BAB XI
KESIMPULAN

11.1 Kesimpulan

Berdasarkan uraian dan hasil perhitungan dari bab–bab sebelumnya pada prarancangan pabrik Propilen glikol dengan kapasitas 70.000 ton/tahun dapat disimpulkan sebagai berikut :

1. Prarancangan pabrik Propilen glikol dari Gliserol dan Hidrogen dengan kapasitas 70.000 ton/tahun direncanakan untuk memenuhi kebutuhan dalam negeri dan sebagiannya di ekspor ke luar negeri.
2. Dari analisisis teknis dan ekonomi yang dilakukan, maka pabrik Propilen glikol dari Gliserol dan Hidrogen dengan kapasitas 70.000 ton/tahun layak didirikan di Pelintung,Kec.Medang Kampai,Kota Dumai Riau.
3. Prarancangan Propilen glikol dari Gliserol dan Hidrogen merupakan perusahaan berbentuk Perseroan Terbatas (PT) dengan struktur organisasi *line and staff* dengan jumlah tenaga kerja 176 orang yang terdiri dari 131 karyawan shift dan 45 orang karyawan non shift.
4. Dari perhitungan analisa ekonomi, maka Prarancangan pabrik Propilen glikol dari Gliserol dan Hidrogen ini layak didirikan dengan :
   - *Fixed Capital Investment (FCI)* = US$ 37.142.918
     = Rp 605.706.285.322
   - *Working Capital Investment (WCI)* = US$ 6.554.633
     = Rp 106.889.344.469
   - *Total Capital Investment (TCI)* = US$ 43.697.551
     = Rp 712.595.629.790
   - *Total Production Cost (TPC)* = US$ 111.826.222
     = Rp 1.823.600.528.530
   - *Total Sales (TS)* = US$ 124.482.600
     = Rp 2.030.000.000.000
   - *Rate of Return (ROR)* = 54,59%.
   - *Pay Out Time (POT)* = 2 tahun 2 bulan
   - *Break Event Point (BEP)* = 36,0 %

11.2 Saran

Berdasarkan pertimbangan dari analisa ekonomi yang telah dilakukan Pabrik Propilen glikol dari gliserol dan Hidrogen ini layak untuk dilanjutkan ke tahap rancangan pabrik. Untuk itu disarankan kepada pengurus dan pemilik modal untuk dapat mempertimbangkan dan mengkaji ulang tentang rancangan pabrik Propilen glikol ini.

DAFTARPUSTAKA

Aditama Muhammad, 2024, Pabrik Kertas-Pulp, Diakses Tanggal 16 Februari 2014, Asosiasi Pulp dan Kertas Indonesia, 2013.

Bassel, William D., 1976, Preliminary Chemical Engineering Plant Design, New York, Elsevier.

Badan Pusat Statistik Provinsi Sumatera Barat, 2012.

Bagaskara, I., 2010, Pengembangan Industri Kertas, Jakarta, Bumi Aksara.

Biermann, Christopher J., 1996, Handbook of Pulping and Papermaking, 2nd Edition, California, Academic Press.

Brown, Royce N., 1997, Compressors : Selection and Sizing, 2nd Edition, United States of America, Butterworth-Heinemann.

Hoval, 2005, Industrial Steam Boiler, Nottinghamshire, Hoval Limited.

Irawadi, T.T., 1991, Produksi Enzim Ekstraseluler (Selulase dan Xilanase) dari Neurospora sp pada Substrat Limbah Padat Kelapa Sawit, Disertasi, Bogor, Institut Pertanian Bogor. ilahi, R. (2024). Pengantar Teknologi Pembuatan Kertas. Jakarta: Penerbit Teknologi Karya. Kern, Donald Q., 1965, Process Heat Transfer, Singapore : McGraw-Hill Book Company Inc.

Kementerian Perhubungan RI. (2023). Laporan tahunan Kementerian Perhubungan Republik Indonesia 2023. Kementerian Perhubungan RI. Kementerian Perindustrian RI. (2023).

Outlook industri kimia dasar Indonesia 2023.

Kementerian Perindustrian RI. Loo, Sjaak van dan Jaap Koppejan, 2008, The Handbook of Biomass Combustion and Co-firing, United Kingdom, Earthscan

LAMPIRAN A
NERACA MASSA

Kapasitas Produksi = 70.000 ton/tahun

$$= 70.000 \frac{\cancel{ton}}{\cancel{tahun}} \times \frac{1\ \cancel{tahun}}{330\ \cancel{hari}} \times \frac{1\ \cancel{hari}}{24\ jam} \times \frac{1000\ kg}{1\ \cancel{ton}}$$

= 7.000.000 kg/jam

Waktu Operasi = 330 hari

Perbandingan Bahan Baku

- Gliserol = 1
- Hidrogen = 5

(Fundamentals and Applications of Chemical Engineering,2020)

Basis Perhitungan = 1000 kg/jam

Kapasitas Produksi = 70000 ton/tahun = 8838,3838 kg/jam

Kapasitas Produksi Basis = 754,0300 kg/jam

Faktor Pengali $= \frac{\text{Kapasitas Produksi Sebenarnya}}{\text{kapasitas Produksi Basis}}$

$$= \frac{8838{,}3838\ kg/jam}{754{,}0300\ kg/jam}$$

= 11,7268

Maka untuk memproduksi Propilen glikol 70.000 ton/tahun dibutuhkan bahan baku :

Jumlah Gliserol pada saat *Start up* = Basis Perhitungan x Faktor Pengali

= 1000 kg/jam x 11,7215

= 11721,5281 kg/jam

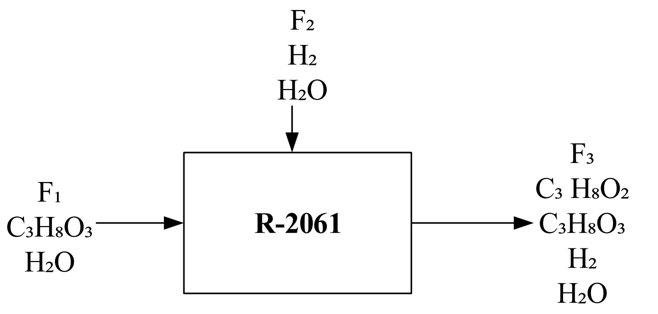
Jumlah Hidrogen pada saat *Start up* = Basis Perhitungan x Faktor Pengali

= 2000kg/jam x 11,7251

= 23443,0562 kg/jam

1. Reaktor Fixed Bed Multitube (R-2061)

Fungsi : Untuk mereaksikan $C_3H_8O_3$



Kondisi Operasi :

Temperatur	: 150°C
Tekanan	: 1 atm
Waktu	: 1 Jam
Konversi	: 98%

- **Input**

Aliran F1

- Jumlah $C_3H_8O_3$ didalam Reaktor Fixed Bed Multitubular adalah :

Mol $C_3H_6O_3$ $= \frac{Massa}{BM} = \frac{11252,667\ Kg/Jam}{92\ Kg/kmol} = 122,3116\ kmol/jam$

- Jumlah H_2O didalam Reaktor Fixed Bed Multitubular adalah :

Mol H_2O $= \frac{Massa}{BM} = \frac{468,8611\ Kg/Jam}{18\ Kg/kmol} = 26,0478\ kmol/jam$

	$C_3H_8O_3$	$\longrightarrow$	$C_3H_8O_2$	+	H_2O
M	: 122,3116				
B	: 119,8645		119,8645		119,8645
S	: 2,4462		119,8645		119,8645 kmol/jam

- **Output**

Aliran F3

Hasil Produk

Massa Produk = mol × BM

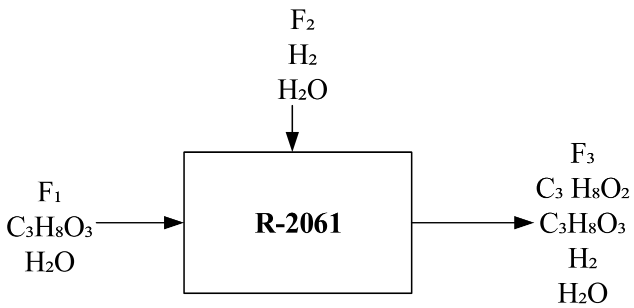
- $C_3H_8O_3$ = 2,4462 kmol/jam x 92 kg/kmol = 225,0533 Kg/Jam
- $C_3H_6O_2$ = 119,8645 kmol/jam x 74 kg/kmol = 8870,04 Kg/Jam
- H_2O = 119,8645 kmol/jam x 18 kg/kmol = 2626,438 Kg/Jam

Tabel A.1 Neraca Massa Reaktor Fix Bad Multitubular

Komponen	BM	Masuk		Keluar	
		F1		F2	
		Kg/jam	%	Kg/jam	%
$C_3H_8O_3$	92	11252,667	96	225,0533	1,92
C3H6O2	74			8870,04	75,67304
H_2O	18	468,86112	4	2626,438	22,40696
		11722	100	11722	100

2. Reaktor Fixed Bed Multitube (R-2061)

Fungsi : Untuk mereaksikan $C_3H_6O_2$ dengan H_2



Kondisi Operasi :

Temperatur : 210°C

Tekanan : 13 atm

Waktu : 1 Jam

Konversi : 98%

- **Input**

Aliran F1

- Jumlah $C_3H_8O_3$ didalam Reaktor Fixed Bed Multitubular adalah :

Mol $C_3H_6O_3$ $= \frac{Massa}{BM} = \frac{8870{,}15\ Kg/Jam}{72\ Kg/kmol} = 119{,}8654\ kmol/jam$

Aliran F2

- Jumlah H_2 didalam Reaktor Fixed Bed Multitubular adalah :

Mol H_2 $= \frac{Massa}{BM} = \frac{1198{,}67\ Kg/Jam}{2\ Kg/kmol} = 599{,}335\ kmol/jam$

$C_3H_6O_2$ + H_2 ⟶ $C_3H_8O_2$

M : 119,8654 599,335

B : 117,4681 117,4681 117,4681

S : 2,3973 481,8669 117,4681 kmol/jam

- **Output**

Aliran F3

Hasil Produk

Massa Produk = mol × BM

- $C_3H_8O_3$ = 2,4462 kmol/jam x 92 kg/kmol = 225,0533 Kg/Jam
- H_2 = 481,8669 kmol/jam x 2 kg/kmol = 963,7338 Kg/Jam
- $C_3H_6O_2$ = 2,3973 kmol/jam x 74 kg/kmol = 177,4007 Kg/Jam
- H_2O = 119,8645 kmol/jam x 18 kg/kmol = 2626,0533 Kg/Jam
- $C_3H_8O_2$ = 117,4681 kmol/jam x 76 kg/kmol = 8927,572 Kg/Jam

Tabel A.1 Neraca Massa Reaktor Fix Bad Multitubular

Komponen	BM	MASUK				KELUAR	
		F2		F3		F4	
		Kg/jam	%	Kg/jam	%	Kg/jam	%
$C_3H_8O_3$	92	225,0563	1,92			225,05632	1,74
H_2O	18	2626,473	22,41			2626,4725	20,33
C3H6O2	74	8870,155	75,67			177,40309	1,37
H_2	2			1198,67	100	963,73032	7,46
$C_3H_8O_2$	76					8927,6908	69,10
SUB		11721,68	100	1198,67	100	12920,353	100
TOTAL		12920,353				12920,353	

3. Flash Drum

Fungsi : Memisahkan Hidrogen dari keluaran reaktor.

