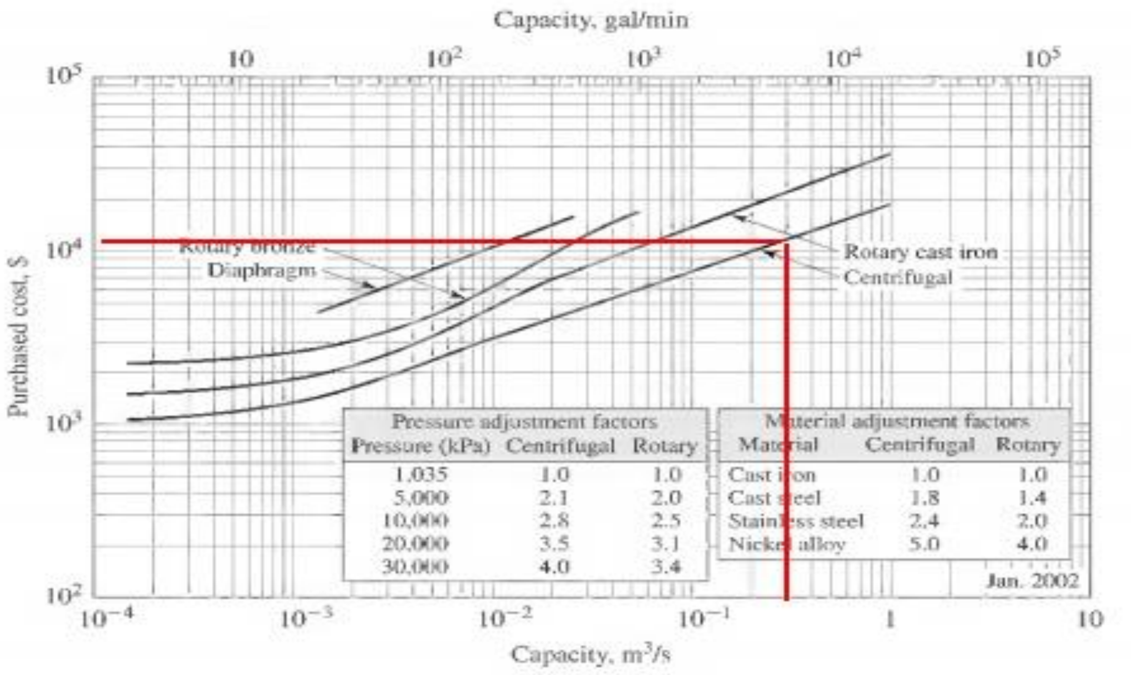
$$\text{Index ratio} = \frac{\text{CEPCI (2021 Oct.)}}{\text{CEPCI (1998)}} = \frac{266.6}{115.1} = \ 2.316$$

At vessel height = 3.5 m the vessel cost = 9900$ but this value need material of constriction correction factor, which is equal 2 for stainless steel, since vessel operates at 0.35 atm , it doesn't need pressure correction factor.

Cost of Vessel at 1998 = 9900*2 = 19800$

Cost of V-101 at 2021= cost of flash in 1998*Index ratio=19800*2.316 =45856.8 $

13.7 PUMP 101 A/B PURCHASE COST



Pressure adjustment factors

Pressure (kPa)	Centrifugal	Rotary
1,035	1.0	1.0
5,000	2.1	2.0
10,000	2.8	2.5
20,000	3.5	3.1
30,000	4.0	3.4

Material adjustment factors

Material	Centrifugal	Rotary
Cast iron	1.0	1.0
Cast steel	1.8	1.4
Stainless steel	2.4	2.0
Nickel alloy	5.0	4.0

Figure 12-23
Purchased cost of diaphragm, centrifugal, and rotary pumps

Pump capacity is 0.3 m^3/s, it delivers the stream pressure to reach 607.95 Kpa , it is made of S.S ,then at these condition the cost of P-102 is =10000*1*2.4 =24,000 $

- $CEPCI(2002) = 128.7$
- $CEPCI(2021\ Oct.) = 266.6$
- $Index\ ratio = \frac{\text{CEPCI (2021 Oct.)}}{\text{CEPCI (2002)}} = \frac{266.6}{128.7} = 2.07$
- $\text{Cost of pump at 2021} = \text{cost of pump at 2002} * \text{Index ratio} = 24{,}000 * 2.07$

  $= 49{,}680\$$

13.8 COMPRESSOR C-101 PURCHASE COST

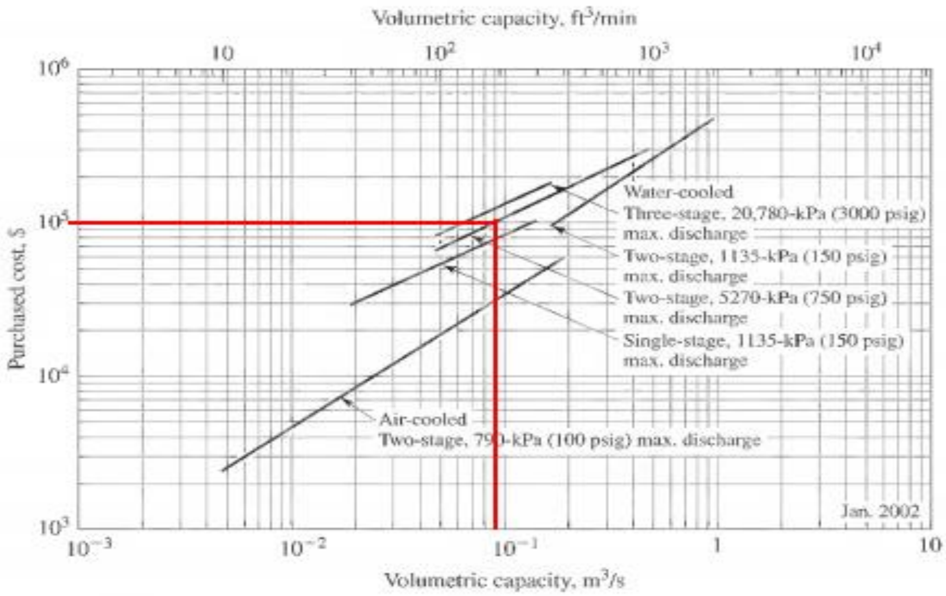


Figure 12-29
Purchased cost of reciprocating compressors

At the oxygen volumetric flow rate is 0.09 m^3/s and for two stage reciprocating compressor the cost of the C-101 in 2020 is 100,000 $

$$\text{Cost of } (C - 101) \text{ at } 2021 = \text{cost at } 2002 * \text{Index ratio} = 100{,}000 * 2.07$$

$$= 207{,}000\$$$

13.9 FLARE PURCHASE COST: [118]

- The cost of the flare can be determined from the following equation,

$$C_F = (78 + 9.14 * D + 0.749 * L)^2$$

$$C_F = (78 + 9.14 * 0.968 + 0.749 * 30)^2$$

$C_F = \$11950$ (April 1992)

$\text{CCI (1992)} = 100$

$\text{CCI(Oct} - 2021) = 266.6$

$$\text{Cost index ratio } = \frac{CCI\ (2021)}{CCI\ (1992)} = \frac{266.6}{100} = 2.67$$

Cost of flare in 2021 = cost of flare in 1992* index ratio = 11950*2.67 = $31906.5

13.10 H_2O_2 STORAGE TANK PURCHASE COST

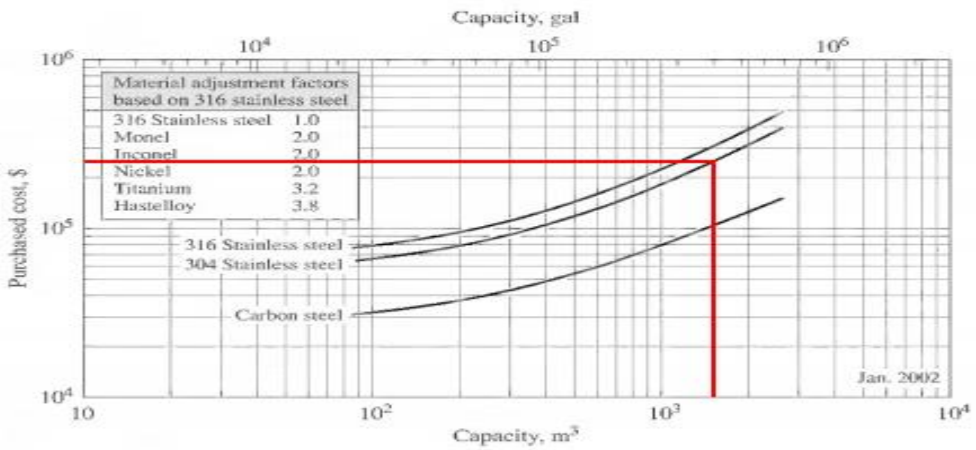


Figure 12-55
Cost of small field-erected tanks, including stairs, platforms, and a normal complement of flow hardware. Carbon-steel tank is for an API Standard 650 tank while the other two are for stainless-steel tanks with flat bottoms operating at atmospheric pressure.

- The Total Volume of Storage tank is 1494.5 m^3 with cost = 250,000$
- $\text{CEPCI (2002)} = 395.6$
- $\text{CEPCI (2021)} = 720.4$
- $\text{Index ratio} = (720.4/395.6\ ) = 1.82$
- $The\ cost\ in\ 2021 = 250{,}000\ \$ * 1.82 = 455{,}000\ \$$

CONCLUSIONS

1. This report presents the design of hydrogen peroxide plant which was chosen to locate in Aqaba . This location was chosen based on several factors, the most important of which is the abundance of raw materials.

2. For the manufacturing of hydrogen peroxide, the enzymatic process was chosen for many reasons such as less expensive raw material & catalyst, and no harmful emissions.

3. This project was done through several mass and energy balances were the key factor that controlled the equipment design parameter as well as mechanical design for the capacity required to produce 50,000 tons/year of H_2O_2.

