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LAMPIRAN 1

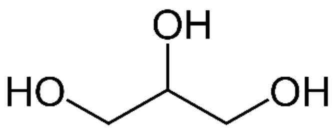
DATA

L1.1 Spesifikasi Bahan Baku dan Produk

Tabel L1. 1 Spesifikasi bahan baku gliserol

Glicerol

Struktur



Identifikasi

EC No	:	200-289-5
Cas No	:	56-81-5
Formula	:	$C_3H_8O_3$
Synonyms	:	1,2,3-Propanetriol, Glycerin

Additional Information

Kemurnian	:	99,50%
Impuritis	:	0,50%

Properties

Berat Molekul	:	92,09 g/mol
Titik didih	:	290 °C
Titik leleh	:	18 °C
Kenampaka	:	Cairan bening
Titik nyala	:	190 °C
Titik *auto-ignition*	:	290 °C
Viskositas	:	954 (25 °C)
Density	:	1,25 g/cm^3

Other Information

Solubility:
Miscible in cold water, hot water and alcohol.
Partially soluble in acetone.
Very slightly soluble in diethyl ether (ethyl ether).
Limited solubility in ethyl acetate.
Insoluble in carbon tetrachloride, benzene, chloroform, petroleum ethers, and oils

Corrosivity : Non-corrosive in presence of glass
Chronic Effects : May cause damage to the following organs: kidneys.
Other Toxic Effects : Slightly hazardous in case of skin contact (irritant), of ingestion, of inhalation.

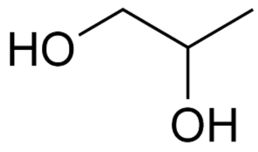
Tabel L1. 2 Spesifikasi bahan baku hydrogen

Hidrogen		
Struktur		
H–H		
Identifikasi		
EC No	:	-
Cas No	:	1333-74-0
Formula	:	H_2
Synonyms	:	Protyum, Dyhidrogen, Parahydrogen,
Additional Information		
Kemurnian	:	100,00%
Impuritis	:	-
Properties		
Berat Molekul	:	2,02 g/mol
Titik didih	:	-252,90 ℃
Titik leleh	:	-259,15 ℃
Density	:	0,082 g/L
Solubility	:	0,0182 v/v at 20
Suhu Kritis	:	-240,15 ℃
Viskositas	:	954 (25 ℃)

Tabel L1. 2 Spesifikasi produk propilen glikol

Propilen Glikol

Struktur



Identifikasi

EC No	:	200-338-0
Cas No	:	57-55-6
Formula	:	$C_3H_8O_2$
Synonyms	:	1,2-Propanediol, 1,2-Dihidroxy Propane

Additional Information

Kemurnian	:	99,00%
Impuritis	:	1,00%

Properties

Berat Molekul	:	76,09 g/mol
Fase	:	Cairan
Titik didih	:	187 ℃
Titik leleh	:	-60 ℃
Density	:	1,05 g/cm^3
Viskositas (20 ℃)	:	56,0 cP
Kapasitas panas	:	2,481 J/(g ℃)

Other Information

Soluble in cold water, hot water, acetone
Non-corrosive in presence of glass.
Chronic Effects: May cause damage to the following organs: central nervous system (CNS).
Other Toxic Effects: Hazardous in case of ingestion. Slightly hazardous in case of skin contact (irritant, permeator), of inhalation.

L1.2 Sumber Literatur

(19) Europäisches Patentamt / European Patent Office / Office européen des brevets

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(54) **PRODUCTION OF PROPYLENE GLYCOL FROM GLYCEROL**

(57) The present invention describes a process for the production of propylene glycol from glycerol, the transformation of purified glycerol to propylene glycol being carried out by means of a reaction of hydrogenolysis, in the liquid phase, where the two stages of the reaction take place simultaneously and in one and the same reactor (1) under specified conditions of temperature and pressure, and the effluent from the fixed-bed reactor (1) is led to subsequent process stages that comprise separation and purification.

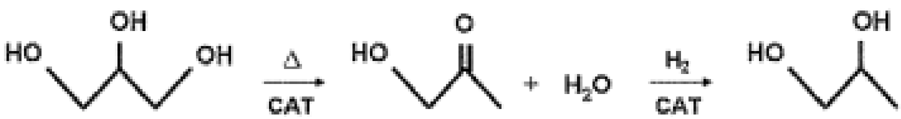

FIG. 1

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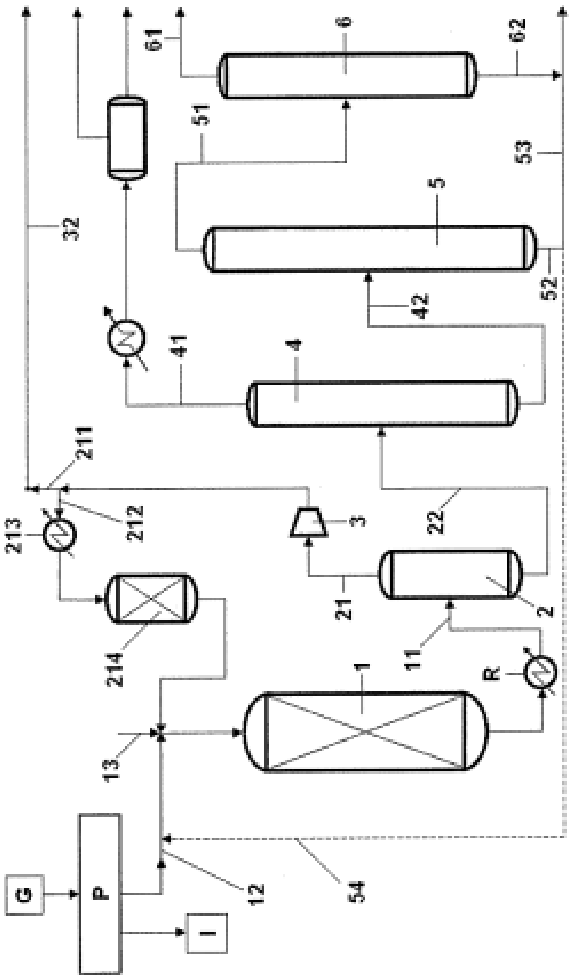

FIG. 2

Description

FIELD OF THE INVENTION

[0001] The present invention relates to processes for the production of propylene glycol from glycerol. Specifically, the invention relates to the production of propylene glycol from glycerol by a reaction of hydrogenolysis.

BACKGROUND OF THE INVENTION

[0002] Currently we are seeing a significant increase in supply of glycerol on the market, largely due to worldwide expansion of biodiesel production. In the process of production of biodiesel, glycerol is a by-product generated at a weight ratio of 10/1 (biodiesel/glycerol).

[0003] However, this excess in the supply of glycerol has a direct impact on reduction of its commercial value, so that its use in various other applications becomes viable.

[0004] Various studies have been initiated with the aim of making use of this surplus glycerol as a raw material for the production of new products.

[0005] The number of technical publications on this subject is increasing as solutions are being devised for utilizing this glycerol surplus.

[0006] One of the products under investigation that is arousing interest is propylene glycol and its production from glycerol, whatever its origin.

PRIOR ART

[0007] From the prior art in this area, we may mention in particular some documents that are representative of the development of processes relating to this subject.

[0008] Document WO 2008/051540 A2 (Archer-Daniels-Midland Company) presents a process for the production of propylene glycol from glycerol that uses a fluidized-bed reactor operating at temperatures in the range from 178°C to 205°C, pressures in the range from 1200 psi to 1600 psi, with hydrogen/glycerol molar ratio in the range from 1:1 to 10:1, and a space velocity in the range from 0.5 h^{-1} to 10 h^{-1}. The feed supplied to the reactor is a solution, predominantly aqueous, containing at least 30 wt.% of glycerol in an alkaline medium selected from metal hydroxides, alkoxides, basic salts and metal oxides.

[0009] Document WO 2007/053705 A2 (University of Missouri Board of Curators) presents a process for producing a mixture of acetol and propylene glycol in any proportions from glycerol in the vapour phase. The glycerol partial pressure is in the range from 0.01 bar to 0.5 bar and the temperature is in the range from 80°C to 300°C. The hydrogen partial pressure in the process is in the range from 0.01 bar to 25 bar. A heterogeneous catalyst is used that has at least one element of group I or VIII of the periodic table, ruthenium, copper, chromium, or a combination of these elements.

[0010] Documents US 2005/0244312 and PI 0507874-1 (Galen J. Suppes, William Rusty Sutterlin and Mohanprasad Dasari) present processes for the production of acetol, propylene glycol or a mixture of glycerol and propylene glycol, mainly for use as antifreeze, from glycerol as raw material. For the production of propylene glycol, the emphasis is on a process conducted in two stages, the first stage being the production of acetol by reactive distillation, in which the glycerol is dehydrated, obtaining acetol with the aid of a catalyst, the acetol being distilled immediately from the reaction mixture with the aid of a catalyst. This first stage employs reactors in batch or semi-batch mode, a temperature in the range from 170°C to 270°C, pressure in the range from 0.2 bar to 25 bar, the glycerol for the feed must contain between 0 and 15 wt.% of water, with a heterogeneous catalyst that can have an element selected from palladium, nickel, rhodium, chromium, copper or a combination of these, preferably a catalyst based on a combination of copper and chromium. In a second stage, the acetol produced is mixed with hydrogen and undergoes hydrogenation to propylene glycol with the aid of a catalyst. In this second stage the reaction time is in the range from 0.1 to 24 hours, the temperatures are in the range from 50°C to 250°C, preferably from 150°C to 220°C, the pressure is between 1 bar and 25 bar, preferably between 10 bar and 20 bar, the acetol feed contains a maximum of 50 wt.% of water, preferably between 0% and 35%, with a catalyst that can be selected in the same way as for the first stage, a catalyst based on a combination of copper and chromium being preferred.

[0011] Document WO 2008/049470 A1 (Clariant International Ltd.) presents an autoclave process for the production of propylene glycol from glycerol, in which said glycerol must have a purity of at least 95%, under hydrogen pressure in the range from 20 bar to 100 bar, at a temperature in the range from 180°C to 240°C and in the presence of a catalyst that comprises from 20% to 0% of copper oxide, 30% to 70% of zinc oxide and 1% to 10% of manganese oxide.

[0012] Document WO 2007/010299 (Davy Process Technology Ltd.) presents a process for the production of propylene glycol from glycerol, carried out in the vapour phase in the presence of a copper-based catalyst. The process temperature is in the range from 160°C to 260°C and the pressure is in the range from 10 bar to 30 bar. The volume ratio of hydrogen to glycerol is between 400:1 and 600:1 with a space velocity between 0.01 h^{-1} and 2.5 h^{-1}.

[0013] Document WO 2008/012244 (Davy Process Technology Ltd.) presents modifications to the process in the aforementioned document by adding various stages to the process with emphasis on reduction of the amount of hydrogen required for the process when operating in this regime with stages.

[0014] Document WO 03/035582 (Battelle Memorial Institute & Michigan State University) presents a process for hydrogenolysis of sugars containing 6 carbon atoms and polyols, including glycerol, using a multi-metal solid catalyst that contains rhenium. The process is conducted at a temperature in the range from 120°C to 250°C, with a pH of the liquid feed in the range from 8 to 13 and a hydrogen pressure during the hydrogenolysis reaction of the glycerol ranging from 600 psi to 1800 psi.

[0015] Document WO 2008/133939 (Archer-Daniels-Midland Company) presents a process of catalytic hydrogenolysis of glycerol in which the space velocity of the feed is in the range from 0.5h^{-1} to 2.5h^{-1}. The feed is composed of a solution that contains between 25% and 40% of glycerol USP (*United States Pharmacopeia*), alkalized with hydroxide or alkoxide of sodium or potassium so that a pH is obtained in the range from 6 to 14. The hydrogen pressure is between 500 psi and 2000 psi, at a temperature in the range from 150°C to 300°C and a hydrogen/glycerol molar ratio in the range from 1:1 to 10:1.

[0016] The document emphasizes that, besides producing propylene glycol as main compound, there is formation of butanediols and a proportion of other diols in the range from 0.04% to 2.31%.

[0017] The section for purification/separation of products proposed in this process comprises a stage of removal of the alkalizing agent by means of an ion exchange resin and by stages of distillation of the products. In the first distillation column, water and light alcohols are removed at the top and are sent to a second distillation column to be separated. The effluent from the bottom of the first distillation column goes to a plurality of small distillation columns, in which the residual water and unconverted glycerol are separated. The stream enriched with propylene glycol is led to a distillation column for separation of other by-products, principally ethylene glycol.

[0018] Document US 2008/0275277 A1 (Peter Kalagias) presents a process for purification of propylene glycol and ethylene glycol, produced from a renewable source. The process uses a polar solvent to aid separation by distillation of the components of the mixture. In this process, a mixture of polyols produced in hydrogenolysis of the renewable raw material, including glycerol, is mixed with a polar solvent that does not form an azeotrope with the components of the mixture, and is then distilled. This type of distillation, called extractive distillation, is used when components of a mixture have a relatively low volatility, reducing the chances of efficient separation by conventional distillation. The polar solvent that is added interacts in different ways with the components of the mixture, causing a change in the relative volatility of the components, thus making it possible to separate them.

[0019] It has been observed that investigations are continuing, although the processes mentioned above are observed to have very little preoccupation with the stage of separation of the products.

[0020] The present invention presents an alternative route for obtaining propylene glycol, with refinement of the process as a whole; as well as recovery, separation and final treatment of the products obtained.

SUMMARY OF THE INVENTION

[0021] The present invention relates to a process for the production of propylene glycol from glycerol, in which the reactions take place in the liquid phase and the glycerol used as raw material is, specifically, the by-product from the process for obtaining biodiesel.

[0022] The glycerol obtained from biodiesel, called crude glycerol, is produced with a purity in the range from 40% to 85%. For it to be used as a raw material for the production of propylene glycol according to the present invention, it must be purified by means of purification processes normally employed in the industry, giving a glycerol with a degree of purity in the range from 90 to 99.9 wt.%.

[0023] The transformation of this purified glycerol to propylene glycol is effected by means of a reaction of hydrogenolysis, which is conducted in the liquid phase, with the two stages of the reaction taking place simultaneously and in one and the same reactor.

[0024] In the first stage the glycerol, under the action of temperature and a catalyst, is transformed to acetol and water. In the second stage the acetol, under the action of a catalyst and hydrogen gas, is transformed to propylene glycol.

[0025] The recycle gas is led to a methanation reactor, for converting the CO and CO_2 impurities present in the hydrogen recycle stream to methane, for subsequent recycling to the reactor, minimizing the effects of catalyst deactivation and maintaining glycerol conversion above 95% and selectivity of propylene glycol above 90%.

[0026] The effluent from the reactor is led to a section for separation and purification, after which a final product is

DETAILED DESCRIPTION OF THE INVENTION

[0028] The present invention relates to a process for the production of propylene glycol from glycerol of any origin, in which the reactions take place in the liquid phase.

[0029] The glycerol stream can come from biodiesel units, soap-works or production of fatty acids.

[0030] For purposes of illustration, the glycerol that will be referred to hereinafter originates from the process for obtaining biodiesel, where it is generated as a by-product.

[0031] The glycerol obtained in the course of the process for production of biodiesel initially has a degree of purity in the range from 40% to 85% due to the presence of numerous reaction by-products, for example:

- salts containing sodium, chlorine, sulphur and phosphorus, which deactivate the catalyst irreversibly;
- fatty acids, phospholipids, glycerides, soaps and biodiesel residues, which deactivate the catalyst by blocking the pores.

[0032] The production of propylene glycol has this glycerol as raw material. However, before it is admitted to the hydrogenolysis reactor, the glycerol is first submitted to a stage of concentration and a stage of purification, for removing impurities such as: salts and non-glycerol organic material (NGOM), resulting in a glycerol stream with a degree of purity normally situated in a range of values between 90% w/w and 100% w/w. The process most commonly used for the purification stage is vacuum fractional distillation.

[0033] After being purified, the glycerol is transformed to propylene glycol by means of a reaction of hydrogenolysis. This reaction is carried out in the liquid phase, in which the two stages of reaction take place simultaneously and in one and the same reactor, i.e. the glycerol, under the action of temperature and a catalyst, is transformed to acetol and water and then, under the action of a catalyst and hydrogen gas, the acetol is transformed to propylene glycol. The chemical reactions involved in the process are shown in Fig. 1.

[0034] The process for the production of propylene glycol from glycerol obtained from biodiesel, according to the present invention, can be described referring to Fig. 2. Basically, it comprises the following steps:

a) submitting the glycerol (G), received from biodiesel production, containing impurities, to a stage of purification (P) for removing the impurities (I) by any known purification process, provided said process supplies a glycerol with a degree of purity in the range from 90 to 99.9 wt.%;

b) proceeding to the reaction of hydrogenolysis, in a fixed-bed adiabatic reactor (1) operating in descending flow, in the liquid phase, where the two stages of reaction take place simultaneously as follows:

- activating a catalyst based on copper chromite by passing hydrogen through it;
- pressurizing the fixed-bed reactor (1) to a pressure that can be in the range from 5 kgf/cm^2 to 50 kgf/cm^2;
- raising the temperature of the catalyst bed to a working temperature that is in the range from 160°C to 260°C;
- starting admission of purified glycerol with a certain water content via a feed line (12) and starting admission of hydrogen via a gas line (13), with a molar ratio of glycerol to hydrogen in the range from 10 mol/mol to 120 mol/mol and a space velocity in the range from 0.1h^{-1} to 10h^{-1} with respect to the limiting reactant (glycerol);

c) cooling by means of a cooler (R) and withdrawing the effluent preferably from the bottom of this reactor and conveying it via a reacted product line (11);

d) sending the effluent from the fixed-bed reactor (1) via the reacted product line (11) to a phase separating vessel (2), to remove a stream containing hydrogen and non-condensable light compounds by means of a first top stream (21), and removal of liquid effluent by means of a first bottom stream (22);

e) sending the first top stream (21) received from the phase separating vessel (2) to a recycling compressor (3), which compresses the contents of this stream, where a first portion (211) is conveyed via a first purge line (32) intended for burning by incineration;

f) sending a second portion (212) of the first top stream (21) received from the phase separating vessel (2) to a heater (213), and then to a methanation reactor (214) for transformation of contaminating compounds CO and CO_2 present in the first top stream (21) to methane, then to be mixed in the feed line (12) and recycled to the hydrogenolysis such as methanol, ethanol, n-propanol and isopropanol, while the bottom effluent is conveyed via a second bottom stream (42);

h) sending the bottom effluent of the second bottom stream (42) received from the primary fractionating tower (4) to a tower for removing heavy fractions (5), which removes unreacted glycerol and other heavy products formed during the hydrogenolysis reaction, which are conveyed via a third bottom stream (52) to destinations that can be selected from: a second burning line (53) for incineration and power generation and a second recycling line (54) where glycerol is returned to the fixed-bed reactor (1), while propylene glycol containing impurities is conveyed via the third top stream (51);

i) sending the effluent of the third top stream (51) received from the tower for removing heavy fractions (5) to a purification tower (6), which removes impurities such as ethylene glycol, 1,3-propylene glycol and a small content of heavy compounds, which are conveyed via a fourth bottom stream (62) as far as the junction with the second burning line (53) for incineration and power generation, while propylene glycol with a high degree of purity is withdrawn by means of the fourth top stream (61).

[0035] The methanation reactor (214) employs a nickel-based catalyst supported on alumina, operates in a temperature range between 200°C and 450°C, at pressure in the range from 5 kgf/cm^2 to 50 kgf/cm^2, and a space velocity in the range from 5000 h^{-1} to 20 000 h^{-1}.

[0036] The CO and CO_2 are converted to methane in the methanation reactor (214) according to the following reactions (1) and (2):

$$CO + 3H_2 \rightarrow CH_4 + H_2O \qquad (1)$$

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O \qquad (2)$$

[0037] Both CO and CO_2 act as reversible poisons for the catalyst, being adsorbed on the metal sites and reducing their hydrogenating function, temporarily. Because of this, the conversion and selectivity of the catalyst are gradually reduced and heavy compounds are deposited on the surface of the catalyst, through condensation of the glycerol that was not converted.

[0038] At the same time, other benefits may be obtained:

a) lower consumption of glycerol: greater selectivity and productivity;
b) lower consumption of hydrogen: reduction of the purge;
c) operation at lower temperatures;
d) longer catalyst campaign;
e) reduction in size of the process equipment;
f) reduction of operating cost;
g) reduction of the amount of liquid and gaseous effluents;
h) increase in operating time of the unit.

[0039] Owing to the exothermic nature of the reactions involved, the fixed-bed reactor (1) can contain more than one catalyst bed and have a cooling system between the beds. The propylene glycol produced according to the present invention has an estimated degree of purity above 99% with low content of impurities.

[0040] Two examples of application of the process according to the present invention will be presented below, purely for purposes of illustration.

EXAMPLE 1

[0041] After activation *in situ* of the catalyst, by passing hydrogen through it, the system was pressurized to approximately 20 kgf/cm^2 with hydrogen and the working temperature was adjusted to 225°C.

[0042] Purified glycerol and hydrogen in a glycerol/hydrogen molar ratio of 1/120 was admitted at a temperature of 225°C with a space velocity of 0.5 h^{-1} with respect to the limiting reactant, glycerol, into the fixed-bed adiabatic reactor (1) containing 557 mL of catalyst. The fixed-bed reactor (1) operates with a temperature gradient of 10°C. Analysis showed that the effluent from the reactor (1) had a conversion of glycerol of about 99% and the selectivity for propylene glycol was 89.8%.

EXAMPLE 2

[0043] The same reaction system as in EXAMPLE 1 was supplemented with a methanation reactor (214) as the means for removing CO and CO_2 formed in the process. The operating conditions were maintained, initially constant and equal to EXAMPLE 1.

[0044] Once complete conversion of the glycerol was observed, the temperature was reduced to 220°C, to give a content of 0.2% of glycerol in the final product.

[0045] It was observed that with the introduction of the aforesaid system for reducing impurities, the catalyst displayed greater activity and selectivity in comparison with the results in EXAMPLE 1.

[0046] The full composition of the effluent from the reactor in each of the two examples can be compared with the aid of the following Table 1:

TABLE 1

COMPONENT	FORMULA	Example 1 T=225°C	Example 2 T=220°C
Methanol	C2H60	0.1%	0.08%
Ethanol	C3H8	0.2%	0.06%
Ethylene glycol	C2H602	0.2%	0.01%
n-Propanol	C3H80	0.8%	0.98%
Isopropanol	C3H80	0.1%	0.73%
Acetol	C3H602	2.3%	1.66%
Propanoic acid	C3H602	0.2%	0.01%
1,2-propylene glycol	C3H8O2	89.8%	93.52%
1,3-Propanediol	C3H802	0.1%	0.01%
2,4-Dimethyl-2-hydroxymethyl-1,3-dioxolane	C6H1204	0.9%	0.70%
2-Methyl-2,4-dihydroxymethyl-1,3-dioxolane	C6H 1203	1.8%	0.71%

[0047] Although the invention has been described in its preferred embodiment, the main concept that guides the present invention, which is a process for the production of propylene glycol from glycerol, in particular from glycerol from biodiesel, is preserved with respect to its innovative character, where a person skilled in the art will be able to envisage and implement variations, modifications, changes, adaptations and the like, conceivable and compatible with the working means in question, but without departing from the scope and spirit of the present invention, which are represented by the following claims.

Claims

1. **PRODUCTION OF PROPYLENE GLYCOL FROM GLYCEROL**, where the glycerol employed as raw material may be of any origin, and the process basically comprises the following steps that are to be carried out:

a) submitting the glycerol (G) received from biodiesel production and with impurities to a stage of purification (P) for removing the impurities (I) by any known purification process, provided said process supplies a glycerol with a degree of purity in the range from 90 to 99.9 wt.%;
b) proceeding to the reaction of hydrogenolysis, in a fixed-bed adiabatic reactor (1) operating in descending flow, in the liquid phase, where the two stages of the reaction take place simultaneously as follows:

■ activating a catalyst based on copper chromite by passing hydrogen through it;
■ pressurizing the fixed-bed reactor (1) to a pressure that can be in the range from 5 kgf/cm^2 to 50 kgf/cm^2;
■ raising the temperature of the catalyst bed to a working temperature that is in the range from 160°C to 260°C;
■ starting admission of purified glycerol with a certain water content via a feed line (12) and starting admission of hydrogen via a gas line (13), with a molar ratio of glycerol to hydrogen in the range from 10

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mol/mol to 120 mol/mol and a space velocity in the range from 0.1 h^{-1} to 10 h^{-1} with respect to the limiting reactant (glycerol);

c) cooling by means of a cooler (R) and withdrawing the effluent preferably from the bottom of this reactor and conveying it via a reacted product line (11);
d) sending the effluent from the fixed-bed reactor (1) via the reacted product line (11) to a phase separating vessel (2), to remove a stream containing hydrogen and non-condensable light compounds by means of a first top stream (21), and removal of the liquid effluent by means of a first bottom stream (22);
e) sending the first top stream (21) received from the phase separating vessel (2) to a recycling compressor (3), which compresses the contents of this stream where a first portion (211) is conveyed via a first purge line (32) intended for burning by incineration;
f) sending the liquid effluent from the first bottom stream (22) received from the phase separating vessel (2) to a primary fractionating tower (4), which discharges, via a second top stream (41), water and light polar compounds such as methanol, ethanol, n-propanol and isopropanol, while the bottom effluent is conveyed via a second bottom stream (42);

e) sending the first top stream (21) received from the phase separating vessel (2) to a recycling compressor (3), which compresses the contents of this stream where a first portion (211) is conveyed via a first purge line (32) intended for burning by incineration;

f) sending the liquid effluent from the first bottom stream (22) received from the phase separating vessel (2) to a primary fractionating tower (4), which discharges, via a second top stream (41), water and light polar compounds such as methanol, ethanol, n-propanol and isopropanol, while the bottom effluent is conveyed via a second bottom stream (42);

g) sending the bottom effluent of the second bottom stream (42) received from the primary fractionating tower (4) to a tower for removing heavy fractions (5), which removes unreacted glycerol and other heavy products formed during the hydrogenolysis reaction, which are conveyed via a third bottom stream (52) to destinations that can be selected from: a second burning line (53) for incineration and power generation and a second recycling line (54) where glycerol is returned to the fixed-bed reactor (1), while propylene glycol with impurities is conveyed via the third top stream (51);

h) sending the effluent of the third top stream (51) received from the tower for removing heavy fractions (5) to a purification tower (6), which removes impurities such as ethylene glycol, 1,3-propylene glycol and a small content of heavy compounds, which are conveyed via a fourth bottom stream (62) as far as the junction with the second burning line (53) for incineration and power generation, while propylene glycol with a high degree of purity is withdrawn by means of the fourth top stream (61).

characterized in that a second portion (212) of the first top stream (21) received from the phase separating vessel (2) is sent to a heater (213), and then goes into a methanation reactor (214) for transformation of contaminating compounds CO and CO_2 present in the first top stream (21) to methane, then to be mixed in the feed line (12) and recycled to the hydrogenolysis reactor (1).

2. **PRODUCTION OF PROPYLENE GLYCOL FROM GLYCEROL**, according to claim 1, **characterized in that** the methanation reactor employs a nickel-based catalyst supported on alumina, operates in a temperature range between 200°C and 450°C, at pressure in the range from 5 kgf/cm^2 to 50 kgf/cm^2, and a space velocity in the range from 5000 h^{-1} to 20 000 h^{-1}.

3. **PRODUCTION OF PROPYLENE GLYCOL FROM GLYCEROL**, according to claim 1, **characterized in that** a conversion of glycerol above 95% and selectivity for propylene glycol greater than 90% are achieved.