Kondisi Operasi:

- Temperatur : 16°C
- Tekanan : 1,2 atm

Tabel A.2 Data-data Komponen pada Flash Drum

Komponen	BM	Massa (Kg/Jam)	Mol (Kmol/Jam)	Konsentrasi (%)	Xi	Titik Didih (°C)	
$C_3H_8O_3$	92,00	225,0563	2,4463	1,7419	0,0033	290	
$C_3H_8O_2$	76,00	8927,6908	117,4696	69,0979	0,1566	188	
H_2O	18,00	2626,4725	145,9151	20,3282	0,1945	100	
C3H6O2	74,00	177,4031	2,3973	1,3731	0,0032	56,55	HK
H2	2,00	963,7303212	481,8651606	7,459009151	0,642406776	-252,87	LK
TOTAL		12920,3531	750,0935	100,0000	1,0000		

- **Penentuan Temperatur Flash Drum**

Persamaan Antoine :

$$\text{Log Pi (Kpa)} = \frac{\text{Ant A} - \text{Ant B}}{\text{Ant C} + \text{T}}$$

T = 16,049°C = 289,199°K

P = 912 mmHg = 1,2 atm

Tabel A.4 Penentuan Operasi Flash Drum

Komponen	BM	Xi	Ant A	Ant B	Ant C	Log Pi	Pi	ki (Pi / Pt)	Yi (Xi × Ki)
H_2O	18,00	0,0033	8,1411	1810,9400	244,4850	1,1902	15,4961	0,0170	0,0001
$C_3H_8O_2$	76,00	0,1566	8,1192	2145,5200	205,3200	-1,5729	0,0267	0,0000	0,0000
$C_3H_8O_3$	92,00	0,1945	7,7891	2131,5200	124,5400	-7,3723	0,0000	0,0000	0,0000
C3H6O2	74,00	0,0032	7,21764	1252,8	234,565	2,2187	165,47	0,1814	0,0006

H2	2,00	0,6424	12,7800	232,3200	8,0800	3,1518	1418,25	1,555	0,9990
TOTAL		1,0000					1599,24		1,000

- **Penentuan Distribusi masing-masing komponen**

Volatilitas

$\alpha \frac{K_f}{K_{Hk}}$ (Pers. 8-10, Hal : 263, Coulson)

Keterangan :

α = Volatilitas relatif

K_f = Ki tiap komponen feed

K_{Hk} = Ki komponen *Heavy Key* (Komponen Berat)

Distribusi Komponen

- ***Heavy Key***

$(C_3H_6O_2)D$ = 1% x 2,3973 kmol = 0,02397 kmol = 1,82198 kg/jam

$(C_3H_6O_2)B$ = 99% x 2,3973 kmol = 2,37337 kmol = 180,37579 kg/jam

Total = 182,19777 kg/jam

$$Log \frac{(C_3H_8O_2)D}{(C_3H_8O_2)B} = -1,9956$$

- ***Light Key***

$(H_2O)D$ = 99% x 481,8652 kmol = 477,0465 kmol = 8586,8372 kg/jam

$(H_2O)B$ = 1% x 481,8652 kmol = 4,8187 kmol = 86,7357 kg/jam

Total = 8673,5729 kg/jam

$$Log \frac{(H_2O)D}{(H_2O)B} = 1,9956$$

Tabel A.5 Distribusi Komponen Flash Drum

Komponen	Ki	ai = Ki/KHk	log ai	log id/ib	id/ib
H_2O	0,0170	0,0936	-1,0285		
$C_3H_8O_2$	0,0000	0,0002	-3,79156999		

$C_3H_8O_3$	0,0000	0,0000	-9,5910		
C3H6O2	0,1814	1	0,0000	-1,9956	0,0101
H2	1,5551	8,5711	0,9330	1,9956	99,0000
TOTAL		9,6649	-13,4780		

- **Persamaan Van Winkle** (Hal. 345)

$$\text{Log } \alpha = m (\log id/ib) + b$$

Dimana :

$$m = \frac{(\log \alpha i)lk - (\log \alpha i)hk}{(\log id/ib)lk - (\log id/ib)hk}$$

$$= \frac{0,9330 - 0,0000}{1,9956 - (-1,9956)} = 0,2338$$

$$b = (\log \alpha i)lk - m (\log id/ib)lk$$

$$= 0,9330 - 0,4665$$

$$= 0,4665$$

Sehingga Persamaan menjadi :

$$\text{Log } (id/ib)I = \frac{(\log \alpha i - b)}{m}$$

$$= \frac{\log \alpha i - (0.4665)}{0,2338}$$

- **Gliserol ($C_3H_8O_3$)**

$$\text{Log } \frac{(C_3H_8O_3))D}{(C_3H_8O_3))B} = \frac{-9,591 - 0,4665}{0,2338}$$

$$= -43,0232$$

$$(C_3H_8O_3))B = \frac{2,4463}{0,00 + 1}$$

$$= 2,4463 \text{ kmol}$$

$$= 255,056 \text{ kg/jam}$$

$$(C_3H_8O_3))D = 2,4463 - 2,4463$$

$$= 0,0000 \text{ kmol}$$

$$= 0,0000 \text{ kg/jam}$$

- **Propilen glikol ($C_3H_8O_2$)**

$$\text{Log } \frac{(C_3H_8O_2)\text{D}}{(C_3H_8O_2)\text{B}} = \frac{-3{,}792 - 0{,}4665}{0{,}2338}$$

$$= -18{,}215$$

$$(C_3H_8O_2)\text{B} = \frac{117{,}4696}{0{,}000 + 1}$$

$$= 117{,}470 \text{ kmol}$$

$$= 8927{,}6908 \text{ kg/jam}$$

$$(C_3H_8O_2)\text{D} = 117{,}4696 - 117{,}4696$$

$$= 0{,}000 \text{ kmol}$$

$$= 0{,}000 \text{ kg/jam}$$

- **Air (H_2O)**

$$\text{Log } \frac{(H_2\text{O})\text{D}}{(H_2\text{O})\text{B}} = \frac{-1{,}0285 - 0{,}4665}{0{,}2338} = -6{,}3953$$

$$(H_2\text{O})\text{B} = \frac{145{,}9151}{0{,}000 + 1}$$

$$= 145{,}9151 \text{ kmol}$$

$$= 2626{,}4715 \text{kg/jam}$$

$$(H_2\text{O})\text{D} = 145{,}9151 - 145{,}9151$$

$$= 0{,}000 \text{kmol}$$

$$= 73{,}9429 \text{ kg/jam}$$

- **Acetol ($C_3H_6O_2$)**

$$\text{Log } \frac{(C_3H_6O_2)\text{D}}{(C_3H_6O_2)\text{B}} = \frac{0{,}000 - 0{,}4665}{0{,}2338}$$

$$= -1{,}9956$$

$$(C_3H_6O_2)\text{B} = \frac{2{,}3973}{0{,}0101 + 1}$$

$$= 2{,}3734 \text{ kmol}$$

$$= 175{,}6291 \text{ kg/jam}$$

$$(C_3H_6O_2)\text{D} = 2{,}3973 - 2{,}3734$$

$$= 0{,}0240 \text{ kmol}$$

$$= 1{,}7740 \text{ kg/jam}$$

- **Hidrogen (H_2)**

$$\text{Log } \frac{(H_2)\text{D}}{(H_2)\text{B}} = \frac{0{,}9330 - 0{,}4665}{0{,}2338} = 1{,}9956$$

$$(H_2)\text{B} = \frac{481{,}8652}{99{,}000 + 1}$$

$$= 4{,}8187 \text{ kmol}$$

$$= 9{,}6373 \text{ kg/jam}$$

$$(H_2)\text{D} = 481{,}8652 - 4{,}8187$$

$$= 477{,}0465 \text{ kmol}$$

$$= 954{,}0930 \text{ kg/jam}$$

Tabel A.8 Neraca Massa Flash Drum (FD-3501)

Komponen	Masuk (kg/jam)		Keluar (kg/jam)			
	F5 (Feed)	%	F6 (Destilat)	%	F7 (Bottom)	%
H_2O	2626,4725	20,32817924	0,0011	0,00011059	2626,4715	22
$C_3H_8O_2$	8927,6908	69,09788577	0,0000	0	8927,69	75
$C_3H_8O_3$	225,0563	1,741874408	0,0000	0	225,0563	2
C3H6O2	177,4031	1,373051436	1,7740	0,185593695	175,6291	1
H2	963,73032	7,459009151	954,0930	99,81429572	9,6373	0
SUBTOTAL	12920,353	100	955,86811	100,00000	11964,48501	100
TOTAL	12920,3531		12920,35311			

4. Distilasi (D-3071)

Fungsi : Memisahkan Hidrogen dari keluaran reaktor.

Kondisi Operasi:

- Temperatur : 121°C
- Tekanan : 1,2 atm

Tabel A.2 Data-data Komponen pada Destilasi

Komponen	BM	Masuk (Kg/jam)	Kmol/jam	Xi	Titik Didih	
$C_3H_8O_2$	76,00	8927,690836	117,4696163	0,437985534	188,2	Hk
$C_3H_8O_3$	92,00	225,0563244	2,446264395	0,009120898	290	
C3H6O2	74,00	175,629063	2,373365716	0,008849095	56,55	
H_2O	18,00	2626,471482	145,9150823	0,544044472	100	Lk
TOTAL		11954,84771	268,2043287	1		

- **Penentuan Temperatur Distilasi**

  Persamaan Antoine :

  $$\text{Log Pi (Kpa)} = \frac{\text{Ant A} - \text{Ant B}}{\text{Ant C} + \text{T}}$$

  T = 121,63 °C = 394,63 °K

  P = 912 mmHg = 1,2 atm

Tabel A.4 Penentuan Operasi Destilasi

Komponen	BM	Xi	Ant A	Ant B	Ant C	Log Pi	Pi	Ki = Pi/Pt	Yi = Ki Xi
$C_3H_8O_2$	76,00	0,437985534	8,1192	2145,52	205,32	1,55697336	36,05565254	0,039534707	0,0173
$C_3H_8O_3$	92,00	0,009	7,7891	2131,52	124,54	-0,869631771	0,135010713	0,000	1,35E-06
C3H6O2	74,00	0,008849095	7,21764	1252,8	234,565	3,700465419	5017,246269	5,501366523	0,04868
H_2O	18,00	0,544044472	8,1411	1810,94	244,485	3,19473069	1565,779814	1,716863831	0,9341
		1,0000							1,0000

- **Penentuan Distribusi masing-masing komponen**

  Volatilitas

  $$\alpha \frac{K_f}{K_{Hk}}$$ (Pers. 8-10, Hal : 263, Coulson)

Keterangan :

α = Volatilitas relatif

K_f = Ki tiap komponen feed

K_{Hk} = Ki komponen *Heavy Key* (Komponen Berat)

Distribusi Komponen

- ***Heavy Key***

$(C_3H_8O_2)D$ = 1% x 117,4696 kmol = 1,1747 kmol = 89,277 kg/jam

$(C_3H_8O_2)B$ = 99% x 117,4696kmol = 116,295 kmol = 8838,414 kg/jam

Total = 8927,691 kg/jam

$$Log\frac{(C_3H_8O_2)D}{(C_3H_8O_2)B} = -1{,}9956$$

- ***Light Key***

$(H_2O)D$ = 99% x 145,9151 kmol = 144,456 kmol = 2600,21 kg/jam

$(H_2O)B$ = 1% x 145,9151 kmol = 1,459 kmol = 26,264 kg/jam

Total = 2626,47 kg/jam

$$Log\frac{(H_2O)D}{(H_2O)B} = 1{,}9956$$

Tabel A.5 Distribusi Komponen Destilasi

Komponen	Ki	ai = Ki/KHk	log ai	log id/ib	id/ib
$C_3H_8O_2$	0,04	1	0,00	-1,99563519	0,01010101
$C_3H_8O_3$	0,0001	0,003744509	-2,426605131		
C3H6O2	6	139,152835	2,143492059		
H_2O	1,72	43,42675014	1,63775733	1,995635195	99
TOTAL		183,5833297	1,35		