4. The materials of construction that were used for designing plant equipment were between these types :304 stainless steel and carbon steel.

5. The total capital investment is about (23,963,098) dollars, and (3.3) years payback period.

6. Teamwork was an important factor; we would have not finished the project in time without a group effort.

RECOMMENDATIONS

It is recommended to use software programs to solve advanced equipment design problems for certain units such as reactor and heat exchanger, distillation. Regular training of the employees and workers and keeping the number of incidences and accidents zero.

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APPENDICES

APPENDIX A: MOLE FRACTIONS OF EACH COMPONENT IN EACH UNIT

A.1 Mole Fractions of Components in the Streams of Mixing Tank (Mix-100)

Streams	M_1	M_2	M_{17}
Units	Mole%	Mole%	Mole%
Components	-	-	-
HCHO	0	0	-
CH_3OH	0.84	0	0.084
H_2O	0.16	0.9995	0.915
H_2O_2	0	0.0004	0.00036
O_2	0	0	0

A.2 Mole Fractions of Components in the Streams of Fermenter (R-101)

Streams	M_3	M_4	M_5	M_6
Units	Mole%	Mole%	Mole%	Mole%
Components	-	-	-	-
HCHO	0	-	-	0.0324
CH_3OH	0	0.084	0.0031	0.049
H_2O	0	0.915	-	0.885
H_2O_2	0	0.00036	-	0.0324
O_2	1	-	0.996	0.00020

A.3 Mole Fractions of Components in the Streams of Flash Column (V-103)

Streams	**M_9**	**M_{11}**	**M_{12}**
Units	**Mole%**	**Mole%**	**Mole%**
Components	**-**	**-**	**-**
HCHO	**0.033**	**-**	**0.033**
CH_3OH	**0.05**	**0.25**	**0.050**
H_2O	**0.9153**	**0.75**	**0.915**
H_2O_2	**0.00039**	**-**	**0.0039**
O_2	**-**	**-**	**-**

A.4 Mole Fractions of Components in the Streams of Distillation Column (T-101)

Streams	**M_8**	**M_9**	**M_{10}**
Units	**Mole%**	**Mole%**	**Mole%**
Components	**-**	**-**	**-**
HCHO	**0.032**	**0.033**	**-**
CH_3OH	**0.049**	**0.050**	**-**
H_2O	**0.886**	**0.915**	**-**
H_2O_2	**0.032**	**0.00039**	**1**
O_2	**-**	**-**	**-**

A.5 Mole Fractions of Components in the Streams of Multicomponent Distillation(T-102)

Streams	**M_{12}**	**M_{13}**	**M_{14}**
Units	**Mole%**	**Mole%**	**Mole%**
Components	**-**	**-**	**-**
HCHO	**0.0334**	**0.000023**	**0.0355**
CH_3OH	**0.050**	**0.741**	**0.0045**
H_2O	**0.915**	**0.2560**	**0.959**
H_2O_2	**0.0039**	**-**	**0.0004**
O_2	**-**	**-**	**-**

A.6 Mole Fractions of Components in the Streams of Multicomponent Distillation(T-103)

Streams	M_{14}	M_{15}	M_{16}
Units	Mole%	Mole%	Mole%
Components	-	-	-
HCHO	0.0355	0.267	-
CH_3OH	0.0045	0.034	-
H_2O	0.9594	0.6984	0.9995
H_2O_2	0.0004	-	0.00046
O_2	-	-	-

A.7 Mole Fractions of Components in the Streams of Flash Column (V-101)

Streams	M_6	M_7	M_8
Units	Mole%	Mole%	Mole%
Components	-	-	-
HCHO	0.032	0.04	0.032
CH_3OH	0.049	0.16	0.049
H_2O	0.8857	0.52	0.886
H_2O_2	0.032	0.04	0.032
O_2	0.0002	0.24	-

A.8 Mole Fractions of Components in the Streams of heat exchanger

Streams	M_{17}	M_4
Units	Mole%	Mole%
Components	-	-
CH_3OH	0.084	0.084
H_2O	0.915	0.915
H_2O_2	0.00036	0.00036

A.9 Mole Fractions of Components in the Streams of mixer (M-101)

Streams	M_1	M_2	M_{17}
Units	Mole%	Mole%	Mole %
Components	-	-	-
CH_3OH	0.84	0.9995	0.08440
H_2O	0.16	------	0.91523
H_2O_2	------	0.0004	0.00036

A.10 Mole Fractions of Components in the Streams of ultrafiltration membrane(UF-101)

Streams	M_6*	M_6
Units	Mole%	Mole%
Components	-	-
HCHO	0.0324	0.0324544
CH_3OH	0.049	0.049433
H_2O	0.885	0.88544
H_2O_2	0.0324	0.032446
O_2	0.00020	0.0002047
Enzyme	0.0011	-

APPENDIX B: ENTHALPY OF FORMATION

Component	Enthalpy of formation
O_2	0 kJ/mol [42]
CH_3OH	-238.42 kJ/mol [43]
H_2O_2	-187.341 kJ/mol [44]
HCHO	-109.15 kJ/mol [44]

APPENDIX C:

a . HEAT CAPACITY FOR DIFFERENT COMPONENTS

Stream number	**Heat capacity for different component in (kJ/kmol. k) [45&46]**		
	Temperature K	**CH_3OH**	**H_2O**
1	**322.45**	**86**	**75.3**
2	**476.55**	**177**	**81.4**
3	**311.9**	**84**	**75.3**
4	**323.15**	**87**	**75.4**
5	**363.15**	**89**	**75.8**
6	**251.15**	**74**	**-------**
7	**341.15**	**92**	**75.5**
8	**341.15**	**92**	**75.5**
9	**339.75**	**91**	**75.5**
10	**358.05**	**97**	**75.7**
11	**341.25**	**92**	**75.5**
12	**341.25**	**92**	**75.5**
13	**337.95**	**91**	**75.5**
14	**386.05**	**106**	**76.2**
15	**480.55**	**183**	**81.8**
16	**512.85**	**------**	**85.9**

To calculate heat capacity for HCHO, O_2 and H_2O_2 Shomate equation will be used:

$Cp=A+BT+CT^2$

Where T is Temperature (K)/1000

So, Shomate constant firstly will be founded:

Table 46: Shomate constant

Component	**Temperature K**	**A**	**B**	**C**
HCHO [52]	**298-1200**	**5.193767**	**93.23249**	**-44.85457**
O_2 [53]	**100-700**	**31.32234**	**-20.23531**	**57.86644**
H_2O_2 [54]	**289-1500**	**34.25667**	**55.18445**	**-35.15443**

b. HEAT CAPACITY FOR DIFFERENT COMPONENTS

Stream number	Heat capacity for different component in (kJ/kmol. k)			
	Temperature K	HCHO	O_2	H_2O_2
1	322.45	30.593	30.81	48.40
2	476.55	39.437	34.82	52.57
3	311.9	29.909	30.64	48.05
4	323.15	30.638	30.83	48.42
5	363.15	33.136	31.61	49.66
6	251.15	25.780	29.89	45.90
7	341.15	31.780	31.15	48.99
8	341.15	31.780	31.15	48.99
9	339.75	31.692	31.13	48.95
10	358.05	32.825	31.50	49.51
11	341.25	31.786	31.16	48.99
12	341.25	31.786	31.16	48.99
13	337.95	31.579	31.09	48.89
14	386.05	34.501	32.13	50.32
15	480.55	39.638	34.96	52.66
16	512.85	41.211	36.16	53.31

APPENDIX D: LATENT HEAT OF VAPORIZATION

Substance	$\Delta\hat{H}_v$ (KJ/Kmol)
CH_3OH [47]	0.03734
O_2 [48]	0.0034009
HCHO [49]	0.0233
H_2O_2 [50]	0.0485
H_2O [51]	0.0408

APPENDIX E :PHYSICAL AND CHEMICAL PROPERTIES OF H_2O_2

Property	**Value**
Molar mass (g/mol)	34.0147
Appearance	Colorless in solution, very light blue in pure state
Odor	Slightly sharp
Density (g/cm^3) at 20ºC	1.45 in pure state, 1.11 in solution
Melting point (ºC)	-0.43
Boiling point (ºC)	150.2
Solubility	Miscible in water, soluble in ether, alcohol and insoluble in petroleum ether
Vapor pressure (mmHg) at 30ºC	5
Refractive index	11.4061
Acidity	11.75
Viscosity (CP) at 20ºC	1.245
Dipole moment	2.26D
Specific heat capacity (J/(g.k)	1.267 at gas state, 2.619 at liquid state
Standard enthalpy of formation (KJ/mol) at 298K	-187.8
Specific gravity (50%)	1.17 to 1.21

APPENDIX F: SAMPLE CALCULATION-MASS BALANCE

The general material balance used :

$$\mathbf{Input + Generation - Output - Consumption = Accumulation}$$

Generation = Consumption = 0 since no material produce or consume during T-101

Assume steady state system so accumulation term also = 0

So the material balance reduced to :

$$\mathbf{Input = Output}$$

F.1 MATERIAL BALANCE ON DISTILLATION COLUMN(T-101)

Basis of mass balance : production of H_2O_2 =1000 kgmole/hr = M_{10}

- Total material balance:

$$M_8 = M_9 + M_{10}$$

$$\boldsymbol{M_8 = M_9 + 1000} \ldots\ldots(112)$$

- By component balance on H_2O we get:

$$\mathbf{0.0324}\boldsymbol{M_8} = \mathbf{0.00039}\boldsymbol{M_9} + \mathbf{1 * 1000} \ldots\ldots(113)$$

- Solving equation (112) & (113) then :

$$\boldsymbol{M_9} = \mathbf{30228.0537\ kg\ mole/hr}$$
$$\boldsymbol{M_8} = \mathbf{31228.0537\ kg\ mole/hr.}$$