LAMPIRAN 2

NERACA MASSA DAN ENERGI

L2.1 Informasi Umum

L2.1.1 Basis Perhitungan

Basis kapasitas	:	Produk
Kapasitas pabrik	:	50.000 ton/tahun
Sifat operasional	:	Kontinyu
Operasional pabrik		
Jumlah hari	:	330 hari/tahun
Jumlah jam	:	24 jam/hari
Kapasitas proses produksi	:	$\frac{50.000}{tahun} \times \frac{1000}{ton} \times \frac{1}{330\ hari} \times \frac{1}{24\ jam}$
	:	6.313,131 kg/jam
Massa bahan baku	:	8.373 kg
Produk yang dihasilkan	:	6.3131,31 kg propilen glikol/ 76960 kg gliserol
Total bahan baku yang dibutuhkan	:	76960 kg/jam (837,02 kmol/jam)

L2.1.2 Komposisi massa bahan baku dan produk

Tabel L2. 1 Komposisi Massa Bahan Baku

Komponen	Gliserol	Hidrogen
	% Massa	
Gliserol	99.5	-
Air (Pengotor)	0.5	-
Hidrogen	-	100

Tabel L2. 2 Komposisi Massa Produk

Komponen	Propilen Glikol
	% Massa
Propilen Glikol	99
Air	1

L2.1.3 Properti Bahan

Tabel L2. 3 Berat Molekul Bahan pada Proses Produksi…*(sebutkan nama produk)*

No.	Komponen	MR (kg/kmol)
1.	Gliserol	92
2.	Air	18
3.	Hidrogen	2
4.	Propilen Glikol	76

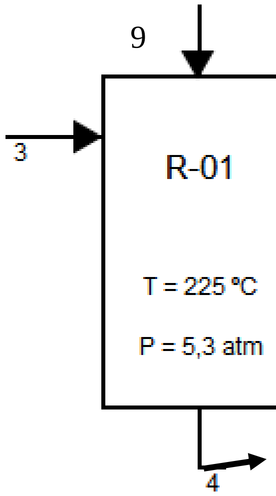
Tabel L2. 4 Panas Spesifik Bahan pada Proses Produksi Propilen Glikol

Komponen	A	B	C
Gliserol	132,14	0,86	-0,002
Air	92,05	-0,04	-0,0002
Hidrogen	25,40	0,02	-0,00004
Propilen Glikol	118,61	0,67	-0,002
Asetol	57,31	0,64	-0,002

Tabel L2. 5 Properti Fisik dan Kimia Bahan pada Proses Produksi 225 ^oC

L2.2 Reaktor *Fixed Bed Reactor (R-01)*

L2.2.1 Diagram Alir Reaktor



L2.2.2 Neraca Mol Reaksi

Reaksi 1 Hidrogenasi

Parameter Reaksi 1

Reaktan pembatas	:	Gliserol
Konversi	:	100% terhadap reaktan pembatas pada suhu 225 ^oC
Rasio reaktan	:	Rasio mol Gliserol: H_2 = 1: 10

kmol/jam	$C_3H_8O_3$	+	H_2	→	$C_3H_8O_2$	+	H_2O
Awal	90,55		418,63		-		-
Reaksi	90,55		90,55		90,55		90,55
Sisa	0,00		328,08		90,55		90,55

L2.2.3 Neraca Massa Fixed Bed Reactor

Tabel L2. 6 Ringkasan Neraca Massa Reaksi Hidrogenasi

Komponen	Masuk (kg/jam)		
	Aliran 3	**Aliran 9**	**Aliran 4**
$C_3H_8O_3$	8.330,73	-	153,15
H_2O	41,86	-	1.671,79
H_2	-	837,26	656,16
$C_3H_8O_2$	-	-	6.881,90
Total		9.209,85	9.209,85

L2.2.4 Neraca Energi Reaktor

Temperatur aliran masuk = 90 ^oC

Temperatur aliran keluar = 225 ^oC

Tabel L2. 7 Panas Reaksi Hidrogenasi di *Fixed Bed Reactor (R-01)*

	Reaktan			Produk		
	n terpakai (kmol/h)	**ΔH_f* (kJ/kmol)**	**ΔH_f (kJ/h)**	**n dihasilkan (kmol/h)**	**ΔH_f* (kJ/kmol)**	**ΔH_f (kJ/h)**
$C_3H_8O_3$	90,55	-582800	-5,28, E+07	-	-	-
H_2O	-	-241800	-	90,55	-241.800	-2,25,E+07
H_2	90,55	0	0	0	0	0
$C_3H_8O_2$	-	-	-	90,55	-421500	-3,82,E+07
	ΔH_f reaktan	-5,28,E+07			**ΔH_f produk**	-6,06,E+07
		ΔH_R (kJ/h)	**7.289.384,72**			
		H reaksi (kJ/h)	**7.289.384,72**			

Tabel L2. 8 Entalpi Aliran Masuk dan Keluar Reaktor *Fixed Bed Reactor (R-01)*

Komponen	Masuk			Keluar		
	n	**H***	**H**	**n**	**H***	**H**
$C_3H_8O_3$	90,55	12.997	1,18,E+06	-	-	-
H_2O	2,33	3.620,27	8,42,E+03	92,88	7.084,32	6,58,E+05
H_2	418,63	1.809,78	7,58,E+05	328,08	5.438,69	1,78,E+06
$C_3H_8O_2$	-	-	-	90,55	26.263,36	2,38,E+06
Total Entalpi Masuk (H_{in})			**1,94,E+06**	**Total Entalpi Keluar (H_{out})**		**4,82,E+06**

n = mol komponen (kmol/h); H* = entalpi spesifik komponen (kJ/kmol); H = entalpi total komponen (kJ/h)

Kalkulasi Neraca Energi *Fixed Bed Reactor* (R-01)

Tujuan: untuk menentukan apakah reaktor membutuhkan sistem pemanas/pendingin

$$H_{in} + \Sigma H_{reaksi} + Q = H_{out}$$

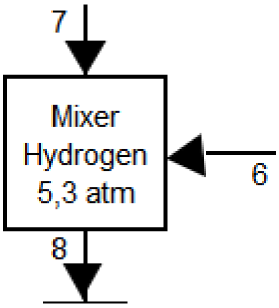
$$1.942.984,41 + 7.289.384,72 + Q = 4.820.469,29$$

$$Q = 10.729.228,45 \text{ kJ/h}$$

Tanda positif pada nilai Q menunjukkan bahwa, untuk menjaga temperatur keluaran reaktor tetap di nilai 225 ^{o}C maka Reaktor R-01 harus dilengkapi dengan sistem pemanas yang memberi panas ke reaktor sebesar 10.729.228,45 *kJ/h*. Besaran nilai panas yang harus diberikan ke reaktor ini kemudian digunakan untuk menghitung kebutuhan air pemanas yang rinciannya akan dijabarkan pada Lampiran 3.

L2.3 Mixing Tank (M-01)

L2.3.1 Diagram Alir Proses Pencampuran



L2.3.2 Neraca Massa Mixing Tank (M-01)

Tabel L2. 9 Ringkasan Neraca Massa Mixing Tank

Komponen	Masuk (kg/jam)		
	Aliran 7	**Aliran 6**	**Aliran 8**
H_2 *fresh feed*	181,10	-	-
H_2 *recycle*	-	656,16	-
H_2 out	-	-	837,26
Total	837,26		837,26

L2.3.3 Neraca Energi Mixing Tank (M-01)

Kalkulasi Neraca Energi

Tujuan: menghitung temperatur produk Mixing Tank T-01 akibat adanya perubahan entalpi pada proses pencampuran gas H_2 *fresh feed* dan gas H_2 *recycle*.

Tabel L2. 10 Entalpi Aliran Masuk Mixing Tank M-01 pada 30 ºC

Komponen	Masuk		
	m	H*	H
H_2 *fresh feed*	181,10	139,90	12668,58
		Total Entalpi Masuk (H_{in})	12668,58

m = massa komponen (kg/h); H* = entalpi spesifik komponen (kJ/kg); H = entalpi total komponen (kJ/h)

Tabel L2. 2 Entalpi Aliran Masuk Mixing Tank M-01 pada 80 ºC

Komponen	Masuk		
	m	H*	H
H_2 *recycle*	656,16	1532,97	502934,26
		Total Entalpi Masuk (H_{in})	502934,26

m = massa komponen (kg/h); H* = entalpi spesifik komponen (kJ/kg); H = entalpi total komponen (kJ/h)

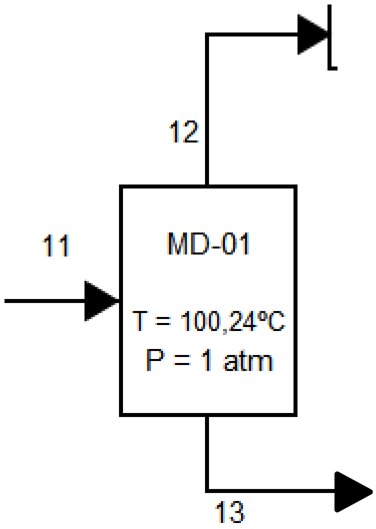
Tabel L2. 2 Entalpi Aliran Keluar Mixing Tank T-01 pada 69.14 ºC

Komponen	Masuk		
	m	H*	H
H_2 out	837,26	1232,67	515602,84
		Total Entalpi Masuk (H_{in})	515602,84

m = massa komponen (kg/h); H* = entalpi spesifik komponen (kJ/kg); H = entalpi total komponen (kJ/h)

L2.4 Kolom Distilasi (Kode Alat)

L2.4.1 Diagram Alir Proses Distilasi



L2.4.2 Spesifikasi Kemurnian Produk

Spesifikasi distilat: 99% air, 1% propilen glikol

Spesifikasi bottom: 100% propilen glikol, 1% air (konsentrasi propilen glikol 99%)

Light key (LK) component: air

Heavy key (HK) component: propilen glikol

Recovery LK di distilat: 99% dari massa LK di feed

Recovery HK di bottom: 99% dari massa HK di feed

L2.4.3 Neraca Massa Kolom Distilasi

Prosedur kalkulasi neraca massa

Tabel L2. 11 Ringkasan Neraca Massa Kolom Distilasi

Komponen	**Masuk (kg/jam)**		**Keluar (kg/jam)**			
	Feed		**Overhead**		**Bottom**	
	kg/jam	%	kg/jam	%	kg/jam	%
H_2O	1.671,79	19,54	1.604,92	85	66,87	1
$C_3H_8O_2$	6.881,90	80,46	275,28	15	6.606,63	99
Sub Total	8.553,69	100	1.880,19	19,71	6.673,50	80,29
Total	8.553,69		8.553,69			

Overhead: aliran vapor dari bagian atas kolom menuju kondenser; Bottom: aliran liquid dari bagian bawah kolom menuju reboiler.

L2.4.4 Neraca Energi Kolom Distilasi V-101

Temperatur umpan: 100,23 ºC

Neraca Energi Condenser

Tujuan: menghitung panas yang harus dibuang di condenser agar overhead vapor dapat terkondensasi menjadi saturated liquid.

A. Diagram alir condenser

B. Penentuan temperatur dan tekanan operasi condenser

1. Temperatur **bubble** saturated liquid hasil kondensi. Dicari dengan cara goal seek total Yd = 1. Yd = ki.xi

P	1	atm	=	760	mmHg
T	100,01		C		

Komponen	XD	Pi º	Ki	Yd	Pi
Propilen Glikol	0,04	747,05	0,98	0,04	29,16
H20	0,96	760,45	1,00	0,96	730,76
JUMLAH	1,00	1507,50	1,98	1,00	759,92

2. Temperatur aliran overhead vapor yang menuju condenser (temperatur **dew**). Dicari dengan cara goal seek total Xd = 1. Xd = yi/ki

P	1	atm	=	760	mmHg
T	100,02		C		

Komponen	XD	Pi º	Ki	Xd	Pi
Propilen Glikol	0,039	747,172	0,983	0,040	29,168
H_2O	0,961	760,585	1,001	0,960	730,893
Jumlah	1,000	1507,757	1,984	1,000	760,061

C. Kalkulasi neraca energi condenser

Tabel L2. 12 Entalpi Aliran Masuk Kondenser di Kolom Distilasi V-101

Komponen	Overhead Vapor @Tdew		
	m	H*	H
H_2O	1604,92	4108,55	366326,53
$C_3H_8O_2$	275,28	10243,41	37102,17
	Total Entalpi Overhead (H_{in})		403.428,70

m = massa komponen (kg/h); H* = entalpi spesifik komponen (kJ/kg); H = entalpi total komponen (kJ/h)

Tabel L2. 14 Entalpi Aliran Keluar Kondenser di Kolom Distilasi V-101

Komponen	Overhead Vapor @Tbubble		
	m	H*	H
H_2O	1604,92	4108,55	37100,07
$C_3H_8O_2$	275,28	10243,41	366304,82
	Total Entalpi Overhead (H_{in})		403404,89

m = massa komponen (kg/h); H* = entalpi spesifik komponen (kJ/kg); H = entalpi total komponen (kJ/h)

Tabel L2. 135 Entalpi Kondensasi Kondenser di Kolom Distilasi V-101

Komponen	Overhead Vapor @Tdew		
	n	H*	H
H_2O	89162,00	39,50	3522051,69
$C_3H_8O_2$	3622,05	61,76	223704,33
	Total Entalpi Overhead (H_{in})		3745756,03

n = mol komponen (mol/jam); H* = entalpi spesifik komponen (kJ/mol); H = entalpi total komponen (kJ/h)

$$H_{in} + H\text{ cond} = H_{out} + Q\text{ pendingin}$$

$$403.428,70 + 3745756,03 = 403404,89 + Q\text{ pendingin}$$

$$Q = 3.745.779,84\text{ kJ/h}$$

Dari hasil kalkulasi diketahui bahwa untuk menghasilkan kondensat saturated liquid pada temperature 100,01ºC, air (atau udara) pendingin harus menyerap panas sebesar 3.745.779,84 kJ/h. Proses kalkulasi massa air (atau udara) pendingin yang dibutuhkan untuk menyerap panas tersebut akan dijabarkan pada lampiran tentang utilitas.

Neraca Energi Reboiler

Tujuan: menghitung panas yang harus disuplai ke reboiler agar aliran liquid dari bottom dapat teruapkan sebagian menjadi saturated vapor (boil-up).

A. Diagram alir Reboiler

B. Penentuan temperatur dan tekanan operasi reboiler

Tujuan dari langkah ini adalah untuk menghitung:

1. Temperatur dew boil-up dari reboiler. Dicari dengan cara goal seek total Yd = 1. Yd = ki.xi

P	1	atm	=	760	mmHg
T	100,55	C			

Komponen	Xb	Pi º	Ki	Yb	Pi
Propilen Glikol	0,96	759,75	1,00	0,96	728,62
H20	0,04	775,20	1,02	0,04	31,77
JUMLAH	1,00	1534,95	2,02	1,00	760,39

2. Temperatur aliran bottom yang menuju reboiler (temperatur **bubble**). Dicari dengan cara goal seek total Xd = 1. Xd = yi/ki

P	1	atm	=	760	mmHg
T	100,54	C			

Komponen	Xb	Pi º	Ki	Xb	Pi
Propilen Glikol	0,96	759,48	1,00	0,960	728,35
H20	0,04	774,88	1,02	0,040	31,76
JUMLAH	1,00	1534,36	2,02	1,00	760,11

C. Kalkulasi neraca energi reboiler

Asumsi: 5,4% massa liquid bottom teruapkan menjadi saturated vapor dan diumpakan kembali ke kolom distilasi sebagai boil-up. Sisanya berfasa saturated liquid dan diambil sebagai produk bottom. *(Diskusikan dengan pembimbing terkait asumsi % liquid bottom yang teruapkan).*

Tabel L2. 1614 Entalpi Aliran Masuk Reboiler di Kolom Distilasi V-101

Komponen	Overhead Vapor @Tdew

	m	H*	H
H_2O	66,87	4133,59	15356,64
$C_3H_8O_2$	6606,63	10303,04	895636,06
	Total Entalpi Overhead (H_{in})		910992,70

m = massa komponen (kg/h); H* = entalpi spesifik komponen (kJ/kg); H = entalpi total komponen (kJ/h)

Tabel L2. 14 Entalpi Aliran Keluar Reboiler di Kolom Distilasi V-101

Komponen	Overhead Vapor @Tbubble		
	m	H*	H
H_2O	66,87	4133,16	15355,02
$C_3H_8O_2$	6606,63	10302,00	895545,61
	Total Entalpi Overhead (H_{in})		910900,62

m = massa komponen (kg/h); H* = entalpi spesifik komponen (kJ/kg); H = entalpi total komponen (kJ/h)

Tabel L2. 155 Entalpi Boil-up Reboiler di Kolom Distilasi V-101

Komponen	Overhead Vapor @Tdew		
	n	H*	H
H_2O	200,61	373,68	39,48
$C_3H_8O_2$	4694,18	373,68	61,72
	Total Entalpi Overhead (H_{in})		297666,04

n = mol komponen (mol/jam); H* = entalpi spesifik komponen (kJ/mol); H = entalpi total komponen (kJ/h)

$$H_{in} + Q \text{ pemanas } = H_{out} + H \text{ boil up}$$

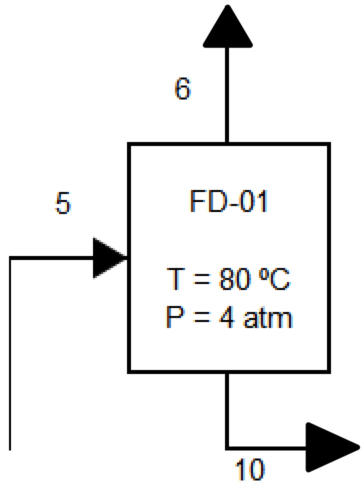
$$910992,70 + Q = 910900,62 + 297666,04$$

$$Q = 297573,97 \text{ kJ/h}$$

Dari hasil kalkulasi diketahui bahwa untuk menghasilkan saturated vapor dengan massa 360,37 kg/h pada temperature 100,53 ^oC, panas yang harus disuplai ke reboiler adalah sebesar 297573,97 kJ/h. Proses kalkulasi massa steam yang dibutuhkan sebagai sumber panas akan dijabarkan pada lampiran tentang utilitas.

L2.5 H_2 Separator (FD-01)

L2.5.1 Diagram Alir Proses Separasi



L2.5.2 Spesifikasi Proses Separasi

Rasio fasa uap terhadap liquid di feed (V/L):

Temperatur operasi: 80 ^oC

Tekanan operasi: 4 bar

L2.7.3 Neraca Massa H_2 Separator (FD-01)

Buat langkah-langkah kalkulasi neraca massa flash separator seperti yang telah diajarkan di kelas OTK 2 (rujuk kembali flash distillation di materi OTK 2).

Tabel L2. 16 Ringkasan Neraca Massa Flash Separator

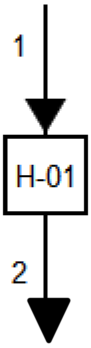
Komponen	**Masuk (kg/jam)**	**Keluar (kg/jam)**	
	Aliran 5	**Aliran 6**	**Aliran 10**
H2	656,16	656,16	-
C3H8O2	6.881,90	-	6.881,90
H2O	1.671,79	-	1.671,79
Total	9.209,85	9.209,85	

L2.5.4 Neraca Energi H2 Separator (FD-01)

Proses pada flash drum separator tidak melibatkan perubahan entalpi karena yang terjadi hanya proses separasi gas-cair tanpa ada perubahan entalpi sehingga tidak perlu dilakukan kalkulasi neraca energi.

L2.6 Heater (H-01)

L2.6.1 Diagram Alir Proses



L2.6.2 Spesifikasi Proses Pertukaran Panas

Temperatur masuk fluida panas: 270 ^oC

Temperatur keluar fluida panas: 270 ^oC

Temperatur masuk fluida dingin: 30 ^oC

Temperatur keluar fluida dingin: 90 ^oC

L2.6.3 Neraca Massa Heater (H-01)

Yang terjadi hanya proses pertukaaran panas tanpa ada perubahan fasa yang menyebabkan terjadinya perubahan massa sehingga kalkulasi neraca massa tidak perlu dilakukan.

L2.6.4 Neraca Energi Heater (H-01)

Stream (Hin)							
Komponen	Fasa	Mr (g/mol)	m (Kg/Jam)	n (Kmol/Jam)	T (K)	∫CpdT (kJ/Kmol K)	H in (kJ/Jam)
Air	Liquid	18	41,86	2,33	303.15	304,83	708,94
Gliserol	Liquid	92	8330,73	90,55	303.15	1061,23	96095,39

			TOTAL				96804,33

Stream (Hout)							
Komponen	Fasa	Mr (g/mol)	m (Kg/Jam)	n (Kmol/Jam)	T (K)	∫CpdT (kJ/Kmol K)	H in (kJ/Jam)
Air	Liquid	18	41,86	2,33	363,15	3620,27	8419,73
Gliserol	Liquid	92	8330,73	90,55	363,15	12.997	1176938,24
			TOTAL				1185357,97

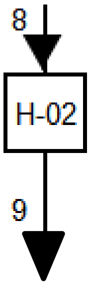
H_{in} + Q pemanas = H_{out}

96804,33 + Q = 1185357,97

Q pemanas = 1088553,64 kJ/jam

L2.7 Heater (H-02)

L2.7.1 Diagram Alir Proses



L2.7.2 Spesifikasi Proses Pertukaran Panas

Temperatur masuk fluida panas: 270 ^oC

Temperatur keluar fluida panas: 270 ^oC

Temperatur masuk fluida dingin: 69,14 ^oC

Temperatur keluar fluida dingin: 90 ^oC

L2.7.3 Neraca Massa Heater (H-02)

Yang terjadi hanya proses pertukaaran panas tanpa ada perubahan fasa yang menyebabkan terjadinya perubahan massa sehingga kalkulasi neraca massa tidak perlu dilakukan.

L2.7.4 Neraca Energi Heater (H-02)

Stream (Hin)							
Komponen	Fasa	Mr (g/mol)	m (Kg/Jam)	n (Kmol/Jam)	T (K)	∫CpdT (kJ/Kmol K)	H in (kJ/Jam)
Hidrogen	Liquid	2	837,26	418,63	342,29	1231,56	515568,53
TOTAL							515568,53

Stream (Hout)							
Komponen	Fasa	Mr (g/mol)	m (Kg/Jam)	n (Kmol/Jam)	T (K)	∫CpdT (kJ/Kmol K)	H in (kJ/Jam)
Hidrogen	Liquid	2	837,26	418,63	363,15	1809,78	757626,44
TOTAL							757626,44

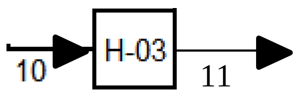
H_{in} + Q pemanas = H_{out}

515568,53 + Q = 757626,44

Q pemanas = 242057,91 kJ/jam

L2.8 Heater (H-03)

L2.8.1 Diagram Alir Proses



L2.8.2 Spesifikasi Proses Pertukaran Panas

Temperatur masuk fluida panas: 270 ^{o}C

Temperatur keluar fluida panas: 270 °C

Temperatur masuk fluida dingin: 69,14 °C

Temperatur keluar fluida dingin: 90 °C

L2.8.3 Neraca Massa Heater (H-03)

Yang terjadi hanya proses pertukaaran panas tanpa ada perubahan fasa yang menyebabkan terjadinya perubahan massa sehingga kalkulasi neraca massa tidak perlu dilakukan.

L2.8.4 Neraca Energi Heater (H-03)

Stream (Hin)							
Komponen	Fasa	Mr (g/mol)	m (Kg/Jam)	n (Kmol/Jam)	T (K)	∫CpdT (kJ/Kmol K)	H in (kJ/Jam)
Air	Liquid	18	1671,79	92,88	353,15	3113,54	289176,50
Propilen Glikol	Liquid	76	6881,90	90,55	353,15	7830,55	709066,64
TOTAL							998243,13

Stream (Hin)							
Komponen	Fasa	Mr (g/mol)	m (Kg/Jam)	n (Kmol/Jam)	T (K)	∫CpdT (kJ/Kmol K)	H in (kJ/Jam)
Air	Liquid	18	1671,79	92,88	373,39	4119,13	382573,09
Propilen Glikol	Liquid	76	6881,90	90,55	373,39	10268,62	929837,13
TOTAL							1312410,22

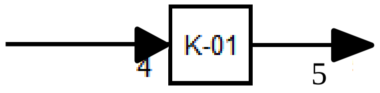
H_{in} + Q pemanas = H_{out}

998243,13 + Q = 1312410,22

Q pemanas = 314167,09 kJ/jam

L2.9 Kondensor (K-01)

L2.9.1 Diagram Alir Proses



L2.9.2 Spesifikasi Proses Pertukaran Panas

Temperatur masuk fluida panas: 225 ^oC

Temperatur keluar fluida panas: 80 ^oC

Temperatur masuk fluida dingin: 30 ^oC

Temperatur keluar fluida dingin: 40 ^oC

L2.9.3 Neraca Massa Kondensor (K-01)

Yang terjadi hanya proses pertukaaran panas tanpa ada perubahan fasa yang menyebabkan terjadinya perubahan massa sehingga kalkulasi neraca massa tidak perlu dilakukan.

L2.9.4 Neraca Energi Kondensor (K-01)

Diketahui stream masuk merupakan produk keluaran dari reaktor, jadi

Hin kondensor = Hout reactor.

Hout reaktor = 4.820.469,29 kJ/jam

Stream (Hin)							
Komponen	Fasa	Mr (g/mol)	m (Kg/Jam)	n (Kmol/Jam)	T (K)	∫CpdT (kJ/Kmol K)	H in (kJ/Jam)
Air	Liquid	18	1671,79	92,88	353,15	3113,54	289176,50
Hidrogen	Gas	2	656,16	328,08	353,15	1532,97	502934,26
Propilen Glikol	Liquid	76	6881,90	90,55	353,15	7830,55	709066,64
TOTAL							1.501.177,40

H_{in} + Q pemanas = H_{out}

4.820.469,29 + Q = 1.501.177,40

Q pemanas = 3319291,89 kJ/jam

LAMPIRAN 3

UTILITAS

L3.1 Air Umpan Boiler

Kebutuhan air umpan boiler dihitung berdasarkan laju alir massa steam yang dibutuhkan pada produksi propilen glikol 99%. Pada pabrik ini steam dibutuhkan sebagai fluida panas pada beberapa unit yaitu *heater* gliserol 99,5% (H-01), *heater* hidrogen (H-02), *fixed bed reactor* (R-01) dan *heater* umpan masuk menara destilasi (H-03), dan *reboiler* menara destilasi (MD-01)

L3.1.1 Kebutuhan Steam untuk Heater H-01

Kebutuhan pemanas di heater (H-01)

Steam		
Media Pemanas	Saturated Steam	
Kebutuhan Panas (ΔH)	1.088.553,64 kJ/jam	
Suhu Steam in	270	°C
Suhu Steam out	270	°C
Tekanan Steam	0,98	bar
ΔHfg (ΔHv)	1604,6	kJ/kg
Laju Alir Steam	678,40	kg/jam

L3.1.2 Kebutuhan Steam untuk Heater H-02

Steam		
Media Pemanas	Saturated Steam	
Kebutuhan Panas (ΔH)	242057,91 kJ/jam	
Suhu Steam in	270	°C
Suhu Steam out	270	°C
Tekanan Steam	5,23	bar
ΔHfg (ΔHv)	1604,6	kJ/kg
Laju Alir Steam	150,85	kg/jam

L3.1.2 Kebutuhan Steam untuk Fixed Bed Reactor (R-01)

Steam		
Media Pemanas	Saturated Steam	
Kebutuhan Panas (ΔH)	10.729.228,45 kJ/jam	
Suhu Steam in	270	°C
Suhu Steam out	270	°C
Tekanan Steam	5,23	bar
ΔHfg (ΔHv)	1604,6	kJ/kg
Laju Alir Steam	6.686,54	kg/jam

L3.1.2 Kebutuhan Steam untuk Heater H-03

Steam		
Media Pemanas	Saturated Steam	
Kebutuhan Panas (ΔH)	314167,09 kJ/jam	
Suhu Steam in	270	°C
Suhu Steam out	270	°C
Tekanan Steam	0,99	bar
ΔHfg (ΔHv)	1604,6	kJ/kg
Laju Alir Steam	195,79	kg/jam

L3.1.2 Kebutuhan Steam untuk Reboiler (RB-01)

Steam		
Media Pemanas	Saturated Steam	
Kebutuhan Panas (ΔH)	297573,97 kJ/jam	
Suhu Steam in	270	°C
Suhu Steam out	270	°C
Tekanan Steam	0,99	bar
ΔHfg (ΔHv)	1604,6	kJ/kg
Laju Alir Steam	185,45	kg/jam

L3.1.4 Total Kebutuhan Air Umpan Boiler

1) Kebutuhan steam dihitung berdasarkan entalpi vaporisasi yang dimiliki steam (ΔHfg/ΔHvap)

Unit Pemanas	Suhu Operasi (°C)	Suhu Steam (°C)	Kebutuhan Steam (kg/jam)
H-01	90	270	678,40
H-02	90	270	150,85
R-01	225	270	6686,54
H-03	100,24	270	195,79
RB-01	100,54	270	185,45
		Total Kebutuhan	1025,04
		Total Kebutuhan Make Up	102,50

L3.2 Air Pendingin

Kebutuhan air pendingin dihitung berdasarkan laju alir massa air yang dibutuhkan pada produksi propilen glikol 99%. Pada pabrik ini air dibutuhkan sebagai fluida dingin pada beberapa unit yaitu kondenser (K-01), kondenser menara destilasi (K-02), dan cooler (C-01).