- **Persamaan Van Winkle** (Hal. 345)

$$\text{Log } \alpha = m\ (\text{log id/ib}) + b$$

Dimana :

$$m = \frac{(\log \alpha i)lk - (\log \alpha i)hk}{(\log id/ib)lk - (\log id/ib)hk}$$

$$= \frac{1{,}6377 - 0{,}0000}{1{,}9956 - (-1{,}9956)} = 0{,}4103$$

$$b = (\log \alpha i)lk - m\ (\log id/ib)lk$$

$$= 1{,}6377 - 0{,}4103$$

$$= 0{,}8189$$

Sehingga Persamaan menjadi :

$$\text{Log } (id/ib)I = \frac{(\log \alpha i - b)}{m}$$

$$= \frac{\log \alpha i - (0.8189)}{0{,}4103}$$

- **Gliserol ($C_3H_8O_3$)**

$$\text{Log}\ \frac{(C_3H_8O_3))\text{D}}{(C_3H_8O_3))\text{B}} = \frac{-2{,}4266 - 0{,}8189}{0{,}4103}$$

$$= -7{,}9093$$

$$(C_3H_8O_3))\text{B} = \frac{2{,}4463}{0{,}00 + 1}$$

$$= 2{,}4463 \text{ kmol}$$

$$= 255{,}056 \text{ kg/jam}$$

$$(C_3H_8O_3))\text{D} = 2{,}4463 - 2{,}4463$$

$$= 0{,}0000 \text{ kmol}$$

$$= 0{,}0000 \text{ kg/jam}$$

- **Propilen glikol ($C_3H_8O_2$)**

$$\text{Log}\ \frac{(C_3H_8O_2)\text{D}}{(C_3H_8O_2)\text{B}} = \frac{0{,}000 - 0{,}8189}{0{,}4103}$$

$$= -1{,}9956$$

$$(C_3H_8O_2)\text{B} = \frac{117{,}4696}{0{,}0101 + 1}$$

$$= 116{,}2949 \text{ kmol}$$

$$= 8838{,}4139 \text{ kg/jam}$$

$$(C_3H_8O_2)\text{D} = 117{,}4696 - 116{,}2949$$

$$= 1{,}1747 \text{kmol}$$

$$= 89{,}2769 \text{ kg/jam}$$

- **Air (H_2O)**

$$\text{Log}\ \frac{(H_2\text{O})\text{D}}{(H_2\text{O})\text{B}} = \frac{1{,}6377 - 0{,}8189}{0{,}4103} = 1{,}9956$$

$$(H_2\text{O})\text{B} = \frac{145{,}9151}{99{,}00 + 1}$$

$$= 1{,}4591 \text{ kmol}$$

$$= 26{,}2647 \text{kg/jam}$$

$$(H_2\text{O})\text{D} = 145{,}9151 - 1{,}4591$$

$$= 144{,}4559 \text{kmol}$$

$= 2600{,}2068$ kg/jam

- **Acetol ($C_3H_6O_2$)**

$$\text{Log } \frac{(C_3H_6O_2)\text{D}}{(C_3H_6O_2)\text{B}} = \frac{2{,}1435 - 0{,}8189}{0{,}4103}$$

$$= 3{,}2281$$

$$(C_3H_6O_2)\text{B} = \frac{2{,}3733}{1690{,}9390 + 1}$$

$= 0{,}0014$ kmol

$= 0{,}1038$ kg/jam

$(C_3H_6O_2)\text{D} = 2{,}3734 - 0{,}0014$

$= 2{,}3720$ kmol

$= 175{,}5252$ kg/jam

Tabel A.8 Neraca Massa Distilasi

Komponen	Masuk (kg/jam)		Keluar (kg/jam)			
	F5 (Feed)	%	F6 (Destilat)	%	F7 (Bottom)	%
H_2O	2626,471482	21,96992841	2600,206767	90,75737017	26,26471482	0,288945882
$C_3H_8O_2$	8927,690836	74,67841545	89,27690836	3,116112735	8838,41	97,23400111
C3H6O2	175,629063	1,469103307	175,5252596	6,126516998	0,103803422	0,001141972
$C_3H_8O_3$	225,0563244	1,882552835	2,77292E-06	9,67856E-08	225,0563216	2,47591104
SUBTOTAL	11954,84771	100	2865,008938	100	9089,838767	100
TOTAL	11954,84771		11954,84771			

LAMPIRAN B
NERACA ENERGI

Kapasitas Produksi : 70.000 ton/tahun

Waktu Operasi : 330 hari/tahun

Temperatur referensi : 25ºC (298 K)

Satuan Operasi : kJ/Jam

- Persamaan yang digunakan untuk menghitung nilai panas (Q)
  - Menggunakan data Cp dalam bentuk konstanta

    $Q = m\ .Cp.\Delta T$ (Himmelblau, Pers. 23.12, Hal. 693)

    Data Cp konstanta dapat diperoleh dari Perry's *Chemical Handbook* Vol. 7 hal 354. Sedangkan data Cp konstanta untuk bahan yang dihitung berdasarkan gugus fungsi dapat dapat dilihat pada buku Perry's *Chemical Handbook* Vol. 7 hal 354.
  - Menggunakan data Cp yang dipengaruhi oleh temperatur

    $Q = m\ \int Cp\ \Delta T$ (Himmelblau, Pers. 23.12, Hal. 693)

    $Cp.\Delta T = A\ (T - T_0) + \frac{B}{2}(T^2 - T_0^2) + \frac{C}{3}(T^3 - T_0^3) + \frac{D}{4}(T^4 - T_0^4)$
- Persamaan yang digunakan untuk menghitung panas reaksi (Qr)

  $Q_r = -\Delta H_R$ (Himmelblau, Pers. 25.1, Hal. 770)

  $\Delta H_R = \Delta H^o{}_R + (\Delta H_{Produk} - \Delta H_{reaktan})$

  $\Delta H^o{}_R = \Delta H^o{}_{f\ produk} - \Delta H^o{}_{f\ reaktan}$

  $\Delta H_{produk} = \sum(m\ x\ cp\ x\ \Delta T)\ Produk$

  $\Delta H_{reaktan} = \sum(m\ x\ cp\ x\ \Delta T)\ reaktan$

Nilai kapasitas panas pada masing-masing komponen dapat dilihat pada Tabel B.1

Tabel B.1 Nilai kapasitas panas komponen

BM	Komponen	A	B	C	D	E
18	H2O	92,053	-0,039953	-0,0002110	0,0000005	
92	C3H8O3	132,145	8,6007E-01	-1,9740E-03	1,8060E-06	
2	H2	25,399	2,0178E-02	-3,8549E-05	3,1880E-08	-8,7585E-12
76	C3H8O2	118,614	6,7283E-01	-1,8377E-03	2,1303E-06	

74	C6H6O2	47,479	8,1081E-01	-2,6421E-03	3,6081E-06	

Sumber : *Perry's* hal 202, *Carl Yaws*, *Himmlebleu* 1049

Nilai pembentukan panas pada masing-masing komponen dapat dilihat pada Tabel B.2

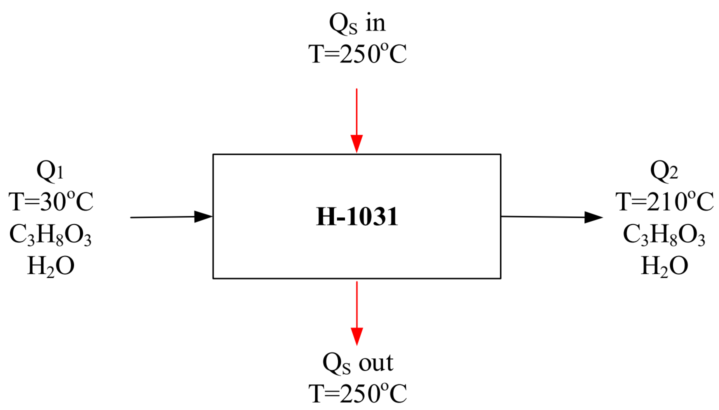
Tabel B.2 Nilai Panas Pembentukan Komponen

KOMPONEN	ΔHf (Kj/kmol)
$C_3H_8O_3$	-676,6
H_2	-4,2
H_2O (L)	-285,8
$C_3H_8O_2$	-421,5
H_2O (G)	-241,83
$C_3H_6O_2$	-384,5

Sumber*: Smith Van Ness App*

1. *Heater*

Fungsi : Tempat memanaskan Gliserol sebelum direaksikan dalam Reaktor



- **Input**

  T in = 30°C (303 K)

  T ref = 25°C (298 K)

  - **Q_1**

Tabel LB.1 Energi Q_1 *Heater* (H-1031)

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	11252,81622	122,3132198	1306,74419	159832,089
air	18	468,8673424	26,04818569	377,5027992	9833,263011
TOTAL					169665,352

- **Output**

T in = 150ºC (483 K)

T ref = 25ºC (298 K)

- **Q_2**

Tabel LB.2 Energi Q_2 *Heater* (H-1031)

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	11252,81622	122,3132198	33780,48926	4131800,407
air	18	468,8673424	26,04818569	9469,148779	246654,1457
TOTAL					4378454,552

- **Beban Panas *Heater***

$\Delta Q = Q_{out} - Q_{in}$ **(Himmelblau, Pers. 5.11, Hal 71)**

= 4378454,552 Kj/jam – 169665,352 Kj/jam

= 4208789,2 Kj/jam

- **Panas *Steam***

Medium pemanas adalah *saturated steam* pada T = 200°C

Sehingga:

H_l = 852,43 Kj/kg

H_v = 2793,2 Kj/kg

λ_s = 1940,77 Kj/kg

(Smith van Ness, Appendix F.1 *Saturated Steam*, Hal. 669)

- **Jumlah *steam* yang dibutuhkan (m_s)**

$$m_s = \frac{\Delta Q}{\lambda_s}$$

$$= \frac{4208789,2 \text{ kj/jam}}{1940,77 \text{ Kj/kg}}$$

= 2168,6182 Kj/kg

- **Panas *steam* masuk ($Q_{s\ in}$)**

$Q_{s\ in} = m_s \times H_v$

= 2168,6182 Kg/jam x 2793,2 Kj/kg

= 6057384,437 Kj/jam

- **Panas kondensat keluar ($Q_{s\ out}$)**

$Q_{s\ out}$ = m_s x H_l

= 2168,6182 Kg/jam x 852,43 Kj/kg

= 1848595,237 Kj/jam

Tabel LB.3 Neraca Energi *Heater* (H-1031)

Energi	Masuk (Kj/jam)	Keluar (Kj/jam)
Q1 in	169665,352	
Q2 out		4378454,552
Qs in	6057384,437	
Qs out		1848595,237
TOTAL	6227049,789	6227049,789

2. **Reaktor**

Fungsi : Tempat terjadinnya reaksi gliserol membentuk Acetol

Kondisi Operasi :

- Temperatur : 150 ˚C
- Tekan : 1 atm

Input

- **Q_3**
- T_{in} :150 ˚C
- T_{ref} : 25 ˚C

Tabel LB Neraca Energi Q_3 Reaktor

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	11252,81622	122,3132198	33780,48926	4131800,407
air	18	468,8673424	26,04818569	10523,31022	274113,1387
TOTAL					4405913,545

- **Q_4**

T_{out} : 150 ˚C

T_{ref} : 25 ˚C

Tabel LB Neraca Energi Q_4 Reaktor

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	225,0563244	2,446264395	33780,48926	82636,00813
air	18	2626,47	145,9151411	10523,31022	1535510,295
acetol	74	8870,15	119,8669554	20886,32933	2503580,706
TOTAL					4121727,009

- **Panas Reaksi Standar (25** ˚C)

ΔHR° = ΔHf° produk - ΔHf° reaktan

Reaksi

$C_3H_8O_3 \longrightarrow C_3H_6O_2 + H_2O$

ΔHR° = ΔHf° produk - ΔHf° reaktan

= (ΔHf° $C_3H_6O_2$ + ΔHf° H_2O) – (ΔHf° $C_3H_8O_3$)