F.2 MATERIAL BALANCE ON FLASH COLUMN (V-103):

- Total material balance:

$$M_9 = M_{11} + M_{12}$$

$$\mathbf{30228.0537} = \boldsymbol{M_{11} + M_{12}} \ldots\ldots(114)$$

- By component balance on H_2O we get:

$$\mathbf{0.9153\ (30228.0537) = 0.75\ \boldsymbol{M_{11}} + 0.9153\boldsymbol{M_{12}}} \ldots \ldots (115)$$

- Solving equation (114) & (115) then :

$$\boldsymbol{M_{11}} = \mathbf{0.02048\ kg\ mole/hr}$$
$$\boldsymbol{M_{12}} = \mathbf{30228.0537\ kg\ mole/hr}$$

F.3 MATERIAL BALANCE ON DISTILLATION COLUMN(T-102)

- Total material balance:

$$M_{12} = M_{13} + M_{14}$$

$$\mathbf{30228.0537} = \boldsymbol{M_{13}} + \boldsymbol{M_{14}} \ldots \ldots (116)$$

- By component balance on HCHO we get:

$$\mathbf{0.0334\ (30228.0537) = 0.000023\boldsymbol{M_{13}} + 0.0355\boldsymbol{M_{14}}} \ldots \ldots (117)$$

- Solving equation (116) & (117) then:

$$\boldsymbol{M_{13}} = \mathbf{1789.298\ kg\ mole/hr}$$
$$\boldsymbol{M_{14}} = \mathbf{28438.756\ kg\ mole/hr}$$

F.4 MATERIAL BALANCE ON DISTILLATION COLUMN(T-103)

- Total material balance:

$$\mathrm{M}_{14} = \mathrm{M}_{15} + \mathrm{M}_{16}$$

$$\mathbf{28438.756} = \boldsymbol{M_{15}} + \boldsymbol{M_{16}}(118)$$

- By component balance on H_2O we get:

$$\mathbf{0.9594\ (28438.756) = 0.6984\boldsymbol{M_{15}} + 0.9995\boldsymbol{M_{16}}} \ldots \ldots (119)$$

- Solving equation (118) & (119) then:

$$\boldsymbol{M_{15} = 3787.426\ kg\ mole/hr}$$
$$\boldsymbol{M_{16} = 24651.33\ kg\ mole/hr}$$

F.5 MATERIAL BALANCE ON FLASH COLUMN (V-101)

- Total material balance:

$$\mathrm{M_6 = M_7 + M_8}$$

$$\boldsymbol{M_6 = M_7 + 31228.0537} \ldots\ldots (120)$$

- By component balance on H_2O we get:

$$\boldsymbol{0.8857M_6 = 0.52M_7 + 0.8860\ (31228.0537)} \ldots\ldots (121)$$

- Solving equation (120) & (121) then:

$$\boldsymbol{M_6 = 31253.67\ kg\ mole/hr}$$
$$\boldsymbol{M_7 = 25.618\ kg\ mole/hr}$$

F.6 MATERIAL BALANCE ON FERMENTER (R-101) :

- Total material balance:

$$\mathrm{M_{3+}M_4 = M_5 + M_6}$$

$$\boldsymbol{M_{3+}M_4 = M_5 + 31253.67} \ldots\ldots\ (122)$$

- M_6 stream contains 3.24% of hydrogen peroxide which is the major product of this process, so:
  $\mathrm{H_2O_2 = 0.324 * (31253.67) = \ 10126.2\ kg\ mole/hr}$

- Chemical Reaction is given by:

$$\boldsymbol{\mathrm{CH_3}OH + O_2 \rightarrow H_2O_2 + \mathrm{HCHO}}$$

- To find number of O_2 , reaction stoichiometry illustrates that:

  $\mathrm{1\ mole\ of\ H_2O_2\ Produced\ by = 1\ mole\ of\ O_2}$

10126.2 mole of H_2O_2 Produced consume 10126.2 mole of O_2=Total mole of O_2 Feed = M_3

- For 90% Conversion:

  Reacted amount of O_2 within reactor = X_A*Total O_2 feed

  = 0.90 *10126.2 = 9113.58 Kgmol/hr

- Since composition of CH_3OH is too small = 0.003 the value of its was neglected which make stream 5 almost pure O_2

- So 10% oxygen is unreacted hence:

$$\boldsymbol{M_5 = 0.10 * M_3 = 0.10\ (10126.2) = 1012.62\ \text{kg mole/hr}}$$

- Now put the values in overall balance equation we get:

$$\boldsymbol{M_4 = 22140.09\ \text{kg mole/hr}}$$

F.7 MATERIAL BALANCE ON HEAT EXCHANGER:(E-101)

- General material balance on shell side :

$$\text{Input} = \text{Output}$$

- So that:

$$\boldsymbol{M_4 = M_{17} = 22140.09\ kg\ mole/hr}$$

- General material balance on Tube side :

$$\text{Input} = \text{Output}$$

$M_{CW_{in}} = M_{CW_{out}} =$ Mass will be calculated in the energy balance section.

F.8 MATERIAL BALANCE ON MIXER (M-101)

- Total material balance:

$$M_1 + M_2 = M_{17}$$

$$\boldsymbol{M_1 + M_2 = 22140.09} \ldots\ldots\ (123)$$

- Material balance on hydrogen peroxide :

$$\mathbf{0 * M_1 + 0.0004 M_2 = 0.00036 * 22140.09} \ldots\ldots (124)$$

- Solving equations (123) & (124) , then:

$$\boldsymbol{M_1 = 2214.009\ kg\ mol/hr}$$

$$\boldsymbol{M_2 = 19926.081\ kg\ mol/hr}$$

F.9 MATERIAL BALANCE ON ULTRAFILTRATION MEMBRANE (UF-101)

- Total mass balance is:

$$M_6^* = M_6 + M_{\text{enzyme}}$$

$$\boldsymbol{M_6^* = 31253.67 + M_{enzyme}} \ldots\ldots (125)$$

- Material balance on the enzyme:

$$\boldsymbol{0.0011 M_6^* = 0 + 1 * M_{enzyme}} \ldots\ldots\ (126)$$

- Solving equations (125) & (126) , then:

$$\boldsymbol{M_6^* = 31288.0869\ \mathbf{Kgmol/hr}}$$

$$\boldsymbol{M_{\mathbf{enzyme}} = 34.416\mathbf{Kgmol/hr}}$$

APPENDIX G: SAMPLE CALCULATION- ENERGY BALANCE

G.1 ENERGY BALANCE ON THE MIXER M-101

- Total heat balance :

$$\boldsymbol{Q = Heat\ out - Heat\ in}$$

$$\boldsymbol{Q} \ = \sum \dot{n}\hat{H}_{\boldsymbol{out}} - \sum \dot{n}\hat{H}_{\boldsymbol{in}} \dots \dots (127)$$

- $Heat\ in\ from\ stream\ Q_1 = \dot{n}_{CH_3OH} C_p \Delta T + \dot{n}_{H_2O} C_p \Delta T$

Component	Cp (kJ/kmol. K)	$\dot{n}$ component Cp ΔT	$Q_{component}$
CH_3OH	86	1859.76 × 86 × (322.45-295.15)	4366344.528
H_2O	75.3	354.241 × 75.3 × (322.45-295.15)	728209.6813
Q_1 total(KJ/hr)	Q_1 total = $Q_{CH3OH} + Q_{H2O}$		5094554.209

- $Heat\ in\ from\ stream\ Q_2 = \dot{n}_{H_2O} C_p \Delta T + \dot{n}_{H_2O_2} C_p \Delta T$

Component	Cp (kJ/kmol. K)	$\dot{n}$ cp ΔT	$Q_{component}$
H_2O	81.4	19916.117 × 81.4 ×(476.55 -295.15)	292635473
H_2O_2	52.57	7.9704324 × 52.57 ×(476.55 -295.15)	76007.6177
Q_2 total(KJ/hr)	Q_2 total = $Q_{CH3OH} + Q_{H2O}$		292711481.1

- So total Q_{Input} to the system:

$$\boldsymbol{Q_{input} = Q_1 + Q_2 = 5094554.209 + 292711481.1 = 297806035.4}\ \textbf{KJ/hr}$$

- The mixing stream which is M_{17} is inter heat exchanger at 412.15K
- Cp_{CH3OH} at 412.15 K = 117 KJ/Kmol.k [55]
- Cp_{H2O} at 412.15 K = 77 KJ/Kmol.k
- Cp_{H2O2} at 412.15 K = 51.03 KJ/Kmol.k

Component	**ni× cp × (Ti-T_R)**
CH_3OH	**1868.6235 × 117× (412.15-298.15) = 37705084.98**
H_2O	**20263.2745 × 77 × (412.15-298.15) = 177871023.6**
H_2O_2	**7.9704324 × 51.03 × (412.15-298.15) =46364.837**
Q_{17} (KJ/hr)	**215622473.4**

$$\boldsymbol{Q = 215622473.4 - 297806035.4 = -82183561.95\ \mathrm{KJ/hr}}$$

- Q value represent the amount of heat that release if stream 1 mixed with stream 2 .