L3.2.1 Kebutuhan Air Pendingin untuk Kondenser (K-01)

Media Pendingin	Air Pendingin	
Panas Yang harus dihilangkan (Q)	3.319.291,89 kJ/jam	
Suhu Air in	30	°C
Suhu Air out	40	°C
Cp Air	1.339,46	kJ/kg°C
Laju Alir Air	247,81	kg/jam

L3.2.2 Kebutuhan Air Pendingin untuk Kondenser Menara Destilasi (K-02)

Media Pendingin	Air Pendingin	
Panas Yang harus dihilangkan (Q)	3.745.779,84 kJ/jam	
Suhu Air in	30	°C
Suhu Air out	40	°C
Cp Air	1.339,46	kJ/kg°C
Laju Alir Air	279,65	kg/jam

L3.2.3 Kebutuhan Air Pendingin untuk Cooler C-01

Media Pendingin	Air Pendingin	
Panas Yang harus dihilangkan (Q)	797147,84 kJ/jam	
Suhu Air in	30	°C
Suhu Air out	40	°C
Cp Air	1.339,46	kJ/kg°C
Laju Alir Air	59,51	kg/jam

L3.2.4 Total Kebutuhan Air Pendingin

Berdasarkan perhitungan air pendingin yang dibutuhkan untuk mendinginkan suatu fluida di Neraca Energi
1) Semua air pendingin yang digunakan adalah air pada suhu ruang sekitar 30°C dengan keluaran 40°C
2) Kebutuhan air pendingin dihitung
berdasarkan Q = m Cp ΔT
3) Cp air adalah 1.339,46 kJ/kg°C

Unit Pendingin	Suhu Operasi (°C)	Suhu Air Pendingin Masuk (°C)	Suhu Air Pendingin Keluar (°C)	Kebutuhan Steam (kg/jam)
K-01	80,00	30	40	247,81
K-02 (MD)	100,01	30	40	279,65
C-01	100,55	30	40	45079,29
			Total Kebutuhan	60349.33

L3.3 Kebutuhan Air Domestik

1) Karyawan

Kebutuhan Air per Orang	120	Liter/hari
Jumlah Karyawan	145	Liter/hari
Kebutuhan Air untuk Karyawan	5600	L/hari
	233	Liter/jam

2) Laboratorium

Kebutuhan Air Laboratorium	20.0000	Liter/jam
	480.0000	L/hari

Total Kebutuhan Air Domestik	6080	L/hari
	6	m^3/hari
	0.3	m^3/jam
Densitas air pada suhu 30°C	996	kg/m^3

Total Kebutuhan	252	kg/jam

L3.4 Listrik

L3.4.1 Listrik Perkantoran

Tabel L3. 4.1 Kebutuhan Listrik Perkantoran

	kW/hari	kW/jam
1) Peralatan Bengkel Pemeliharaan dan perbaikan peralatan pabrik	40.0000	1.6667
2) Instrumentasi Alat instrumen yang berupa alat kontrol dan alat pendeteksi	10.0000	0.4167
3) Penerangan dan Pendingin Ruangan, Alat dibutuhkan di lingkungan pabrik, perkantoran dan lingkungan sekitar pabrik	50.0000	2.0833
4) Penunjang Kegiatan Pabrik (Alat perkantoran seperti komputer, intercom, pengeras suara dan lainnya)	40.0000	1.6667
Total (kWh)		5.8333

L3.4.2 Listrik Proses

Tabel L3. 1. 2 Kebutuhan Listrik Proses

Alat	Fungsi	Daya (HP)
C-01	Mengalirkan gas hidrogen dari mixing tank ke reaktor	3200,00
C-02	Mengalirkan hidrogen dari H2 separator (FD-01) ke mix tank (T-02)	102,00
P-01	Memompa gliserol dari tangki penyimpanan ke heater	3,00
	Total (HP)	3305,00
	Total (kWh)	2464,54

L3.4.3 Listrik Penunjang

Alat	Fungsi	Daya (HP)
PU-01	Mengalirkan air dari tangki air bersih ke tangki air domestik	0,25
PU-02	Mengalirkan air dari tangki air bersih ke tangki demineralisasi	2,00
PU-03	Mengalirkan air dari tangki demineralisasi ke tangki air umpan boiler	0,50
PU-04	Mengalirkan air dari tangki demineralisasi ke tangki air pendingin	2,00
	Total	4,75

Total (kWh)	3,54

L3.4.4 Total Kebutuhan Listrik untuk Produksi Propilen Glikol 99%

Deskripsi	Kebutuhan	Satuan
Listrik Alat Proses	2464,54	kWh
Listrik Alat Penunjang	3,54	kWh
Listrik Perkantoran	5,83	kWh
Total Kebutuhan	2473,91	kWh
Safety Factor 10%	2721,30	kWh

L3.5 Bahan Bakar

Bahan bakar yang digunakan adalah jenis bahan bakar solar yang digunakan sebagai bahan bakar untuk menjalankan Boiler dan Generator listrik

Tabel L3. 2 Bahan Bakar yang Digunakan pada Produksi

No.	Bahan Bakar	Low Heating Value	Fungsi	Lokasi Unit
1.	Solar	19.200 BTU/lb	Bahan bakar Boiler & Generator	Utilitas

L3.5.1 Kebutuhan Solar sebagai Bahan Bakar Proses pada Boiler

Bahan Bakar	Solar	
Heating Value (Hv)	19200.00	BTU/lb
	44659.20	kJ/kg
Efisiensi Pembakaran	0.85	
Densitas Bahan Bakar	850.00	kg/m^3
	0.85	kg/L

$$m\ solar = \frac{m\ steam\ x\ (Hv\ - Hf)}{E\ x\ Heating\ Value}$$

Suhu (°C)	Tekanan (bar)	Hv (kJ/kg)	Hf (kJ/kg)	m steam (kg/jam)	m solar (kg/jam)	V solar (Liter/jam)
270,00	55,04	1604,38	1185,28	1025,04	11,32	13,31
			Total	1025,04	11,32	13,31
			Total per hari	24600,95	271,61	319,54

L3.5.2 Kebutuhan Bahan Bakar untuk Generator Listrik

Asumsi kebutuhan listrik Generator Selama 1 tahun:

Kebutuhan Listrik	2721,30	kW	9796697,37	kJ/jam
Generator	700,00	kW	2520000,00	kJ/jam

$$m\ solar = \frac{Generator}{E\ x\ Heating\ Value}$$

m solar	66,39	kg/jam
V solar	78,10	Liter/jam

Total kebutuhan bahan bakar

Deskripsi	Kebutuhan (per jam)	Kebutuhan (per hari)	Satuan
Boiler	13,31	319,54	Liter
Generator	78,10	78,10	Liter
Total kebutuhan	91,41	397,64	Liter
Safety factor 10%	100,56	437,40	Liter

LAMPIRAN 4

SPESIFIKASI ALAT

L3.1. Tangki Penyimpanan Produk Propilen Glikol

Fungsi : Menyimpan produk propilen glikol 99%

Bahan : Stainless Steel SA 167 Grade 11, Type 316

Bentuk *Shell* : Silinder vertikal

Bentuk *Head* : *Thorisperical*

Jumlah : 1 unit

Kondisi operasi :

-Tekanan = 1 atm

-Temperatur = 30°C

a. Perhitungan ukuran tangki:

Densitas Campuran : 383,79 kg/m^3

Densitas (ρ)

Liquid	A	B	n	Tc	T	Fraksi	Densitas (g/ml)	Dmix
Gliserol	0,34908	0,24902	0,1541	723	303,15	0,9950	415,53917	413,461474
Air	0,3471	0,274	0,28571	647,13	303,15	0,0050	446,6614605	2,2333073
Asetol	0,32119	0,25855	0,2745	506,8	303,15	0	383,7519093	0
Propilen Gliko	0,31839	0,26106	0,20459	626	303,15	0	383,1556694	0
Hidrogen	0,03125	0,3473	0,2756	33,18	303,15	0	12,85563867	0
								415,694781

a. Volume tangki

Laju alir massa = 6313,13 kg/jam

$$laju\ alir\ massa\left(\frac{kg}{jam}\right) \times waktu\ tinggal \times \frac{24\ jam}{1\ hari}$$

$$6313{,}13\,\frac{\text{kg}}{\text{jam}}\,x\ 1\ x\,\frac{24\ jam}{1\ hari}$$

Massa total bahan = 151.515,12 kg

Volume bahan = $\dfrac{\text{Massa total bahan}}{\rho\ \text{campuran}}$

$\dfrac{151.515,12\ \text{kg}}{383,79\ \text{kg/m}^3}$

= 394,78 m^3

Dengan faktor keamanan 10 %, maka volume tangki total menjadi :

Volume Tangki (VT) = (1 + 0,1) x 394,78 m^3

= 434,26 m^3

b. Ukuran tangki

1. Menentukan diameter dalam tangki

   Volume Tangki = $\frac{\pi}{4} \times Di^2 \times H$

   Diambil H = 3Di, maka

H = $\frac{3}{4} \times \pi \times Di^3$

Di^3 $\dfrac{4 \text{ x VT}}{3 \text{ x } \pi}$ =

Di^3 = $\dfrac{4 \text{ x } 434,26\ m^3}{3 \text{ x } 22/7}$

Di^3 = 184,23 m^3

Di = 5,69 m

= 224,02 in

r = 2,84 m

= 112,01 in

H = 3D

= 3 x 5,69 m

= 17,07 m

= 672,063 in

2. Menentukan tekanan design tangki

Tinggi cairan dalam tangki (hL)

VT = (3.14/4) x D^2 x hL

hL $\frac{17,07\ m^3}{((3,14/4) \times 5,69^2}$ =

hL = 15,52 m

= 610, 97 in

P operasi = 1 atm

= 14.6959 psi

1 atm =	101.325	kN/m2
	101,325	N/m2
1 N/m2 =	0.000009869232667	atm

P hidrostatik = ρ x g x hL

= 383,79 kg/m^3 x 9.8 m/dt^2 x 15.52 m

= 58367,73788 kg/ms^2

= 0,58 atm

= 8,47 psi

Faktor keamanan diambil 10%

P design = (1+0.1) x (P operasi + P hidrostatik)

= (1,1 x (0,58 atm + 1 atm)

= 1,1 x 1,58

= 1,73 atm

= 25,48 psi

3. Menentukan tebal tangki

Untuk clylindrical vessel pada tekanan atmosfer dan dihitung berdasarkan tekanan internal: (Peter, tabel 4, hal 537)

Dimana:

Bahan = Stainless steel SA-167 grade 316 (Brownell & Young, item 4, hal 3)

P	= Tekanan Design	= 25,4799	psi
ID	= Diameter bagian dalam tangki	= 1710.8873	in
f	= max allowable stress	= 18750	psi
E	= welded joint efficiency	= 0.8	
C	= Corrosion Allowance	= 0.0125	in/tahun

(Tabel 13.2 hal 254, Brownell & Young)

Umur Alat 10 tahun, maka:

C = Corrosion Allowance = 0.125 in/10tahun

Sehingga tebal tangki adalah :

$$Ts = \frac{P \times r}{f \times E - 0,6 \times P} + c$$

$$Ts = \frac{25,48\ psi \times 112,01}{18750 \times 0,8 - 0,6 \times 25,48\ psi} + 0,0125$$

Ts = 0,2030 in

Diambil t standar = 1/4 in

= 0,25 in

(Tabel 5.8 hal 93, Brownell & Young)

4. Menentukan Diameter Luar dan Tinggi Tangki Sesungguhnya

Diameter luar = OD = Di + (2 x t standar)

= 224,02 in + (2 x 0.250 in)

= 225 inch

OD Standar = 228 in (Tabel hal 59, Brownell & Young)

b. Perhitungan ukuran head tangki:

Bentuk : Torispherical head (Flange dan dishead head)

Bahan : Stainless Steel SA - 167 grade 11 tipe 316 (Brownell : hal 342)

1. Tebal Head (th)

Karena tebal tangki diambil ts = 0.250 in

ID = OD - (2 x ts)

228 in - (2 x 0.2030 in)

= 227.75 in

> Berdasarkan Brownell & Young, Hal 87

icr (inside corner radius) = 13,75 in

r (Crown radius) = 180 in

icr / r = 7,64 %

Jika icr/r > 6%, maka :

Tebal head, th = w + C (Persamaan 7.77, Brownell & Young)

(2 x f x E) - (0.2 x P design)

Keterangan :

ts = Tebal tangki

P = Tekanan design

f = Tegangan maksimum yang diijinkan (max. Allowable Stress). Bahan yang digunakan adalah stainless steel SA 167 Grade 11 Type 316 = 18750

C = faktor korosi. Diperkirakan umur pabrik alat 10 tahun sehingga

= 0.0125 in/tahun (Tabel 6, Peters hal 574)

= 0.125 in/10 tahun

E = Welded Joint Efficiency untuk single welded butt joint dengan backing stripfull radiography dan dengan stress relieve = 80%

W = faktor stress intensification untuk Torispherical Dished Head

= 1/4 x (3 + (r/icr)1/2) (Persamaan 7.76, Brownell & Young)

= 1/4 x (3 + (180/13.75)1/2)

= 1,6545 in

sehingga:

th = $\frac{25{,}48\ psi \times 180\ in \times 1{,}6545\ in}{(2 \times 18750\ lb/in2 \times 0{,}8) - (0{,}2 \times 25{,}48\ psi)} + 0{,}0125$

= 0,27 in

diambil tebal standar = 0,3125 in

= 5/16 in

Diambil tebal standar, th = 5/16 in, dengan straight flange, sf = 1 1/2 - 3 in, diambil

Sf = 2 in (Tabel hal 88, Brownell & Young)

2. Volume Head

Adapun volume head dihitung dengan cara :

> Bagian lengkung torispherical head (Vh')

Vh' = 0,000049 x D^3 (Pers. 5.11, Brownell & Young, hal 88)

= 0,000049 x $(227{,}75\ in)^3$

= 578,86 in^3

> Bagian straight flange (Vsf)

Volume toripheriscal head bagian straight flange (Vsf) dihitung sebagai bentuk suatu silinder dengan ketinggian

Vsf = $\frac{\pi}{4} \times ID^2 \times sf$

= $\frac{22/7}{4} \times 227{,}75^2 \times 2$

= 81510,10 in^3

> Volume Total Head (Vh)

Vh = Vh' + Vsf

= (578,86 + 81510,10) in^3

= 82088,96 in^3

3. Tinggi Tangki

Untuk menghitung tinggi head, dijelaskan melalui Brownell & Young fig 5.8, hal 87

ID = 227.75 in

a = ID / 2

= 113,88 in

AB = a - icr

= 113,88 in - 13,75 in

= 100,13 in

BC = r - icr

= 180 in – 13,75 in

= 166,25 in

AC = $(BC^2 - AB^2)^{1/2}$

= $((166{,}25\ in)^2 – (100{,}13\ in)^2)^{1/2}$

= 132,72 in

b = r - AC

= 180 in – 132,72 in

= 47,28 in

Jadi tinggi penutup tangki :

OA = th + b + sf

= 0,27 in + 47,28 in + 2 in

= 49,59 in

4. Tinggi *Shell*

Volume *Shell* = $V\ total - V\ head$

= 26.500.347,07 in^3 - 82088,96 in^3

= 26.418.258,12 in^3

Volume *Shell* = $\frac{\pi}{4} \times ID^2 \times H\ shell$

H *Shell* = $\frac{Volume\ Shell}{\frac{\pi}{4}\ x\ ID^2}$

= $\frac{26.418.258{,}12\ \text{in3}}{\frac{22/7}{4}\ x\ 227{,}75^2}$

= 648,22 in

5. Tinggi Tangki (HT)

HT = H + 2OA

= 648,22 in + (2 x 49,59 in)

= 747,41 in

6. Tinggi Cairan

Tinggi cairan = H *shell* + OA

= 648,22 in + 49,59 in

L3.2. Pompa Gliserol

Fungsi : Mengalirkan gliserol 99.5% dari tangki penyimpanan (T-01) ke heater (HE-01)

Tipe	: Pompa sentrifugal	
T	: 30 °C	
Densitas Campuran	: 415,6948	kg/m^3
	: 25.95	lb/ft^3
Viskositas	: 533,56	cp
	: 0,3735	lb/ft.det
Laju Alir Massa	: 8372,5883	kg/jam
	: 18458,3757	lb/jam
Jumlah pompa	: 1 unit	

Densitas (ρ)

Liquid	A	B	n	Tc	T	Fraksi	Densitas (g/ml)	Dmix
Gliserol	0,34908	0,24902	0,1541	723	303,15	0,9950	415,53917	413,4614742
Air	0,3471	0,274	0,28571	647,13	303,15	0,0050	446,6614605	2,233307303
Asetol	0,32119	0,25855	0,2745	506,8	303,15	0	383,7519093	0
Propilen Gliko	0,31839	0,26106	0,20459	626	303,15	0	383,1556694	0
Hidrogen	0,03125	0,3473	0,2756	33,18	303,15	0	12,85563867	0
								415,6947815

Viskositas (μ) $\log\eta = A + \frac{B}{T} + C\,T + D\,T^2$

Liquid	BM	A	B	C	D	Fraksi	Viskositas (centipoise)	Viskositas mix (centipoise)
Gliserol	92,0000	-18,2152	4,23E+03	2,87E-02	-1,86E-05	0,9950	536,2345761	533,5534032
Air	18,0000	-10,2158	1,79E+03	1,77E-02	-1,26E-05	0,0050	0,788285672	0,003941428
Asetol	74,0000	-7,0933	9,31E+02	1,75E-02	-1,90E-05	0	0,344202003	0
Propilen Gliko	76,0000	-29,492	5,25E+03	5,82E-02	-4,23E-05	0	38,20521124	0
Hidrogen	2,0000	-7,0154	4,08E+01	2,37E-01	-4,08E-03	0	0	0
							μ campuran =	533,5573446

a. Perhitungan dimensi pompa

Menghitung diameter optimal pipa. Diasumsikan aliran turbulen

D optimum = 3,9 x Qf ^{0,45} x $\rho^{0.13}$ (Peters, pers 15 hal 496)

Dimana :

Qf (laju alir volumetrik)	=	11,8546	ft^3/menit
	=	0,1976	ft^3/detik
ρ (densitas)	=	25.95	lb/ft3

D optimum = 3,9 x Qf ^{0,45} x $\rho^{0.13}$
= 3,9 x 0,1976 ^{0,45} x 25,95 $^{0.13}$
= 3,9091 in

Dipilih pipa dengan spesifikasi sebagai berikut :

(Sumber: Process Equipment Design, Brownell & Young, Hal. 387)

Bahan	=	Stainless steel
Nominal pipe size	=	3 in
Schedule number	=	40
Diameter luar	=	4.500 in
	=	0.3750 ft
Diameter dalam	=	4.026 in
	=	0.3355 ft
Ketebalan pipa	=	0.2370 in
	=	0.0198 ft

Luas Bagian dalam penampang pipa

A = ¼ x π x ID^2
= ¼ x 22/7 x 0,3355^2
= 0.0884 ft^2

Kecepatan Linier Fluida

V = Qf / A

= 0,1976 ft^3/detik / 0.0884 ft^2

= 2.2340 ft/detik

Penentuan bilangan Reynold

$$Re = \frac{\rho \cdot V \cdot L}{\mu} \text{ or } Re = \frac{V \cdot L}{\nu}$$

Dimana :

ρ	=	25,9510	lb/ft^3
miu	=	0,3735	lb/ft.s
L/ID	=	0,3355	ft
V	=	2,2340 ft/s	
Re	=	52,08	

(Asumsi aliran laminar, Nre < 2100)

Pada Moody Diagram, untuk steel structural or forged (Ɛ = 0.00015)

Dimana :

Ɛ	=	0.00015	
D	=	4.026 in	
Relative roughness	=	Ɛ/D	
	=	0.0004	
Faktor friksi (f)	=	1.23	(Brown, hal 141)

1. Penentuan sistem perpipaan

Diperkirakan pipa yang digunakan mempunyai :

Panjang lurus (L)	=	20	m
	=	65.6168 ft	
ΔZ	=	10	m
	=	32.8084 ft	
Standard Elbow 90	=	3	buah
Gate valve	=	2	buah
Globe valve	=	0	buah

Dari fig.127, brown, hal 141, diperoleh panjang ekuivalen (Le)

ID	=	4.026	ft

Standard Elbow 90	=	8	ft
Gate valve	=	1.60	ft
Globe valve	=	80	ft

Maka total panjang pipa

Lt = L + Le

= 92,8168 ft

2. Penentuan tenaga yang hilang karna friksi:

Energi mekanik yang hilang akibat friksi dapat ditentukan dengan persamaan Fanning. Friksi yang terjadi adalah sebagai berikut:

Dimensional Constant, gc = 32,1740 lbm.ft/lbf.detik2

$$F = \frac{2\ x\ f\ x\ L\ x\ V^2}{g_C\ x\ ID}$$

$$= \frac{2\ x\ 1.23\ x\ 92{,}8168\ x\ 2{,}2340^2}{32{,}1740\ x\ 0{,}3355}$$

= 105,47727 lbf.ft/lbm

3. Menghitung Velocity Head

$$\frac{\Delta V^2}{2gc} = \frac{V^2 - V_1}{2gc}$$

karena diameter tangki jauh lebih besar dari diameter pipa maka V1 = 0 (diabaikan)

$$\frac{\Delta V^2}{2gc} = \frac{V^2}{2gc}$$

$$\text{Velocity Head} = \frac{V^2}{2gc}$$

$$= \frac{2.2340^{\wedge}2}{2\ x\ 32.1740}$$

= 0,0776 lbf.ft/lbm

4. Menghitung Daya Pompa

W = $\left(\Delta Z \frac{g}{gc}\right) + (F) + \left(\frac{\Delta P}{\rho}\right) + \left(\frac{\Delta V^2}{2gc}\right)$

= 32,8084 + 105,47727 + 0,0776

= 138,3632 lbf.ft/lbm

Daya Pompa

P = $w \; x \; \rho \; x \; Qf \; x \left(\frac{1 \; HP}{550 \; \left(\frac{lbf.ft}{detik}\right)}\right)$

= (138,3632 lbf.ft/lbm x 25,9510 lb/ft^3 x 0,1976 ft^3/detik) x 1 / 550

= 1,2899 Hp

Qf = 88,6788 gpm

Efisiensi pompa = 63%

(Sumber: Plant Design and Economics for Chemical Engineers, Peters & Timmerhaus, Hal. 520)

Broke Horse Power = $\frac{\text{P}}{\text{Efisiensi pompa}}$

= $\frac{1{,}2899 \text{ Hp}}{0{,}63}$

= 2,0474 Hp

Efisiensi motor = 82% (Peters, hal 521, Fig 14-38)

daya pompa $\frac{\text{Broke Horse Power}}{\text{Efisiensi motor}}$ sebenarnya =

= $\frac{2.0474 \text{ Hp}}{0{,}82}$

= 2.50 Hp

Yang digunakan = 3.00 Hp

L3.3. Heater (H-03)

Fungsi : Memanaskan feed kolom distilasi dari suhu 80℃ menjadi 100,27℃.

Tipe : Double pipe, SS SA 167, Grade 3, Type 304

Jumlah : 1

Fluida panas

Temperatur awal (T1) = 270°C = 518°F

Temperatur akhir (T2) = 270 ºC = 518 ºF

Fluida dingin

Laju alir fluida dingin = 8553,69 kg/jam = 18857,64 lbm/jam

Temperatur awal (t1) = 80 ºC = 176 ºF

Temperatur akhir (t2) = 100 ºC = 212 ºF

Panas yang diserap (Q) = 314167,09 kJ/jam = 297956,07 Btu/jam

Menetukan True Temperatur Difference (Δttrue)

$$LMTD = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \frac{T_1 - t_2}{T_2 - t_1}}$$

$$= \frac{(518-212)\ ºF - (518-176)\ ºF}{\text{Ln } (518-212)\ ºF / (518-176)}$$

= 323,44 ºF

Menetukan Luas Permukaan Transfer Panas (A')

U = 50 Btu/jam.ft².ºF

$$A' = \frac{Q}{U_{o,asm} \times LMTD}$$

A' = $\frac{297956{,}0674 \text{ btu/jam}}{50 \text{ Btu/jam.ft}^2.^\circ\text{F} \times 323{,}44\ ^\circ\text{F}}$

= 18,42 ft^2

= 1,71 m^2

Tipe HE : **Double Pipe** karena,

tipe HE **shell & tube** jika **A' > 200 ft²**

tipe HE **double pipe** jika **A' ≤ 200 ft²**

Menentukan Diameter Inner Pipe, Annulus, dan Panjang Pipa

Fluida panas : Steam — Anulus (Outer)

Fluida dingin : Campuran propilen glikol dan air — Tube (Inner)

Asumsi awal

	Inner Pipe		Annulus	
	m	ft	m	ft
OD	0,0375	0,12300	0,05	0,16400
BWG	16	16	14	14
ID	0,03425	0,11234	0,04585	0,1504

Effective lengths untuk penukar panas double pipe dengan panjang :
Asumsi L : 12 ft
3,6576 m
(Sumber: Kern, Hal 103)

a. Inner Pipe
Flow area, a inner

$$a_{inner} = \frac{\pi (ID_{inner})^2}{4}$$

A inner = $\frac{3{,}14 \times (0{,}11234)^2}{4}$

= 0,00991 ft^2

$$G_{inner} = \frac{W_{inner}}{a_{inner}}$$

G inner = $\frac{18857{,}63839 \text{ lb/jam}}{0{,}00991 \text{ ft}^2}$

= 1.903.482 lb/h. ft^2

$$Re_{inner} = \frac{ID_{inner} G_p}{\mu}$$

Re inner $\frac{0,11234 \text{ ft x } 1.903.482 \text{ lb/h. ft}^2}{(2,5969 \text{ x } 2,42)}$ =

= 34.025,92

L/D = 12 ft / 0,11234
= 106,82
jh inner = 0,0035
jH inner = Re inner x jh inner
= 34025,92 x 0,0035
= 119,0907

$$h_i = jH_{inner}\frac{k}{ID_{inner}}\left(\frac{Cp\mu}{k}\right)^{1/3}\left(\frac{\mu}{\mu_w}\right)^{0.14}$$ Dengan mengasumsikan $\frac{\mu}{\mu_w} = 1$, maka

hi = 413,98 Btu/h.ft².°F

b. Annulus
Flow area

$$a_a = \frac{\pi\left\{(ID_a)^2 - \left(OD_p\right)^2\right\}}{4};$$

Aa = 0,005877727 ft²

Mass Velocity, Ga

$$G_a = \frac{w_a}{a_a}$$

Ga $\frac{431,65 \text{ lb/jam}}{0,005877727 \text{ ft}^2}$ =

= 73437,56 lb/h. ft²

Equivalent Diameter, De

$$D_e = \frac{ID_a{}^2 - OD_{inner}{}^2}{OD_{inner}}$$

De = 0,06087 ft

Bilangan Reynold, Rea

$$Re_a = \frac{D_e G_a}{\mu}$$

Rea $= \dfrac{0{,}06087 \text{ ft} \times 73437{,}56 \text{ lb/h. ft}^2}{0{,}0191 \text{ cp} \times 2{,}42}$ =

= 96.512,06

L/D = 12 ft / 0,1504 ft

= 79,79

jh,a = 0,003

jH,a = Rea x jh,a

= 96.512,06 x 0,003

= 289,54

Jadi,

$$h_o = jH_a \frac{k}{D_e}\left(\frac{Cp\mu}{k}\right)^{1/3}\left(\frac{\mu}{\mu_w}\right)^{0.14}$$ Dengan mengasumsikan $\frac{\mu}{\mu_w} = 1$, maka

ho = 69,73 Btu/h. ft².°F

c. Koefisien Perpindahan Panas Dalam-Luar (h_{io})

$$h_{io} = h_i \frac{ID_{inner}}{OD_{inner}}$$

h_{io} = 413,98 Btu/h. ft².°F x $\dfrac{0{,}11234 \text{ ft}}{0{,}12300 \text{ ft}}$

= 378,10 Btu/h.ft².°F

d. Koefisien Perpindahan Panas Menyeluruh Bersih (U_c)

$$U_c = \frac{h_{io} h_o}{h_i + h_o}$$

U_c $= \dfrac{378{,}10 \text{ Btu/h. ft}^2.°\text{F} \times 69{,}73 \text{ Btu/h. ft}^2.°\text{F}}{413{,}98 \text{ Btu/h.ft}^2.°\text{F} + 69{,}73 \text{ Btu/h. ft}^2.°\text{F}}$ =

= 54,51 Btu/h.ft².°F

e. Koefisien Perpindahan Panas Menyeluruh Kotor (U_D)

$$\frac{1}{U_D} = \frac{1}{U_C} + R_{D,total}$$

$1/U_D$ = (1 / 54,51 Btu/h.ft².°F) + (0,0010+0,0005) ft².jam.°F /Btu

= 0,01985 h. ft².°F/Btu

U_D = 1 / 0,01985 h. ft².°F/Btu

= 50,39 Btu/h. ft².°F

Kesimpulan : Memenuhi syarat. Karena $U_D < U_c$

f. Luas area perpindahan panas

$$A = \frac{Q}{U_D \times LMTD}$$

A = $\frac{297956{,}0674 \text{ btu/jam}}{50{,}39 \text{ Btu/h. ft}^2\text{.°F} \times 323{,}44 \text{ °F}}$

= 18,28 ft²

g. Panjang yang dibutuhkan

Ext surface area = 0,2618 ft²/ft

Total panjang yang dibutuhkan

$$= \frac{A}{ext\ surface\ area}$$

Panjang = $\frac{18{,}28 \text{ ft}^2}{0{,}2618 \text{ ft}^2\text{/ft}}$

= 69,82 ft

= 21,30 m

$$Hairpin = \frac{Total\ panjang}{2L}$$

Hairpin = $\frac{69{,}82 \text{ ft}}{2 \times 12 \text{ ft}}$

= 2,91 set

= 3 set

h. Revisi U_D dan R_D

A aktual

$$A_{aktual} = jumlah\ hairpin \times 2 \times L \times ext.surface\ area$$

= 3 set x 2 x 12 ft x 0,2618 ft^2/ft

= 18,85 ft^2

= 1,75 m^2

$$U_D = \frac{Q}{A \times LMTD}$$

$$U_D = \frac{297956{,}0674\ btu/jam}{18{,}85\ ft \times 323{,}44\ ^oF}$$

= 48,87 $Btu/h.ft^2.^oF$

R_D min = Total fouling

= 0,0010 + 0,0005

= 0,0015 $Btu/h.ft^2.^oF$

$$R_D = \frac{U_C - U_D}{U_C \times U_D}$$

$$R_D = \frac{54{,}51\ Btu/h.ft^2.^oF - 48{,}87\ Btu/h.ft^2.^oF}{54{,}51\ Btu/h.ft^2.^oF \times 48{,}87\ Btu/h.ft^2.^oF}$$

= 0,0021 $h.ft^2.^oF/btu$

Kesimpulan : Memenuhi syarat. Karena R_D > R_D min

i. Pressure Drop Annulus

$$D'_e = ID_a - OD_{inner}$$

= 0,1505 ft – 0,12300 ft

D'e = 0,02739 ft

$$Re'_a = \frac{D_e' G_a}{\mu}$$

$$\frac{0{,}02739\ ft \times 73437{,}56\ lb/h.ft^2}{(0{,}0191\ cp \times 2{,}42)}$$

Re'a =

= 43421,74

$$f = 0.0035 + \frac{0.264}{Re^{0.42}}$$

f = $\frac{0,264}{43421,74^{0,42}}$ 0,0035 +

= 0,006476918

$$\Delta F_a = \frac{4 f G_a^2 L}{2 g \rho^2 D_e'}$$

ΔFa $\frac{4 \times 0,006476918 \times 73437,56^2 \times 12}{2 \times 32,174 \times 46,3790^2 \times 0,02739}$ =

= 0,1023 ft

$$F_t = 3\left(\frac{\left(G_a/\rho\right)^2}{2g}\right)$$

Ft = $\frac{(73437,56/46,3790)^2}{2 \times 32,174}$ 3

= 0,009019343

$$\Delta P_a = \rho(F_t + \Delta F_a) \times \frac{g}{g_c}$$

ΔPa = 46,3790 (0,009019343+0,1023 $\frac{32,174}{32,174 \times 144}$ ft) x

= 0,0358798

Kesimpulan : Memenuhi syarat. Karena ΔPa < 10 psi

j. Pressure Drop Inner Pipe

$$\Delta F_{inner} = \frac{4 f G_{inner}^2 L}{2 g \rho^2 ID_{inner}}$$

ΔF_{inner} = $\frac{4 \times 0{,}006476918 \times 1.903.482 \text{ lb/h. ft}^2 \times 12}{2 \times 32{,}174 \times 46{,}3790^2 \times 0{,}11234}$ =