= (-384,5 + - 285,8) – (-676,6)

= -670,3 - -676,6

= 6,3

- **Panas Reaksi Operasi (150** ˚C)

ΔH = n∫Cp dt

$\Delta H_{C_3H_8O_3}$ = 4131800,407 kJ

ΔH_{H_2O} =1261397,157 kJ

$\Delta H_{C_3H_6O_2}$ = 2503580,71 kJ

$\Delta HR_{T\ operasi}$ = ΔHR° + ($\Delta H_{Produk} - \Delta H_{Reaktan}$)

= 6,30 + (3764977,86 – 4131800,4067)

= 6,30 + -366822,544

= - 366816,244

Qr = -1 x $\Delta HR_{T\ operasi}$

= -1 x – 366816,244

= 366816,244

- **Beban Panas Reaktor**

ΔQ = Q_{out} - Q_{in}

= 651022,780 kJ/jam

Sistem butuh air pendingin dengan kalor sebesar 100650741,7480 kJ/jam

- **Kebutuhan Air Pendingin**

  T_{reff} = 25 °C

  T_{in} = 25 °C

  T_{out} = 45 °C

  $m_w = \frac{\Delta Q}{Cp \times \Delta T}$

  m_w = 43400,185 kg/jam

- **Panas Air Pendingin Masuk**

  $$Q_{w\,in} = m_w \,.Cp \,.\Delta T$$

  = 217000,927 kJ/jam

- **Panas Air Pendingin Keluar**

  $$Q_{w\,in} = m_w \,.Cp \,.\Delta T$$

  = 868003,707 kJ/jam

Tabel LB.8 Neraca Energi *Reactor*

Energi	Masuk (Kj/jam)	Keluar (Kj/jam)
Q1 In	4405913,5454	
Q2 Out		4121727,0095
QR	366816,2441	
Qw in	217.000,9267	
Qw out		868003,7066
TOTAL	4989730,7161	4989730,7161

3. Coller

Fungsi : Tempat mendinginkan produk keluaran reaktor

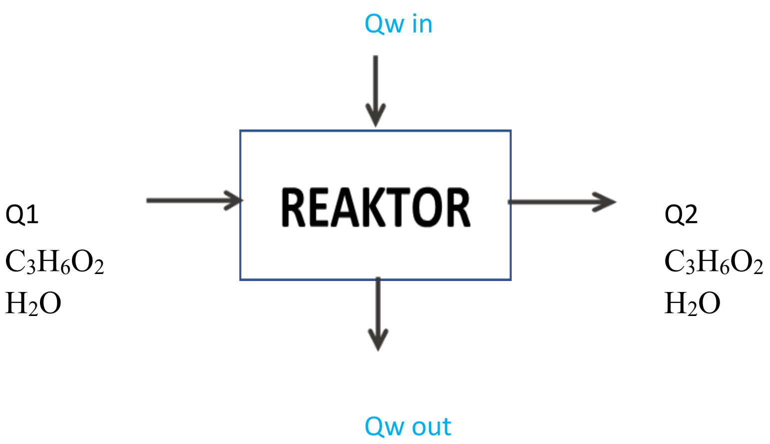
Kondisi Operasi :

T_{reff} = 25 °C

T_{in} = 150 °C

T_{out} = 60 °C

Tekanan = 1 atm

- 

- **Q_5**

Tabel Neraca Energi Coller Masuk

Komponen	BM (Kmol)	Massa (kg/jam)	n(Kmol)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
Gliserol	92	225,0563244	2,446264395	33780,48926	82636,00813
air	18	2626,472539	145,9151411	10523,31022	1535510,295
acetol	74	8870,15	119,8669554	20886,32933	2503580,706
TOTAL					4121727,009

- **Q_6**

Tabel Neraca Energi Coller Keluar

Komponen	BM (Kmol)	Massa (kg/jam)	n(Kmol)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
Gliserol	92	225,0563244	2,446264395	12789,92023	31287,52648
air	18	2626,472539	145,9151411	2634,235862	384374,8973
acetol	74	8870,15	119,8669554	5378,987723	644762,8814
TOTAL					1060425,305

- **Panas yang diserap pendingin**

ΔQc = Q_{out} - Q_{in}

= 3061301,704 Kj/Jam

Tabel Neraca Energi Coller

Energi	Masuk(kj/jam)	Keluar (kj/jam)
Q masuk	4121727,009	
Q keluar		1060425,305
QC		3061301,704
Total	4121727,009	4121727,009

4. **Heater**

Fungsi : Untuk memanaskan produk sebelum direaksikan di Reaktor

- **Input**

T in = 60ºC (333 K)

T ref = 25ºC (298 K)

- Q_6

Tabel Neraca Energi Heater Masuk

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	225,0563244	2,446264395	12789,92023	31287,52648
air	18	2626,472539	145,9151411	2634,235862	384374,8973
acetol	74	8870,15	119,8669554	5378,987723	644762,8814
TOTAL					1060425,305

- **Output**

T in = 210ºC (483K)

T ref = 25ºC (298 K)

- Q_7

Tabel Neraca Energi Heater Keluar

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	225,0563244	2,446264395	54423,29165	133133,7606
air	18	2626,472539	145,9151411	14299,64855	2086535,235
acetol	74	8870,15	119,8669554	33407,09538	4004406,811
TOTAL					6224075,807

- **Beban Panas *Heater***

$\Delta Q = Q_{out} - Q_{in}$ **(Himmelblau, Pers. 5.11, Hal 71)**

= 6224075,807 Kj/jam – 1060425,305 Kj/jam

= 5163650,502 Kj/jam

- **Panas *Steam***

Medium pemanas adalah *saturated steam* pada T = 200°C

Sehingga:

H_l = 1085,35 Kj/kg

H_v = 2801,5 Kj/kg

λ_s = 1716,15 Kj/kg

(Smith van Ness, Appendix F.1 *Saturated Steam*, Hal. 669)

- **Jumlah *steam* yang dibutuhkan (m_s)**

$$m_s = \frac{\Delta Q}{\lambda_s}$$

$$= \frac{5163650{,}502 \text{ kj/jam}}{1716{,}15 \text{ Kj/kg}}$$

= 3008,857 Kj/kg

- **Panas *steam* masuk ($Q_{s\ in}$)**

$Q_{s\ in}$ = m_s x H_v

= 3008,857 Kg/jam x 2801,5 Kj/kg

= 8429313,801 Kj/jam

- **Panas kondensat keluar ($Q_{s\ out}$)**

$Q_{s\ out}$ = m_s x H_l

= 3008,857 Kg/jam x 1085,35 Kj/kg

= 3265663,300 Kj/jam

Tabel LB.3 Neraca Energi *Heater* (H-1031)

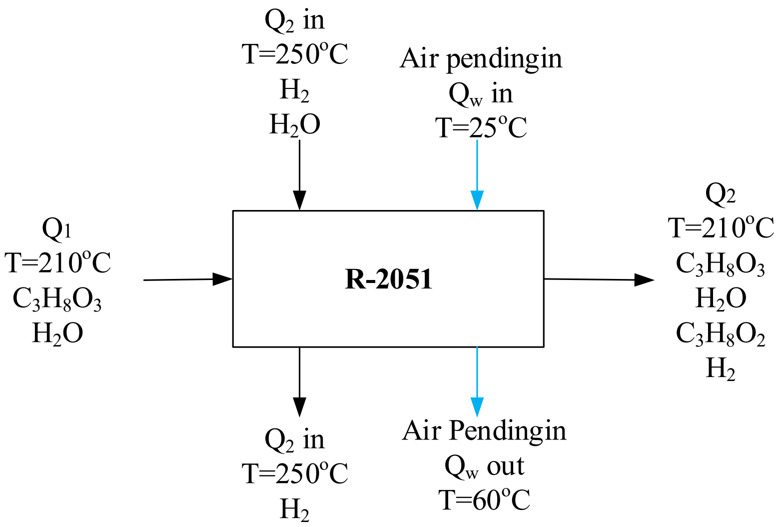
Energi	Masuk (Kj/jam)	Keluar (Kj/jam)
Q1 in	1060425,305	
Q2 out		6224075,807
Qs in	8429313,801	
Qs out		3265663,300
TOTAL	9489739,106	9489739,106

5. Reaktor

Fungsi : Tempat terjadinya reaksi antara gliserol untuk membentuk Acetol

Kondisi Operasi:

- Temperatur : 210 ˚C
- Tekanan : 13 atm



Input

- **Q_1**

T_{in} =210˚C

T_{ref} =25˚C

Tabel LB.4 Neraca Energi Q_1 Reaktor

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	225,0563244	2,446264395	54423,29165	133133,7606
air	18	2626,472539	145,9151411	14299,64855	2086535,235
acetol	74	8870,15	119,8669554	33407,09538	4004406,811
TOTAL					6224075,807

- **Q_2**

T_{in} =210˚C

T_{ref} =25˚C

Tabel LB.5 Neraca Energi Q_2 Reaktor

Komponen	Bm	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
hidrogen	2	1198,67	599,3347768	5377,6652	3223021,773
Total					**3223021,773**

Output

- **Q_3**

T_{out} =210˚C

T_{ref} =25˚C

Tabel LB.6 Neraca Energi Q_3 Reaktor

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	225,0563244	2,446264395	54423,29165	133133,7606
air	18	2626,472539	145,9151411	14299,64855	2086535,235
acetol	74	177,40	2,397339107	33407,09538	80088,13621
propilen G	76	8927,690836	117,4696163	42522,84528	4995142,317
					7294899,449

- **Q_4**

T_{out} =210˚C

T_{ref} =25˚C

Tabel LB.7 Neraca Energi Q_4 Reaktor

Komponen	Bm	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
hidrogen	2	963,73	481,8651606	5377,6652	2591309,505
Total					**2591309,505**

- **Panas Reaksi :**

$\Delta HR°$ = $\Delta Hf°$ produk - $\Delta Hf°$ reaktan

Reaksi :

$C_3H_6O_2 + H_2 \longrightarrow C_3H_8O_2$

$\Delta HR°$ = $\Delta Hf°$ produk - $\Delta Hf°$ reaktan

= (-421,50) – (-388,70)

= -32,80 Kj/mol

= -27442,82Kj/jam

ΔH = $n\int Cp\ dt$

$\Delta H_{C_3H_6O_2}$ =4004406,811 kJ

$\Delta H_{C_3H_8O_2}$ = 4995142,3168 kJ

ΔH_{H2} = 3223021,773 kJ

$\Delta HR_{T\ operasi}$ = $\Delta HR^o + (\Delta H_{Produk} - \Delta H_{Reaktan})$

= (-32,80) + ((4995143,32) - (7227428,5834))