G.2 ENERGY BALANCE ON THE HEAT EXCHANGER E-101

- Total energy balance:

$$Q_{cooling\ water} + \sum \dot{n}\hat{H}_{out} = \sum \dot{n}\hat{H}_{in}$$

$Q_{17} = 215622473.4$ From the energy balance on the mixture

$$Q_4 = \dot{n}_{CH_3OH}C_p\Delta T + \dot{n}_{H_2O_2}C_p\Delta T$$

Component	**ni× cp × (Ti-TR)**
CH_3OH	**1859.76 × 87 × (323.15-298.15) = 4044978**
H_2O	**20258.1823× 75.4 x (323.15-298.15)=38186673.64**
H_2O_2	**7.970432 × 48.42 × (323.15-298.15) =9648.208**
Q_4 (KJ/hr)	**42241299.85**

- So:

$$\boldsymbol{Q_{cooling\ water} = 215622473.4 - 42241299.85 = 173381173.6\ KJ/hr}$$

$$Q_{cooling\ water} = nC_p(T_{out} - T_{in}) + n\lambda$$

- Water was assumed to inter the heat exchanger at 293.15 K and out at average temperature between stream 4 & 16 which equal 367.65 K , at this temperature specific heat of water is = 75.9 KJ/Kmol.K .

$$173381173.6 = \mathrm{n} * (\ 75.9\,(367.65\ -\ 293.15) +\ 35.3\ )$$

$$\boldsymbol{n} = \mathbf{30472.0113\ Kmol/hr}$$

- This value represents amount of water needed to cool down steam 17 to steam 4 temperature.

G.3 ENERGY BALANCE IN THE FERMENTER (R-101)

- **Input - output + generation - consumption = accumulation**

Assume steady state system so energy balance will be:

- **Input - output + generation - consumption = 0**

$$\dot{\mathrm{n}}\mathrm{Cp}\Delta\mathrm{T}_{\boldsymbol{input}} - \dot{\mathrm{n}}\mathrm{Cp}\Delta\mathrm{T}_{\boldsymbol{output}} + \text{heat of reaction}\ -\ \text{heat of refrigerant R} - 134\text{a} = 0$$

$$\boldsymbol{Q_3} + \boldsymbol{Q_4} - \boldsymbol{Q_5} + \boldsymbol{Q_6} + \Delta\boldsymbol{H_R} - \boldsymbol{Q_{R-134a}} = \mathbf{0} \dots\dots\ (128)$$

- Heat of reaction is -ΔH$_R$

$$\boldsymbol{CH_3OH} + \boldsymbol{O_2} \rightarrow \boldsymbol{H_2O_2} + \boldsymbol{HCHO}$$

$$\begin{aligned} -\Delta H_R &= \Delta H_{pro} - \Delta H_{REA} \\ &= -109.15 - 187.341 - 0 + 238.42 \\ &= \mathbf{-58.071\ KJ/hr} \end{aligned}$$

- Since -ΔH$_R$ in negative sign that give indication that the previous reaction is exothermic reaction which was cold by using R-134a .

- Energy input to the system equals Q$_3$+Q$_4$

$$\boldsymbol{Q_4} = \mathbf{42241299.85}\ \boldsymbol{KJ/hr} \qquad \textit{from energy balance on } (E-101)$$

$$\boldsymbol{Q_3} = \boldsymbol{\dot{n}_3\lambda} \qquad (Q_3 \textit{ is in the vapor phase })$$

- *Where λ is the latent heat of vaporization in KJ/Kmol*

$$= 11251.33\ \frac{\cancel{Kmole}}{hr} \times 0.0034009\ \frac{KJ}{\cancel{Kmol}}$$

$$\mathbf{Q_3} = \mathbf{38.366\ KJ/hr}$$

- $\textit{Input heat to the system} = 42241299.85 + 38.366 =\ 42241338.22\ KJ/hr$

- Energy output from the system equal $Q_5 + Q_6$

$$Q_5 = \dot{n}_{O_2}\lambda = 1120.632 \times 0.0034009 = 3.81116\ KJ/hr$$

- To calculate Q_6 first Cp of mixture should be found:

$$C_{P_{mix}} = \sum yi\ Cpi$$

$$= \sum(y_{CH_3OH}C_{P_{CH_3OH}} + y_{O_2}C_{P_{O_2}} + y_{HCHO}C_{P_{HCHO}} + y_{H_2O}C_{P_{H_2O}} + y_{H_2O_2}C_{P_{H_2O_2}}$$

$$= 0.049 * 74 * 0.00020 * 29.89 + 0.032 * 25.780 + 0.885 * 0 + 0.032 * 45.9$$

$$\boldsymbol{C_{P_{mix}} = 5.925738}$$

$$Q_6 = \dot{n}_6 C_p \Delta T \quad (Q_6\ is\ in\ liquid\ phase)$$

$$= 31253.67\ \frac{\text{Kmole}}{\text{hr}} \times 5.925738\ \frac{\text{KJ}}{\text{Kmol.K}} \times (\ 251.15 - 298.15)\ \text{K}$$

$$\boldsymbol{Q_6 = -8704449.818\ \text{kJ/hr}}$$

- To find $Q_{R\text{-}134a}$ substitute both Q_3 & Q_4 & Q_5 & Q_6 and ΔH_R in equation (128) :

$$Q_{R-134a} = 42241338.22 - (3.81116\ + -8704449.818) - \ 55.071$$

$$\boldsymbol{Q_{R-134a} = 50945729.15\ \text{KJ/hr}}$$

- Inlet temperature of R-134a = -28 C
- Outlet temperature of R-134a = 35 C

$$\boldsymbol{Q_{R-134a} = \dot{m}_{R-134a} C_P}\Delta + \dot{\text{m}}\ \boldsymbol{\lambda} \ldots \ldots (129)$$

- Cp R-134a at average temperature and pressure 8 bar = 1.34849 KJ/Kg.K [58]

$\lambda = 216.97$ KJ/Kg [59]

$$50945729.15 = \dot{\text{m}} * (1.34849 * (35 - -28) + 216.97\ )$$

- By solving equation (129) amount of refrigerate **R-134 a = 168736.44 Kg/hr .**

G.4 ENERGY BALANCE AROUND FLASH COLUMN (V-101)

$$\boldsymbol{q} = \sum(\dot{n}\hat{H})_{\boldsymbol{in}} - \sum(\dot{n}\hat{H})_{\boldsymbol{out}} \quad \ldots\ldots(130)$$

$$\sum(\dot{n}\hat{H})_{\boldsymbol{in}} = \boldsymbol{Q_6} = \boldsymbol{-8704449.818 KJ/hr}$$

$$\sum(\dot{n}\hat{H})_{\boldsymbol{out}} = \boldsymbol{\dot{n}_7}\hat{H}_7 + \boldsymbol{\dot{n}_8}\hat{H}_8$$

$$\hat{H} = \int_{298.15}^{T} Cp\Delta T$$

- Cp for different component at different temperature was found either directly from online site or its was calculated by using Shomate equation . The value of heat capacity is at appendix section.
- Firstly, Q_7 was calculated by using the following equation:

$$Q_7 = \dot{n}_7\hat{H}_7 = \dot{n}_7(C_P\Delta T + \lambda) = (\dot{n}_{7_{HCHO}} + \dot{n}_{7_{CH_3OH}} + \dot{n}_{7_{H_2O}} + \dot{n}_{7_{H_2O_2}} + \dot{n}_{7_{O_2}}) * (C_P\Delta T + \lambda)$$

Substance	**$\dot{n}_{7i}$ (CpΔT + λ)**
CH_3OH	**4.09888 ×92 ×(341.15-298.15) + 4.09888 × 0.03734 = 16215.32**
O_2	**6.14832 × 31.15 ×(341.15-298.15) +6.14832× 0.00340 = 8235.39**
H_2O	**13.32 × 75.5×(341.15-298.15)+13.32 × 0.0408 = 43243.92346**
H_2O_2	**1.02472 × 48.99 × (341.15-298.15)+1.02472× 0.0485 =2158.694109**
HCHO	**1.02472×31.78×(341.15-298.15)+1.02472×0.0233 = 1400.34**
Q_7	**71253.66756 kJ/hr**

- Secondly, Q_8 was calculated by using the following equation:

$$Q_8 = \left(\dot{n}_8\hat{H}_8 = \dot{n}_{8_{HCHO}} + \dot{n}_{8_{CH_3OH}} + \dot{n}_{8_{H_2O}} + \dot{n}_{8_{H_2O_2}}\right) * (C_P\Delta T)$$

Substance	**$\dot{n}_{8i}$ CpΔT**
CH_3OH	**1530.1746 ×92 ×(341.15-298.15) + 1530.1746 × 0.03734 = 345869609.5**
H_2O	**27668.055 × 75.5×(341.15-298.15)+ 27668.05 × 0.0408 = 89825469.41**
H_2O_2	**999.23× 48.99 × (341.15-298.15)+999.2× 0.0485 =2104996.402**
HCHO	**999.23×31.78×(341.15-298.15)+999.23×0.0233 = 1365511.046**
Q_8	**437800075.3 kJ/hr**

- By substitute the value of Q_7 , Q_8 and Q_6 in equation 130 :

$$q = -8704449.818 - 71253.66756 - 437800075.3$$

$$\boldsymbol{q = -446575778.8\ \text{kJ/h}}$$

- The steam enters V-101 at 20 bars so λ = 1888.65 KJ / Kg [60]

$$\boldsymbol{q = \dot{m}_s \lambda} \ldots \ldots (131)$$

$$446575778.8 = \dot{m}_s * 1888.65$$

- Solve equation 131 :

$$\boldsymbol{\dot{m}_s = 236452.37\ \text{Kg /hr}}$$

G.5 ENERGY BALANCE AROUND DISTILLATION COLUMN (T-101)

- Total energy balance :

$$\boldsymbol{Q - W_s = \Delta H + \Delta E_k + \Delta E_p} \ldots \ldots (132)$$

- $W_s = 0 (no\ moving\ part)$
- $\Delta E_k \cong 0$ (neglect kinetic energy change)
- $\Delta E_P = 0$ (no change of elevation)

$$Q = \Delta H$$

$$\boldsymbol{Q_c + Q_r} = \sum \dot{n}\hat{H}_{\boldsymbol{out}} - \sum \dot{n}\hat{H}_{\boldsymbol{in}} \ldots \ldots (133)$$

$$\sum \dot{n}\hat{H}_{\boldsymbol{in}} = \boldsymbol{Q_8} = \boldsymbol{437800075.3\ \text{kJ/hr}}$$

$$\sum \dot{n}\hat{H}_{\boldsymbol{out}} = \boldsymbol{Q_9 + Q_{10}}$$

- $Q_9 = \dot{n}_9 \hat{H}_9 = (\dot{n}_{9_{HCHO}} + \dot{n}_{9_{CH_3OH}} + \dot{n}_{9_{H_2O}} + \dot{n}_{9_{H_2O_2}})(C_P \Delta T)$