10,3117 ft

$$\Delta P_{inner} = \rho \Delta F_{inner} \times \frac{g}{g_c}$$

ΔP_{inner} = 46,3790 x $\frac{32{,}174}{32{,}174 \times 144}$ 10,3117 x

= 4,3388

Kesimpulan : Memenuhi syarat. Karena ΔP_{inner} < 10 psi

L3.4. H_2 Separator

Fungsi : Memisahkan gas H_2 dari campuran cairan propilen glikol dan air

Tipe : SA-167 Grade 11 tipe 316

Kondisi operasi

P	=	1 atm	= 14,69 psi
T	=	80 °C	= 353,15 K
Laju Alir gas, F gas	=	656,16	kg/jam
Laju alir cairan	=	8.553,69	kg/jam
n gas	=	328,08	kmol/jam
n Cairan	=	183,43	kmol/jam

1. **Menghitung densitas**

$$\rho_{gas} = \frac{BMav \times P}{R \times T}$$

2 x 14,69 psi

0,082 x 353,15 K

Densitas gas =

= $1{,}0146\ kg/m^3$

= $0{,}0633\ lbm/ft^3$

ρ cairan = $353{,}38\ kg/m^3$

= $22{,}0606\ lbm/ft^3$

2. Menghitung volume

$$V_{gas} = \frac{BMav\ N}{\rho}$$

$$V_{gas} = \frac{2 \times 328{,}08\ kmol/jam}{1{,}0146\ kg/m^3} =$$

= $646{,}74\ m^3/jam$

$$V_{cairan} = \frac{F}{\rho}$$

$$V_{cairan} = \frac{8.553{,}69\ kg/jam}{353{,}38\ kg/m^3} =$$

= $24{,}21\ m^3/jam$

= $0{,}2374\ ft^3/detik$

3. Menghitung kecepatan linear yang diizinkan

$$\mu = 0{,}14 \sqrt{\frac{\rho}{\rho\ gas} - 1}$$ (Wallas, 1988)

μ = 2,6091

4. Menghitung diameter tangki

$$D = \sqrt{\frac{V\ gas}{\left(\frac{\pi}{4}\right)\mu} - 1}$$ (Wallas, 1988)

D = 2,66 ft
= 0,81 m

Hasil Goal seek

D = 4 ft
= 1,22 m

Tinggi kolom uap minimum = 5,5 ft

Waktu tinggal = 60 detik

5. Menghitung tinggi cairan

$$L_{cairan} = \frac{Vt}{\left(\frac{\pi}{4}\right)D^2}$$

$$L\ cairan = \frac{0{,}237 \times 60\ \text{detik}}{(3{,}14/4) \times 4^2}$$

= 1,13 ft
= 0,3457 m

6. Menghitung panjang kolom

L = L cairan + L uap
= (1,13 + 5,5) ft
= 6,634 ft
= 2,02 m

L/D= 6,63/4
= 1,66

7. Menghitung tebal *Shell* tangki

P hidrostatik = p x g x l

= 353,38 kg/m^3 x 9,8 x 0,89 m

= 2705,09 kPa

Po = Tekanan operasi

= 101,325 kPa

Faktor kelonggaran 20%

P design = 1,2 x (Phidrostatik + Po)

= 1,2 x (2705,09 kPa + 101,325 kPa)

= 3367,70 kPa

Joint efficiency € = 0,8

Allowable stress = 18750 psi = 129276,75 kPa

Faktor korosi = 0,125 in

Double welded butt joint (Brownel & Young, 1959)

(Brownel & Young, 1959)

(Brownel & Young, 1959)

$$t = \frac{PD}{2SE-1,2\,P}$$

Tebal shell = $\frac{3367,70 \text{ kPa x } 1,2192 \text{ m}}{2 \text{ x } 18750 \text{ x } 0,8 - (1,2 \text{ x } 3367,70 \text{ kPa})}$

= 0,040 m

= 1,56 in

Maka, Tebal shell tangki + faktor korosi

= 1,688 in

8. Menghitung tutup tangki

Diameter tutup = Diameter tangki = 0,8111 m

Ratio axis = Lh : D

= 1 : 4

$L_h = \frac{Lh}{D}D$

$= \frac{1}{4}$ x 0,8111 M

= 0,2028 m

L (panjang tangki) = Ls + 2Lh

= 2,4575 m + 2(0,2028)

= 2,8631 m

L3.5. Reboiler Kolom Destilasi 1

Fungsi : Memberikan panas pada kolom distilasi 1

Tipe : Double Pipe

Fluida panas

Temperatur awal (T1) = 270°C = 518°F

Temperatur akhir (T2) = 270 ºC = 518 ºF

Fluida dingin

Laju alir fluida dingin = 300,14 kg/jam = 661,70 lbm/jam

Temperatur awal (t1) = 100,55 ºC = 213 ºF

Temperatur akhir (t2) = 100,55 ºC = 213 ºF

Panas yang diserap (Q) = 298338,15 kJ/jam = 282943,90 Btu/jam

Kebutuhan steam (Wa) = 86,58 kg/jam = 190,88 lb/jam

Menetukan True Temperatur Difference (Δttrue)

$$\text{LMTD} = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \frac{T_1 - t_2}{T_2 - t_1}}$$

$$= \frac{(518\text{-}213)\ ºF - (518\text{-}213)\ ºF}{\text{Ln}\ (518\text{-}213)\ ºF\ /\ (518\text{-}213)}$$

= 305,02 ºF

Menetukan Luas Permukaan Transfer Panas (A')

U = 100 Btu/jam.ft².°F

$$A' = \frac{Q}{U_{o.asm} \times LMTD}$$

$$A' = \frac{282943{,}90 \text{ btu/jam}}{100 \text{ Btu/jam.ft}^2.°F \times 305{,}02\ °F} =$$

= 9,28 ft²
= 0,86 m²

Tipe HE : **Double Pipe** karena,

tipe HE **shell & tube** jika **A' > 200 ft²**
tipe HE **double pipe** jika **A' ≤ 200 ft²**

Menentukan Diameter Inner Pipe, Annulus, dan Panjang Pipa

Fluida panas : Steam — Anulus (Outer)

Fluida dingin : Bottom Product Kolom Destilasi 1 — Tube (Inner)

Asumsi awal

	Inner Pipe		Annulus	
	m	ft	m	ft
OD	0,01875	0,061500	0,25	0,82
BWG	12	12	14	14
ID	0,01355	0,044444	0,02085	0,068388

Effective lengths untuk penukar panas double pipe dengan panjang :
Asumsi L : 12 ft
3,6576 m
(Sumber: Kern, Hal 103)

k. Inner Pipe
Flow area, a inner

$$a_{inner} = \frac{\pi (ID_{inner})^2}{4}$$

$$A\ inner = \frac{3{,}14 \times (0{,}0444)^2}{4} =$$

= 0,001551 ft²

$$G_{inner} = \frac{W_{inner}}{a_{inner}}$$

$$G\ inner = \frac{661{,}70 \text{ lb/jam}}{0{,}001551 \text{ ft}^2}$$

= 426.744 lb/h. ft²

$$Re_{inner} = \frac{ID_{inner} G_p}{\mu}$$;

Re inner $= \frac{0{,}0444 \text{ ft} \times 426.744 \text{ lb/h. ft}^2}{(0{,}670 \times 2{,}42)}$ =

= 11.702

L/D = 12 ft / 0,0444
= 270,00
jh inner = 0,0032
jH inner = Re inner x jh inner
= 11.702 x 0,0032
= 35,44715

$$h_i = jH_{inner} \frac{k}{ID_{inner}} \left(\frac{Cp\mu}{k}\right)^{1/3} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$ Dengan mengasumsikan $\frac{\mu}{\mu_w} = 1$, maka

hi = 143,27 Btu/h.ft².°F

l. Annulus
Flow area

$$a_a = \frac{\pi\left\{(ID_a)^2 - (OD_p)^2\right\}}{4}$$;

Aa = 0,000702315 ft²

Mass Velocity, Ga

$$G_a = \frac{w_a}{a_a}$$

Ga $= \frac{190{,}88 \text{ lb/jam}}{0{,}000702315 \text{ ft}^2}$ =

= 271,785 lb/h. ft²

Equivalent Diameter, De

$$D_e = \frac{ID_a^{\,2} - OD_{inner}^{\,2}}{OD_{inner}}$$

De = 0,01455 ft

Bilangan Reynold, Rea

$$Re_a = \frac{D_e G_a}{\mu}$$

Rea = $\dfrac{0{,}01455 \text{ ft x } 271{,}785 \text{ lb/h. ft}^2}{0{,}0191 \text{ cp x } 2{,}42}$ =

= 85.357

L/D = 12 ft / 0,0684 ft

= 175,47

jh,a = 0,003

jH,a = Rea x jh,a

= 85.357 x 0,003

= 256,07

Jadi,

$$h_o = jH_a \frac{k}{D_e}\left(\frac{Cp\mu}{k}\right)^{1/3}\left(\frac{\mu}{\mu_w}\right)^{0.14}$$ Dengan mengasumsikan $\frac{\mu}{\mu_w} = 1$, maka

ho = 258,08 Btu/h. ft².°F

m. Koefisien Perpindahan Panas Dalam-Luar (h_{io})

$$h_{io} = h_i \frac{ID_{inner}}{OD_{inner}}$$

h_{io} = 143,27 Btu/h.ft².°F x $\dfrac{0{,}0444 \text{ ft}}{0{,}0615 \text{ ft}}$

= 103,54 Btu/h.ft².°F

n. Koefisien Perpindahan Panas Menyeluruh Bersih (U_c)

$$U_C = \frac{h_{io} h_o}{h_i + h_o}$$

U_c = $\frac{103{,}54 \text{ Btu/h. ft}^2.^\circ\text{F} \times 258{,}08 \text{ Btu/h. ft}^2.^\circ\text{F}}{143{,}27 \text{ Btu/h.ft}^2.^\circ\text{F} + 258{,}08 \text{ Btu/h. ft}^2.^\circ\text{F}}$

= 66,58 Btu/h.ft².°F

o. Koefisien Perpindahan Panas Menyeluruh Kotor (U_D)

$$\frac{1}{U_D} = \frac{1}{U_C} + R_{D,total}$$

$1/U_D$ = (1 / 66,58 Btu/h.ft².°F) + (0,0010+0,0005) ft².jam.°F /Btu

= 0,01652 h. ft².°F/Btu

U_D = 1 / 0,01652 h. ft².°F/Btu

= 60,53 Btu/h. ft².°F

Kesimpulan : Memenuhi syarat. Karena $U_D < U_c$

p. Luas area perpindahan panas

$$A = \frac{Q}{U_D \times LMTD}$$

A = $\frac{282943{,}90 \text{ btu/jam}}{60{,}53 \text{ Btu/h. ft}^2.^\circ\text{F} \times 305{,}02\ ^\circ\text{F}}$

= 15,32 ft²

q. Panjang yang dibutuhkan

Ext surface area = 0,13 ft²/ft

Total panjang yang dibutuhkan

$$= \frac{A}{ext\ surface\ area}$$

= $\frac{15{,}32 \text{ ft}^2}{0{,}13 \text{ ft}^2/\text{ft}}$

Panjang = 117,06 ft

= 35,70 m

$$Hairpin = \frac{Total\ panjang}{2L}$$

Hairpin $= \dfrac{117{,}06 \text{ ft}}{2 \times 12 \text{ ft}}$ =

= 4,88 set

= 5 set

r. Revisi U_D dan R_D

A aktual

$$A_{aktual} = jumlah\ hairpin \times 2 \times L \times ext.surface\ area$$

= 5 set x 2 x 12 ft x 0,13 ft²/ft

= 15,71 ft^2

= 1,46 m^2

$$U_D = \frac{Q}{A \times LMTD}$$

U_D $= \dfrac{282943{,}90 \text{ btu/jam}}{15{,}71 \text{ ft x } 305{,}02 \text{ °F}}$ =

= 59,05 Btu/h.ft².°F

R_D min = Total fouling

= 0,0010 + 0,0005

= 0,0015 Btu/h.ft².°F

$$R_D = \frac{U_C - U_D}{U_C \times U_D}$$

R_D $= \dfrac{66{,}58 \text{ Btu/h.ft}^2\text{.°F} - 59{,}05 \text{ Btu/h.ft}^2\text{.°F}}{66{,}58 \text{ Btu/h.ft}^2\text{.°F x } 59{,}05 \text{ Btu/h.ft}^2\text{.°F}}$ =

= 0,00192 h.ft².°F/btu

Kesimpulan : Memenuhi syarat. Karena R_D > R_D min

s. Pressure Drop Annulus

$$D'_e = ID_a - OD_{inner}$$

= 0,068388 ft – 0,06150 ft

D'e = 0,006888 ft

$$Re'_a = \frac{D_e' G_a}{\mu}$$

Re'a = $\frac{0,006888 \text{ ft x } 271.785 \text{ lb/h.ft}^2}{(0,0191 \text{ cp x } 2,42)}$

= 40415,47

$$f = 0.0035 + \frac{0.264}{Re^{0.42}}$$

f = 0,0035 + $\frac{0,264}{40415,47^{0,42}}$

= 0,00656799

$$\Delta F_a = \frac{4 f G_a^2 L}{2 g \rho^2 D_e'}$$

ΔFa = $\frac{4 \text{ x } 0,00656799 \text{ x } 271.785^2 \text{ x } 12}{2 \text{ x } 32,174 \text{ x } 46,3790^2 \text{ x } 0,006888}$

= 9,4236 ft

$$F_t = 3\left(\frac{\left(G_a/\rho\right)^2}{2g}\right)$$

Ft = 3 $\frac{(271.785/46,3790)^2}{2 \text{ x } 32,174}$

= 0,123534997

$$\Delta P_a = \rho(F_t + \Delta F_a) \times \frac{g}{g_c}$$

ΔPa = 46,3790 (0,123534997+9,4236 ft) x $\frac{32,174}{32,174 \text{ x } 144}$

= 3,0749

Kesimpulan : Memenuhi syarat. Karena ΔPa < 10 psi

t. Pressure Drop Inner Pipe

$$\Delta F_{inner} = \frac{4fG_{inner}^2 L}{2g\rho^2 ID_{inner}}$$

$$\Delta F_{inner} = \frac{4 \times 0{,}00656799 \times 426.744 \text{ lb/h. ft}^2 \times 12}{2 \times 32{,}174 \times 46{,}3790^2 \times 0{,}0444} =$$

= 2,8243 ft

$$\Delta P_{inner} = \rho \Delta F_{inner} \times \frac{g}{g_c}$$

$$\Delta P_{inner} = 46{,}3790 \times 2{,}8243 \times \frac{32{,}174}{32{,}174 \times 144} \times$$

= 1,1796

Kesimpulan : Memenuhi syarat. Karena ΔP_{inner} < 10 psi

L3.6. Kondenser Kolom Destilasi 1

Fungsi : Mendinginkan Fluida yang keluar dari top destilasi

Tipe : Tube and Shell

Fluida dingin (Cooling water)		
t1 =	30,00	C
t2 =	40,00	C
Fasa =	Cair	
Temperatur (C) =	35,00	C
Tekanan =	1,00	Atm
=	101,33	kPa
Laju alir massa =	279,58	kg/h
Densitas =	444,49	kg/m3
=	27,75	lb/ft3
Kapasitas Panas =	1.339,46	kj/kg.K
=	319,92	BTU/lb.F
Viskositas =	0,70935	cP
Konduktivitas thermal =	3,00	W/m.K
=	1,74	Btu/hr ft F
Fouling =	0,001	ft2.jam F/Btu

Fluida panas (Air dan PG)		
T1 =	100,01	C
T2 =	100,02	C
Fasa =	Liquid	
Temperatur (C) =	100,02	C
Tekanan =	1,00	Atm
=	101,33	kPa

Laju alir massa =	1.880,19	kg/h
Densitas =	411,06	kg/m3
=	25,66	lb/ft3
Kapasitas Panas =	57,42	J/mol.K
=	13,72	BTU/lb.°F
Viskositas =	0,60	cP
Konduktivitas thermal =	2,59	W/m.K
=	1,50	Btu/hr ft F
Fouling =	0,001	ft2.jam F/Btu

Laju alir massa fluida panas (Wt)	=	1.880,19	kg/jam
	=	4.145,11	lb/jam
Kebutuhan panas yg diberikan (Q)	=	3.744.820,38	kj/jam
	=	3.549.592,77	btu/jam
Laju alir massa fluida dingin (Ws)	=	279,58	kg/jam
	=	616,36	lb/jam

1. **Menentukan True Temperature Difference (DeltaT)**

$$LMTD = \frac{(T1-t2)-(T2-t1)}{(\ln(T1-t2)/(T2-t1))}$$

$$= \frac{(100-40) - (100-30)}{(\ln(100-40)/(100-30))}$$

= 64,89 °C

= 148,80 °F

Dicari Ft = 1 (Figure 18 kern)

LMTD terkoreksi

= 64,89 °C

= 148,80 °F

2. Menentukan Luas Permukaan Transfer Panas (A)

Dari table 8 kern diambil Ud = 100 btu/Batch.ft^2.°F

A	=	Q/UD x Delta LMTD
	=	3.744.820,38 kj/jam / 148,80 °F x 100
	=	238,56 ft^2

3. Menentukan Tc average dan tc average

tc average	=	(t1+t2)/2
	=	(30+40) / 2
	=	35 °C
	=	95 °F
TC average	=	(T1+T2)/2
	=	(100,01+100,02) / 2
	=	100,02 °C
	=	212,03 °F

4. Menentukan dimensi *Shell and Tube*

A. Menentukan jumlah tube (Nt)
(Kern, tabel 10 hal 728)

ODt	=	0,75	in
	=	0,02	m
ao	=	0,20	ft2/ft
Panjang tube	=	35,00	ft
Nt	=	A/(ao x P tube)	
	=	34,77	tube

Tube : (Kern, Tabel 9, hal 841)

ODt	=	0,75	in
Nt standard	=	52,00	tube
Pt	=	1,00	in
n	=	1,00	passes
Susunan tube	=	Square Pitch, 1 in	

Shell : (Kern, Tabel 9, hal 841)

IDs	=	10,00	in	
Baffle (B)	=	10,00	in	B = 1 IDs
Passes	=	2,00	passes	
L tube/ID Shell	=	42,00	ft	

Koreksi harga A dan Ud

Nt	=	52,00 tube
A	=	Nt x ao x L
	=	356,72 ft^2
Ud	=	Q/A x Delta LMTD
	=	66,87 Btu/jam.ft^2.°F

5. **Menentukan faktor pengotor**

Shell side (cooling water)

Flow area, as = (IDs x c" x B)/(Pt x 144)

Dimana :

Diameter dalam shell (IDs)	=	10,00	in
Clearense antar tube (C")	=	0,25	in
Tube pitch (Pt)	=	1,00	in
Jarak antar baffle (B)	=	10,00	in

as = (IDs x c" x B)/(Pt x 144)
= (10 x 0,25 x 10) / (1 x 144)
= 0,17 ft^2

Laju alir massa, gs

Gs = Ws/as

Dimana

Laju alir massa feed (Ws)	=	616,36	lb/jam
Flow area shell (as)	=	0,17	ft^2

Gs = Ws/as
= 616,36 lb/jam / 0,17 ft^2
= 3.550,23 lb/jam.ft^2

Bilangan Reynold

Res = De x Gs/µ

Dimana :

Dari figure 28 kern, diperoleh:

Diameter equivalent shell (De) = 0,95 in
= 0,08 ft

tc	=	35,00	C
µ	=	0,7094	cp
	=	1,7166	lb/ft.jam

Res = De x Gs/µ
= 0,95 x 3.550,23 lb/jam.ft^2 / 1,7166 lb/ft.jam

= 163,73 lb/ft.jam

Koefisien Transfer Panas

Untuk nilai Re	=	163,73 lb/ft.jam
Jh	=	800,00
Pada Tc	=	95,00 °F
C	=	319,92 btu/lb.°F
k	=	1,74 btu/ft.hr.°F
ho/φs	=	$((Jh \times k) / De) \times (((C \times \mu) / k)^{1/3}))$
	=	$((800 \times 1,74) / 0,95) \times (((319,92 \times 1,7166) / 1,74)^{1/3}))$
	=	119.494,28 btu/jam.ft^2.°F

Tube side (propilen glikol dan air)

Flow area, at = (Nt x a't)/n

Dimana :

a't	=	0,182	in^2
n	=	1	passes
Jumlah tube (Nt)	=	52,00	tube

at = (Nt x a't)/n
= (52 x 0,182) / 1
= 0,06572 ft^2

Laju alir massa, gs

Gt = Wt/at

Dimana

Laju alir massa feed (Wt)	=	4.145,11	lb/jam
Flow area shell (at)	=	0,06572	ft^2

Gt = Wt/at
= 4.145,11 lb/jam / 0,06572 ft^2
= 363.070 lb/jam.ft^2

Bilangan Reynold

Res = IDt x Gt/μ

Dimana :

IDt	=	0,75 in
	=	0,0625 ft
tc	=	100,01 C
μ	=	0,6023 cp
	=	1,45691 lb/ft.jam

Ret = IDt x Gt/μ
= 0,75 x 363.070 lb/jam.ft^2 / 1,45691 lb/ft.jam
= 2,706

L/D	=	355 / 0,0625
	=	560

Koefisien Transfer Panas

Untuk nilai Re	=	2,706
Jh	=	800,00
Tc	=	212,03 °F
C	=	13,72 btu/lb.°F
k	=	1,74 btu/ft.hr.°F
ho/φs	=	((Jh x k) / De) x (((C x μ) / k)$^{1/3}$))
	=	((800 x 1,74) / 0,95) x (((13,72 x 1,45691) / 1,74) $^{1/3}$))
	=	50155,3656 btu/jam.ft^2.°F

Tw adalah temperatur dinding tube bagian luar (F)

Tw SS	=	tc + ((ho ts/(ho ts+ho ss)) x (Tc ss-tc ts))
	=	35 + ((50155,3656/(50155,3656+119.494,28)) x 212,03-95))
	=	54,22 °F
	=	12,35 °C

μw	=	0,1789 Cp
	=	0,433 lb/ft.jam
μ/μw	=	1,7166 / 0,433
	=	3,97
φs	=	$3{,}97^{0,14}$
	=	1,21
ho/φs terkoreksi	=	119.494,28 x 1,21
	=	144.920,76 btu/jam.ft^2.°F

Tube side

Tw	=	54,22°F
	=	12,35 °C

μw = 18,0377 Cp

= 43,634 lb/ft.jam

μ/μw = 1,45691 / 43,634

= 0,0334

φt = $0,0334^{0,14}$

= 0,6213

ho/φt terkoreksi = 50155,3656 x 0,6213

= 31.162 btu/jam.ft^2.°F

6. Menentukan Uc

Uc = $\frac{\text{Hio x ho}}{\text{Hio + ho}}$

= $\frac{31.162 \text{ x } 144.920,76}{31.162 + 144.920,76}$

= 25.646,99 btu/jam.ft2.°F

7. Menentukan nilai fouling (Rd)

Rd $\frac{\text{Uc – Ud}}{\text{Uc x Ud}}$ =

= $\frac{25.646,99 – 100}{25.646,99 \text{ x } 100}$

= 0,01

Rd total dari tabel (Rd min) = 0,001 + 0,001

= 0,002

8. Menentukan *Pressure drop*

a. Shell Side

D =	0,08	Ft
Gs =	3.550,23	lb/jam ft
μ =	1,72	lb/jam ft
Res = D x Gs/μ =	163,73	lb/jam ft
f =	0,00	(lihat table 29)
ρ =	444,49	kg/m3
s =	7,12	
Ds =	0,83	Ft
N +1 =	42,00	
ΔPs =	0,00	Psi

b. Tube Side

D =	0,0625	Ft
Gs =	63.070	lb/jam.ft
μ =	1,45691	lb/jam.ft
Res = IDt x Gt/μ =	2.706	
f =	0,00046	
s =	0,55	
ΔPt =	0,000574454	Psi
v2/2g =	0,02	Fig 27
ΔPr =	0,145	Psi
ΔPT =	0,146	Psi

L3.7. Tangki Akumulator Kolom Distilasi 1

Fungsi : Menampung kondensat yang dihasilkan kondensor kolom distilasi 1

Tipe : Silinder horizontal dengan torispherical head

Bahan : SS SA 167, Grade 3, Type 304

Kondisi operasi

P	=	760 mmHg	= 1	atm
T	=	100,55 °C		
Kapasitas	=	6313,13 kg/jam		
Densitas Campuran	=	969,90 kg/m^3		

1. Menghitung Volume Cairan dan Volume Tangki Akumulator

$$\text{Volume Cairan} = \frac{\text{Kapasitas Penyimpanan}}{\rho \text{ cairan}}$$

$$= \frac{6313,13}{969,90}$$

$= 6,51\ m^3$

Digunakan overdesign sebesar 20%

V cairan = $1,2 \times 6,51\ m^3$

= $7,81\ m^3$

2. Menentukan Diameter Dalam Tangki (ID) dan Tinggi Tangki (H)

L/ID = 5 (Treyball, hal 397)

L = 5 ID

Volume tangki silinder horisontal = luas alas x panjang

$$= \left(\frac{\pi\, ID^2}{4}\right) x\ L = \frac{\pi\, ID^2}{4}\ x\ 5\ ID$$

ID = 1,26 m

= 49,52 in

r = 0,63 m

= 24,763 in

L = 6,29 m

= 247,60 in

3. Menentukan Tebal Tangki Akumulator

$$\text{Volume Cairan} = \frac{1}{4} \cdot \pi \cdot D^2 \cdot L$$

Karena tangki berbentuk silinder horizontal, maka Diameter = H_{cairan}

$$H_{cairan} = \sqrt{\frac{\text{Volume Cairan}}{\frac{1}{4} \times \pi \times L}}$$

Hcairan = 1,15 m

= 45,20 in

$P_{desain} = P_{operasi} + P_{hidrostatik}$

P desain

P operasi = 1 atm

P hidrostatik = 10991,90 N/m2

= 0,11 atm

P desain = 1,11 atm

Safety factor 20%, maka:

P desain = 1,2 x 1,11atm

= 1,33 atm

= 19,55 psi

Tebal tangki akumulator

$$ts = \frac{P_{desain} \times ri}{(f \times E) - (0{,}6 \times P_{desain})} + C$$

Interpolasi	
T	f
100	18750
100,55	18740,38
200	17000

r = 24,76 in

f = 18740,375 psi

C = 0,125 in

E = 80%

Ts = 0,1573 in

Tebal standar yang dipilih adalah **0,1875 in = 0,0047625 m** (Brownell and Young, hal 93)

4. Menentukan Tebal Tangki Akumulator

$OD = ID + (2 \times ts)$ (Brownell and Young, hal 89)

OD = 49,90 in

OD standar = 12 in

= 0,3048 m

$ID = OD - (2 \times ts)$

ID sesungguhnya = 11,625 in

= 0,295 m

5. Menentukan Ukuran Head Tangki Akumulator

Tebal Head

OD = 12 in

icr = 0,75 in

r = 12 in

icr/r = 6,25% > 6% maka memenuhi syarat untuk torispherical head

(Brownell and Young, hal 88)

$$W = \frac{1}{4}\left(3 + \sqrt{\frac{rc}{ri}}\right)$$

W = 1,75 in

$$th = \frac{P_{desain} \times ri \times W}{(2 \times f \times E) - (0,2 \times P_{desain})} + C$$

Th = 0,1387 in

th standar = 0,1875 in

sf = 1,75 (1,5 - 2)

a = ID/2

= 11,625 in / 2

= 5,8125 in

AB = (ID/2) - icr

= (11,625 in / 2) – 0,75

= 5,0625 in

BC = r – icr

= 12 – 0,75

= 11,25 in

AC = 10,05 in

b = r – AC

= 12 – 10,05

= 1,95 in

OA = th standar + sf + b

= 0,1875 in + 1,75 in + 1,95 in

= 3,89 in

= 0,0988 m

6. Menentukan Tinggi Tangki Akumulator

$H_{tangki} = ID_s + (2 \times t_s)$

H tangki = 0,295 + (2 x 0,0047625)

= 0,3048 m

= 12 in

7. Menentukan Panjang Tangki Akumulator

L total = L + (2 x OA)

= 6,289 + (2 x 0,0988 m)

= 6,4868 m

L3.8. Tangki Utilitas

Fungsi : Menyimpan air bersih dari

Bentuk : Shell Silinder Vertikal

Bentuk : Head Torispherical

Bahan : Stainless Steel SA 167 Grade 3, Type 304

Jumlah : 1 buah

Kondisi Penyimpanan

Suhu	=	30	°C
Tekanan	=	1	atm
Lama Penyimpanan	=	1	jam
Densitas Bahan	=	1022,87 kg/m^3	

1. Menghitung Volume Tangki

Laju Alir Massa = 51685,1124 kg/jam

Laju Alir Volume = 50,5292 m^3/jam

Massa Bahan = $laju\ alir\ massa \left(\frac{kg}{jam}\right) \times waktu\ tinggal$

= 51685,1124 kg/jam x 1

= 51685,1124 kg

Volume Bahan = $\frac{massa\ total\ bahan\ (kg)}{densitas\ campuran\ \left(\frac{kg}{m^3}\right)}$