= -2232319,0666kJ/jam

QR = $(-\Delta HR_{operasi})$

QR = 2232319,0666 kJ/jam

- **Beban Panas Reaktor**

ΔQ = $Q_{out} - Q_{in}$

= 1793207,6916 kJ/jam

Sistem butuh air pendingin dengan kalor sebesar 1793207,6916 kJ/jam

- **Kebutuhan Air Pendingin**

T_{reff} = 25 °C

T_{in} = 25 °C

T_{out} = 45 °C

$m_w = \frac{\Delta Q}{Cp \times \Delta T}$

m_w = 89660,3846 kg/jam

- Panas air pendingin masuk

$$Q_{w\ in} = m_w . Cp . \Delta T$$

= 0,0000 kJ/jam

- panas air pendingin keluar

$$Q_{w\ in} = m_w . Cp . \Delta T$$

=1793207,6916 Kj/jam

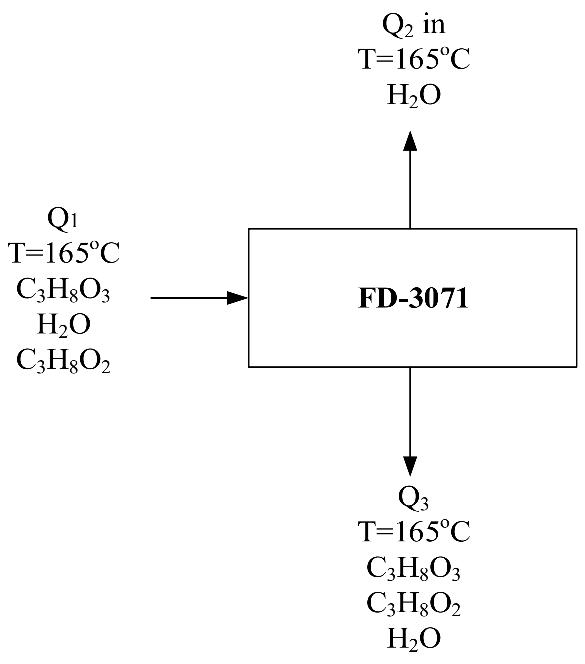
Tabel LB.8 Neraca Energi *Reactor*

Energi	Masuk (Kj/jam)	Keluar (Kj/jam)
Q1 In	6224075,8067	
Q2 In	3223021,7728	
Q3 Out		7294899,4491
Q4 Out		2591309,505
QR	2232319,0666	
Qw in	0,0000	
Qw out		1793207,6916
TOTAL	11679416,6461	11679416,6461

6. FLASH DRUM

Kondisi Operasi :

- Temperatur : 16 ˚C
- Tekanan : 1,2 atm



Input

- **Q_1**

T_{in} =16˚C

T_{ref} =25˚C

Tabel LB.15 Energi Q_1 *Flash Drum*

Komponen	massa (kg/jam)	Bm	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	225,06	92	2,446264395	1218,897579	2981,745749
propilen glikol	8927,69	76	117,4696163	-1904,814828	-223757,8669
air	2626,47	18	145,9151411	-681,0154442	-99370,46461
acetol	**177,4030939**	**74,00**	**2,397339107**	**-1341,734382**	**-3216,592306**
hidrogen	963,7303212	2	481,8651606	-258,6731928	-124645,5996
total					-448008,7776

Output

- **Bottom Q_2**

T_{in} =165˚C

T_{ref} =25˚C

Tabel LB.16 Energi Q_2 Flash Drum

Komponen	massa (kg/jam)	Bm	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	0,00	92	0,00	959,556	0,000
propilen glikol	0,00	76	0,00	-2115,738	0,000
air	0,00	18	0,00	-756,817	0,000
acetol	1,774	74,00	0,024	-1489,864	-35,717
hidrogen	954,093	2	477,046	-287,390	-137098,333
total					-137134,095

- **Top Q_3**

T_{in} =165˚C

T_{ref} =25˚C

Tabel LB.17 Energi Q_3 Flash Drum

Komponen	massa (kg/jam)	Bm	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	225,06	92	2,446	52394,732	128171,368
propilen glikol	8927,69	76	117,470	40747,819	4786630,628
air	2626,47	18	145,915	13716,369	2001425,178
acetol	175,63	74,00	2,373	31821,022	75522,924
hidrogen	9,64	2	4,819	5172,826	24926,044

total		7016676,142

- **Beban Panas**

  ΔQ = $Q_{out} - Q_{in}$
  = -7327550,83

  Q Laten penguap = Qf-Qpa-QPb
  = - 7327551,83

Tabel LB.18 Neraca Energi *Flash Drum*

Energi	Panas Masuk(kJ/Jam)	Panas keluar (kJ/Jam)
Q IN	-448.009	
Q OUT		6.879.542
Q Laten Penguapan		-7.327.551
Total	-448.009	-448.009

7. Colller

Fungsi : Tempat mendinginkan produk keluaran reaktor

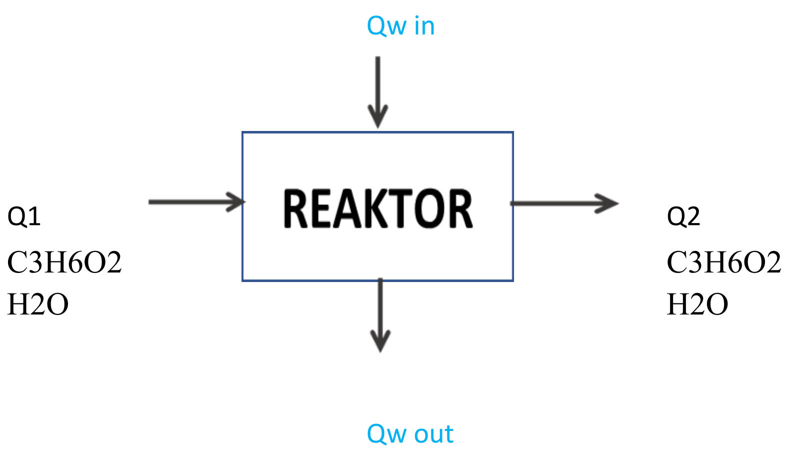
Kondisi Operasi :

T_{reff} = 25 °C

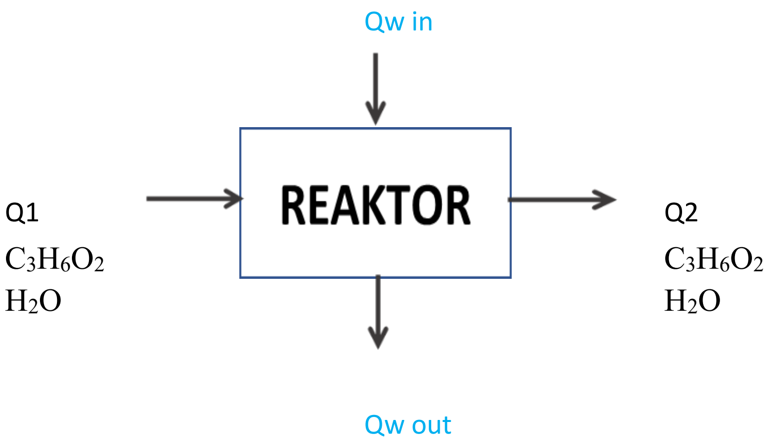
T_{in} = 150 °C

T_{out} = 60 °C

Tekanan = 1 atm



-



- **Q_{13}**

Tabel Neraca Energi Coller Masuk

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	225,0563244	2,446264395	54423,29165	133133,7606
air	18	2626,472539	145,9151411	14299,64855	2086535,235
acetol	74	177,40	2,397339107	33407,09538	80088,13621
propilen G	76	8927,690836	117,4696163	42522,84528	4995142,317
TOTAL					7294899,449

- **Q_{14}**

Tabel Neraca Energi Coller Keluar

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	225,0563244	2,446264395	6465,23326	15815,66994
air	18	2626,472539	145,9151411	6703,458371	978136,0738
acetol	74	177,40	2,397339107	6936347,511	16628777,15
propilen G	76	8927,690836	117,4696163	19518,25935	2292802,436
TOTAL					19915531,33

- **Panas yang diserap pendingin**

$\Delta Qc \quad = Q_{out} - Q_{in}$

$= -12620631,9$ Kj/Jam

Tabel Neraca Energi Coller

Energi	Masuk(kj/jam)	Keluar (kj/jam)
Q masuk	7294899,449	
Q keluar		19915531,33
QC		-12620631,88
Total	7294899,449	7294899,449

8. *DESTILASI*

Kondisi Operasi :

- Temperatur : 120 ˚C
- Tekanan : 1 atm

Input

- **Q_1**

T_{in} =˚C

T_{ref} =25˚C

Tabel LB.15 Energi Q_1 *Flash Drum*

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	225,0563244	2,446264395	6465,23326	15815,66994
air	18	2626,472539	145,9151411	6703,458371	978136,0738
acetol	74	177,40	2,397339107	6936347,511	16628777,15
propilen G	76	8927,690836	117,4696163	19518,25935	2292802,436
TOTAL					19915531,33

Output

- **Bottom Q_2**

T_{in} =165˚C

T_{ref} =25˚C

Tabel LB.16 Energi Q_2 Flash Drum

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
air	18	2600,206767	144,4559315	6551,578396	946414,3601
acetol	74	175,53	2,371962967	13972,72364	33142,78303
gliserol	92	2,77292E-06	3,01404E-08	23272,08087	0,00070143
propilen G	76	89,27690836	1,174696163	19064,1885	22394,62907
TOTAL					1001951,773

- **Top Q_3**

T_{in} =165˚C

T_{ref} =25˚C

Tabel LB.17 Energi Q_3 Flash Drum

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	225,0563216	2,446264365	49702,24303	121584,826
propilen G	76	8838,41	116,2949201	41506,26752	4826968,064
air	18	26,26471482	1,459150823	13965,58844	20377,89987
acetol	74	0,103803422	0,001402749	32496,1784	45,58398009
TOTAL					4968976,374

- **QR**

T_{in} =165˚C

T_{ref} =25˚C

Tabel LB Neraca Energi Destilasi

Komponen	BM	Massa	n (Kmol/Jam)	∫ Cp dt (Kj/Kmol)	Q (Kj/jam)
gliserol	92	297,5996468	3,23477877	49702,24303	160775,7606
propilen G	76	11687,34	153,7807376	41506,26752	6382864,436
air	18	34,73072784	1,92948488	13965,58844	26946,39174

acetol	74	0,137262804	0,001854903	32496,1784	60,27725096
TOTAL					6570646,865

- **Beban Panas yang diserap kondensor**

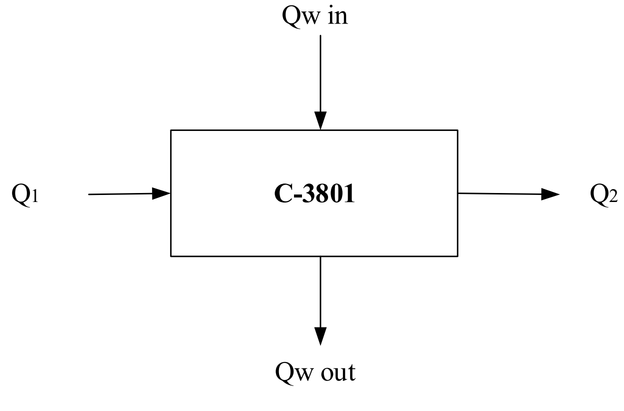
= Panas masuk + panas dari reboiler = Panas produk bawah + beban panas diserap kondensor + panas produk atas

= 19915531,33 + 6570646,865 = 4968976,374 + Qc + 1001951,773
= 26486178,19 = 5970928,147 + Qc
QC = 20515250,05

Tabel LB.18 Neraca Energi *Flash Drum*

Energi	Masuk (kj/jam)	Keluar (kj/jam)
Q1 (panas Masuk)	19915531,33	
Q3 (produk atas)		1001951,77
Q2 (produk bawah)		4968976,37
QR (Panas dari reboiler)	6570646,87	
QC (panas diserap kondensor)		20515250,05
TOTAL	26486178,19	26486178,19

9. Kondensor



Kondisi operasi :

T = 50ºC

P = 1 atm

T_{in} = 130ºC (403 K)

T_{out} = 30ºC (303 K)

T_{ref} = 25ºC (298 K)

- Beban panas yang diserap Kondensor (Qc)

  Qc = 20511898,17

- Menghitung jumlah air pendingin yang dibutuhkan

$$m = \frac{Qc}{Cp\ x\ \Delta T}$$

$$= \frac{20515250{,}05}{1732{,}014}$$

= 11844,7385 Kg/jam

- Menghitung panas air masuk (Qw in)

  QW in = m.Cp.dT

  = 11844,7385 x 151,0672

  = 1789351,33 kg/jam

- Menghitung panas air keluar (QW out)

  QW in = m.Cp.dT

  = 11844,7385 x 1883,0809

  = 22304601,4 kg/jam

Tabel LB.21 Neraca Energi Cooler (C-3901)