Substance	$\dot{n}_{9i}$ CpΔT
CH_3OH	1511.402685×91 ×(339.75-298.15) = 5721566.004
H_2O	27658.66914× 75.5× (339.75-298.15) = 86870348.03
H_2O_2	11.788 × 48.95 × (339.75-298.15) = 24004.14016
HCHO	997.525×31.692× (339.75-298.15)= 1315124.192
Q_9	92615918.17 kJ/hr

$$Q_{10} = \dot{n}_{10}\hat{H}_{10} = \dot{n}_{10_{H_2O_2}} C_P \Delta T = 1000 * 49.51 * (358.05 - 298.15) = 2965649 KJ/hr$$

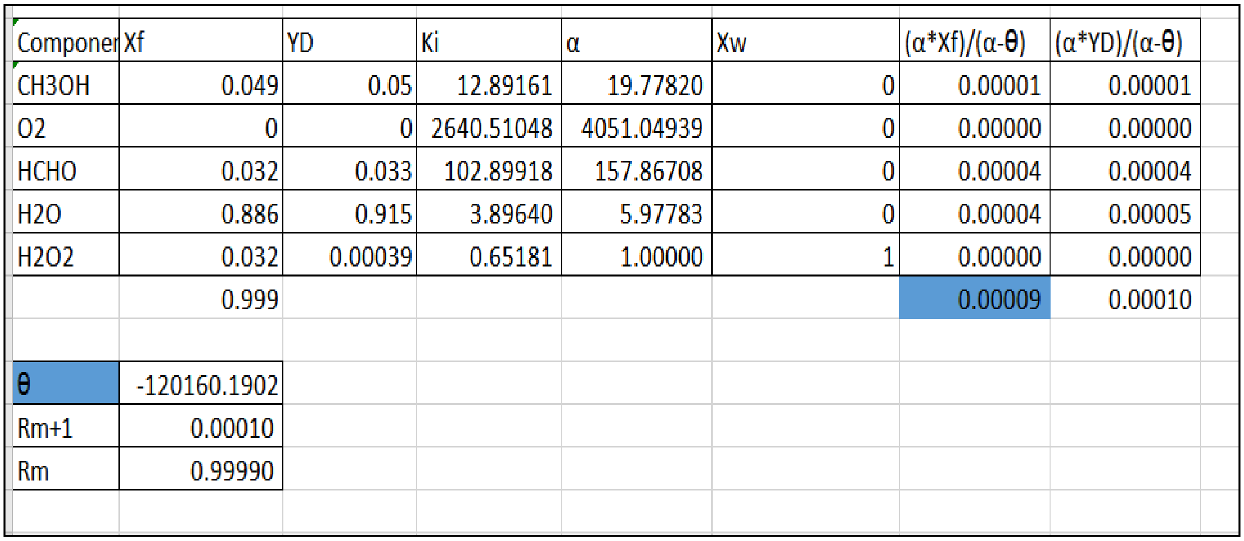
$$\sum \dot{n}\hat{H}_{out} = 92615918.17 + 2965649 = 95581567.17 \text{ kJ/hr}$$

- Know heat of condenser will be calculated according to the following equation:

  $\boldsymbol{Q_c = V_1 \Delta H_{evaporation}}$ … … (134)

- Where $V_1 = L_0 + D$, so reflux ratio needed to be calculated to find the value of V_1.

- Trial and error were done to know the minimum reflux ratio at average temperature.



Componer	Xf	YD	Ki	α	Xw	(α*Xf)/(α-θ)	(α*YD)/(α-θ)
CH3OH	0.049	0.05	12.89161	19.77820	0	0.00001	0.00001
O2	0	0	2640.51048	4051.04939	0	0.00000	0.00000
HCHO	0.032	0.033	102.89918	157.86708	0	0.00004	0.00004
H2O	0.886	0.915	3.89640	5.97783	0	0.00004	0.00005
H2O2	0.032	0.00039	0.65181	1.00000	1	0.00000	0.00000
	0.999					0.00009	0.00010
θ	-120160.1902						
Rm+1	0.00010						
Rm	0.99990						

- So that R_{min} = 0.999 and actual Reflux ratio is 1.2 R_{min} = 1.1988

  $R = \frac{L_0}{D} \rightarrow L_0 = 1.1988 * D \rightarrow V_1 = 1.1988\, D + D \rightarrow V_1 = 2.1988\, D$

  $\boldsymbol{V_1 = 36179.1 Kgmol/hr}$

- $\Delta H_{evaporation} = \sum y_i * H_i$

- Where y_i is the mole fraction and H_i is the heat of evaporation of the individual component in the Q_9.

$$\Delta H_{evaporation} = 0.033 * 0.0233 + 0.05 * 0.03734 + 0.915 * 0.0408 + 0.00039 * 0.0485$$
$$= \mathbf{0.03998\ \ KJ/Kmol}$$

- By substitute the value of ΔHevaporation as well as V_1 in equation 134:

$$\boldsymbol{Q_c} = \mathbf{1446.440KJ/hr}$$

- Qr can be found know by using equation 133:

$$1446.440\ +\ \mathrm{Qr} = 95581567.17\ -\ 437800075.3$$

$$\boldsymbol{Q_r} = -\ \mathbf{342219954.6\ KJ/hr}$$

G.6 ENERGY BALANCE ON FLASH V-103 :

- Total energy balance around distillation column is :

$$\boldsymbol{Q} = \sum(\dot{\mathrm{n}}\hat{\mathrm{H}})_{\boldsymbol{in}} - \sum(\dot{\mathrm{n}}\hat{\mathrm{H}})_{\boldsymbol{out}}\ \ldots\ldots(135)$$

$$\sum(\dot{\mathrm{n}}\hat{\mathrm{H}})_{\boldsymbol{in}} = \boldsymbol{Q_9} = \mathbf{92615918.17\ \ kJ/hr}$$

$$\sum(\dot{\mathrm{n}}\hat{\mathrm{H}})_{\boldsymbol{out}} = \dot{\mathrm{n}}_{\mathbf{11}}\hat{\mathrm{H}}_{\mathbf{11}} + \dot{\mathrm{n}}_{\mathbf{12}}\hat{\mathrm{H}}_{\mathbf{12}}$$

- Firstly, Q_{11} was calculated by using the following equation:

$$Q_{11} = \dot{\mathrm{n}}_{\mathbf{11}}\hat{\mathrm{H}}_{\mathbf{11}} = \dot{\mathrm{n}}_{\mathbf{11}}(\boldsymbol{C_p}\Delta\boldsymbol{T} + \lambda) = \left(\dot{\mathrm{n}}_{\mathbf{11}_{CH_3OH}} + \dot{\mathrm{n}}_{\mathbf{11}_{H_2O}}\right) * (\boldsymbol{C_p}\Delta\boldsymbol{T} + \lambda)$$

Substance	$\dot{\mathbf{n}}_{11i}$ **(CpΔT + λ)**
CH_3OH	**0.00512 ×92 ×(341.25-298.15) + 0.00512 × 0.03734 = 20.30201518**
H_2O	**0.01536 × 75.5×(341.25-298.15)+0.01536× 0.0408 = 49.98283469**
Q_{11}	**70.28484987 kJ/hr**

- Secondly, Q_{12} was calculated by using the following equation:

$$Q_{12} = \dot{\mathrm{n}}_{\mathbf{12}}\hat{\mathrm{H}}_{\mathbf{12}} = \dot{\mathrm{n}}_{\mathbf{12}_{\boldsymbol{HCHO}}} + \dot{\mathrm{n}}_{\mathbf{12}_{\boldsymbol{CH_3OH}}} + \dot{\mathrm{n}}_{\mathbf{12}_{\boldsymbol{H_2O}}} + \dot{\mathrm{n}}_{\mathbf{12}_{\boldsymbol{H_2O_2}}}(\boldsymbol{C_p}\Delta\boldsymbol{T})$$

Substance	$\dot{n}_{12i}$ **CpΔT**
CH_3OH	$\mathbf{1511.5257721} \times 92 \times (341.25-298.15) + 1511.5257721 \times 0.03734 = 5993558.432$
H_2O	$\mathbf{27658.66914} \times 75.5 \times (341.25-298.15) + 27658.66914 \times 0.0408 = 90003820.79$
H_2O_2	$117.8894094 \times 48.89 \times (341.25-298.15) + 117.8894094 \times 0.0485 = 248417.4477$
HCHO	$\mathbf{997.5257721} \times 31.786 \times (341.25-298.15) + 997.5257721 \times 0.0233 = 1366610.208$
Q_{12}	**97612406.88 kJ/hr**

- By substitute the value of Q_{11}, Q_{12} and Q_9 in equation 135 :

$$Q = 92615918.17 - 70.28484987 - 97612406.88$$

$$\boldsymbol{Q = -4996558.995}$$

- The steam enters V-101 at 20 bars so λ = 1888.65 KJ / Kg [60]

$$q = \dot{m}_s \lambda$$

$$\mathbf{4996558.995} = \dot{m}_{\boldsymbol{s}} * \mathbf{1888.65} \dots\dots \ (136)$$

- Solve equation 136:

$$\dot{m}_{\boldsymbol{s}} = \mathbf{2645.572\ Kg\ /hr}$$

G.7 ENERGY BALANCE ON DISTILLATION COLUMN T-102 :