= $\frac{51685{,}1124\ \text{kg}}{1022{,}87\ \text{kg/m3}}$

= 50,5292 m^3

Untuk faktor keamanan diambil over design 10%

= 1,1 x 50,5292 m^3

= 55,5822 m^3

= 3391831,92 $inch^3$

2. Menghitung Dimensi Tangki

Volume Tangki = $\frac{\pi}{4} \times Di^2 \times H$

H = 3 Di

Volume Tangki = $\frac{3}{4} \times \pi \times Di^3$

Di^3 = 23,58 m^3

Di = 2,87 m

= 112,90 inch

r = 1,43 m

= 56,45 inch

H = 8,60 m

= 338,69 inch

3. Menghitung Tinggi Cairan

Volume Cairan = $\frac{\pi}{4} \times Di^2 \times H\ cairan$

H cairan = 7,82 m

= 307,90 inch

4. Menghitung Tekanan Desain

Tekanan Udara	=	1	atm
Percepatan Gravitasi	=	9,8	m/s^2
Tekanan Operasi	=	1	atm
Tekanan Hidrostatik	=	$\rho\ cairan\ x\ H\ cairan\ x\ g$	
	=	1022,87 kg/m^3 x 7,82 m x 9,8 m/s^2	
	=	78395,99	kg/ms2
	=	78395,99	N/m2
	=	0,77	atm
	=	11,37	psi

Untuk faktor keamanan diambil over design 10%

Tekanan desain	=	$1{,}1\ x\ (P\ operasi\ +\ P\ hidrostatik)$	
	=	1,1 x (1 atm + 0,77 atm)	
	=	1,95	atm
	=	28,68	psi

5. Menghitung Tebal Desain

Tegangan Maksimum (f)

Bahan Tangki	=	SA-167 Grade 3, Type 304	
Suhu Operasi	=	30	°C
	=	86	°F
f	=	18750,0000	$lb/inch^2$

Efisiensi Penyambungan (E)

Jenis Sambungan	=	Double-welded butt Joint with Backing Strip
E	=	80%
Faktor Korosi	=	0,0125 inch/tahun

Tebal Shell = $\frac{P \times r}{f \times E - 0{,}6 \times P} + c$

= $\frac{28{,}6760 \times 56{,}45}{18750 \times 0{,}8 - 0{,}6 \times 28{,}6760} + c$

= 0,1205 inch

ts sesuai Standar = 3/16 inch

= 0,1875 inch

6. Menghitung Tebal Desain Sebenarnya

OD = $Di + (2\,x\,ts)$

= 113,2721 inch

OD sesuai standar = 114,0000 inch

ID = $OD - (2\,x\,ts)$

= 113,8125 inch

7. Menghitung Tebal Head Tangki

OD Standar = 114,0000 inch

icr = 7,8750 inch

r = 108,0000 inch

icr/r = 7,29%

Radius of Crown r rc = 108,0000 inch

Inside Corner Radius icr r1 = 7,8750 inch

Faktor Intensifikasi Stress untuk Torispherical Dished Heads (inch)

Stress Intensification Factor = $\frac{1}{4} \times \left(3 + \sqrt{\frac{rc}{r1}}\right)$

W = 1,6758 inch

$$Th = \frac{P \times rc \times W}{(2 \times f \times E) - (0{,}2 \times P)} + c$$

$$= \frac{28{,}6760 \times 108 \times 1{,}6758}{(2 \times 18750 \times 0{,}8) - (0{,}2 \times 28{,}6760)} + 0{,}0125$$

= 0,1855 inch

th sesuai Standar = 3/16 inch

= 0,1875 inch

Sf = 2 inch

8. Menghitung Tinggi Head Tangki

a = ID/2
= 113,81 in / 2
= 56,91 in

AB = (ID/2) - icr
= (113,81 in / 2) – 7,8750
= 49,03 in

BC = r – icr
= 108 – 7,8750
= 100.1250 in

AC = 87,2981 in

b = r – AC
= 108 – 87,2981
= 20,7019 in

OA = th standar + sf + b
= 0,1875 in + 2 in + 20,7019 in
= 22,8894 in
= 0,0988 m

9. Menghitung Volume Head Tangki

Vh’ = 0,000049 x ($113,8125^3$)
= 72,2380 $inch^3$

Vsf = $\frac{\pi}{4} \times ID^2 \times sf$

= 20355,16 $inch^3$

V Head Total = $Vh' + Vsf$

= 72,2380 $inch^3$ + 20355,16 $inch^3$

= 20427,4004 $inch^3$

10. Menghitung Tinggi Shell

Volume shell = $V\ total\ -\ V\ head$

= 3371404,52 $inch^3$

Volume shell = $\frac{\pi}{4} \times\ ID^2 \times H\ shell$

Tinggi Shell = 331,2579 inch

11. Menghitung Tinggi Tangki

Tinggi tangki = $H\ shell\ +\ 2OA$

= 377,0368 inch

Tinggi Cairan = $H\ shell\ +\ OA$

= 354,1474 inch

L3.9. Reaktor

Fungsi : Mereaksikan gliserol 99,5% dan gas hidrogen menjadi propilen glikol 99%

Jenis : Fixed Bed Reactor

Bentuk Shell : Silinder Vertikal

Bentuk Head : Torispherical (Flanged and Dished Head)

Bahan : Stainless Steel SA 167 Grade 11, Type 316

Jumlah : 1 buah

Kondisi Operasi:

Suhu : 225 °C = 498,15 K

Tekanan : 5,3 atm = 77,8885 psia

Densitas Campuran = 322,8110 kg/m^3

1. Menghitung Volume Tangki

Volume Bahan Total = 231,1355 m^3

Untuk faktor keamanan diambil over design 10%

Volume Tanki + Over Design = 1,1 x 231,1355 m^3

= 254,2491 m^3

= 15.515.229 $inch^3$

2. Menghitung Dimensi reactor awal

Volume Tangki = $\frac{\pi}{4} \times Di^2 \times H$

H = 3 Di

V Tangki Vertikal = $\frac{3}{4} \times \pi \times Di^3$

Di3 = 107,8632 m^3

Di = 4,7602 m

= 187,4091 inch

= 15,6174 ft

r = 2,3801 m

= 93,7046 inch

= 7,8087 ft

H = 14,2806 m

= 562,2274 inch

= 46,8523 ft

Luas Penampang Reaktor

Ar = 22/7 x $2,3801^2$

= 17,8038 m^2

= 191,6389 ft^2

= 27596,0025 $inch^2$

3. Menghitung Inert Ball

Spesifikasi Inert Ball

Inert Ball	:	Alumina Ceramic Balls	
Composition	:	Al2O3	26%
		SiO2	72%
		Others	2%
Bulk Density	:	1400	kg/m3
Specific Density	:	2400	kg/m3
Porosity	:	1%	/ball
Free Volume	:	45%	https://asenergi.com/en/products/ceramic-balls

Penyusunan Innert Ball

Lokasi	Lapisan	Tebal		Ukuran Inert Ball	
Atas Katalis	1	6	inch	1	inch
	2	6	inch	1/2	inch
Bawah Katalis	1	3	inch	1/4	inch
	2	4	inch	1/2	inch
	3	5	inch	3/4	inch

Volume Innert Ball

Total Tebal Lapisan Inert Ball	tb	=	24,0000	inch
Luas Penampang Reaktor	Ar	=	27596,0025	$inch^2$
Volume Inert Ball	Vb	=	662304,0601	$inch^3$
		=	383,2778	ft^3
		=	10,8532	m^3
Massa Inert Ball	mb	=	8356,9786	kg

4. Menghitung Penyangga Katalis

Penyangga	*Grid Support*			
Bentuk	*Perforated Plate*			
Bahan	*Stainless Steel SA 167 Grade 11, Type 316*			
Ukuran Lubang	120	mesh	0,125	mm

Luas Penyangga	(Ap)	Luas Penampang Reaktor	
	=	17,8038	m^2
	=	191,6389	ft^2
	=	27596,0025	$inch^2$

Tinggi Katalis	(z)	=	1,2466 m	
		=	4,0898 ft	
		=	49,0776 inch	

Gaya Berat yang disebabkan oleh massa katalis dan inert ball

$$F = m\,x\,g$$

Massa Katalis	m	=	14453,6014	kg
Percepatan Gravitasi	g	=	9,8000 m/s2	
Gaya	F	=	141645,2941	N

Tekanan yang diterima penyangga

$$P = \frac{F}{Ap}$$

Gaya	F	=	141645,2941	N
Luas Penyangga	Ap	=	17,8038	m/s2
Tekanan	P	=	7955,8858	Pa
		=	1,1539 psi	

Tegangan Maksimum

Bahan Penyangga		=	SA-167 Grade 11, Type 316	
Suhu Operasi		=	225	°C
		=	437	°F
f		=	18750,0000	$lb/inch^2$

Tebal Penyangga =

$$tp = Di\,x\left(\frac{3\,x\,P}{16\,x\,f}\right)^{\frac{1}{2}}$$

Diameter Reaktor	Di	=	15,6174	ft

Tekanan Penyangga	P	=	1,1539 psi	
Tegangan Maksimum	f	=	18750,0000	psi
Tebal Penyangga	tp	=	0,0531 ft	
		=	0,6366 inch	
		=	0,0162 m	
Volume Penyangga	Vp	=	0,28789	m^3
		=	10,1667	ft^3
		=	17567,9755	$inch^3$

5. Menghitung Pressure Drop

Pressure drop yang disebabkan oleh tumbukan aliran umpan dengan partikel katalis

$$\frac{dP}{dz} = -\frac{G}{\rho g_c D_p}\left(\frac{1-\Phi}{\Phi^3}\right)\left[\frac{150(1-\Phi)\mu}{D_p} + 1.75G\right]$$

Kecepatan Massa Superfisial Gas G = 4,3689 $lbm/ft^2.jam$

= 4,3689 $lbm/ft^2.jam$

Densitas Gas ρ = 5,2918 kg/m^3

= 0,3304 lb/ft^3

Dimensional Constant gc = 4,1700,E+08 $lbm.ft/lbf.jam^2$

= 4,1700,E+08 $lbm.ft/lbf.jam^2$

Diameter Partikel Katalis	Dp	=	0,1470 mm
		=	0,0005 ft
Porositas Katalis	Φ	=	0,5900
Viskositas Gas	η	=	0,0111 cP
		=	0,0268 lb/ft.jam

Fungsi Pressure Drop terhadap Katalis yang Dilewati dP/dz -0,4493

Tinggi Katalis dz = 1,2466 m

		=	4,0898 ft
Pressure Drop		=	-0,5601 psi
Pressure Drop < 10% Tekanan Operasi			**= Memenuhi Kriteria**

6. Menghitung Nozzle Liquid Feed

Aliran masuk dan keluar tangki diasumsikan turbulen (NRe > 2100)

Laju Alir Volumetrik	Qf	=	22,1937 m^3/jam
			= 0,2177 ft^3/detik
Densitas Cairan	ρ	=	377,2498 kg/m^3
		=	23,5510 lb/ft^3
Din Optimum	Di	=	2,9612 inch
		=	0,0752 m
Din Nozzle/ID Pipa	ID	=	3,0680 inch
		=	0,0779 m
Din Nozzle/ID Pipa	OD	=	3,5000 inch
		=	0,0889 m
Nomimal Size Pipe		=	3,00
Schedule Number		=	40

7. Menghitung Nozzle Liquid Product

Laju Alir Volumetrik	Qf	=	28,5302 m^3/jam
		=	0,2799 ft3/detik
Densitas Cairan	ρ	=	322,8110 (kg/m^3)
			= 20,1524 lb/ft^3
Din Optimum	Di	=	3,2491 inch
		=	0,0825 m

Din Nozzle/ID Pipa	ID	=	3,5480 inch
		=	0,0901 m
Din Nozzle/ID Pipa	OD	=	4,0000 inch
		=	0,1016 m
Nomimal Size Pipe		=	3,50
Schedule Number		=	40

8. Menghitung Sparger (Gas Feed)

Jenis Sparger : Perforated Plate

Bahan : Stainless Steel SA 167 Grade 11, Type 316

Standard Cubic Feet per Minute (SCFM)

SCFM Liquid	=	22,1937	m^3/jam
	=	47025,8905	ft^3/menit
SCFM Gas	=	158,2179	m^3/jam
	=	335244,6457	ft^3/menit

Actual Cubic Feet per Minute (ACFM)

Tekanan Operasi = Tekanan Liquid = Tekanan Gas

Temperatur Operasi = Temperatur Liquid = Temperatur Gas

$$ACFM = SCFM \times \frac{14.7}{(14.7 + P)} \times \frac{(460 + T)}{520}$$

Tekanan Liquid	=	5,3000 atm	
	=	77,8885	psi
Temperatur Liquid	=	225,0000	°C
	=	437,0000	°F
ACFM Liquid	=	12879,1227	ft^3/menit
Tekanan Gas	=	5,3000 atm	
	=	77,8885	psi
Temperatur Gas	=	225,0000	°C
	=	437,0000	°F

ACFM Gas	=	91814,4638	ft^3/menit

Gas Exit Velocity (yang diinginkan)

In-Tank, Not Agitated:

5 – 10 FPM Design
25 FPM Maximum

Feet per Minute (FPM Gas) = 25,0000 ft/menit

Luas Keseluruhan Lubang Sparger yang diperlukan

$$A = \frac{ACFM}{FPM}$$

Total Luas Lubang Sparger		At' =	3672,5786 ft^2
Direncanakan :			
Diameter Sparger	Dsp	=	0,7500 Di Reaktor
		=	11,7131 ft
Luas Plate Sparger	Asp	=	107,7969 ft^2
Open Area		=	3407%
Close Area		=	-3307%
Diameter Lubang	Dt	=	0,0984 ft
		=	30,0000 mm
Luas Lubang	At	=	0,0076 ft^2
Jumlah Lubang	nt	=	482496 buah
Volume Sparger			
Luas Plate Sparger	Asp	=	107,7969 ft^2
Tebal Ruang	tsp	=	0,2917 ft
Volume Sparger	Vsp	=	31,4408 ft^3

9. Menghitung Tekanan Design

Tekanan Design = $1{,}2\ x\ (P\ operasi\ +\ P\ hidrostatik)$

= 1,2 x 5,3 atm

= 6,36 atm

= 93,4662 psi

10. Menghitung Tebal Design

Tegangan Maksimum (f)

Bahan Tangki = SA-167 Grade 11, Type 316

Suhu Operasi = 225 °C

= 437 °F

f = 18750,0000 lb/inch^2

Jenis Sambungan = Double-welded butt Joint with Backing Strip

E = 80%

Faktor Korosi (c) = 0,0125 inch/tahun

Tebal Shell = $\frac{P \times r}{f \times E - 0{,}6 \times P} + c$

= $\frac{93{,}4663 \times 93{,}7046}{18750 \times 0{,}8 - 0{,}6 \times 93{,}4663} + 0{,}0125$

= 0,5986 inch

ts yang direncanakan = 0,4000 inch

karena tebal plate sheet standar maksimal 3 inch, maka akan digunakan 2 lapis plate sheet dengan:

ts yang akan digunakan = 0,4375 inch

11. Menghitung Tebal Design Sebenarnya

Outside Diameter = $Di + (2\,x\,ts)$

OD = 188,21 inch

OD sesuai Standar = 126,00 inch

ID = $OD - (2\,x\,ts)$

= 125,56 inch

12. Menghitung Tebal Head Reaktor

OD Standar = 126,00 inch

icr	=	7,6250 inch		
r	=	120,0000 inch		
icr/r	=	6,35%		
Radius of Crown	r	rc	=	120,0000 inch
Inside Corner Radius	icr	r1	=	7,6250 inch

Faktor Intensifikasi Stress untuk Torispherical Dished Heads (inch)

$$\frac{1}{4} \times \left(3 + \sqrt{\frac{rc}{r1}}\right)$$

W = 1,7418 inch

$$Th = \frac{P \times rc \times W}{(2 \times f \times E) - (0{,}2 \times P)} + c$$

$$= \frac{93{,}4662 \times 120 \times 1{,}7418}{(2 \times 18750 \times 0{,}8) - (0{,}2 \times 93{,}4662)} + 0{,}01250$$

= 0,6641 inch

ts yang direncanakan = 0,7000 inch

karena tebal plate sheet standar maksimal 3 inch, maka akan digunakan 3 lapis plate sheet dengan:

ts yang akan digunakan = 0,7500 inch

Sf = 2 inch

13. Menghitung Tinggi Head Reaktor

a = ID/2
= 125,56 in / 2
= 62,78 in

AB = (ID/2) - icr
= (125,56 in / 2) – 7,6250
= 55,16 in

BC = r – icr

	=	120 – 7,6250
	=	112,38 in
AC	=	97,91 in

b	=	r – AC
	=	120 – 97,91 in
	=	22,0922 in

OA	=	th standar + sf + b
	=	0,700 in + 2 in + 22,0922 in
	=	24,7922 in

14. Menghitung Volume Head Reaktor

Vh'	=	0,000049 x $(125,5625)^3$
	=	97,0009 $inch^3$
Vsf	=	$\frac{\pi}{4} \times ID^2 \times sf$
	=	24775,0508 $inch^3$
V Head Total	=	$Vh' + Vsf$
	=	97,0009 $inch^3$ + 24775,0508 $inch^3$
	=	24872,05 $inch^3$

15. Menghitung Tinggi Shell

Volume Tangki + over design	=	$V\ shell\ + (2\ x\ V\ head) + V\ accessories$	
V inert ball	=	662304,0601	$inch^3$
V penyangga katalis	=	17567,9755	$inch^3$
V sparger	=	54329,6299	$inch^3$
Volume shell	=	$V\ total\ - (2\ x\ V\ head)$	
Tinggi Shell	=	1307,7420	inch

16. Menghitung Tinggi Reaktor

Tinggi tangki = $H\ shell\ +\ 2OA$

= 1357,33 inch

17. Menghitung Sistem Pemanas

Sistem Pemanas : Reactor Jacket

Jenis : Single Jacket

Bahan : Stainless Steel SA 167 Grade 11, Type 316

Kondisi Operasi

Fluida Dingin t1	30,00	°C	86,00	°F
Fluida Dingin t2	40,00	°C	104,00	°F
Fluida Panas T1	270,00	°C	518,00	°F
Fluida Panas T2	270,00	°C	518,00	°F

Menghitung True Temperature Difference (ΔTtrue)

$$\Delta LMTD = \frac{(T_1\text{-}t_2) - (T_2 - t_1)}{ln \frac{T_1\text{-}t_2}{T_2 - t_1}}$$

ΔLMTD = 422,9362 °F

Menghitung Luas Permukaan Transfer Panas (A)

Laju Alir Fluida Panas Ws 113477,65 kg/jam = 250175,39 lb/jam

Panas yang diserap Q 10.729.228 kJ/jam =10169346,41 btu/jam

Laju Alir Fluida Dingin Wt 10671,36 kg/jam = 23526,33 lb/jam

Reaktor	Fluida Panas	PG
Jaket	Fluida Dingin	Air
UD	50,0000	BTU/jam.ft^2.°F

LP reaktor (L) = 105.665

L > A = jaket

L < A = koil

$$A = \frac{Q}{U_D \ x \ \Delta LMTD}$$

= 480,8927 ft^2 (luas permukaan transfer panas minimal yang harus dimiliki jaket)

Volume Jaket

Rata-Rata Suhu Air Pemanas	t average	=	270,00 °C
Densitas Air Pendingin	ρ	=	28,0900 kg/m^3
Laju Alir Volumetrik	Qf	=	379,8990 m^3

Untuk faktor keamanan diambil over design 20%

Volume Jaket	Vj	=	455,8788	m^3
		=	27819429	$inch^3$
Tinggi Jaket	(H shell + OA)	=	1332,5342	inch
		=	33,84636868	m

Bagian jaket menutupi bagian reaktor yang meliputi:

1. Bagian Shell Reaktor
2. Bagian Head Bawah Reaktor

Asumsi:

Panjang straight flange pada head bawah untuk volume jaket dihitung sebagai tinggi shell

$$\left(\frac{1}{4}\ x\ \frac{22}{7}\ x\ IDj^2\ x\ H\ shell + sf\right) + (0{,}00049\ x\ IDj^3)\ -$$
$$\left(\frac{1}{4}\ x\ \frac{22}{7}\ x\ ODs\ x\ H\ shell\right) + (0{,}00049\ x\ ODj^3) + \left(\frac{1}{4}\ x\ \frac{22}{7}\ x\ sf\right)$$

Jaket bagian atas	=	1029,0830	=	IDj^2
Jaket bagian bawah	=	0,000049	=	IDj^3
V shell	=	16199686,69	$inch^3$	
	=	265,4653	m^3	
V head bawah	=	97,0009	$inch^3$	
	=	0,0016 m^3		
V straight flange	=	24775,0508	$inch^3$	
	=	0,4060 m^3		

0,000049 IDj3 + 1029,0830 IDj2 - 16224558,74

= 27819429,4730

IDj	=	127,5268 inch (hasil goal seek)
	=	3,2392 m
Target	=	511607,4076 $inch^3$
	=	8,3837 m^3

Tekanan desain jaket

Tekanan Udara	=	1	atm
Percepatan Gravitasi	=	9,8	m/s^2
Tekanan Operasi	=	1	atm
Densitas Cairan	=	1004,2874	kg/m^3
Tinggi Jaket	=	33,8464	m
Tekanan Hidrostatik	=	$\rho\ cairan\ x\ H\ jaket\ x\ g$	
	=	333116,522	kg/ms2
	=	333116,522	N/m2
	=	3,29	atm
	=	48,33	psi
Tekanan desain	=	$1,2\ x\ (P\ operasi\ +\ P\ hidrostatik)$	
	=	5,1461 atm	
	=	75,6268 psi	

Tebal Bahan Jaket

Tegangan Maksimum (f)

Bahan Tangki	=	SA-167 Grade 11, Type 316	
Suhu Operasi	=	225	°C
	=	437	°F
f	=	18750,0000	$lb/inch^2$
E	=	80%	
Faktor Korosi (c)	=	0,0125 inch/tahun	

$$\frac{P \times r}{f \times E - 0,6 \times P} + c$$

Ts =

$$= \frac{75,6268 \times 63,7634}{18750 \times 0,8 - 0,6 \times 75,6268} + 0,0125$$

= 0,3350 inch

ts sesuai Standar = 3/16 inch

= 0,1875 inch

OD = $Di + (2\,x\,ts)$

= 127,9018 inch

= 3,2487 m

Tebal jaket (ODj – IDs) = 1,9018 inch

= 0,0483 m

Uji Jaket

Asumsi: luas permukaan dianggap silinder

Luas Permukaan Reaktor ke Jaket

$$Aj = \frac{22}{7}\ x\ ODs\ x\ H\ jaket$$

Luas Transfer Panas = 527683,5432 $inch^2$

= 3664,4691 ft^2

Luas Transfer Panas Minimal yang Dierlukan

= 480,8927 ft^2

Kesimpulan: Luas Jaket > Luas Transfer Panas Minimal = Memenuhi Kriteria

L3.10. Menara Destilasi

Fungsi	: Memisahkan propilen glikol dengan asetol
Bahan konstruksi	: Carbon Steel SA-283 Grade C
Bentuk	: Silinder vertical dengan alas dan tutup elipsoidal
Jenis	: Sieve-tray
Jumlah	: 1 unit

<u>Data Bobot Molekul dan Densitas</u>

Menentukan Bobot Molekul (BM) dan Densitas (ρ) Produk Atas Kolom Distilasi

P = 765 mmHg
= 14,76 psia

T = 100,02^{o}C
= 80,01^{o}R

R = 10,73 psia.cuft/lbmol.R

Keterangan:
PG : Propilen glikol
A : Air

BM gas rata-rata = (BM PG x fraksi PG gas produk atas) + (BM A x fraksi A gas produk atas)
= (76 kg/kmol x 0,15) + (18 kg/kmol x 0,85)
= 26,49 kg/kmol
= 26,49 lb/lbmol

BM *liquid* rata-rata = (BM PG x fraksi PG liquid produk atas) + (BM A x fraksi A liquid produk atas)
= (76 kg/kmol x 0,15) + (18 kg/kmol x 0,85)
= 26,49 kg/kmol
= 26,49 lb/lbmol

ρ gas campuran $= \frac{\text{P x BM gas rata-rata}}{\text{T x R}}$
= 0,45 kg/m^3

Untuk ρ *liquid* campuran, setiap komponennya dapat dihitung dengan menggunakan persamaan:

$\rho \qquad = \left(A \times B^{-\left(1-\frac{T}{Tc}\right)^{n}}\right) \times \text{fraksi}$

$\rho\ PG \qquad = \left(0{,}25156 \times 0{,}24290^{-\left(1-\frac{568{,}85}{714{,}00}\right)^{0{,}27840}}\right) \times 0{,}8043$

= 411,06 kg/m3

Dengan menggunakan cara yang sama, maka senyawa lainnya diperoleh nilai densitas sebagai berikut :

Tabel L4. 1 Densitas Liquid Campuran untuk Produk Atas (D-101)

Komponen	A	B	n	Tc	T Operasi (K)	Densitas (kg/m^3)	x	ρ.x (kg/m)
Propilen Glikol	0,31839	0,26106	0,20459	626	373,17	364,36	0,85	53,35
Air	0,3471	0,274	0,28571	647,13	373,17	419,07	0,15	357,71
TOTAL								411,06

Menentukan Bobot Molekul (BM) dan Densitas (ρ) Produk Bawah Kolom Distilasi

P = 770 mmHg
= 14,86 psia

T = 100.54°C
= 80.43°R

R = 10,73 psia.cuft/lbmol.R

Keterangan:
PG : Propilen glikol
A : Air

ρ gas campuran $= \frac{\text{P x BM gas rata-rata}}{\text{T x R}}$

= 1.28 kg/m^3

Untuk ρ *liquid* campuran, setiap komponennya dapat dihitung dengan menggunakan persamaan:

ρ = $\left(A \times B^{-\left(1-\frac{T}{Tc}\right)^{n}}\right)$x fraksi

ρ PG = 364,24

Dengan menggunakan cara yang sama, maka senyawa lainnya diperoleh nilai densitas sebagai berikut :

Tabel L4. 2 Densitas Liquid Campuran untuk Produk Bawah (D-101)

Komponen	A	B	n	Tc	T Operasi (K)	Densitas (kg/m³)	x	ρ.x (kg/m)
Propilen Glikol	0,31839	0,26106	0,20459	626	373,17	364,36	0.99	360.59
Air	0,3471	0,274	0,28571	647,13	373,17	419,07	0,01	4.20
TOTAL								364.79

Tahap 2. Menentukan Laju Alir Massa *Overhead Vapor* dan *Bottom Liquid* Maksimum

Top

Laju alir massa *overhead liquid* ($L_{w,top,max}$) = 0.49 kg/s

Laju alir massa *overhead vapour* ($V_{w,top,max}$) = 0.03 kg/s

Bottom

aju alir massa *overhead liquid* ($L_{w,bottom,max}$) = 1.75 kg/s

Laju alir massa *overhead vapour* ($V_{w,bottom,max}$) = 0.10 kg/s

Tahap 3. Menentukan *Liquid-Vapor Flow Factor* (Parameter)

Top

$$F_{lv} = \frac{L_{w,top.max}}{V_{w,top.max}} \sqrt{\frac{\rho_{v,top}}{\rho_{L,\ top}}}$$

= 0,58

Bottom

$$F_{lv} = \frac{L_{w,bottom.max}}{V_{w,bottom.max}} \sqrt{\frac{\rho_{v,bottom}}{\rho_{L,\ bottom}}}$$

$= 1.04$

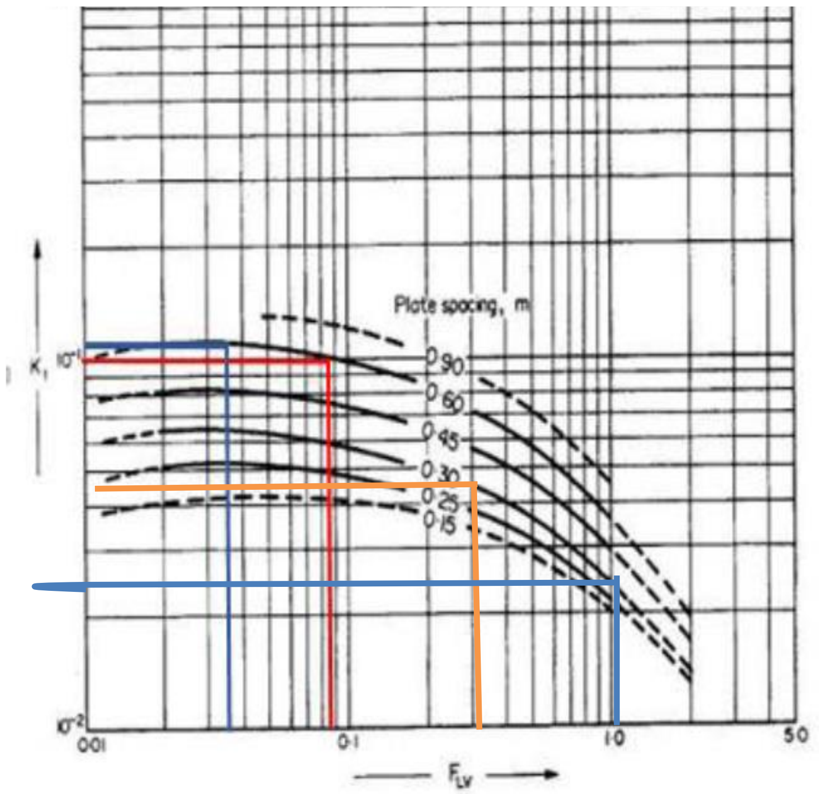
Tahap 4. Menentukan Kecepatan Maksimum *Vapor* (μ_s)

Top

Asumsi jarak antar tray = 0,3 m

= 12 in

Gambar L4. 1 *Fair's Entrainment Flooding Correlation Top* (D-101)



Keterangan :

Plot antara $K_{1,top}$ dan F_{lv} ditandai dengan garis orange.

$K_{1,top}$ = 0,06 m/s

$$\mu_{s,top} = K_{1,top}\sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$$

= 1.80 m/s

Bottom

Asumsi jarak antar tray = 0,3 m

= 12 in

Keterangan :

Plot antara $K_{1,bottom}$ dan F_{lv} ditandai dengan garis biru.