Panas	Masuk (kj)	Keluar (kj)
QWIN	1789351,335	
QWOUT		22304601,3822
QCD	20515250,0472	
TOTAL	22304601,3822	22304601,3822

10. Reboliler

- Beban panas yang diberikan RB

  QRB = 20515250,05

- Menghitung jumlah steam yang dibutuhkan

  Media pemanas adalah saturated steam pada temperatur 250 ˚C

  Hl = 1085,35 Kj/Kg

  Hv = 2801,5 Kj/Kg

  λ_s = 1716,15 Kj/Kg

$$m_s = \frac{QRB}{\lambda_s}$$

= 11954,229

- Menghitung panas steam masuk (Qw in)

  QW in = m.Hv

  = 11954,229 x 2801,5

  = 33489772,46 kg/jam

- Menghitung panas steam keluar (QW out)

  QW in = m.Hl

  = 11954,229 x 1085,35

  = 12974522,41 kg/jam

Tabel LB Neraca Energi Reboiler

panas	masuk (kj)	keluar (kj)
Qin	33489772,4600	
Qout		12974522,4128
QRB		20515250,0472
TOTAL	33489772,4600	33489772,4600

LAMPIRAN C

SPESIFIKASI PERALATAN

A. Spesifikasi Peralatan Utama

1. Tangki Penyimpanan Gliserol (ST-1101)

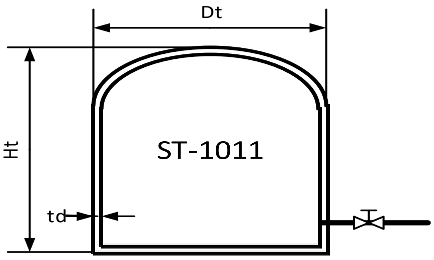
Fungsi : Tempat menyimpan Bahan Baku Gliserol

Bahan konstruksi : *Carbon Steel*

Jumlah : 1 unit

Lama Penyimpanan : 7 Hari

Gambar :



Parameter	M (kg/jam)	ρ (kg/m3)	V (m3/jam)	%	Xi / p (kg/m3)	densitas mix
$C_3H_8O_3$	11252,667	1261	8,924	96	0,0008	1247,7838
H_2O	468,861	997	0,470	4	0,0000	
Total	**11721,528**		**9,394**	**100**	**0,0008**	**1247,7838**

Data :

Laju alir = 11721,684kg/jam = 25841,624 lb/jam

Volumetrik = 9,394 m^3/jam = 0,092 ft^3/s

Densitas mix = 1247,7838 kg/m^3 = 77,899 lb/ft^3

T = 30^0C

t = 7 hari = 168 jam

P = 1 atm = 14,7 psi

Kapasitas tangki, V_t

$V_b = \frac{m \times t}{\rho}$ $= \frac{11721{,}528 \frac{kg}{jam} x\ 168\ jam}{1247{,}7838 ^{kg}/_{m^3}}$

= 1.578,171 m^3

= 5.573,547 ft^3

= 96.305.931,743 in^3

= 14.722.714,306 gal

Faktor keamanan 10%

Maka,

Vp = 0,9 V_t

$V_t = \frac{V_p}{0{,}9}$

$= \frac{1578{,}171\ m^3}{0{,}9}$

= 1753,524 m^3

= 61923,941 ft^3

= 107006590,825 in^3

= 463231,904 gal

Dimensi tangki

- **Volume silinder, V_s**

$V_s = \frac{\pi}{4} \times D_t^2 \times H_t$ $H_t = 1{,}5D_t$

Maka,

$V_s \quad = \frac{\pi}{4} \times D_t^3$

- **Volume ellipsoidal, V_e**

$V_e \quad = \frac{\pi}{24} \times D_t^3$ He = ¼ Dt (peter's, Tabel 4)

$V_e \quad = 0,1308 \text{ x } D_t^3$

- **Diameter tangki, D_t**

$V_t \quad = V_s + V_e$

$= \left(\frac{1,5\pi}{4} \times D_t^3\right) + (\, 0,1308 \times D_t^3)$

$V_t \quad = 1,3083 \text{ x } D_t^3$

$D_t^3 \quad = \frac{Vt}{1,3083}$

$D_t^3 \quad = \frac{1753,524}{1,3083}$

$D_t^3 \quad = 1340,307 \text{ m}^3$

$D_t \quad = \sqrt[3]{1340,307}$

= 10,999 m

= 36,077 ft

= 433,036 in

- **Tinggi tangki, H_t**

Tinggi silinder, Hs $= 1,5\ D_t$

= 16,499 m

= 54,116 ft

= 649,555 in

Tinggi ellipsoidal, H_e $= 1/4 D_t$

= 2,750 m

= 9,019 ft

= 108,259 in (walas, Tabel 18.5)

Tinggi total, H_t = tinggi silinder + tinggi ellipsoidal

H_t = 16,499 m + 2,750 m

= 19,249 m

= 19248,509 mm

- **Tinggi cairan dalam tangki, H_c**

H_c $= \frac{volume\ bahan}{volume\ tangki} \times H_t$

$= \frac{1578,171\ \ m^3}{1753,524\ m^3} \times 19,249\ m$

= 17,324 m

= 56,822 ft

= 682,032 in

- **Tekanan cairan dalam tangki, P_c**

P_c $= \rho \times g \times H_c$

= 1247,784 kg/m^3 x 9,81 m/dt^2 x 17,324 m

= 211838,555 kg/m s^2

= 2,055 atm

= 30,19 psi

- **Tekanan desain, P_d**

Pd = Pop + Pc

= (1 + 2,055) atm

= 3,055 atm

= 44,906 psi

- **Tebal dinding tangki, t_d**

t_d $= \frac{PR}{SE-0,6P} + C$ (Walas, Tabel 18.4)

- Tekanan desain, P_d : 44,90 psi

- Jari-jari, R : 216,518 in
- *Allowable stress*, S : 13700 psi (Walas, Tabel 18.4)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)
- Lama tahun digunakan : 10 tahun

Maka,

td $= \frac{44{,}90\ psi \times 216{,}518\ in}{(13700\ psi \times 0{,}85) - (0{,}6 \times 44{,}90 psi)} + 0{,}002 \frac{in}{tahun} \times 10 tahun$

= 0,856 in

= 0,021 m

= 0,071 ft

= 21,76 mm

- **Tebal dinding ellipsoidal,t_e**

$$t_e = \frac{PD_t}{2SE - 0{,}2P} + C$$

(Walas-Chemical Process Equipment, Tabel 18.4)

t_e $= \frac{44{,}90\ psi \times 649{,}555\ in}{2(13700\ psi \times 0{,}85) - (0{,}2 \times 44{,}90 psi)} + 0{,}002 \frac{in}{tahun} x 10 tahun$

= 1,727 in

= 0,032 m

= 0,106 ft

= 32,331 mm

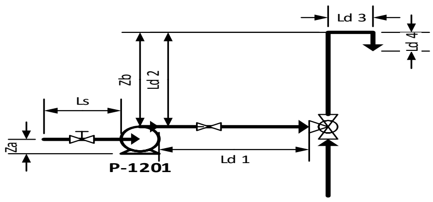
2. Pompa (P-1201)

Fungsi : Mengalirkan gliserol dari tangki penampung menuju heater

Tipe : *Centrifugal Pumps*

Bahan : *Carbon Steel*

Gambar :



Data :

- Laju alir massa, m : 11721,5281 kg/jam = 7,1781 lb/dt
- Densitas campuran, ρ : 1248 kg/m^3 = 77,8991 lb/ft^3
- Viskositas Campuran, μ : 1,4664 cP = 0,00099 lb/ft.s
- Tinggi pompa terhadap cairan masuk, Za : 0 m = 0 ft
- Tinggi pompa terhadap cairan keluar, Zb : 8 m = 26,24 ft
- Panjang pipa hisap, Ls : 5 m = 16,4 ft
- Panjang pipa buang, Ld : 10 m = 32,8 ft
- Faktor keamanan 10%

Pemilihan faktor keamanan pada pompa *centrifugal pump* dapat dilihat pada Gambar 10.1

TABLE 6
Factors in equipment scale-up and design

Type of equipment	Is pilot plant usually neces-sary?	Major variables for operational design (other than flow rate)	Major variables characterizing size or capacity	Maximum scale-up ratio based on indi-cated char-acterizing variable	Approxi-mate recom-mended safety or over-design factor, %
Agitated batch crystallizers	Yes	Solubility-temperature relationship	Flow rate Heat transfer area	>100:1	20
Batch reactors	Yes	Reaction rate Equilibrium state	Volume Residence time	>100:1	20
Centrifugal pumps	No	Discharge head	Flow rate Power input Impeller diameter	>100:1 >100:1 10:1	10

Sumber : (Peter, Pers 14.15 Hal 496)

Laju alir volumetrik, Q_v

$$Q_p = \frac{m}{0,9}$$

$$= \frac{7,1781\ lb/s}{0,9} = 7,9757 \text{ lb/s}$$

$$Q_v = \frac{Q_p}{\rho}$$

$$= \frac{7{,}9757\ lb/s}{77{,}899\ lb/ft^3}$$

$= 0{,}1024\ ft^3/s$

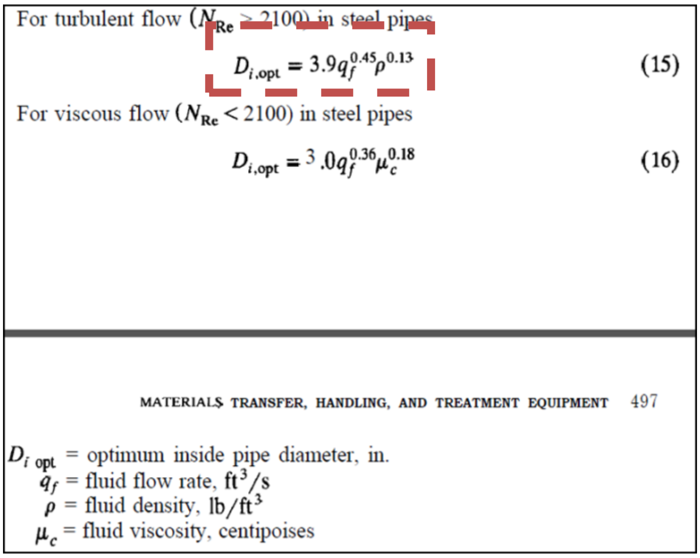
$= 45{,}9565$ gal/min

Diameter optimum, D_{opt}

Asumsi aliran turbulen

$D_{opt} = 3{,}9 * Q_v^{0,45} * \rho^{0,13}$ (Peter, Pers 14.15 Hal 496)

Untuk mendapatkan rumus diameter optimum pada aliran turbulen dapat dilihat pada Gambar 10.2



For turbulent flow ($N_{Re} > 2100$) in steel pipes

$$D_{i,opt} = 3.9 q_f^{0.45} \rho^{0.13} \quad (15)$$

For viscous flow ($N_{Re} < 2100$) in steel pipes

$$D_{i,opt} = 3.0 q_f^{0.36} \mu_c^{0.18} \quad (16)$$

MATERIALS TRANSFER, HANDLING, AND TREATMENT EQUIPMENT 497

$D_{i\ opt}$ = optimum inside pipe diameter, in.
q_f = fluid flow rate, ft^3/s
ρ = fluid density, lb/ft^3
μ_c = fluid viscosity, centipoises

Sumber : (Peter, Pers 14.15 Hal 496)

$D_{opt} = 3{,}9 * Q_v^{0,45} * \rho^{0,13}$

$= 3{,}9 * (0{,}1024)^{0,45} * (77{,}8991)^{0,13}$

$= 2{,}4636$ in

Berdasarkan Tabel 11 Kern, diperoleh pipa baja dengan ukuran sebagai berikut,

	Suction (a)	Discharge (b)
IPS	2,5 in sch 40	

ID	2,4690	in	0,2058	ft	2,4690	in	0,2058	ft
OD	2,8800	in	0,2400	ft	2,8800	in	0,2400	ft
a"	0,0333				ft^2			