- Overall energy balance around distillation column:

$$Q = \Delta H$$

$$\boldsymbol{Q_c} + \boldsymbol{Q_r} = \sum \dot{n}\hat{H}_{\boldsymbol{out}} - \sum \dot{n}\hat{H}_{\boldsymbol{in}} \dots\dots (137)$$

$$\sum \dot{n}\hat{H}_{in} = Q_{12}$$

$$\sum \dot{n}\hat{H}_{out} = Q_{13} + Q_{14}$$

$$Q_{13} = \dot{n}_{13}\hat{H}_{13} = \left(\dot{n}_{13_{HCHO}} + \dot{n}_{13_{CH_3OH}} + \dot{n}_{13_{H_2O}}\right)(C_P\Delta T)$$

Substance	**$\dot{n}_{13i}$ CpΔT**
CH_3OH	**$1325.869818 \times 91 \times (337.95-298.15) = 4802035.307$**
H_2O	**$458.060288 \times 75.5 \times (337.95-298.15) = 1376425.359$**
HCHO	**$0.041153854 \times 31.579 \times (337.95-298.15) = 51.72398271$**
Q_{13}	**6178512.39 kJ/hr**

Substance	**$\dot{n}_{14i}$ CpΔT**
CH_3OH	**$127.974402 \times 106 \times (386.05-298.15) = 1192388.693$**
H_2O	**$27272.767 \times 76.2 \times (386.05-298.15) = 182672447.9$**
HCHO	**$1009.575838 \times 34.501 \times (386.05-298.15) = 1198101.91$**
H_2O_2	**$11.3755024 \times 50.32 \times (386.05-298.15) = 50315.30318$**
Q_{14}	**185113253.8kj/hr**

$$\sum \dot{n}\hat{H}_{out} = 185113253.8 + 6178512.39 = 191291766.2 \text{ kJ/hr}$$

$$\boldsymbol{Q_c = V_1 \Delta H_{evaporation}} \dots\dots (138)$$

- R_{min} = 0.999 and actual Reflux ratio is 1.2 R_{min} = 1.1988

$$R = \frac{L_0}{D} \rightarrow L_0 = 1.1988 * D \rightarrow V_1 = 1.1988\, D + D \rightarrow V_1 = 2.1988\, D$$

$$\boldsymbol{V_1 = 3922.6 \text{Kgmole/hr}}$$

$$\Delta H_{evaporation} = \sum y_i * H_i$$

$$\Delta H_{evaporation} = 0.000023 \times 0.0233 + 0.741 \times 0.03734 + 0.256 \times 0.0408 = 0.03811 \text{ KJ/Kmol}$$

- By substitute the value of $\Delta H_{evaporation}$ as well as V_1 in equation 138 :

$$\boldsymbol{Q_c = 149.5 \text{KJ/hr}}$$

- Qr can be found Know by using equation 137:

$149.5 + Q_r = 191291766.2 - 97612406.88$

$\boldsymbol{Q_r} = \mathbf{935679209.81KJ/hr}$

G.8 ENERGY BALANCE ON DISTILLATION COLUMN T-103:

- Overall energy balance around distillation column:

$$\boldsymbol{Q_c} + \boldsymbol{Q_r} = \sum \dot{n}\hat{H}_{\boldsymbol{out}} - \sum \dot{n}\hat{H}_{\boldsymbol{in}} \dots\dots (139)$$

$$\sum \dot{n}\hat{H}_{in} = Q_{14} = 185113253.8 \text{ kJ/hr}$$

$$\sum \dot{n}\hat{H}_{out} = Q_{15} + Q_{16}$$

$$Q_{15} = \dot{n}_{15}\hat{H}_{15} = (\dot{n}_{15_{HCHO}} + \dot{n}_{15_{CH_3OH}} + \dot{n}_{15_{H_2O}})(C_P\Delta T)$$

Substance	**$\dot{n}_{15i}$ CpΔT**
CH_3OH	**128.772484 × 183 ×(480.55-298.15) = 4298322.498**
H_2O	**2645.138318 × 81.8× (480.55-298.15) = 39466310.15**
HCHO	**1011.242742 × 39.638 × (480.55-298.15)= 7311255.901**
Q_{15}	**51075888.55 KJ/hr**

Substance	**$\dot{n}_{16i}$ CpΔT**
H_2O	**24639.00434 × 85.9× (512.85-298.15) = 173458918.8**
H_2O_2	**11.3396118× 53.31 × (512.85-298.15)= 129789.3072**
Q_{16}	**173588708.1 *KJ*/hr**

$$\sum \dot{n}\hat{H}_{out} = 51075888.55 + 173588708.1 = 275740485.2\text{kJ/hr}$$

- Know heat of condenser will be calculated according to the following equation:

$$\boldsymbol{Q_c} = \boldsymbol{V_1}\Delta\boldsymbol{H_{evaporation}} \dots\dots (140)$$

R_{min} = 0.999 and actual Reflux ratio is 1.2 R_{min} = 1.1988

$$R = \frac{L_0}{D} \rightarrow L_0 = 1.1988 * D \rightarrow V_1 = 1.1988\, D + D \rightarrow V_1 = 2.1988\, D$$

$\boldsymbol{V_1}$ **= 8322.8 Kgmole/hr**

$$\Delta H_{evaporation} = \sum y_i * H_i$$

$$\Delta H_{evaporation} = 0.267 \times\ 0.0233\ +\ 0.034 \times\ 0.03734\ + 0.6984\ \times\ 0.0408$$

$$= 0.036\ \text{KJ/Kmol}$$

- By substitute the value of ΔHevaporation as well as V_1 in equation 140:

$$\boldsymbol{Q_c = 8322.8 *.036 = 299.5\text{KJ/hr}}$$

- Qr can be found Know by using equation 139:

$$299.5\ +\ \text{Qr} = 275740485.2 - 185113253.8$$

$$\boldsymbol{Q_r = 90626931.9\ \text{KJ/hr}}$$

G.9 TOTAL ENERGY BALANCE AROUND COMPRESSOR C-101

$$\boldsymbol{Q - W_s = \Delta H + \Delta E_K + \Delta E_P} \ldots\ldots (141)$$

- $Q = 0$ (assume adiabatic system)
- $\Delta E_K \cong 0$(neglect kinetic energy change)
- $\Delta E_P = 0$(no change of elevation)

$$-W_s = \Delta H = \dot{\text{n}} C_P (T - T_i)$$

$$C_{P_{average}} = \frac{\text{Cp at } (-22\ ^\circ\text{C})\ +\ \text{Cp at } (90.2\ \ ^\circ\text{C})}{2} = \frac{29.89 +\ 31.6}{2} = 30.745\text{kJ/kmol. K}$$

$$-W_S = 1124.12038 * 30.745 * (90 - (-22))$$

$$-W_s = 3870841.081\ \text{KJ/hr}$$

$$\boldsymbol{W_s = -3870841.081}\ \mathbf{KJ/hr}$$

G.10 ENERGY BALANCE AROUND PUMP (P-102)

Total energy balance on (P-102):

$$\boldsymbol{Q + W_s = \Delta H + \Delta E_K + \Delta E_P} \ldots\ldots (142)$$

- Assume no heat generation, no elevation change, and no temperature change, pump (P-102) is used to pump a mixture of hydrogen peroxide, methanol, some water, and small amounts of formaldehyde from 0.29 bar up to 1.22 bar.

  $W_s = \Delta H$

  $$\boldsymbol{\Delta H = v_{avg} * (P_2 - P_1)} \ldots\ldots (143)$$

  $$\mathbf{v}_{avg} = \frac{1}{978.7} + \frac{1}{760} + \frac{1}{1152} + \frac{1}{760} = \frac{0.0009}{4} \mathrm{m3/kg}$$

  $$\mathbf{v}_{avg} = 0.000225 \mathrm{m3/kg}$$

  $$MW_{avg} = 0.915 * 18 + 0.0039 * 34 + 0.05 * 32 + 0.0334 * 30$$

  $$\boldsymbol{MW_{avg} = 18.2}\ \mathbf{g/mol}$$

  $$H = (1.22 - 0.29) bar * 0.000225 \frac{m3}{kg} * \ 18.2 \frac{g}{mol} (\frac{100000\ pa}{bar} * \frac{kj}{pa.m\hat{}3} * \frac{kg}{1000g}) \ * \text{molar flow}$$

  $$H = (1.22 - 0.29)\ \text{bar} * \frac{m\hat{}3}{kg}$$

  $$H = \ 0.381 \frac{kj}{mol} \ * 30297.57822 \frac{kgmol}{hr} \ (\frac{1hr}{3600s} * \frac{1000gmol}{kgmol} * \frac{1mol}{1gmol})$$

  $$\boldsymbol{H = W_s = 3206.5}\ \mathbf{kJ/s} \ = \mathbf{3206\ kw}$$

APPENDIX H: MASS BALANCE FOR NEW PLANT CAPACITY

50000 tons / year of H_2O_2= 91.8 Kmol/hr

H.1 MATERIAL BALANCE ON DISTILLATION COLUMN (T-101):

Streams	**M_8**	**M_9**	**M_{10}**
Units	**Flowrate**	**Flowrate**	**Flowrate**
Components	**-**	**-**	**-**
HCHO	**91.7466283**	**91.5843104**	**0.15953534**
CH_3OH	**140.487025**	**138.764107**	**1.69338287**
H_2O	**2540.23477**	**2539.38315**	**0.83701929**
H_2O_2	**91.7466283**	**1.08236003**	**89.1100625**
Stream flow rate Kmol/hr	**2864.21505**	**2770.81393**	**91.8**

H.2 MATERIAL BALANCE ON DISTILLATION COLUMN(T-102)

Streams	**M_{12}**	**M_{13}**	**M_{14}**
Units	**Flowrate**	**Flowrate**	**Flowrate**
Components	**-**	**-**	**-**
HCHO	**92.6944233**	**0.00378126**	**92.6862187**
CH_3OH	**138.764107**	**121.822326**	**11.7489573**
H_2O	**2539.38315**	**42.0870653**	**2503.83335**
H_2O_2	**10.8236003**	**-**	**1.04435176**
Stream flow rate kmol/hr	**2775.282**	**164.4026**	**2610.879**