$K_{1,bottom}$ = 0,01 m/s

$\mu_{s,bottom}$ = $K_{1,top}\sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$

= 0,04 m/s

Tahap 4. Menentukan Kecepatan Aktual *Vapor*

Nilai %*flooding* yang dipilih = 80% (*range %flooding yang disarankan adalah 80-90%)*

$\mu_{act,top}$ = %*flooding* x $\mu_{s,top}$

= 90% x 1.80 m/s

= 1.44 m/s

$\mu_{act,bottom}$ = %*flooding* x $\mu_{s,bottom}$

= 90% x 0,04 m/s

= 0,03 m/s

Tahap 5. Menentukan Laju Alir Volumetrik *Vapor* Maksimum

$Q_{top,max}$ = $\frac{V_{w,top,max}}{\rho_{v,top}}$

= 0,06 m^3/s

$Q_{bottom,max}$ = $\frac{V_{w,bottom,max}}{\rho_{v,top}}$

= 0,08 m^3/s

Tahap 6. Menentukan *Bubbling Area* Untuk Kontak Uap – Cair

$$SF_{Koch} = \begin{pmatrix} \frac{2.915}{\rho v^{0.32}} \; jika\ \rho v > 29.1\ kg/m^3 \\ 1 \quad jika\ \rho v \leq 29.1\ kg/m^3 \end{pmatrix}$$

Gambar L4. 3 Pemilihan Nilai SF_{Koch} (D-101)

Karena nilai densitas vapor untuk top dan bottom nilainya lebih kecil dari 29.1 kg/m^3 sehingga nilai SF sama dengan 1. Karena persen flooding yang diambil pada tahap 4 adalah 90% maka persamaan untuk menentukan bubbling area adalah sebagai berikut:

$$An_{top} = \frac{Q_{top,max}}{SF_{top} \times \%flooding \times \mu_{act,top}}$$

$$= 0{,}04\ m^2$$

$$An_{bottom} = \frac{Q_{bottom,max}}{SF_{bottom} \times \%flooding \times \mu_{act,bottom}}$$

$$= 2.44\ m^2$$

Tahap 7. Menentukan Luas Downcomers (Ad), Luas penampang Lintang Utara dan diameter menara berdasarkan kecepatan flooding

Menentukan Luas Penampang Lintang Utara					
Ac top $Ac = \frac{An}{0,88}$	0,05	m2	Ac bottom	2,77	m2
Menentukan Diameter Menara berdasarkan Kec Flooding					
Dc Top	0,25	m	Dc Bottom	1,88	m
Overdesign 20%	0,05	m	Overdesign 20%	0,38	m
	0,16	ft		1,24	ft

Perancangan Tray				
Diameter menara	Dc		0,05	m
Luas Menara	Ac	$Ac = Ad \times 0{,}12\ Ac$	0,05	m2
Luas Downcorner	Ad	An = Ac - Ad	0,01	m2
Luas Netto	An	Aa = Ac - 2Ad	0,04	m2
Luas Aktif	Aa	Ah = 0,1 Aa	0,04	m2
Luas Hole	Ah		0,00	m2

Tahap 8. Laju Alir Volumetrik *Liquid* Maksimum

$$Q_{liq,bottom,max} = \frac{L_{w,bottom,max}}{\rho_{L,bottom}}$$

$$= 0{,}005\ m^3/s$$

Tahap 9. Menentukan Jenis Aliran

Jenis aliran ditentukan dengan menghitung laju alir volumetrik liquid maksimum pada bagian bawah kolom. Sesuai dengan perhitungan yang dilakukan pada tahap 7 nilai laju alir volumetrik liquid maksimum pada bagian bawah kolom adalah 0.005 m^3/s. Karena laju alir volumetrk liquid terdapat pada range 0.003 – 0.03 maka jenis alirannya adalah *cross flow.*

Tahap 11. Menentukan Tebal Kolom Distilasi

Referensi : Appendix D Process Equipment Design by Brownell Young (hal.342)				
P kolom	760,00	mmHg	14,70	psia
P desain ditambah 20% dari tekanan operasi				
P desain	912,00	mmHg	17,64	psia
ID	0,38	m	14,80	in
ri	0,19	m	7,40	in
f (stress value)	18740,37			
E (eff welding)	0,80			
C (Corrotion allow)	0,02			
Waktu pakai alat	10,00			
C (Corrotion allow)	0,15			
ts	**0,16**			
Tebal yang digunakan adalah 3/16	0,19	in	0,005	m

Tahap 12. Menghitung Diameter Luar (OD) dan Diameter Dalam (ID) dan tebal head

Diameter luar = OD = Di + 2 x ts				
OD	0,76	m	29,79	in
Digunakan diameter luar standar menurut referensi Brownell (tabel 5.7 ,hal.89)				
OD	30	in		
	0,76	m		
ID kolom	29,63	in		
	0,75	m		

$$th = \frac{Pdesain\ x\ ri\ x\ W}{(2\ x\ f\ x\ E) - (0.2\ x\ Pdesain)} + C$$

icr	1,88			
r	30,00			
icr/r	6,25%	> 6% maka memenuhi syarat untuk torispherical head		
W	1,75			
th	0,1759			
Tebal yang digunakan adalah 3/16	0,1875	in	0,0048	m

Berdasarkan referensi Bronell and young (tabel 5.6) untuk th = 3/16 in diperoleh :		
sf (*standard straight flange)*	1,5 - 3,5	in
Diambil sf	1,5	in
a	14,81	in
AB	12,94	in
BC	28,13	in
AC	24,97	in
b	5,03	in
OA	6,71	in
	0,17	m

Tahap 13. Menentukan Tinggi Kolom

1. Tinggi Column Sump Level

DATA				
TOP			BOTTOM	
Flow rate distillate	280,89	kg/jam	8272,80	kg/jam
Temperatur	100,02	°C	100,55	°C
Tekanan	760,00	mmHg	760,00	mmHg
BM gas	3,03	kg/kmol	91,27	kg/kmol

Tinggi *Column Sump Level*

Column sump level ditentukan berdasarkan *liquid draw – off rate* (volume liquid produk bawah) yang nilainya bergantung pada tujuan dialirkan kemana liquid tersebut setelahnya, misalnya ke tangki penyimpanan atau ke unit proses lain. Sehingga dari tujuan aliran tersebut *liquid draw – off rate*nya memiliki rumus yang berbeda–beda. Karena produk bawah diumpankan ke unit lain maka:

Reflux	0,3766	Kg/Jam
$P = \frac{B}{\rho_{L,bottom}} \times \frac{1\,h}{60\,min}$		
P	0,0000	m^3/min
$R = \frac{R}{\rho_{L,bottom}} \times \frac{1\,h}{60\,min}$		
R	0,0000172	m^3/min
V	2P + 2R	
	6,87137E-05	m^3/min

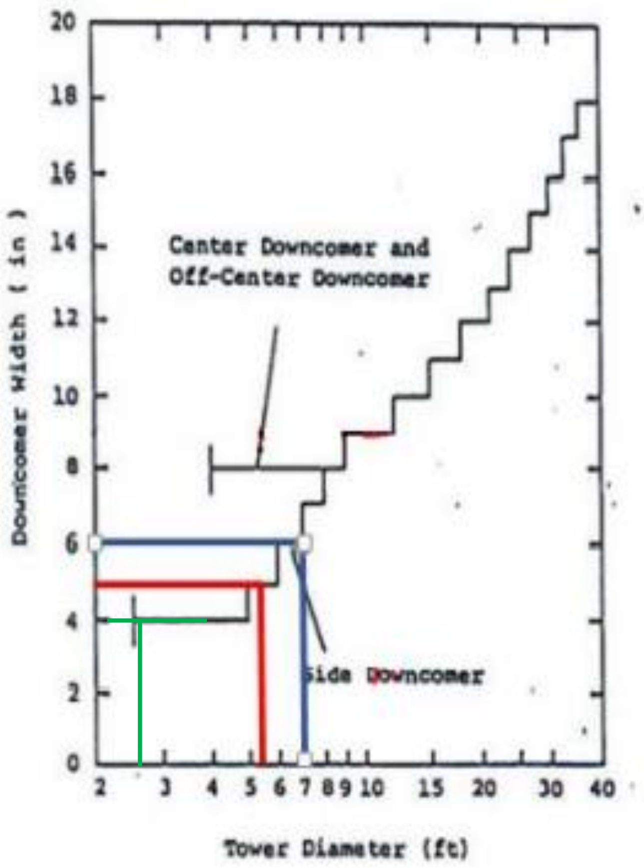
Level	Holding Time (menit)
HHLL – TL	5
HLL – TL	4
NLL – TL	2,5
LLL – TL	2
HLLL – TL	1

Volume HHLL-TL		
$Volume\ HHLL - TL = Liquid\ Draw\ off - rate \times Holding\ time$	0,0003	m^3
	0,00310	m
Volume HLL-TL		
$Volume\ HLL - TL = Liquid\ Draw\ off - rate \times Holding\ time$	0,0003	m^3
$H_{HLL-TL} = \frac{4 \times Volume\ HLL - TL}{\pi ID^2}$	0,00248	m
Volume NLL-TLL		
$Volume\ NLL - TL = Liquid\ Draw\ off - rate \times Holding\ time$	0,0002	m^3
$H_{NLL-TLL} = \frac{4 \times Volume\ NLL - TL}{\pi ID^2}$	0,00155	m
Volume LLL-TL		
$Volume\ LLL - TL = Liquid\ Draw\ off - rate \times Holding\ time$	0,0001	m^3
$H_{LLL-TL} = \frac{4 \times Volume\ LLL - TL}{\pi ID^2}$	0,00124	m

Volume LLLL-TL			
$Volume\ LLL - TL = Liquid\ Draw\ off - rate \times Holding\ time$		0,0001	m^3
$H_{LLL-TL} = \frac{4 \times Volume\ LLL - TL}{\pi ID^2}$		0,00062	m
Maka, ketinggian untuk masing-masing level adalah			
$H_{HHLL-HLL} = H_{HHLL-TL} - H_{HLL-TL}$		0,00062	m
$H_{HLL-NLL} = H_{HLL-TL} - H_{NLL-TLL}$		0,00093	m
$H_{NLL-LLL} = H_{NLL-TLL} - H_{LLL-TL}$		0,00031	m
$H_{LLL-LLLL} = H_{LLL-TL} - H_{LLLL-TL}$		0,00062	m
$H_{LLLL-TL}$		0,00062	m
Maka tinggi column sump = $H_{bs} = H_{HHLL-TL}$		0,00310	m
2. Tinggi Total Kolom			
Jumlah tray aktual +1	5		
Tray spacing	0,30	m	
Tinggi total tray	1,64	m	
Tinggi penutup kolom	0,17	m	
Tinggi total kolom	1,98	m	

Tahap 14. Mendesain Geometri *Tray*

Hole diameter (d_h)	6	mm
Hole area	5%	
Hole area top	0,072217628	m^2
Hole area bottom	0,001597181	m^2
Hole pitch	0,162040773	m
Weir height (h_w)	0,05	m
Weir length (L_w)	1	
Weir length top	0,451485903	m
Weir length bottom	0,752476505	m
Dc	0,75	m
	2,468755017	ft
Downcomer widht	4	in
	0,1016	m



Gambar L4. 4 Minimum Downcomer Width (D-101)

Keterangan:

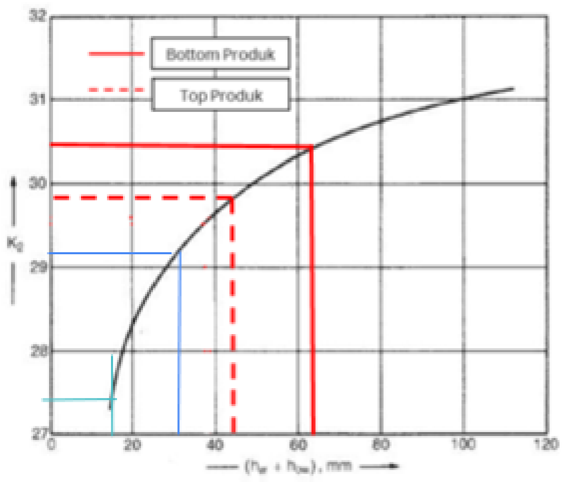
Berdasarkan plot garis berwarna hijau didapatkan *downcomer width* sebesar 4 in atau 0,1016 in.

Tahap 15. Evaluasi Desain

Cek *Weeping Top*

Top

Cek Weeping Top		
Turndown ratio	80%	
$L_{w,top,max}$	0,52	Kg/s
$L_{w,top,min}$	0,42	Kg/s
h_{ow}	14,95	m
$h_w + h_{ow}$	15,40	m
K2	23	



K2

μ_h	0,304413674	m/s
μ_{am}	0,688935184	m/s
Nilai $\mu am > \mu h$ sehingga weeping tidak terjadi		DITERIMA

Bottom

Cek Weeping Bottom		
Turndown ratio	80%	
$L_{w,bottom,max}$	1,85	Kg/s
$L_{w,bottom,min}$	1,48	Kg/s
h_{ow}	23,09	mm
$h_w + h_{ow}$	23,84	mm
K2	19	

μ_h	1,358994297	m/s
μ_{am}	39,04580995	m/s
Nilai $\mu am > \mu h$ sehingga weeping tidak terjadi		DITERIMA

Cek *Pressure Drop*

Top

Hole Area	0,072217628	m^2
Active area	0,043058449	m^2
(Hole Area/Active area) x 100%	168%	
Tebal stainles steel (tp)	3	mm
tp/dh	0,5	
CO	0,66	

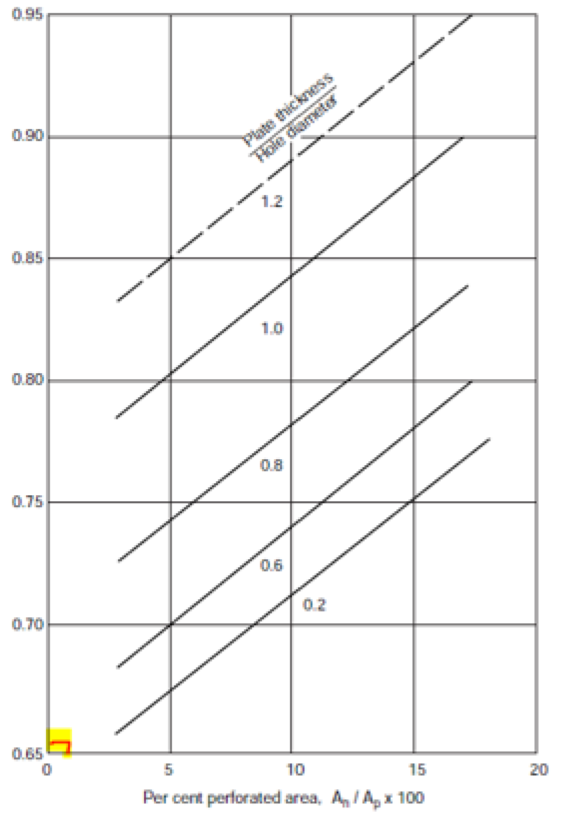


Figure 11.34. Discharge coefficient, sieve plates (Liebson *et al.*, 1957)

u_h	1,16	m/s
Dry plat pressure drop (h_d)	0,17	mm
Residual head (h_r)	0,03	mm
Pressure drop total (h_t)	15,61	mm
ΔP_t	**0,00**	**Pa**
	0,00	**mmHg**
Nilai ΔPt < 10 mmHg maka memenuhi kriteria design		**DITERIMA**

Bottom

Cek Pressure Drop		
Hole Area	0,00	m^2
Active area	2,44	m^2
(Hole Area/Active area) x 100%	0,00	
Tebal stainles steel (tp)	3,00	mm
tp/dh	0,50	
CO	0,66	

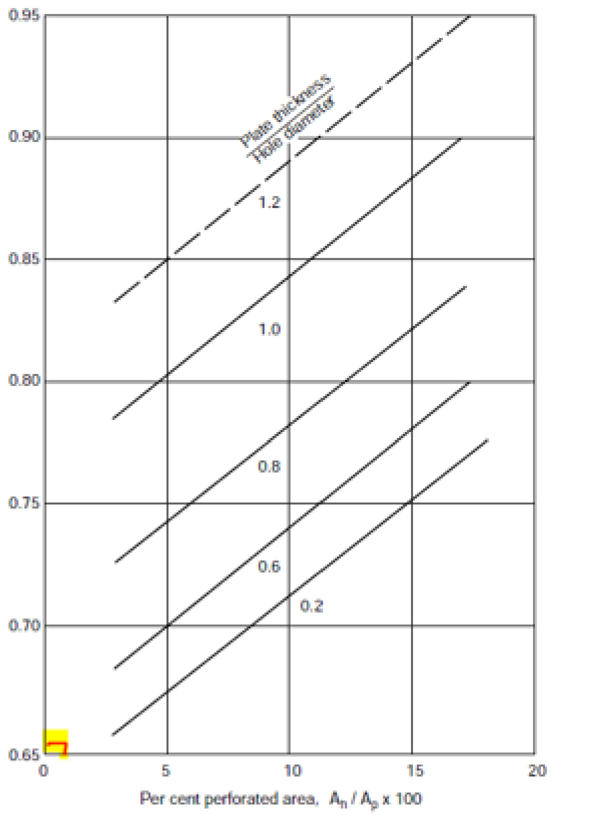


Figure 11.34. Discharge coefficient, sieve plates (Liebson *et al.*, 1957)

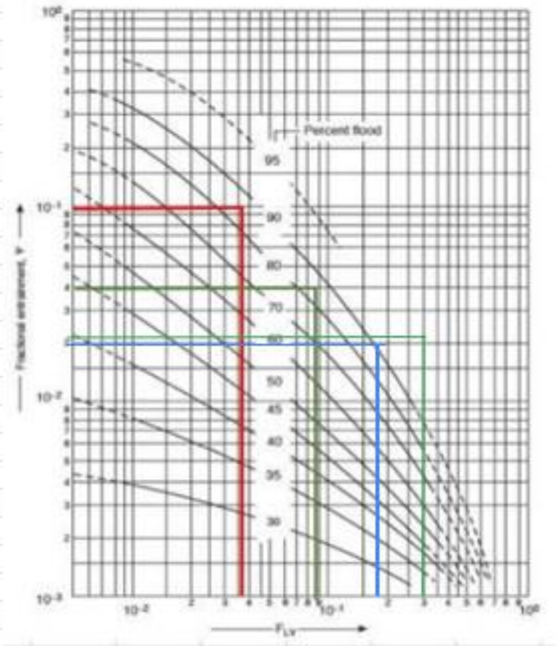
CO

u_h	0,02	m/s
Dry plat pressure drop (h_d)	0,00	mm
Residual head (h_r)	9,73	mm
Pressure drop total (h_t)	33,58	mm
ΔP_t	**0,00**	**Pa**
	0,00	**mmHg**
Nilai ΔPt < 10 mmHg maka memenuhi kriteria design	**DITERIMA**	

Cek *Downcomer Area Liquid Backup*

Top

Cek Entraintment	
Flooding	0,80
Flv	0,58
Fraksional entrainment (ψ)	**0,02**

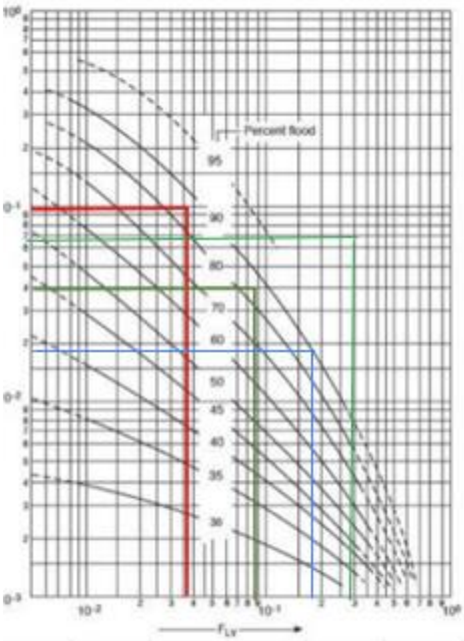


Fraksional entrainment (ψ)

Nilai fraksional entrainment (ψ) <0,1 sehingga sesuai dengan param	**DITERIMA**

Bottom

Cek Entraintment	
Flooding	0,80
Flv	1,04
Fraksional entrainment (ψ)	**0,07**



Fraksional ent

Nilai fraksional entrainment (ψ) <0,1 sehingga sesuai dengan pa	**DITERIMA**

Cek *Enterainment*

Top

Cek Downcomer Area Liquid Backup			
Selisih antara downcomwe clearance dg weir heigh	5,00	mm	
h_{ap}	45,00	mm	
A_{ap}	0,02	m^2	
h_{dc}	0,65	mm	
h_b	**31,66**	**mm**	
	0,03	m	
Tray spacing (l_t)	0,60	m	
	600,00	mm	
$1/2(l_t+h_w)$	**300,23**	**mm**	
Nilai hb < $1/2(l_t+h_w)$ maka sesuai dengan parameter desain			DITERIMA

Bottom

Cek Downcomer Area Liquid Backup		
Selisih antara downcomwe clearanc	5,00	mm
h_{ap}	45,00	mm
A_{ap}	0,03	m^2
h_{dc}	3,74	mm
h_b	**61,16**	**mm**
	0,06	m
Tray spacing (l_t)	0,60	m
	600,00	mm
$1/2(l_t+h_w)$	**300,23**	**mm**
Nilai hb < $1/2(l_t+h_w)$ maka sesuai dengan parameter de	**DITERIMA**	

LAMPIRAN 5

ANALISIS EKONOMI

Untuk merancang suatu pabrik, diperlukan analisis terhadap berbagai parameter ekonomi guna melihat potensi mengenai jumlah investasi modal. Parameter analisis ekonomi yang lakukan yaitu struktur kepemilikan modal, besarnya keuntungan yang didapat, lama investasi modal kembali dan *break even point* (BEP).

Pada perancangan pabrik proilen glikol ini, perkiraan mengenai perhitungan ekonomi dilakukan berdasarkan pada kapasitas yang telah ditentukan. Sedangkan perkiraan harga peralatan alat-alat produksi dan penunjang diambil dari situs jual beli alibaba.com dan matche.com, kenaikkan harga alat pada tahun pabrik didirikan yaitu 2026, diprediksi menggunakan *Chemical Engineering Plant Cost Index.*

L5.1. Ketetapan yg diambil

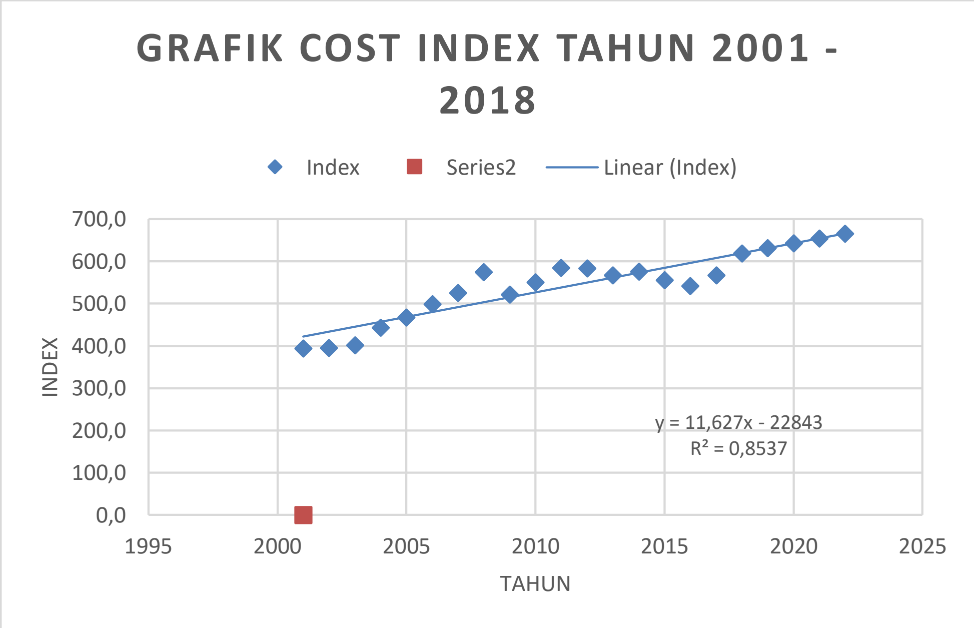
1	Pembangunan fisik pabrik	2025	*masa kontruksi adalah 1 tahun
2	Pabrik mulai beroperasi	2026	
3	Hari kerja per tahun	330	hari
4	Shut down alat per tahun	35	hari
5	Umur teknis pabrik	10	tahun
6	*Salvage value* (Nilai Rongsok)	10%	DFCI (tanpa harga tanah dan bangunan)
7	Nilai tukar rupiah	15.000,00	rupiah
8	Tingkat suku bunga bank	8,06%	per tahun
9	Kenaikan harga bahan baku dan produksi	10%	per tahun

L5.2. Index harga

Metode perkiraan harga peralatan dalam perancangan pabrik propilen glikol ini menggunakan *Chemical Engineering Plant Cost Index*. Indeks harga untuk tahun 2001 sampai 2018, dan proyeksinya untuk tahun 2019 sampai 2026, ditampilkan pada Tabel L5.2

Tahun	Index
2001	394,3
2002	395,6
2003	401,7
2004	444,2
2005	468,2
2006	499,6
2007	525,4
2008	575,4
2009	521,9
2010	550,8
2011	585,7
2012	584,6
2013	567,3
2014	576,1
2015	556,8
2016	541,7
2017	567,5
2018	619,5
2019	631,36
2020	642,99
2021	654,62
2022	666,24
2023	677,87
2024	689,50
2025	701,12
2026	712,75

(Sumber: Chemical Engineering Plant Cost Index, www.chemengonline.com/pci)



L5.3. Daftar Harga peralatan

Harga peralatan pada kapasitas yang sesuai dengan tahun pendirian pabrik. dapat dihitung dengan menggunakan rumus sebagai berikut:

$$H_2 = H_1 x \left[\frac{I_2}{I_1}\right] x \left[\frac{K_2}{K_1}\right]^{exp} x \left[\frac{IDR\ 14.510}{1\$}\right]$$

$$\text{Harga Total} = H_2 \times n$$

Dimana:

H_2 : Harga alat yang dicari (IDR)

H_1 : Harga referensi (harga terpasang atau harga yang telah diketahui) ($)

I_1 : Index harga referensi atau terpasang

I_2 : Index harga terhitung pada tahun tertentu

K_2 : Kapasitas alat yang dicari

K_1 : Kapasitas alat terpasang

n : Jumlah alat

Exp : Eksponen

L.5.3.1 Harga Peralatan Utama

Kode	Nama Alat	Bahan/Material	Spesifikasi Desain		Harga Satuan	Harga Total
					Rp	Rp
T-01	Tangki Penyimpanan Gliserol 9	SS, SA-167 Grade 11 Type 316	531,73	m3	Rp 8.078.103.953,67	Rp 8.078.103.953,67
T-02	Tangki Penyimpanan Hidrogen	Carbon Steel SA 285 Grade A	10.758,64	m3	Rp 15.077.843.613,17	Rp 15.077.843.613,17
T-03	Tangki Penyimpanan PG 99%	SS, SA-167 Grade 11 Type 316	434,26	m3	Rp 7.291.796.655,84	Rp 7.291.796.655,84
R-01	Fixed Bed Reactor	SS, SA-167 Grade 11 Type 316	64,87	m3	Rp 17.536.257.450,48	Rp 17.536.257.450,48
K-01	Kompresor Hidrogen	SS, SA-167 Grade 11 Type 316	102,00	m3	Rp 1.445.842.602,75	Rp 1.445.842.602,75
FD-01	H2 Separator	Commercial Steel	70,00	m3/h	Rp 322.546.462,99	Rp 322.546.462,99
H-01	Heater Gliserol 99,5%	SS SA 167, Grade 3, Type 304	10,51	m2	Rp 60.978.933,30	Rp 60.978.933,30
H-02	Heater Hidrogen 100%	SS SA 167, Grade 3, Type 304	10,51	m2	Rp 60.978.933,30	Rp 60.978.933,30
H-03	Heater umpan kolom destilasi	SS SA 167, Grade 3, Type 304	1,75	m2	Rp 56.164.806,99	Rp 56.164.806,99
MD-01	Kolom destilasi	SS SA 167, Grade 3, Type 304	18.857,66	pounds	Rp 13.012.253.087,11	Rp 13.012.253.087,11
K-01	Condenser keluaran reaktor	Stainless steel AISI tipe 316	356,72	ft2	Rp 3.241.511.717,60	Rp 3.241.511.717,60
K-02	Condensor top destilasi	Stainless steel AISI tipe 316	356,72	ft2	Rp 3.257.558.805,31	Rp 3.257.558.805,31
CL-01	Cooler produk bottom distilasi	SS SA 167, Grade 3, Type 304	16,78	m2	Rp 11.380.594.604,58	Rp 11.380.594.604,58
AC-01	Accumulator	SS SA 167, Grade 3, Type 304	7,81	m3	Rp 57.769.515,76	Rp 57.769.515,76
RB-01	Reboiler	SS SA 167, Grade 3, Type 304	15,71	ft2	Rp 365.873.599,81	Rp 365.873.599,81
P-01	Pompa	Commercial steel	-	HP	Rp 1.091.201,96	Rp 2.182.403,93
P-02	Pompa	Commercial steel	-	HP	Rp 363.733,99	Rp 727.467,98
P-03	Pompa	Commercial steel	-	HP	Rp 363.787,48	Rp 727.574,96
P-04	Pompa	Commercial steel	-	HP	Rp 363.787,48	Rp 727.574,96
P-05	Pompa	Commercial steel	-	HP	Rp 363.787,48	Rp 727.574,96
Total						Rp 81.251.167.339,45
Biaya Peralatan Utama					Rp	81.251.167.339
Biaya Pengangkutan + Asuransi				10%	Rp	8.125.116.734
Biaya Administrasi pelabuhan				5%	Rp	4.062.558.367
SUBTOTAL					Rp	93.438.842.440
Bea Masuk				10%	Rp	9.343.884.244
TOTAL					**Rp**	**102.782.726.684**