Kecepatan aliran, V

Va = Vb, karena ukuran pipa hisap dan pipa buang sama

$$V = \frac{Q_v}{a"}$$

$$= \frac{0{,}1024 \, ft^3/dt}{0{,}0333 \; ft^2} = 3{,}0780 \text{ ft/s}$$

$$\frac{V^2}{2g_c} = \frac{(3{,}0780)^2 \; ft/s}{2 \times 32{,}17 \; ftlbm/lbfs^2} = 0{,}1472 \text{ ft-lb}_f\text{/lb.m}$$

Bilangan Reynolds, N_{Re}

$$N_{Re} = \frac{\rho \times V \times D}{\mu}$$ (Mc Cabe, pers 3.8)

$$N_{Re} = \frac{\rho \times V \times D}{\mu}$$

$$= \frac{77{,}8991 \; \frac{lb}{ft^3} \times 3{,}0780 \frac{ft}{dt} \times 0{,}2058 ft}{0{,}00099 \; \frac{lb}{ft.hr}} = 50.064{,}0489$$

- **Rugi Gesek pada Pipa hisap (*suction*)**

  Pada pipa hisap, rugi gesek timbul akibat gesekan dengan kulit pipa

The hydraulic radius is a useful parameter for generalizing fluid-flow phenomena in turbulent flow. Equation (5.7) can be so generalized by substituting $4r_H$ for D or $2r_H$ for r_w:

$$h_{fs} = \frac{\tau_w}{\rho r_H} \Delta L = \frac{\Delta p_s}{\rho} = f \frac{\Delta L}{r_H} \frac{\bar{V}^2}{2g_c} \quad (5.56)$$

$$N_{Re} = \frac{4 r_H \bar{V} \rho}{\mu} \quad (5.57)$$

The simple hydraulic-radius rule does not apply to laminar flow through noncircular sections. For laminar flow through an annulus, for example, f and N_{Re} are related by the equation[3]

$$f = \frac{16}{N_{Re}} \phi_a \quad (5.58)$$

$$r_H \equiv \frac{S}{L_p} \qquad (5.54)$$

where S = cross-sectional area of channel
L_p = perimeter of channel in contact with fluid

Thus, for the special case of a circular tube, the hydraulic radius is

$$r_H = \frac{\pi D^2/4}{\pi D} = \frac{D}{4}$$

The equivalent diameter is $4r_H$, or simply, D.

An important special case is the annulus between two concentric pipes. Here the hydraulic radius is

$$r_H = \frac{\pi D_o^2/4 - \pi D_i^2/4}{\pi D_i + \pi D_o} = \frac{D_o - D_i}{4} \qquad (5.55)$$

h_{fsa} $= f \dfrac{\Delta L}{r_H} \dfrac{V^2}{2g_c}$ (Mc Cabe, Pers 5.56)

r_H $= \dfrac{ID}{4}$ (Mc Cabe, Hal 103)

$= \dfrac{0{,}2058 ft}{4}$

$= 0{,}0514$ ft

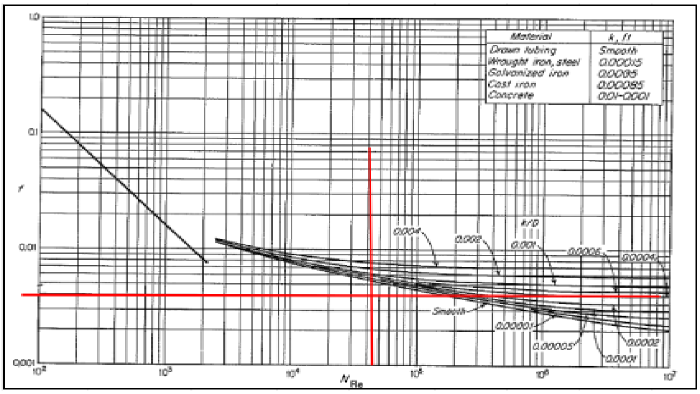
N_{Re} = 50.064,04

Material yang digunakan untuk konstruksi pipa adalah *carbon steel pipe*, dimana

K = 0,00015 ft (Mc Cabe, Fig. 5.9)

$\dfrac{k}{ID}$ $= \dfrac{0{,}00015\, ft}{0{,}2058\, ft} = 0{,}0007$

f = 0,004 (Mc Cabe, Fig. 5.9)



Maka,

h_{fsa} $= 0{,}004 \times \dfrac{16{,}4000 ft}{0{,}0514\, ft} \times 0{,}1060 ft.\, lb_f/lb$

$= 0{,}1878$ ft-lb_f/lb

- Rugi gesek akibat *fitting* dan *valve*

$h_{ff} \quad = K_f \frac{V^2}{2g_c}$ (Mc Cabe, Pers 5.67)

EFFECT OF FITTINGS AND VALVES. Fittings and valves disturb the normal flow lines and cause friction. In short lines with many fittings, the friction loss from the fittings may be greater than that from the straight pipe. The friction loss h_{ff} from fittings is found from an equation similar to Eqs. (5.59) and (5.65):

$$h_{ff} = K_f \frac{\bar{V}_a^2}{2g_c} \qquad (5.67)$$

where K_f = loss factor for fitting
$\bar{V}_a$ = average velocity in pipe leading to fitting

Factor K_f is found by experiment and differs for each type of connection. A short list of factors is given in Table 5.1.

K_f (*gate valve*) = 0,2 = 1 unit (Mc Cabe, Tabel 5.1)

TABLE 5.1
Loss coefficients for standard threaded pipe fittings†

Fitting	K_f
Globe valve, wide open	10.0
Angle valve, wide open	5.0
Gate valve	
Wide open	0.2
Half open	5.6
Return bend	2.2
Tee	1.8
Elbow	
90°	0.9
45°	0.4

† From J. K. Vennard, in V. L. Streeter (ed.), *Handbook of Fluid Dynamics*, McGraw-Hill Book Company, New York, 1961, p. 3-23.

Total K_f = (1 x 0,2) = 0,2

Maka,

$h_{ff} \quad = 0,2 \times 0,1472\ ft.lb_f/lb$ = 0,0294 ft-lbf/lb

- **Rugi Gesek pada Pipa buang (*discharge*)**

Pada pipa buang, rugi gesek timbul akibat gesekan dengan kulit pipa, pengaruh *fitting* dan *valve*.

- Pada pipa hisap, rugi gesek timbul akibat gesekan dengan kulit pipa

h_{fsb} $= f \dfrac{\Delta L}{r_H} \dfrac{V^2}{2g_c}$ (Mc Cabe, Pers 5.56)

r_H $= \frac{ID}{4}$ (Mc Cabe, Hal 103)

$= \dfrac{0{,}2058\ ft}{4} = 0{,}0514$ ft

N_{Re} = 50.064,04

Material yang digunakan untuk konstruksi pipa adalah *carbon steel pipe*, dimana

k = 0,00015 ft (Mc Cabe, Fig. 5.9)

$\frac{k}{ID}$ $= \dfrac{0{,}00015ft}{0{,}2058ft} = 0{,}0007$

f = 0,004 (Mc Cabe, Fig. 5.9)

Maka,

h_{fsb} $= 0{,}004 \times \frac{32{,}8000\ ft}{0{,}0514\ \ ft} \times 0{,}1060 ft.lb_f/lb$

= 0,3756 ft-lb_f/lb

- Rugi gesek akibat *fitting* dan *valve*

h_{ffb} $= K_f \dfrac{V^2}{2g_c}$ (Mc Cabe, Pers 5.67)

EFFECT OF FITTINGS AND VALVES. Fittings and valves disturb the normal flow lines and cause friction. In short lines with many fittings, the friction loss from the fittings may be greater than that from the straight pipe. The friction loss h_{ff} from fittings is found from an equation similar to Eqs. (5.59) and (5.65):

$$h_{ff} = K_f \frac{\bar{V}_a^2}{2g_c} \qquad (5.67)$$

where K_f = loss factor for fitting
$\bar{V}_a$ = average velocity in pipe leading to fitting

Factor K_f is found by experiment and differs for each type of connection. A short list of factors is given in Table 5.1.

K_f (*elbow* 90°) = 0,9 = 2 unit (Mc Cabe, Tabel 5.1)

K_f (*globe valve*) = 10 = 1 unit (Mc Cabe, Tabel 5.1)

TABLE 5.1
Loss coefficients for standard threaded pipe fittings†

Fitting	K_f
Globe valve, wide open	10.0
Angle valve, wide open	5.0
Gate valve	
Wide open	0.2
Half open	5.6
Return bend	2.2
Tee	1.8
Elbow	
90°	0.9
45°	0.4

† From J. K. Vennard, in V. L. Streeter (ed.), *Handbook of Fluid Dynamics*, McGraw-Hill Book Company, New York, 1961, p. 3-23.

Total K_f = (2x0,9) + (1x10) = 11,8

Maka,

h_{ffb} = $11{,}8 \times 0{,}1472\ ft.lb_f/lb$ = 1,7375 ft-lb_f/lb

Sehingga, total rugi gesek adalah

= $h_{fsa} + h_{ff} + h_{fsb} + h_{ff}$

h_f = 2,3303 ft-lb_f/lb

Daya pompa (BHP)

Daya pompa dihitung menggunakan Persamaan Bernoulli :

(Mc.Cabe, pers 4.32)

$$\frac{P_a}{\rho} + \frac{gZ_a}{g_c} + \frac{\alpha_a V_a^2}{2g_c} + \eta W_p = \frac{P_b}{\rho} + \frac{gZ_b}{g_c} + \frac{\alpha_b V_b^2}{2g_c} + h_f$$

fluid is $W_p - h_{fp}$. In practice, in place of h_{fp} a pump efficiency denoted by η is used, defined by the equation

$$W_p - h_{fp} \equiv \eta W_p$$

or

$$\eta = \frac{W_p - h_{fp}}{W_p} \qquad (4.31)$$

The mechanical energy delivered to the fluid is, then, ηW_p, where $\eta < 1$. Equation (4.29) corrected for pump work is

$$\frac{p_a}{\rho} + \frac{gZ_a}{g_c} + \frac{\alpha_a \bar{V}_a^2}{2g_c} + \eta W_p = \frac{p_b}{\rho} + \frac{gZ_b}{g_c} + \frac{\alpha_b \bar{V}_b^2}{2g_c} + h_f \qquad (4.32)$$

Equation (4.32) is a final working equation for problems on the flow of incompressible fluids.

Atau

$$\eta W_p = \left(\frac{P_b}{\rho} + \frac{gZ_b}{g_c} + \frac{\alpha_b V_b^2}{2g_c}\right) - \left(\frac{P_a}{\rho} + \frac{gZ_a}{g_c} + \frac{\alpha_a V_a^2}{2g_c}\right) + h_f$$

Dimana

$P_a = P_b$

$V_a = Vb$

$\rho_a = \rho_b$

$g/g_c = 1$

$\alpha_a = \alpha_b$

$\eta = \frac{58+8}{2} = 43\ \%$ (Peters, Fig. 14.37)

FIGURE 1436
Characteristic curves for a typical centrifugal pump showing effect of viscosity.

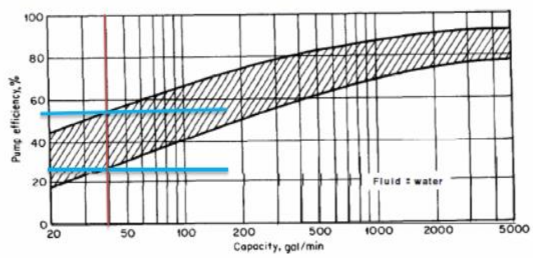


FIGURE 1437
Efficiencies of centrifugal pumps.