H.3 MATERIAL BALANCE ON DISTILLATION COLUMN (T-103)

Streams	**M_{14}**	**M_{15}**	**M_{16}**
Units	**Flowrate**	**Flowrate**	**Flowrate**
Components	**-**	**-**	**-**
HCHO	**92.6862187**	**92.8395272**	**-**
CH_3OH	**11.7489573**	**11.8222619**	**-**
H_2O	**2504.8777**	**242.843168**	**2262.03383**
H_2O_2	**1.04435176**	**-**	**1.04105609**
Stream flow rate Kmol/hr	**2610.879**	**347.7136**	**2263.165**

H.4 MATERIAL BALANCE ON FLASH COLUMN (V-101)

Streams	M_6	M_7	M_8
Units	Flowrate	Flowrate	Flowrate
Components	-	-	-
HCHO	101.86736	12.6509201	91.7466283
CH_3OH	155.984395	50.6036804	140.487025
H_2O	2819.49753	164.461961	2540.23477
H_2O_2	101.86736	12.6509201	91.7466283
O_2	0.636671	75.9055206	-
Stream flow rate Kmol/hr	2263.165	316.2730	2867.082

H.5 MATERIAL BALANCE ON BIOREACTO (R-101)

Streams	M_3	M_4	M_5	M_6
Units	Flow rate	Flow rate	Flow rate	Flow rate
Components	-	-	-	-
HCHO	0	-	-	103.140702
CH_3OH	0	259.700648	0.031579	155.984395
H_2O	0	2828.88206	-	2817.26918
H_2O_2	0	1.11300278	-	103.140702
O_2	101.86736	-	10.14599	0.636671
Stream flow rate Kmol/hr	101.86736	3091.67438	10.18674	2263.16541

H.6 MATERIAL BALANCE ON HEAT EXCHANGER (E-101)

Streams	M_{17}	M_4
Units	Mole%	Mole%
Components	-	-
CH_3OH	259.7006478	259.700648
H_2O	2828.882056	2828.88206
H_2O_2	1.113002776	1.11300278
Stream flow rate Kmol/hr	3091.674379	3091.67438

H.7 MATERIAL BALANCE ON MIXER (M-101)

Streams	**M_1**	**M_2**	**M_{17}**
Units	**Mole%**	**Mole%**	**Mole %**
Components	**-**	**-**	**-**
CH_3OH	**266.006478**	**------**	**260.937318**
H_2O	**50.66790057**	**2773.6125**	**2829.59314**
H_2O_2	**------**	**1.11**	**1.11300278**
Stream flow rate Kmol/hr	**316.6743786**	**2775**	**3091.67438**

APPENDIX I: ENERGY BALANCE FOR NEW PLANT CAPACITY

I.1 ENERGY BALANCE ON THE MIXER M-101

$Q_1 = 728687.5056\text{KJ/hr}$

$Q_2 = 40965636.41\text{KJ/hr}$

$Q_{input} = Q_1 + Q_2 = 728687.5056 + 40965636.41 = \ 41694323.9\text{kJ/hr}$

$Q_{out} = Q_{17} = 28319561.84\text{ KJ /hr}$

$Q_{\text{realize from mixing process}} = Q_{out} - Q_{in} = 28319561.84 \ - \ 41694323.9 = \ -13374762.9\text{KJ/hr}$

I.2 ENERGY BALANCE ON THE HEAT EXCHANGER E-101

$Q_4 = 5898638.875\text{ KJ/hr}$

$Q_{cooling\ water} = Q_{17} - Q_4 = 22420922.965\text{ KJ/hr}$

Amount of water need to cool the streams = 3940.719911 Kmol/hr (Sample calculation is on Graduation Project 1)

I.3 ENERGY BALANCE IN THE BIOREACTOR (R-101)

$Heat\ \text{of reaction} = -58.071\ \text{KJ/hr}$

$Q_{\text{R134}-\text{a}} = 5012043.985\text{KJ/hr}$

Amount of R-134 a need =16598.88182 Kg/hr

I.4 ENERGY BALANCE AROUND FLASH COLUMN (V-101)

$Q_7 = 879733.3701\text{KJ/hr}$

$Q_8 = 9121400.461\ \text{KJ/hr}$

$\text{Q added by steam} = Q_6 - Q_7 - Q_8 = -11621988.5\ \text{KJ/hr}$

Amount of steam needed = 6153.59567 Kmol/hr

I.5 ENERGY BALANCE AROUND DISTILLATION COLUMN (T-101)

$Q_9 = 8626562.606\text{KJ/hr}$

$Q_{10} = 272246.5782\ \text{KJ/hr}$

$Q_c = 243.96\ \text{KJ/hr}$

$Q_r = -222835.2469\ \text{KJ/hr}$

I.6 ENERGY BALANCE ON DISTILLATION COLUMN T-102

$Q_{13} = 567688.2744\ \text{KJ/hr}$

$Q_{14} = 17172793.5\ \text{KJ/hr}$

$Q_{out} = Q_{13} + Q_{14} = 17740481.77 \text{ KJ/hr}$

$Q_{in} = Q_{12} = 8963414.89 \text{ KJ/hr}$

$Q_c = 13.7\ KJ/hr$

$Q_r = 8777053.098 \text{ KJ/hr}$

I.7 ENERGY BALANCE ON DISTILLATION COLUMN T-103:

$Q_{15} = 4689142.527 \text{ KJ/hr}$

$Q_{16} = 41729994.79 \text{ KJ/hr}$

$Q_c = 27.52389472\ KJ/hr$

$Q_r = 29246316.29\text{KJ/hr}$

I.8 ENERGY BALANCE AROUND PUMP (P-102)

$$H = W_s = 3206.5 \frac{KJ}{s} = 3206\ Kw$$

I.9 TOTAL ENERGY BALANCE AROUND COMPRESSOR C-101

$H = -353735.776 \text{ Kw}$

APPENDIX J: PHYSICAL PROPERTIES OF STREAMS AROUND T-101 /T-102 /T-103

Streams	Mw	Component fraction M_9	MW of M_9=X*Mw	M_{10}	MW of M_{10}=X*Mw	Surface Tension of component for M_9	Surface tension for M_9
Units	-	-	-	Mole%	-	mN/m	-
Components	-	-	-	-	-	-	-
HCHO	30	0.033	0.99	0.00173786	0.052135733	27.38	0.90354
CH_3OH	32	0.05	1.6	0.01844644	0.590285967	18	0.9
H_2O	18	0.915	16.47	0.00911786	0.16412143	64.59	59.09985
H_2O_2	34	0.00039	0.01326	0.97069785	33.00372685	72	0.02808
Sum	-	-	19.07326	-	33.81026998	-	60.93147

Streams	Density Kg/m^3	M_{10}	Density of M10=X*ρi Kg/m^3	Surface Tension of component for M10	Surface tension for M_{10}	Density of M_9=X*ρi Kg/m^3
Units		Mole%	-	N/m	-	-
Components		-	-	-	-	-
HCHO	815	0.00173786	1.416354079	27.38	0.047582546	26.895
CH_3OH	1.2	0.01844644	0.022135724	15	0.276696547	0.06
H_2O	979.6	0.00911786	8.931852944	58.93	0.537315327	896.334
H_2O_2	1200	0.97069785	1164.837418	80.4	78.04410702	0.468
Sum	-	-	1175.207761	-	78.90570144	923.757

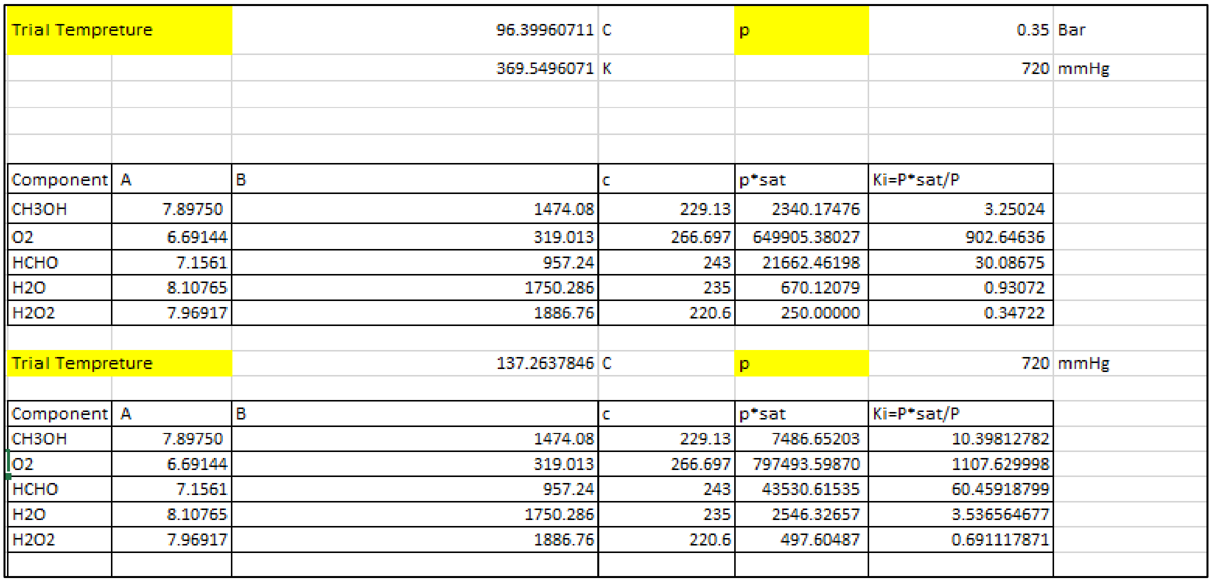
Streams	M_8	Boiling point	Molar average boiling point	Latent heat of	Latent heat of feed	Cp of component	Cp of feed
Units	Mole%	C	-	Kj/mol	-	-	(kJ/kmol. k
Components	-	-	-	-	-	-	
HCHO	0.032	96	3.072	24	0.768	31.78	1.01696
CH_3OH	0.049	6.74E+01	3.3026	53.2	2.6068	92	4.508
H_2O	0.886	100	88.6	42.18	37.37148	75.5	66.893
H_2O_2	0.032	150	4.8	48.95	1.5664	48.99	1.56768
Sum	-	-	99.7746	-	42.31268	-	73.98564