L.5.3.2 Harga peralatan penunjang

Kode	Nama Alat	Bahan	Spesifikasi Desain		Harga Satuan	Harga Total
					Rp	Rp
TU-02	Tangki Air Domestik	Stainless Steel SA 167 Grade 3, Type 304	6,51	m3	100.294.298	100.294.298
TU-03	Tangki Air Pendingin	Stainless Steel SA 167 Grade 3, Type 304	49,05	m3	625.836.421	625.836.421
TU-04	Tangki Demineralisasi	Stainless Steel SA 167 Grade 3, Type 304	50,26	m3	667.558.849	667.558.849
TU-05	Tangki Air Umpan Boiler	Stainless Steel SA 167 Grade 3, Type 304	1,21	m3	701.257.733	701.257.733
6	Tangki Bahan Bakar	Stainless Steel SA 167 Grade 3, Type 304	437,40	m3	702.862.442	702.862.442
1	Cooling tower	Carbon Steel	45.606,75	kg/hr	1.025.408.905	1.025.408.905
P-01	Pompa Utilitas 01	Commercial Steel	0,25	HP	560.043.361	1.120.086.722
P-02	Pompa Utilitas 02	Commercial Steel	2,00	HP	702.862.442	1.405.724.883
P-03	Pompa Utilitas 03	Commercial Steel	0,50	HP	768.655.501	1.537.311.003
P-04	Pompa Utilitas 04	Commercial Steel	2,00	HP	587.323.410	1.174.646.820
-	Generator	-	2.721,30	kW	587.323.410	587.323.410
-	Boiler	Carbon Steel	437,40	l/hr	702.862.442	702.862.442
TU-01	Tangki Air bersih		55,58	m3	144.423.789	144.423.789
	Kendaraaan Pemasaran	Colt box mitsubishi			367.500.000	1.102.500.000
	Mobil Direktur + Komisaris	Toyota Rush			500.000.000,00	1.500.000.000
	Mobil Kepala Divisi	Toyota Avanza			197.700.000,00	790.800.000
	Forklift	REDDOT 4-5ton diesel forklift truck with Japanese engine			369.083.017	1.476.332.069
Total						Rp15.365.229.787
Biaya Peralatan Penunjang					Rp	15.365.229.787
Biaya Pengangkutan + Asuransi				10%	Rp	1.536.522.979
Biaya Administrasi pelabuhan				5%	Rp	768.261.489
SUBTOTAL					Rp	17.670.014.255
bea masuk				10%	Rp	1.767.001.425
TOTAL					**Rp**	**19.437.015.680**
TOTAL ALAT UTAMA + ALAT PENUNJANG					**Rp**	**122.219.742.365**

L5.4. Daftar Gaji Karyawan

No.	Jabatan	Jumlah	Jenjang Pendidikan	Total/Orang	Total
			Minimum	(IDR)	(IDR)
1	Dewan Komisaris	1	-	24.000.000,00	24.000.000,00
2	Direktur Utama	1	S2	28.800.000,00	28.800.000,00
3	Direktur Produksi dan Teknik	1	S1	19.200.000,00	19.200.000,00
4	irektur Administrasi dan Keuangan	1	S1	19.200.000,00	19.200.000,00
5	Direktur SDM dan Umum	1	S1	19.200.000,00	19.200.000,00
6	Sekretaris	1	S1	14.400.000,00	14.400.000,00
7	Kasie. LITBANG (R&D)	1	S1	14.400.000,00	14.400.000,00
8	Karyawan LITBANG (R&D)	2	S1	12.000.000,00	24.000.000,00
9	Kepala Bagian Produksi	1	S1	14.400.000,00	14.400.000,00
10	Kepala Bagian Teknik	1	S1	14.400.000,00	14.400.000,00
11	Kepala Bagian Quality Control	1	S1	14.400.000,00	14.400.000,00
12	Kepala Bagian SDM	1	S1	14.400.000,00	14.400.000,00
13	Kepala Bagian Pemasaran	1	S1	14.400.000,00	14.400.000,00
14	Kepala Bagian Keuangan dan Akuntansi	1	S1	14.400.000,00	14.400.000,00
15	Kepala Bagian Umum	1	S1	14.400.000,00	14.400.000,00
16	Kasie. Produksi	1	D3	11.040.000,00	11.040.000,00
17	Kasie. Bahan Baku	1	D3	11.040.000,00	11.040.000,00
18	Kasie. Utilitas	1	D3	11.040.000,00	11.040.000,00
19	Kasie. Bengkel dan Perawatan	1	D3	11.040.000,00	11.040.000,00

20	Kasie. Jaminan Mutu	1	D3	11.040.000,00	11.040.000,00
21	Kasie. Pengendaliam proses	1	D3	11.040.000,00	11.040.000,00
22	Kasie. Personalia	1	D3	11.040.000,00	11.040.000,00
23	Kasie. Ketenagakerjaan	1	D3	11.040.000,00	11.040.000,00
24	Kasie. Pembelian	1	D3	11.040.000,00	11.040.000,00
25	Kasie. Penjualan	1	D3	11.040.000,00	11.040.000,00
26	Kasie. Promosi dan Periklanan	1	D3	11.040.000,00	11.040.000,00
27	Kasie. Research dan Marketing	1	D3	11.040.000,00	11.040.000,00
28	Kasie. Keuangan	1	D3	11.040.000,00	11.040.000,00
29	Kasie. Akuntansi	1	D3	11.040.000,00	11.040.000,00
30	Kasie. Humas	1	D3	11.040.000,00	11.040.000,00
31	Kasie. Keamanan dan Keselamatan	1	D3	11.040.000,00	11.040.000,00
32	Kasie. Administrasi	1	D3	11.040.000,00	11.040.000,00
33	Karyawan Produksi	12	SMA	5.500.000,00	66.000.000,00
34	Karyawan Divisi Gudang Produksi dan Bahan Baku	8	SMA	5.500.000,00	44.000.000,00
35	Karyawan Divisi Utilitas	8	SMA	5.500.000,00	44.000.000,00
36	Karyawan Divisi Bengkel dan Perawatan	8	SMA	5.500.000,00	44.000.000,00
37	Karyawan Divisi Quality Control	8	SMA	5.500.000,00	44.000.000,00
38	Karyawan Divisi Personalia dan Ketenagakerjaan	3	SMA	4.800.000,00	14.400.000,00
39	Karyawan Divisi Penjualan dan Pembelian	4	SMA	4.800.000,00	19.200.000,00
40	Karyawan Divisi Promosi dan Periklanan	2	SMA	4.800.000,00	9.600.000,00
41	Karyawan research dan Marketing	2	SMA	4.800.000,00	9.600.000,00
42	Karyawan Divisi Keuangan	2	SMA	4.800.000,00	9.600.000,00
43	Karyawan Divisi Akuntansi	2	SMA	4.800.000,00	9.600.000,00

44	Karyawan Divisi Humas	2	SMA	4.800.000,00	9.600.000,00
45	Karyawan Keamanan	8	SMA	5.500.000,00	44.000.000,00
46	Kepala HSE	1	S1	14.400.000,00	14.400.000,00
47	Karyawan HSE	2	SMA	4.800.000,00	9.600.000,00
48	Dokter	1	S1	14.400.000,00	14.400.000,00
49	Karyawan Administrasi	2	SMA	4.800.000,00	9.600.000,00
50	Karyawan Kebersihan dan Logistik	12	SMA	4.800.000,00	57.600.000,00
Total					924.880.000,00
Gaji / tahun (a)			Rp		11.098.560.000,00
Tunjangan Hari Raya (1 bulan gaji)			Rp		924.880.000,00
Tunjangan kesehatan (0,025 a)			Rp		277.464.000,00
Tunjangan Makan + Transport (0,05a)			Rp		554.928.000,00
TOTAL GAJI PER TAHUN			Rp		12.855.832.000,00

L5.5 Perhitungan Modal Investasi (TCI)

L5.5.1 Modal Tetap *(Fixed Capital investment*/FCI*)*

Fixed Capital Investment (FCI) atau modal investasi tetap adalah modal yang diperlukan untuk membeli peralatan yang diperlukan. Fixed Capital Investment dibagi menjadi 2 yaitu:

a. Modal Investasi Tetap Langsung/*Direct Fixed Capital Investment* (DFCI)

A. Modal Investasi Tetap Langsung / Direct Fixed Capital Investment (DFCI)					
No	**Komponen**	**Persen**			**Biaya**
1	Pengadaan alat (Peralatan proses dan utilitas	100%	A	Rp	122.219.742.365
2	Instrumentasi dan control	25,0%	A	Rp	30.554.935.591
3	Instalasi	8,0%	A	Rp	9.777.579.389
4	Perpipaan terpasang	60,0%	A	Rp	73.331.845.419
5	Pelistrikan terpasang	18,0%	A	Rp	21.999.553.626
6	Pemasangan alat	40,0%	A	Rp	48.887.896.946
7	Bangunan pabrik	60,0%	A	Rp	73.331.845.419
8	Service facilities and yard improvement	50,0%	A	Rp	61.109.871.182
9	Harga Tanah (Land survey & cost)			Rp	70.000.000.000
Sub Total			A	Rp	511.213.269.936
DFCI tak terduga		15%	A'	Rp	76.681.990.490
Total Modal Investasi Tetap Langsung (DFCI)			B	**Rp**	**587.895.260.427**

Keterangan:

Luas tanah	: 20.000 m^2
Harga tanah	: Rp 3.500.000,00/ m^2
Harga tanah keseluruhan	: Rp 70.000.000.000,00

b. Modal Investasi Tetap Tidak Langsung / *Indirect Fixed Capital Investment* (IFCI)

Tabel L5. 1 *Indirect Fixed Capital Investment* (IFCI)

B. Modal Investasi Tetap Tidak Langsung / Indirect Fixed Capital Investment (IFCI)					
No	**Komponen**	**Persen**			**Biaya**
10	Engineering and supervision	15,0%	B	Rp	88.184.289.064,00
11	Contactor's fee	6%	B	Rp	35.273.715.625,60
12	Biaya tak terduga (Cotingency)	15%	B	Rp	88.184.289.064,00
13	Trial Run			Rp	58.594.516.661,25
	Sub Total		B'	Rp	270.236.810.414,86
	IFCI tak terduga	15%	B'	Rp	40.535.521.562,23
Total Modal Investasi Tetap Tidak Langsung (IFCI)			C	**Rp**	**310.772.331.977,09**
Keterangan :	Trial run dilakukan selama 2 minggu				
Total Modal Investasi Tetap (FCI) = DFCI + IFCI				**Rp**	**898.667.592.403,78**

L5.5.2 Pengeluaran untuk *Trial Run*

C. Perhitungan Biaya Trial Run

Perhitungan biaya trial run untuk masa 2 minggu dengan jumlah hari kerja 14 hari =

Perhitungan : (14 hari x 24 jam/hari x harga x kebutuhan/jam)

Komponen		**Kebutuhan**	**Harga/satuan**		**Biaya/2 minggu**	
Persediaan bahan baku						
Gliserol	Kg/jam	8.330,725	Rp	18.000	Rp	50.384.227.191
Hidrogen	Kg/jam	837,259	Rp	12.314	Rp	3.464.203.971
Katalis Raney Nickel	Kg/jam	6.096,623	Rp	149.460	Rp	911.201.242
Total persediaan bahan baku	a				**Rp**	**54.759.632.404**

Komponen	**Kebutuhan**	**Harga/satuan**		**Biaya/2 minggu**	
Persediaan sarana penunjang					
1.Solar (liter/hari)	437,402	Rp	9.850	Rp	60.317.733
2. Listrik (kWh)	2.721,305	Rp	1.115	Rp	1.019.271.907
3. Resin (Liter)	26.240,901	Rp	7.500,0	Rp	2.755.294.618
Total persediaan bahan penunjang				**Rp**	**3.834.884.257**
Total biaya trial run	b			**Rp**	**58.594.516.661**

L5.6. Modal Kerja (Working Capital)

Modal kerja dihitung untuk masa 3 bulan dengan jumlah hari kerja = 90 hari

Komponen		**Kebutuhan**	**Harga/satuan**	**Biaya/3 bulan**
Persediaan bahan baku				
Gliserol	Kg/jam	8.330,73	Rp 18.000	Rp 323.898.603.368
Hidrogen	Kg/jam	837,26	Rp 12.314	Rp 22.269.882.673
Katalis Raney Nickel	Kg/jam	6.096,62	Rp 149.460	Rp 911.201.242
Total persediaan bahan baku		a		**Rp 347.079.687.282**

Komponen	**Kebutuhan**	**Harga/satuan**	**Biaya/3 bulan**
Persediaan sarana penunjang			
1.Solar (liter/hari)	437,40	Rp 9.850	Rp 387.756.855
2. Listrik (kWh)	2.721,30	Rp 1.115	Rp 6.552.462.258
3. Resin (Liter)	26.240,90	Rp 7.500	Rp 17.712.608.257
Abonemen/bulan	-	Rp 87.081.754	Rp 261.245.263,28
Total persediaan bahan penunjang	b		**Rp 24.914.072.633**
Total biaya trial run (a + b)			**Rp 371.993.759.915**

c. Biaya pengemasan & distribusi produk	2%	bahan baku	Rp	6.941.593.745,65
d. Biaya pengawasan mutu	1%	bahan baku	Rp	3.470.796.872,82
e. Biaya pemeliharaan dan perbaikan	2%	DFCI	Rp	11.757.905.208,53
f. Gaji karyawan	3	x gaji/bulan	Rp	2.774.640.000,00
Sub Total WCI (a s/d f)			**Rp**	**396.938.695.742,05**
WCI tak terduga	20%	sub total WCI	Rp	79.387.739.148,41
Total Modal Kerja (WCI)			**Rp**	**476.326.434.890,45**
Total Modal Investasi (TCI) = FCI + WCI =			**Rp**	**1.374.994.027.294,23**

L5.7. Struktur Permodalan

Yang dapat dijaminkan	DFCI	Rp	587.895.260.427
Jika bank memberikan pinjam	75% DFCI	**Rp**	**440.921.445.320**
Dibulatkan menjadi		Rp	440.000.000.000
Persentasi permodalan			
Total Capital Investment (TCI)		Rp	1.374.994.027.294
Modal Sendiri		Rp	934.994.027.294
Modal Sendiri = (TCI - pinjaman bank)/TCI x 100%			68,00%
Pinjaman Bank = Pinjaman Bank/TCI x 100%			32,00%

L5.8. Angsuran Pokok Pinjaman dan Bunga Bank

Jangka Waktu Pinjaman	10 Tahun
Grace Period	1 Tahun
Bunga Bank per tahun (diasumsikan tetap selama	8,06%

Tahun	Pokok Pinjaman (Rp)	Angsuran Pokok	Bunga	Jumlah	Sisa
0	Rp 440.000.000.000	Rp -	Rp 35.464.000.000	Rp 35.464.000.000	Rp 440.000.000.000
1	Rp 440.000.000.000	Rp 44.000.000.000	Rp 35.464.000.000	Rp 79.464.000.000	Rp 396.000.000.000
2	Rp 396.000.000.000	Rp 44.000.000.000	Rp 31.917.600.000	Rp 75.917.600.000	Rp 352.000.000.000
3	Rp 352.000.000.000	Rp 44.000.000.000	Rp 28.371.200.000	Rp 72.371.200.000	Rp 308.000.000.000
4	Rp 308.000.000.000	Rp 44.000.000.000	Rp 24.824.800.000	Rp 68.824.800.000	Rp 264.000.000.000
5	Rp 264.000.000.000	Rp 44.000.000.000	Rp 21.278.400.000	Rp 65.278.400.000	Rp 220.000.000.000
6	Rp 220.000.000.000	Rp 44.000.000.000	Rp 17.732.000.000	Rp 61.732.000.000	Rp 176.000.000.000
7	Rp 176.000.000.000	Rp 44.000.000.000	Rp 14.185.600.000	Rp 58.185.600.000	Rp 132.000.000.000
8	Rp 132.000.000.000	Rp 44.000.000.000	Rp 10.639.200.000	Rp 54.639.200.000	Rp 88.000.000.000
9	Rp 88.000.000.000	Rp 44.000.000.000	Rp 7.092.800.000	Rp 51.092.800.000	Rp 44.000.000.000
10	Rp 44.000.000.000	Rp 44.000.000.000	Rp 3.546.400.000	Rp 47.546.400.000	Rp -

L5.9. Biaya Bahan Baku dan Penunjang

a. Persediaan bahan baku :

Tahun pertama

Komponen	Kebutuhan (kg/h)	Harga/satuan		Biaya
a. Persediaan bahan baku				
Gliserol	8330,7254	Rp	18.000,00	Rp 1.187.628.212.349
Hidrogen	181,1027	Rp	12.314,15	Rp 17.662.604.712
Raney Nickel	6096,6228	Rp	149.460,00	Rp 911.201.242
Total persediaan bahan baku	a			Rp 1.206.202.018.303

Tahun pertama

Komponen	Kebutuhan (kg/h)	Harga/satuan		Biaya
a. Persediaan bahan baku				
Hidrogen	103590,1200	Rp	12.314,15	Rp 10.102.947.550.640
Total persediaan bahan baku	a			Rp 10.102.947.550.640

Total pembelian bahan baku tahun pertama = Rp 1.206.202.018.303
Total Pembelian + Recycle Rp 11.309.149.568.943

Tahun	Kapasitas Produksi	Biaya Bahan Baku
1	80%	Rp 964.961.614.642,24
2	90%	Rp 1.194.139.998.119,77
3	100%	Rp 1.459.504.442.146,38
4	100%	Rp 1.605.454.886.361,02
5	100%	Rp 1.766.000.374.997,12
6	100%	Rp 1.942.600.412.496,83
7	100%	Rp 2.136.860.453.746,52
8	100%	Rp 2.350.546.499.121,17
9	100%	Rp 2.585.601.149.033,29
10	100%	Rp 2.844.161.263.936,62

Keterangan : Kenaikan biaya bahan baku *10% per tahun*

Komponen	**Kebutuhan**	**Harga/satuan**	**Biaya**
b. Persediaan sarana penunjang			
1. Solar (liter/hari)	437,402	Rp 9.850,00	Rp 1.421.775.135
2. Listrik (kWh)	2721,305	Rp 1.114,74	Rp 24.025.694.945
3. Resin (Liter)	26240,901	Rp 7.500,00	Rp 64.946.230.274
Abonemen/bulan	-	Rp 87.081.754,43	Rp 1.044.981.053
Total persediaan bahan penunjang	b		**Rp 91.438.681.407**

Tahun	Kapasitas Produksi	Solar	Listrik *variabel cost*	Abonemen	Bahan Penunjang
1	80%	Rp 1.137.420.107,74	Rp 19.220.555.956,33	Rp 1.044.981.053,14	Rp 72.105.964.073
2	90%	Rp 1.407.557.383,33	Rp 23.785.437.995,96	Rp 1.149.479.158,45	Rp 88.081.651.382
3	100%	Rp 1.720.347.912,95	Rp 29.071.090.883,95	Rp 1.264.427.074,29	Rp 106.390.924.614
4	100%	Rp 1.892.382.704,25	Rp 31.978.199.972,35	Rp 1.390.869.781,72	Rp 115.639.147.294
5	100%	Rp 2.081.620.974,67	Rp 35.176.019.969,58	Rp 1.529.956.759,89	Rp 125.673.105.264
6	100%	Rp 2.289.783.072,14	Rp 38.693.621.966,54	Rp 1.682.952.435,88	Rp 136.557.463.354
7	100%	Rp 2.518.761.379,36	Rp 42.562.984.163,20	Rp 1.851.247.679,47	Rp 148.361.962.010
8	100%	Rp 2.770.637.517,29	Rp 46.819.282.579,52	Rp 2.036.372.447,42	Rp 161.161.785.764
9	100%	Rp 3.047.701.269,02	Rp 51.501.210.837,47	Rp 2.240.009.692,16	Rp 175.037.954.648
10	100%	Rp 3.352.471.395,92	Rp 56.651.331.921,22	Rp 2.464.010.661,38	Rp 190.077.739.451

Keterangan : Kenaikan biaya penunjang *10% per tahun*

L5.10. Hasil Penjualan Produk

Hasil produksi	=		50.000.000	kg/th
Harga produk	=	Rp	37.000,00	/kg
Hasil penjualan produk per tahun	=	Rp	1.850.000.000.000,00	

Tahun	Kapasitas Produksi	Hasil penjualan produksi (Total Sales)
1	80%	Rp 1.480.000.000.000,00
2	90%	Rp 1.831.500.000.000,00
3	100%	Rp 2.238.500.000.000,00
4	100%	Rp 2.462.350.000.000,00
5	100%	Rp 2.708.585.000.000,00
6	100%	Rp 2.979.443.500.000,00
7	100%	Rp 3.277.387.850.000,00
8	100%	Rp 3.605.126.635.000,00
9	100%	Rp 3.965.639.298.500,00
10	100%	Rp 4.362.203.228.350,00

Keterangan : Terjadi kenaikan harga produk sebesar 10% /tahun

L.5.11. Salvage Value

Salvage value untuk masing-masing barang modal adalah sebagai berikut :

a. Kendaraan (mobil dan forklift)

=		10%	x	Rp	3.393.300.000
=	Rp	339.330.000			

b. DFCI selain kendaraan, bangunan, dan tanah

=		10%	x	Rp	30.554.935.591
=	Rp	3.055.493.559			

c. Bangunan

=		10%	x	Rp	73.331.845.419
=	Rp	7.333.184.542			

Catatan: Tanah tidak didepresiasi. Pada akhir tahun ke-10 harga tanah tetap :	Rp	70.000.000.000,00
Sehingga total nilai salvage value yang akan diperhitungkan pada akhir tahun ke-10 :	Rp	80.728.008.101

L5.12. Depresiasi

- Depresiasi digolongkan pada masing-masing alat sesuai periode depresiasinya
- Metode yang dipakai adalah Metode Garis Lurus
- Periode depresiasi menurut SK Menteri Keuangan No. 96/PMK-03/2009 adalah :

a. 8 tahun atau 12.5% / tahun untuk kendaraan
b. 16 Tahun atau 6.25% /tahun untuk mesin-mesin industri kimia
c. 20 tahun atau 5% / tahun untuk bangunan
d. 8 tahun atau 12.5% / tahun untuk IFCI tanpa salvage value (amortisasi)

rumus

$$= \frac{HP - NS}{n}$$

HP = Harga perolehan
NS = Nilai sisa
n = taksir umur kegunaan

Tahun	Kendaraan	DFCI tanpa tanah, bangunan &kendaraan	Bangunan	Nilai depresiasi IFCI	Jumlah Nilai Depresiasi
1	Rp 381.746.250	1.718.715.127	3.299.933.044	38.846.541.497	44.246.935.918
2	Rp 381.746.250	1.718.715.127	3.299.933.044	38.846.541.497	44.246.935.918
3	Rp 381.746.250	1.718.715.127	3.299.933.044	38.846.541.497	44.246.935.918
4	Rp 381.746.250	1.718.715.127	3.299.933.044	38.846.541.497	44.246.935.918
5	Rp 381.746.250	1.718.715.127	3.299.933.044	38.846.541.497	44.246.935.918
6	Rp 381.746.250	1.718.715.127	3.299.933.044	38.846.541.497	44.246.935.918
7	Rp 381.746.250	1.718.715.127	3.299.933.044	38.846.541.497	44.246.935.918
8	Rp 381.746.250	1.718.715.127	3.299.933.044	38.846.541.497	44.246.935.918
9	-	1.718.715.127	3.299.933.044	-	5.018.648.171
10	-	1.718.715.127	3.299.933.044	-	5.018.648.171
Total					Rp 364.012.783.685,56

11	-	Rp 1.718.715.127,00	Rp 3.299.933.043,84	-	Rp 5.018.648.170,85
12	-	Rp 1.718.715.127,00	Rp 3.299.933.043,84	-	Rp 5.018.648.170,85
13	-	Rp 1.718.715.127,00	Rp 3.299.933.043,84	-	Rp 5.018.648.170,85
14	-	Rp 1.718.715.127,00	Rp 3.299.933.043,84	-	Rp 5.018.648.170,85
15	-	Rp 1.718.715.127,00	Rp 3.299.933.043,84	-	Rp 5.018.648.170,85
16	-	Rp 1.718.715.127,00	Rp 3.299.933.043,84	-	Rp 5.018.648.170,85
17	-	-	Rp 3.299.933.043,84	-	Rp 3.299.933.043,84
18	-	-	Rp 3.299.933.043,84	-	Rp 3.299.933.043,84
19	-	-	Rp 3.299.933.043,84	-	Rp 3.299.933.043,84
20	-	-	Rp 3.299.933.043,84	-	Rp 3.299.933.043,84
Total					Rp 407.324.404.886,01

L5.13. Perhitungan Biaya Produksi Total (total production cost)

Biaya Produksi adalah biaya yang harus dikeluarkan untuk mengolah bahanbaku menjadi produk jadi yang siap untuk dijual. Biaya produksi dihitung pertahun selama 10 tahun kedepan dengan jumlah 330 hari dan sesuai dengan kapasitas produksi.

TAHUN				I		II	
KAPASITAS PRODUKSI				80%		90%	
BIAYA PRODUKSI (PRODUCT COST)				*Fixed Cost*	*Variable Cost*	*Fixed Cost*	*Variable Cost*
A.	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
a.	Biaya Bahan Baku			-	Rp 964.961.614.642	-	Rp 1.194.139.998.120
b.	Gaji Karyawan			Rp 12.855.832.000		Rp 14.141.415.200	
c.	Biaya Pemeliharaan dan Perbaikan (kenaikan 5% per tahun)	2%	DFCI	Rp 11.757.905.209	-	Rp 11.757.905.209	-
d.	Biaya Royalti dan Paten	0,5%	TS	-	Rp 7.400.000.000	-	Rp 9.157.500.000

TAHUN				I		II	
KAPASITAS PRODUKSI				80%		90%	
BIAYA PRODUKSI (PRODUCT COST)				*Fixed Cost*	*Variable Cost*	*Fixed Cost*	*Variable Cost*
A.	**Biaya Manufacturing (Manufacturing Cost)**						
e.	Biaya Laboratorium	0,5%	BB	-	Rp 4.824.808.073	-	Rp 5.970.699.991
f.	Biaya pengemasan produk	0,5%	BB	-	Rp 4.824.808.073	-	Rp 5.970.699.991
g.	Biaya sarana penunjang			Rp 1.044.981.053	Rp 72.105.964.073	Rp 1.149.479.158	Rp 88.081.651.382
h.	Biaya Start Up			Rp 58.594.516.661	-	Rp 64.453.968.327	-
	Total Biaya Manufacturing Langsung (DMC)			**Rp 84.253.234.923**	**Rp 1.054.117.194.862**	**Rp 91.502.767.894**	**Rp 1.303.320.549.483**
	Biaya Plant Overhead	20%	(b+c)	**Rp 4.922.747.442**	-	**Rp 5.179.864.082**	-
2	**Biaya Manufacturing Tetap (FMC)**						
a.	Depresiasi			Rp 44.246.935.918	-	Rp 44.246.935.918	-
b.	Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th	0,1%		Rp 143.331.845	-	Rp 157.665.030	-
c.	Biaya asuransi (kenaikan 10 %) pertahun	0,5%	DFCI	Rp 2.939.476.302	-	Rp 3.233.423.932	-
	Total Biaya Manufacturing Tetap (FMC)			**Rp 47.329.744.066**	-	**Rp 47.638.024.880**	-
B.	**Pengeluaran Umum (General Expenses)**						
1	Biaya administrasi	5%	b	Rp 642.791.600	-	Rp 707.070.760	-
2	Biaya distribusi dan penjualan	10%	f	-	Rp 482.480.807	-	Rp 597.069.999
3	Bunga Bank + Cicilan Pokok			Rp 79.464.000.000	-	Rp 75.917.600.000	-
	Total Pengeluaran Umum			**Rp 80.106.791.600**	**Rp 482.480.807**	**Rp 76.624.670.760**	**Rp 597.069.999**
Total Biaya				**Rp 216.612.518.030**	**Rp 1.054.599.675.669**	**Rp 220.945.327.616**	**Rp 1.303.917.619.482**
Total Biaya Produksi (TPC)				**Rp 1.271.212.193.699**		**Rp 1.524.862.947.098**	

TAHUN			III		IV	
KAPASITAS PRODUKSI			**100%**		**100%**	
BIAYA PRODUKSI (PRODUCT COST)			***Fixed Cost***	***Variable Cost***	***Fixed Cost***	***Variable Cost***
Biaya Manufacturing (Manufacturing Cost)						
Biaya Manufacturing Langsung (DMC)						
Biaya Bahan Baku			-	Rp 1.459.504.442.146,38	-	Rp 1.605.454.886.36
Gaji Karyawan			Rp 15.555.556.720,00	-	Rp 17.111.112.392,00	-
Biaya Pemeliharaan dan Perbaikan (	2%	DFCI	Rp 11.757.905.208,53	-	Rp 11.757.905.208,53	-
Biaya Royalti dan Paten	0,5%	TS	-	Rp 11.192.500.000,00	-	Rp 12.311.750.00
Biaya Laboratorium	0,5%	BB	-	Rp 7.297.522.210,73	-	Rp 8.027.274.43
Biaya pengemasan produk	0,5%	BB	-	Rp 7.297.522.210,73	-	Rp 8.027.274.43
Biaya sarana penunjang			Rp 1.264.427.074,29	Rp 106.390.924.614,47	Rp 1.390.869.781,72	Rp 115.639.147.29
Biaya Start Up			Rp 70.899.365.160,11	-	Rp 77.989.301.676,12	-
Total Biaya Manufacturing Langsung (DMC)			**Rp 99.477.254.162,94**	**Rp 1.591.682.911.182,31**	**Rp 108.249.189.058,38**	**Rp 1.749.460.332.518,**
Biaya Plant Overhead	20%	(b+c)	**Rp 5.462.692.385,71**	-	**Rp 5.773.803.520,11**	-
Biaya Manufacturing Tetap (FMC)						
Depresiasi			Rp 44.246.935.917,98	-	Rp 44.246.935.917,98	-
Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th			Rp 173.431.532,96	-	Rp 190.774.686,25	-
Biaya asuransi (kenaikan 10 %) perta	0,5%	DFCI	Rp 3.556.766.325,58	-	Rp 3.912.442.958,14	-
Total Biaya Manufacturing Tetap (FMC)			**Rp 47.977.133.776,52**	-	**Rp 48.350.153.562,37**	-
Pengeluaran Umum (General Expenses)						
Biaya administrasi	5%	b	Rp 777.777.836,00	-	Rp 855.555.619,60	-
Biaya distribusi dan penjualan	10%	f	-	Rp 729.752.221,07	-	Rp 802.727.44

3	Bunga Bank + Cicilan Pokok			Rp 72.371.200.000,00	-	Rp 68.824.800.000,00	-
	Total Pengeluaran Umum			**Rp 73.148.977.836,00**	**Rp 729.752.221,07**	**Rp 69.680.355.619,60**	**Rp 802.727.443,18**
Total Biaya				**Rp 226.066.058.161,17**	**Rp 1.592.412.663.403,38**	**Rp 232.053.501.760,46**	**Rp 1.750.263.059.962,00**
Total Biaya Produksi (TPC)				**Rp 1.818.478.721.564,55**		**Rp 1.982.316.561.722,46**	

TAHUN				**V**		**VI**	
KAPASITAS PRODUKSI				**100%**		**100%**	
BIAYA PRODUKSI (PRODUCT COST)				***Fixed Cost***	***Variable Cost***	***Fixed Cost***	***Variable Cost***
A.	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
a.	Biaya Bahan Baku			-	Rp 1.766.000.374.997	-	Rp 1.942.600.412.497
b.	Gaji Karyawan			Rp 18.822.223.631	-	Rp 20.704.445.994	-
c.	Biaya Pemeliharaan dan Perbaikan (	2%	DFCI	Rp 11.757.905.209	-	Rp 11.757.905.209	-
d.	Biaya Royalti dan Paten	0,5%	TS	-	Rp 13.542.925.000	-	Rp 14.897.217.500
e.	Biaya Laboratorium	0,5%	BB	-	Rp 8.830.001.875	-	Rp 9.713.002.062
f.	Biaya pengemasan produk	0,5%	BB	-	Rp 8.830.001.875	-	Rp 9.713.002.062
g.	Biaya sarana penunjang			Rp 1.529.956.760	Rp 125.673.105.264	Rp 1.682.952.436	Rp 136.557.463.354
h.	Biaya Start Up			Rp 85.788.231.844	-	Rp 94.367.055.028	-
	Total Biaya Manufacturing Langsung (DMC)			**Rp 117.898.317.443**	**Rp 1.922.876.409.011**	**Rp 128.512.358.667**	**Rp 2.113.481.097.476**
	Biaya Plant Overhead	20%	(b+c)	**Rp 6.116.025.768**	**-**	**Rp 6.492.470.241**	**-**
2	**Biaya Manufacturing Tetap (FMC)**						
a.	Depresiasi			Rp 44.246.935.918	-	Rp 44.246.935.918	-
b.	Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th			Rp 209.852.155	-	Rp 230.837.370	-
c.	Biaya asuransi (kenaikan 10 %) perta	0,5%	DFCI	Rp 4.303.687.254	-	Rp 4.734.055.979	-

	Total Biaya Manufacturing Tetap (FMC)			**Rp 47.977.133.776,52**	-	**Rp 48.350.153.562,37**	-
B.	**Pengeluaran Umum (General Expenses)**						
1	Biaya administrasi	5%	b	Rp 777.777.836,00	-	Rp 855.555.619,60	-
2	Biaya distribusi dan penjualan	10%	f	-	Rp 729.752.221,07	-	Rp 802.727.443,18
3	Bunga Bank + Cicilan Pokok			Rp 72.371.200.000,00	-	Rp 68.824.800.000,00	-
	Total Pengeluaran Umum			**Rp 73.148.977.836,00**	**Rp 729.752.221,07**	**Rp 69.680.355.619,60**	**Rp 802.727.443,18**
	Total Biaya			**Rp 226.066.058.161,17**	**Rp 1.592.412.663.403,38**	**Rp 232.053.501.760,46**	**Rp 1.750.263.059.962,00**
	Total Biaya Produksi (TPC)			**Rp 1.818.478.721.564,55**		**Rp 1.982.316.561.722,46**	

TAHUN				**V**		**VI**	
KAPASITAS PRODUKSI				**100%**		**100%**	
BIAYA PRODUKSI (PRODUCT COST)				***Fixed Cost***	***Variable Cost***	***Fixed Cost***	***Variable Cost***
A.	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
a.	Biaya Bahan Baku			-	Rp 1.766.000.374.997	-	Rp 1.942.600.412.497
b.	Gaji Karyawan			Rp 18.822.223.631	-	Rp 20.704.445.994	-
c.	Biaya Pemeliharaan dan Perbaikan (	2%	DFCI	Rp 11.757.905.209	-	Rp 11.757.905.209	-
d.	Biaya Royalti dan Paten	0,5%	TS	-	Rp 13.542.925.000	-	Rp 14.897.217.500
e.	Biaya Laboratorium	0,5%	BB	-	Rp 8.830.001.875	-	Rp 9.713.002.062
f.	Biaya pengemasan produk	0,5%	BB	-	Rp 8.830.001.875	-	Rp 9.713.002.062
g.	Biaya sarana penunjang			Rp 1.529.956.760	Rp 125.673.105.264	Rp 1.682.952.436	Rp 136.557.463.354
h.	Biaya Start Up			Rp 85.788.231.844	-	Rp 94.367.055.028	-
	Total Biaya Manufacturing Langsung (DMC)			**Rp 117.898.317.443**	**Rp 1.922.876.409.011**	**Rp 128.512.358.667**	**Rp 2.113.481.097.476**
	Biaya Plant Overhead	20%	(b+c)	**Rp 6.116.025.768**	-	**Rp 6.492.470.241**	-

2	**Biaya Manufacturing Tetap (FMC)**						
a.	Depresiasi			Rp 44.246.935.918	-	Rp 44.246.935.918	-
b.	Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th			Rp 209.852.155	-	Rp 230.837.370	-
c.	Biaya asuransi (kenaikan 10 %) perta	0,5%	DFCI	Rp 4.303.687.254	-	Rp 4.734.055.979	-
	Total Biaya Manufacturing Tetap (FMC)			**Rp 48.760.475.327**	-	**Rp 49.211.829.268**	-
B.	**Pengeluaran Umum (General Expenses)**						
1	Biaya administrasi	5%	b	Rp 941.111.182	-	Rp 1.035.222.300	-
2	Biaya distribusi dan penjualan	10%	f	-	Rp 883.000.187	-	Rp 971.300.206
3	Bunga Bank + Cicilan Pokok			Rp 65.278.400.000	-	Rp 61.732.000.000	-
	Total Pengeluaran Umum			**Rp 66.219.511.182**	**Rp 883.000.187**	**Rp 62.767.222.300**	**Rp 971.300.206**
Total Biaya				**Rp 238.994.329.720**	**Rp 1.923.759.409.198**	**Rp 246.983.880.475**	**Rp 2.114.452.397.682**
Total Biaya Produksi (TPC)				**Rp 2.162.753.738.918**		**Rp 2.361.436.278.157**	

TAHUN				**VII**		**VIII**	
KAPASITAS PRODUKSI				**100%**		**100%**	
BIAYA PRODUKSI (PRODUCT COST)				***Fixed Cost***	***Variable Cost***	***Fixed Cost***	***Variable Cost***
A.	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
a.	Biaya Bahan Baku			-	Rp 2.136.860.453.747	-	Rp 2.350.546.499.121
b.	Gaji Karyawan			Rp 22.774.890.594	-	Rp 25.052.379.653	-
c.	Biaya Pemeliharaan dan Perbaikan (	2%	DFCI	Rp 11.757.905.209	-	Rp 11.757.905.209	-
d.	Biaya Royalti dan Paten	0,5%	TS	-	Rp 16.386.939.250	-	Rp 18.025.633.175

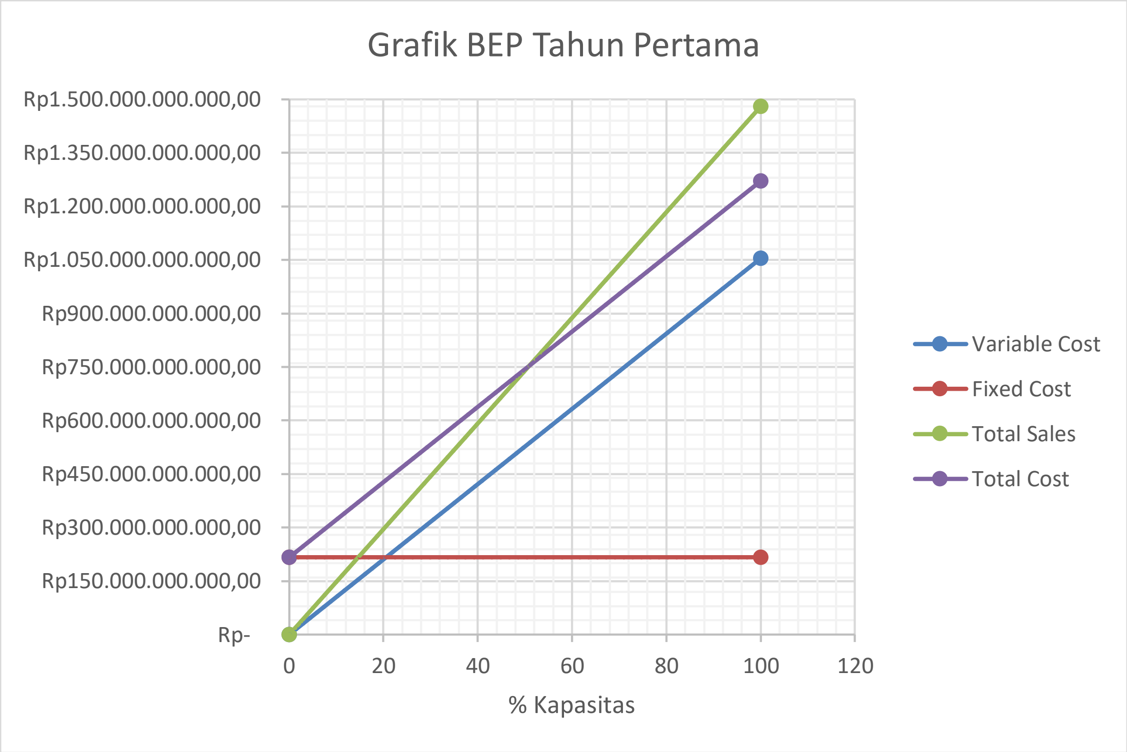
e.	Biaya Laboratorium	0,5%	BB	-	Rp 10.684.302.269	-	Rp 11.752.732.496
f.	Biaya pengemasan produk	0,5%	BB	-	Rp 10.684.302.269	-	Rp 11.752.732.496
g.	Biaya sarana penunjang			Rp 1.851.247.679	Rp 148.361.962.010	Rp 2.036.372.447	Rp 161.161.785.764
h.	Biaya Start Up			Rp 103.803.760.531	-	Rp 114.184.136.584	-
	Total Biaya Manufacturing Langsung (DMC)			**Rp 140.187.804.013**	**Rp 2.322.977.959.544**	**Rp 153.030.793.893**	**Rp 2.553.239.383.051**
	Biaya Plant Overhead	20%	(b+c)	**Rp 6.906.559.160**	-	**Rp 7.362.056.972**	-
2	**Biaya Manufacturing Tetap (FMC)**						
a.	Depresiasi			Rp 44.246.935.918	-	Rp 44.246.935.918	-
b.	Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th			Rp 253.921.107	-	Rp 279.313.218	-
c.	Biaya asuransi (kenaikan 10 %) perta	0,5%	DFCI	Rp 5.207.461.577	-	Rp 5.728.207.735	-
	Total Biaya Manufacturing Tetap (FMC)			**Rp 49.708.318.603**	-	**Rp 50.254.456.871**	-
B.	**Pengeluaran Umum (General Expenses)**						
1	Biaya administrasi	5%	b	Rp 1.138.744.530	-	Rp 1.252.618.983	-
2	Biaya distribusi dan penjualan	10%	f	-	Rp 1.068.430.227	-	Rp 1.175.273.250
3	Bunga Bank + Cicilan Pokok			Rp 58.185.600.000	-	Rp 58.185.600.000	-
	Total Pengeluaran Umum			**Rp 59.324.344.530**	**Rp 1.068.430.227**	**Rp 59.438.218.983**	**Rp 1.175.273.250**
	Total Biaya			**Rp 256.127.026.305**	**Rp 2.324.046.389.771**	**Rp 270.085.526.719**	**Rp 2.554.414.656.301**
	Total Biaya Produksi (TPC)			**Rp 2.580.173.416.076**		**Rp 2.824.500.183.020**	

TAHUN				IX		X	
KAPASITAS PRODUKSI				100%		100%	
BIAYA PRODUKSI (PRODUCT COST)				*Fixed Cost*	*Variable Cost*	*Fixed Cost*	*Variable Cost*
A.	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
a.	Biaya Bahan Baku			-	Rp 2.585.601.149.033	-	Rp 2.844.161.263.937
b.	Gaji Karyawan			Rp 27.557.617.618	-	Rp 30.313.379.380	-
c.	Biaya Pemeliharaan dan Perbaikan (	2%	DFCI	Rp 11.757.905.209	-	Rp 11.757.905.209	-
d.	Biaya Royalti dan Paten	0,5%	TS	-	Rp 19.828.196.493	-	Rp 21.811.016.142
e.	Biaya Laboratorium	0,5%	BB	-	Rp 12.928.005.745	-	Rp 14.220.806.320
f.	Biaya pengemasan produk	0,5%	BB	-	Rp 12.928.005.745	-	Rp 14.220.806.320
g.	Biaya sarana penunjang			Rp 2.240.009.692	Rp 175.037.954.648	Rp 2.464.010.661	Rp 190.077.739.451
h.	Biaya Start Up			Rp 125.602.550.242	-	Rp 138.162.805.267	-
	Total Biaya Manufacturing Langsung (DMC)			**Rp 167.158.082.762**	**Rp 2.806.323.311.664**	**Rp 183.873.891.038**	**Rp 3.084.491.632.169**
	Biaya Plant Overhead	20%	(b+c)	**Rp 7.863.104.565**	-	**Rp 8.414.256.918**	-
2	**Biaya Manufacturing Tetap (FMC)**						
a.	Depresiasi			Rp 5.018.648.171	-	Rp 5.018.648.171	-
b.	Pajak Bumi dan Bangunan diperkirakan 0,1% x (tanah +			Rp 307.244.540	-	Rp 337.968.994	-
c.	Biaya asuransi (kenaikan 10 %) perta	0,5%	DFCI	Rp 6.301.028.509	-	Rp 6.931.131.359	-
	Total Biaya Manufacturing Tetap (FMC)			**Rp 11.626.921.219**	-	**Rp 12.287.748.524**	-
B.	**Pengeluaran Umum (General Expenses)**						
1	Biaya administrasi	5%	b	Rp 1.377.880.881	-	Rp 1.515.668.969	-
2	Biaya distribusi dan penjualan	10%	f	-	Rp 1.292.800.575	-	Rp 1.422.080.632
3	Bunga Bank + Cicilan Pokok			Rp 51.092.800.000	-	Rp 47.546.400.000	-
	Total Pengeluaran Umum			**Rp 52.470.680.881**	**Rp 1.292.800.575**	**Rp 49.062.068.969**	**Rp 1.422.080.632**
Total Biaya				**Rp 239.118.789.427**	**Rp 2.807.616.112.239**	**Rp 253.637.965.449**	**Rp 3.085.913.712.801**
Total Biaya Produksi (TPC)				**Rp 3.046.734.901.666**		**Rp 3.339.551.678.250**	

L5.14. Break Even Point

$$\text{Rumus Umum} = \frac{\text{Total } \textit{Fixed Cost}}{\text{Total } \textit{Sales} - \text{Total } \textit{Variabel Cost}} \times 100\ \%$$

Tahun	Hasil Penjualan Produksi (*Total Sales*)		Total *Fixed Cost*		Total *Variabel Cost*		Pengeluaran (Total Cost)		BEP (%)
1	Rp	1.480.000.000.000	Rp	216.612.518.030	Rp	1.054.599.675.669	Rp	1.271.212.193.699	50,92%
2	Rp	1.831.500.000.000	Rp	220.945.327.616	Rp	1.303.917.619.482	Rp	1.524.862.947.098	41,88%
3	Rp	2.238.500.000.000	Rp	226.066.058.161	Rp	1.592.412.663.403	Rp	1.818.478.721.565	34,99%
4	Rp	2.462.350.000.000	Rp	232.053.501.760	Rp	1.750.263.059.962	Rp	1.982.316.561.722	32,59%
5	Rp	2.708.585.000.000	Rp	238.994.329.720	Rp	1.923.759.409.198	Rp	2.162.753.738.918	30,45%
6	Rp	2.979.443.500.000	Rp	246.983.880.475	Rp	2.114.452.397.682	Rp	2.361.436.278.157	28,55%
7	Rp	3.277.387.850.000	Rp	256.127.026.305	Rp	2.324.046.389.771	Rp	2.580.173.416.076	26,87%
8	Rp	3.605.126.635.000	Rp	270.085.526.719	Rp	2.554.414.656.301	Rp	2.824.500.183.020	25,71%
9	Rp	3.965.639.298.500	Rp	239.118.789.427	Rp	2.807.616.112.239	Rp	3.046.734.901.666	20,65%
10	Rp	4.362.203.228.350	Rp	253.637.965.449	Rp	3.085.913.712.801	Rp	3.339.551.678.250	19,87%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Rp900.000.000.000,00
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Rp600.000.000.000,00
Rp450.000.000.000,00
Rp300.000.000.000,00
Rp150.000.000.000,00
Rp-
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100
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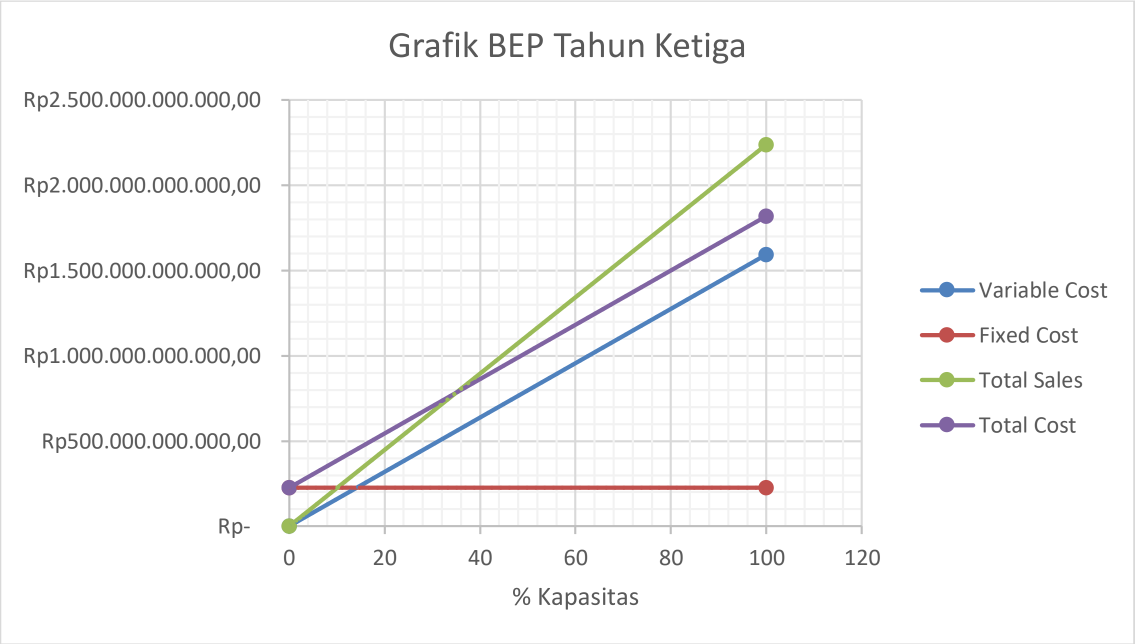
% Kapasitas
Variable Cost
Fixed Cost
Total Sales
Total Cost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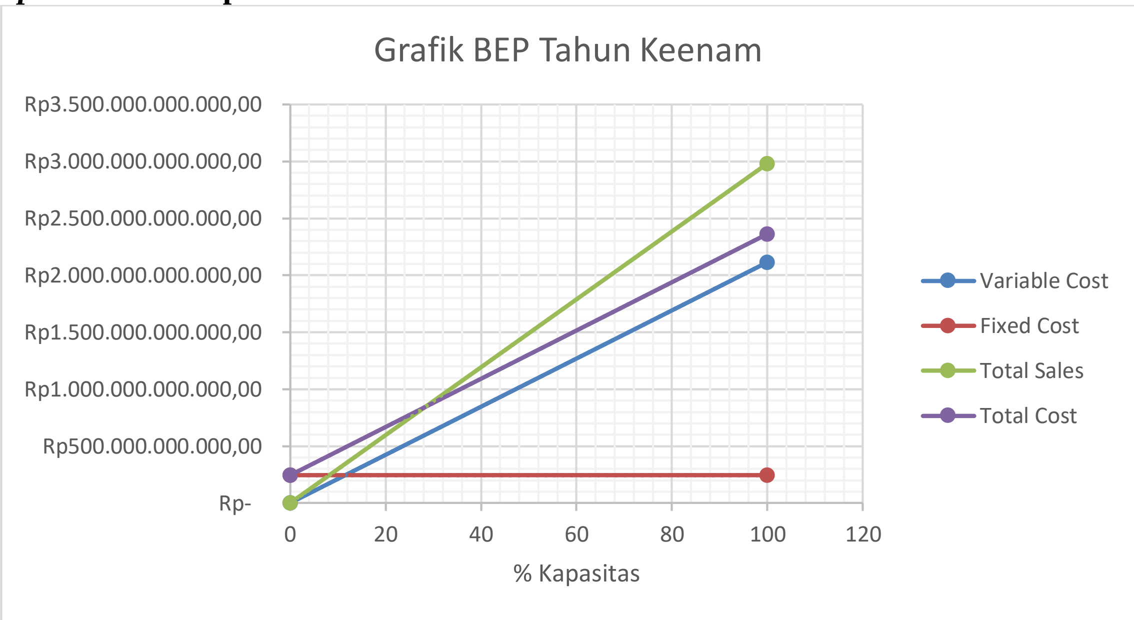
100
120
% Kapasitas
Variable Cost
Fixed Cost
Total Sales
Total Cost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Rp2.500.000.000.000,00
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Rp500.000.000.000,00
Rp-
0
20
40
60
80
100
120
% Kapasitas
Variable Cost
Fixed Cost
Total Sales
Total Cost

L5.15. Laba Rugi dan Pajak

Berdasarkan UU No.36 tahun 2008 sebagai berikut:
Wajib pajak badan usaha dalam negeri adalah 25%

Tahun	Penjualan	Biaya Produksi	Keuntungan kotor	PPH (25%)	Laba Setelah Pajak	Depresiasi	*Salvage Value*	*Cash in Nominal*
1	Rp 1.480.000.000.000	Rp 1.271.212.193.699	Rp 208.787.806.301	Rp 52.196.951.575	Rp 156.590.854.726	Rp 44.246.935.918	Rp -	Rp 200.837.790.644
2	Rp 1.831.500.000.000	Rp 1.524.862.947.098	Rp 306.637.052.902	Rp 76.659.263.225	Rp 229.977.789.676	Rp 44.246.935.918	Rp -	Rp 274.224.725.594
3	Rp 2.238.500.000.000	Rp 1.818.478.721.565	Rp 420.021.278.435	Rp 105.005.319.609	Rp 315.015.958.827	Rp 44.246.935.918	Rp -	Rp 359.262.894.745
4	Rp 2.462.350.000.000	Rp 1.982.316.561.722	Rp 480.033.438.278	Rp 120.008.359.569	Rp 360.025.078.708	Rp 44.246.935.918	Rp -	Rp 404.272.014.626
5	Rp 2.708.585.000.000	Rp 2.162.753.738.918	Rp 545.831.261.082	Rp 136.457.815.271	Rp 409.373.445.812	Rp 44.246.935.918	Rp -	Rp 453.620.381.729
6	Rp 2.979.443.500.000	Rp 2.361.436.278.157	Rp 618.007.221.843	Rp 154.501.805.461	Rp 463.505.416.382	Rp 44.246.935.918	Rp -	Rp 507.752.352.300
7	Rp 3.277.387.850.000	Rp 2.580.173.416.076	Rp 697.214.433.924	Rp 174.303.608.481	Rp 522.910.825.443	Rp 44.246.935.918	Rp -	Rp 567.157.761.361
8	Rp 3.605.126.635.000	Rp 2.824.500.183.020	Rp 780.626.451.980	Rp 195.156.612.995	Rp 585.469.838.985	Rp 44.246.935.918	Rp 339.330.000	Rp 630.056.104.903
9	Rp 3.965.639.298.500	Rp 3.046.734.901.666	Rp 918.904.396.834	Rp 229.726.099.209	Rp 689.178.297.626	Rp 5.018.648.171	Rp -	Rp 694.196.945.797
10	Rp 4.362.203.228.350	Rp 3.339.551.678.250	Rp 1.022.651.550.100	Rp 255.662.887.525	Rp 766.988.662.575	Rp 5.018.648.171	Rp 70.000.000.000	Rp 842.007.310.746

L5.16. Minimum Payback Period (MPP)

Tahun	NCF Nominal	Suku Bunga	*Disc.Factor*	NCF PV	Akumulasi
0	-Rp 1.374.994.027.294	8,06%	1,00	-Rp 1.374.994.027.294	-Rp 1.374.994.027.294
1	Rp 200.837.790.644	8,06%	0,93	Rp 185.857.663.005	-Rp 1.189.136.364.289
2	Rp 274.224.725.594	8,06%	0,86	Rp 234.842.494.152	-Rp 954.293.870.137
3	Rp 359.262.894.745	8,06%	0,79	Rp 284.719.672.385	-Rp 669.574.197.752
4	Rp 404.272.014.626	8,06%	0,73	Rp 296.492.577.742	-Rp 373.081.620.010
5	Rp 453.620.381.729	8,06%	0,68	Rp 307.870.263.715	-Rp 65.211.356.295
6	Rp 507.752.352.300	8,06%	0,63	Rp 318.905.614.175	Rp 253.694.257.880
7	Rp 567.157.761.361	8,06%	0,58	Rp 329.647.007.188	Rp 583.341.265.069
8	Rp 630.056.104.903	8,06%	0,54	Rp 338.890.596.600	Rp 922.231.861.668
9	Rp 694.196.945.797	8,06%	0,50	Rp 345.539.761.999	Rp 1.267.771.623.667
10	Rp 842.007.310.746	8,06%	0,46	Rp 387.852.174.133	Rp 1.655.623.797.800
Total				**Rp 1.655.623.797.800**	**LAYAK**

$$MPP = n + \frac{(a-b)}{(c-b)\ x\ 1\ tahun}$$

MPP = 5 tahun 3 bulan 4 hari

L5.17. Internal Rate of Return

Tahun	Net Cash Flow		Bunga $1/(1+i)^n$	Present Value	
0	-Rp	1.374.994.027.294	1,00	-Rp	1.374.994.027.294
1	Rp	200.837.790.644	0,80	Rp	160.653.656.211
2	Rp	274.224.725.594	0,64	Rp	175.467.612.885
3	Rp	359.262.894.745	0,51	Rp	183.885.676.058
4	Rp	404.272.014.626	0,41	Rp	165.521.492.340
5	Rp	453.620.381.729	0,33	Rp	148.565.665.563
6	Rp	507.752.352.300	0,26	Rp	133.021.859.926
7	Rp	567.157.761.361	0,21	Rp	118.855.731.728
8	Rp	630.056.104.903	0,17	Rp	105.618.659.969
9	Rp	694.196.945.797	0,13	Rp	93.087.058.348
10	Rp	842.007.310.746	0,11	Rp	90.316.614.267
Total				Rp	-

IRR = 25,01 %

L5.18. Kelayakan Proyek

Tahun	Net Cash Flow Nominal		Bunga 8,06%	Net Cash Flow Present Value	
			$1/(1+i)^n$		
0	-Rp	1.374.994.027.294	1,00	-Rp	1.374.994.027.294
1	Rp	200.837.790.644	0,93	Rp	185.857.663.005
2	Rp	274.224.725.594	0,86	Rp	234.842.494.152
3	Rp	359.262.894.745	0,79	Rp	284.719.672.385
4	Rp	404.272.014.626	0,73	Rp	296.492.577.742
5	Rp	453.620.381.729	0,68	Rp	307.870.263.715
6	Rp	507.752.352.300	0,63	Rp	318.905.614.175
7	Rp	567.157.761.361	0,58	Rp	329.647.007.188
8	Rp	630.056.104.903	0,54	Rp	338.890.596.600
9	Rp	694.196.945.797	0,50	Rp	345.539.761.999
10	Rp	842.007.310.746	0,46	Rp	387.852.174.133
Total				Rp	1.655.623.797.800

Parameter Analisis	Nilai
NCFPV di tahun ke-10	Rp 1.655.623.797.800
IRR	25,01%
MPP	5 tahun 3 bulan 4 hari

Berdasarkan Analisa parameter-parameter tersebut, maka pra perancangan pabrik propilen glikol dengan kapasitas 50.000 ton/tahun ini feasible (layak).