Streams	**M_8**	**Viscosity**	**Average viscosity**
Units	**Mole%**	**-**	**-**
Components	**-**	**-**	**-**
HCHO	**0.032**	**2.46**	**0.07872**
CH_3OH	**0.049**	**3.21E-04**	**0.000015729**
H_2O	**0.886**	**0.001**	**0.000886**
H_2O_2	**0.032**	**1.25**	**0.04**
			0.119621729

Streams	**M_{12}**	**Viscosity**	**Average viscosity**
Units	**Mole%**	**-**	**-**
Components	**-**	**-**	**-**
HCHO	**0.0334**	**2.46**	**0.082164**
CH_3OH	**0.05**	**3.21E-04**	**1.605E-05**
H_2O	**0.915**	**0.001**	**0.000915**
H_2O_2	**0.0039**	**1.25**	**0.004875**
Sum	**1.0023**	**-**	**0.0879701**

Streams	**M_{14}**	**Viscosity**	**Average viscosity**
Units	**Mole%**		
Components	**-**		
HCHO	**0.0355**	**2.46**	**0.08733**
CH_3OH	**0.0045**	**3.21E-04**	**1.4445E-06**
H_2O	**0.9594**	**0.001**	**0.0009594**
H_2O_2	**0.0004**	**1.25**	**0.0005**
Sum	**0.9998**		**0.088790845**

APPENDIX K: DEW AND BABBLE POINT FOR T-101



Trial Tempreture		96.39960711	C	p	0.35	Bar
		369.5496071	K		720	mmHg
Component	A	B	c	p*sat	Ki=P*sat/P	
CH3OH	7.89750	1474.08	229.13	2340.17476	3.25024	
O2	6.69144	319.013	266.697	649905.38027	902.64636	
HCHO	7.1561	957.24	243	21662.46198	30.08675	
H2O	8.10765	1750.286	235	670.12079	0.93072	
H2O2	7.96917	1886.76	220.6	250.00000	0.34722	
Trial Tempreture		137.2637846	C	p	720	mmHg
Component	A	B	c	p*sat	Ki=P*sat/P	
CH3OH	7.89750	1474.08	229.13	7486.65203	10.39812782	
O2	6.69144	319.013	266.697	797493.59870	1107.629998	
HCHO	7.1561	957.24	243	43530.61535	60.45918799	
H2O	8.10765	1750.286	235	2546.32657	3.536564677	
H2O2	7.96917	1886.76	220.6	497.60487	0.691117871	

APPENDIX L: PHYSICAL PROPERTIES OF STREAMS AROUND V-101

Streams	**M6**	**Density**	**X*density**
Units			
Components	**-**	**-**	**-**
HCHO	**0.032**	**815**	**26.08**
CH_3OH	**0.049**	**1.2**	**0.0588**
H_2O	**0.8857**	**979.6**	**867.63172**
H_2O_2	**0.032**	**1200**	**38.4**
O2	**0.0002**	**1.314**	**0.0002628**
Sum	**-**	**-**	**932.1707828**

Streams	**M_7**	**Mw**	**Average MW**
Units	**Mole%**	**-**	**-**
Components	**-**	**-**	**-**
HCHO	**0.04**	**30**	**1.2**
CH_3OH	**0.16**	**32**	**5.12**
H_2O	**0.52**	**18**	**9.36**
H_2O_2	**0.04**	**34**	**1.36**
O_2	**0.24**	**32**	**7.68**
	-	**-**	**24.72**

Streams	M_6	Mw	Average MW	Density	Average density
Units	Mole%	-	-	-	-
Components	-	-	-	-	-
HCHO	0.032	30	0.96	1090	34.88
CH_3OH	0.049	32	1.568	791.1	38.7639
H_2O	0.8857	18	15.9426	998.19	884.096883
H_2O_2	0.032	34	1.088	1200	38.4
O_2	0.0002	32	0.0064	1.134	0.0002268
sum	-	-	19.565	-	996.14101

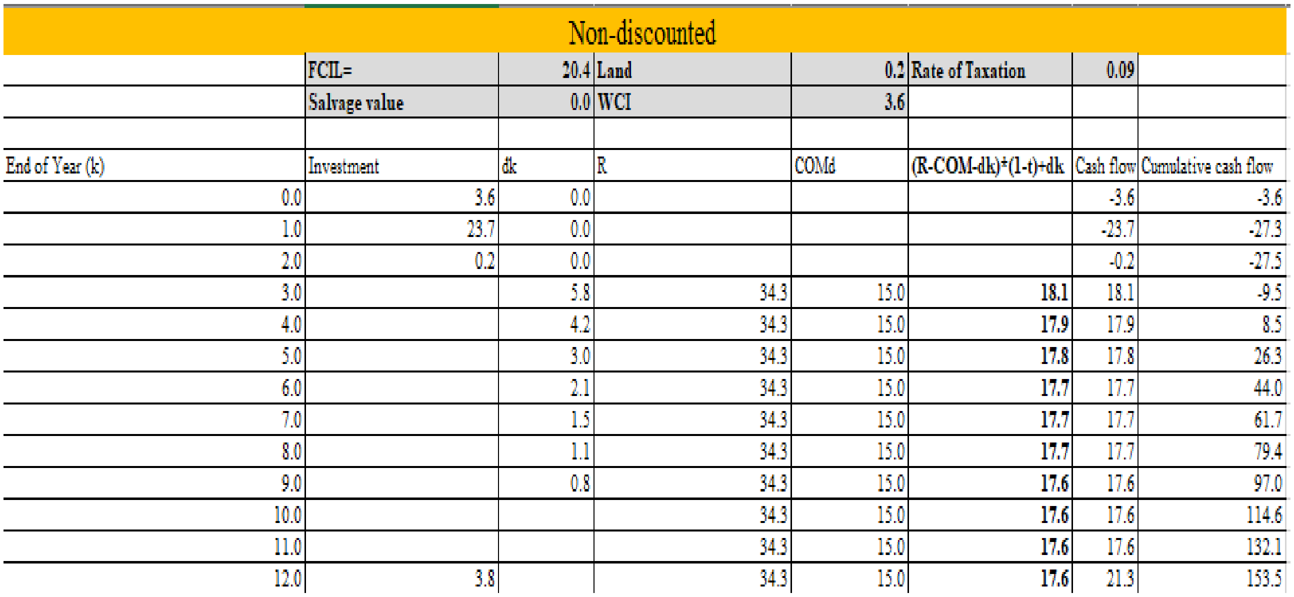
APPENDIX M: PHYSICAL PROPERTIES OF STREAMS AROUND P-101

Streams	M_{13}	Density kg/m^3	Density of M_{13}	Mw	Average MW
Units	Mole%	-	-	-	-
Components	-	-		-	-
HCHO	0.000023	815	0.018745	30	0.00069
CH_3OH	0.741	1.2	0.8892	32	23.712
H_2O	0.256	979.6	250.7776	18	4.608
H_2O_2	0.002977	1200	3.5724	34	0.101218
Sum	1		255.257945		28.421908

APPENDIX N: PHYSICAL PROPERTIES OF STREAMS AROUND HEAT EXCHANGER

	Cp_i at T_{avg} = 109.2 C (KJ/Kmol. K)	Density at T_{avg} = 109.2 C(Kg/ m^3)	(μ) (Pa.s)	K_i	X_i	X_i*Cp_i	ρ_i*X_i	μ_i*X_i	K_i*X_i
CH_3OH	150	689.966 [125]	0.000227 [126]	0.197 [68]	0.084	12.6	57.957	1.9068E-05	0.016548
H_2O	76.1	951.6	0.000256	0.67837	0.915	69.631	870.7	0.0002342	0.62070855
H_2O_2	50.3	962.2 [127]	0.0001	0.4186 [128]	0.00036	0.01749	0.346	0.000000036	0.000151
Cp avg for (mixture) = Σ X_i*Cp_i=82.249KJ/Kmol.K									
Density avg for (mixture) =Σ $X_i*\rho_i$= 929.0175Kg/m^3									
Viscosity avg for (mixture) =Σ $X_i*\mu_i$ = 0.00025334 Pa.s									
Thermal conductivity (K) for avg (mixture) = Σ(K_i*X_i) = 0.6374(W/m. K)									

Appendix O : PROFITABILITY ANALYSIS



Non-discounted							
	FCIL=	20.4	Land	0.2	Rate of Taxation	0.09	
	Salvage value	0.0	WCI	3.6			
End of Year (k)	Investment	dk	R	COMd	(R-COM-dk)*(1-t)+dk	Cash flow	Cumulative cash flow
0.0	3.6	0.0				-3.6	-3.6
1.0	23.7	0.0				-23.7	-27.3
2.0	0.2	0.0				-0.2	-27.5
3.0		5.8	34.3	15.0	**18.1**	18.1	-9.5
4.0		4.2	34.3	15.0	**17.9**	17.9	8.5
5.0		3.0	34.3	15.0	**17.8**	17.8	26.3
6.0		2.1	34.3	15.0	**17.7**	17.7	44.0
7.0		1.5	34.3	15.0	**17.7**	17.7	61.7
8.0		1.1	34.3	15.0	**17.7**	17.7	79.4
9.0		0.8	34.3	15.0	**17.6**	17.6	97.0
10.0			34.3	15.0	**17.6**	17.6	114.6
11.0			34.3	15.0	**17.6**	17.6	132.1
12.0	3.8		34.3	15.0	**17.6**	21.3	153.5