Sehingga persamaan di atas dapat disederhanakan menjadi :

$\eta W_p = (Z_b - Z_a) + h_f$

$0{,}43\ W_p = (26{,}24 - 0)ft + 2{,}3303\ ft\text{-}lb_f/lb$

$W_p = 66{,}4426\ ft\text{-}lb_f/lb$

$BHP = \frac{W_p \times m}{550}$

$= \frac{66{,}4226\ ft.lb_f/lb \times 6{,}0906\ lb/dt}{550}$

$= 0{,}8672$ Hp

Daya motor (MHP)

MPH $= \frac{BHP}{\eta}$

η = 80 % (Peters, Fig 14.38)

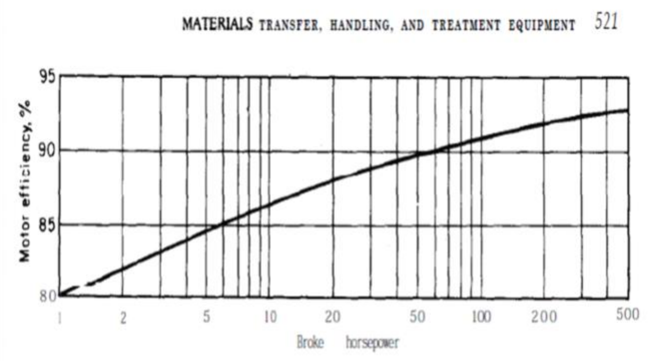


FIGURE 1438
Efficiencies of three-phase motors.

MPH $= \frac{0{,}8672\ hp}{0{,}8}$

= 1,0839 Hp

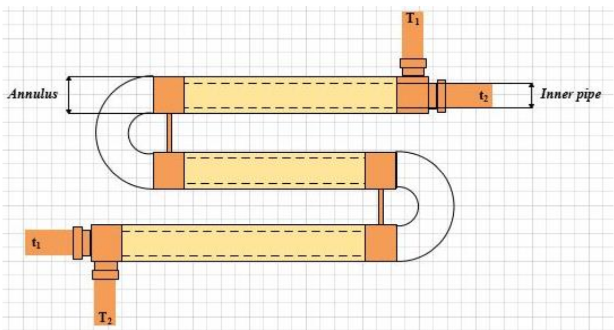
3. Heater (H-1031)

Fungsi : Tempat menaikkan suhu gliserol

Tipe : *Centrifugal Pumps*

Bahan : *Carbon Steel*

Gambar :



Data :

Fluida Panas : ***Steam***

- Laju alir massa : 2168,58 kg/jam = 4780,9157 lb/jam
- Temperatur Masuk (T1) : 200°C = 392 F

- Temperatur Keluar (T2) : 200°C = 392 F

Fluida Dingin : Gliserol

- Laju alir massa : 11721,52 kg/jam = 25841,51529 lb/jam
- Temperatur Masuk (t1) : 30°C = 86 F
- Temperatur Keluar (t2) : 150°C = 302 F

Menghitung ΔL LMTD *(Log Mean Temperature Difference)*

Fluida Panas		Temperatur (°F)	Fluida Dingin		Selisih
T1	392	Temperatur tinggi	t2	302	90
T2	392	Temperatur rendah	t1	86	306

$$LMTD = \frac{\Delta t2 - \Delta t1}{ln\Delta t2 - \Delta t1}$$

LMTD = 176,5029 F

Faktor Koreksi LMTD

$$R = \frac{T_1 - T_2}{t_2 - t_1} = 0{,}000$$

$$S = \frac{t_2 - t_1}{T - t_1} = 0{,}7059$$

Menentukan Luas Area Perpindahan Panas (A)

Dari tabel 8 D.Q KERN maka didapatkan koefisien perpindahan panas sebesar :

Ud : 150 Btu/jamft^{2}F

Q : 4208733,38 Kj/jam = 3989879,25 Btu/Jam

Sehingga :

$$A = \frac{Q}{U_D \ X \ \Delta t \ LMTD}$$

A = 150,70 ft^2

Karena nilai A<200 ft^2, maka tipe *Heater* yang digunakan adalah ***Double Pipe*** (D.Q KERN, Hal.103)

Spesifikasi *Double Pipe*

TABLE 6.2. FLOW AREAS AND EQUIVALENT DIAMETERS IN DOUBLE PIPE EXCHANGERS

Exchanger, IPS	Flow area, in.²		Annulus, in.	
	Annulus	Pipe	d_e	d'_e
2 × 1¼	1.19	1.50	0.915	0.40
2½ × 1¼	2.63	1.50	2.02	0.81
3 × 2	2.93	3.35	1.57	0.69
4 × 3	3.14	7.38	1.14	0.53

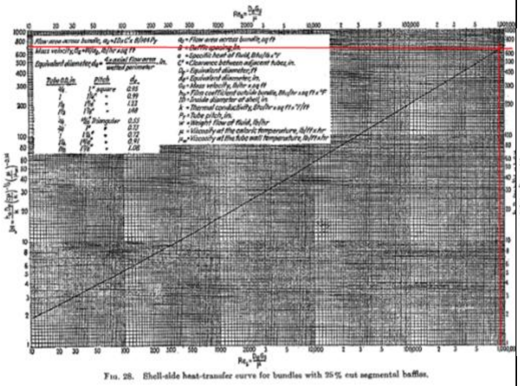
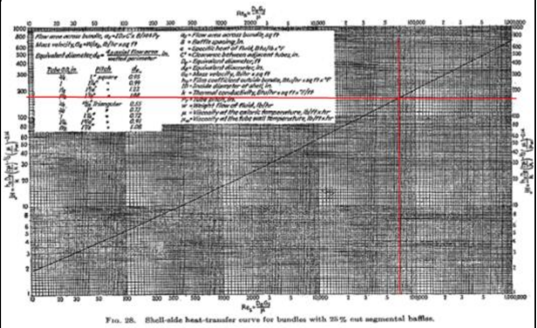
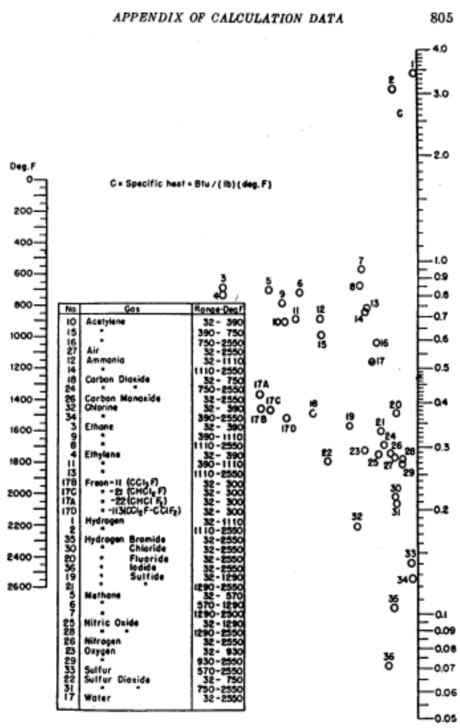
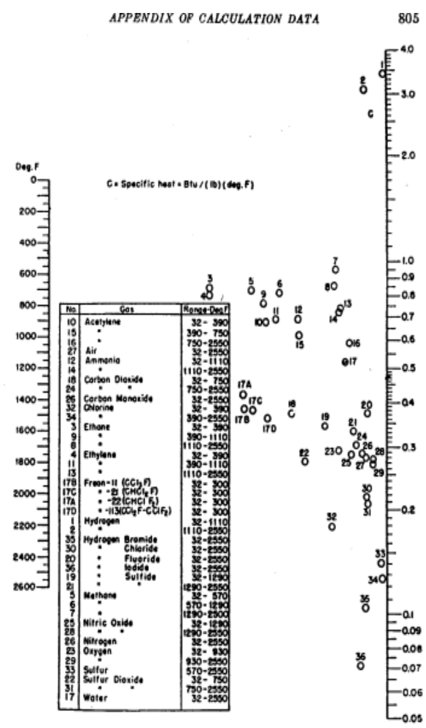
Dari Tabel 6.2 D.Q Kern Hal.110 diperoleh *flow area* dan De pada *Double Pipe Heat Exchanger*, yaitu :

Exchanger, IPS	Flow Area, in^2		Annulus, in	
	Annulus	Pipe	De	De'
2 X 1,25	1,19	1,5	0,915	0,4

	Anulus (in)	(ft)	Inner Pipe (in)	(ft)
IPS	2	0,1666	1,25	0,1041
SC	40	3,3320	50	3,3320
OD (D2)	2,38	0,1983	1,66	0,1383
ID (D1)	2,067	0,1722	1,38	0,1150
a" (ft^2/ft)	0,6220	0,0518	0,435	0,0362

***Annulus* (Steam)**	***Inner Pipe* (Fluida Dingin)**
1. *Flow Area,* (aa) D2 = 0,1983 in = 0,1983 ft D1 = 2,067n = 0,1722 ft	**5. *Flow Area,* a_s** D1 = 0,1150 ft $ap = \frac{\pi(D1^2)}{4}$

$$aa = \frac{\pi(D2^2 - D1^2)}{4}$$ $$aa = \frac{3{,}14\ (0{,}1983^2 - 0{,}1722^2)}{4}$$ $$aa = 0{,}0076\ ft^2$$	$$ap = \frac{3{,}14(0{,}1150^2)}{4}$$ $$ap = 0{,}0104 ft^2$$
2. *Diameter Ekivalen (De)* $$De = \frac{(D2^2 - D1^2)}{D1}$$ $$De = \frac{(0{,}1983^2 - 0{,}1722^2)}{0{,}1983}$$ $$De = 0{,}561\ ft$$	
3. *Mass Velocity* (Ga) $$\text{Ga} = \frac{\text{W}}{aa}$$ $$\text{Ga} = \frac{4780{,}9157\ lb/jam}{0{,}0076\ ft^2}$$ $$\text{Ga} = 630579{,}66 \frac{lb}{h}.ft^2$$	***6. Mass Velocity* (Gp)** $$\text{Gp} = \frac{\text{W}}{ap}$$ $$\text{Gp} = \frac{25841{,}5153\ lb/jam}{0{,}0104\ ft^2}$$ $$\text{Gp} = 2491151{,}52 \frac{lb}{h}.ft^2$$
4. Penentuan nilai μ liquid $$Rep \frac{De}{\mu}\ X\ Ga$$ $$Rep = \frac{0{,}561\ ft\ X\ 630579{,}66}{0{,}0363}$$ $$Rep = 974.427{,}04$$	**5. Penentuan nilai** μ liquid $$Rep = \frac{De}{\mu}\ X\ Gp$$ $$Rep = \frac{0{,}1150\ X\ 2491151}{3{,}5486}$$ $$Rep = 80.699{,}02$$

6. **Faktor Perpindahan Panas (jH)**	7. **Faktor Perpindahan Panas (jH)**
 Masukkan nilai Re = 974,427,04 ke fig. 4 Maka, didapatkan nilai factor perpindahan panas : 700	 Masukkan nilai Re = 80.699,02 ke fig. 4 Maka, didapatkan nilai factor perpindahan panas : 190
8. **Koefisien Perpindahan Panas (k)** Tc = 120 F  **Sumber : DQ Kern. Fig.3, Hal. 805** Dari Fig.3 di atas, maka didapatkan : c = 0,48 Btu/lb. ºF	9. **Koefisien Perpindahan Panas (k)** Tc = 86 F  **sumber : D.Q Kern. Fig. 3, Hal.805** dari Fig. 3 di atas, maka didapatkan : c = 0,57 Btu/lb.ºF

k = 0,0163 Btu/hr.ft. ºF $\frac{C\mu}{k} = (\frac{0,48\frac{Btu}{lb}.°F\ X\ 0,0363\frac{lb}{ft}.hr}{0,0163\frac{Btu}{hr}.ft.°F})$ $\frac{C\mu}{k} = 1,0690$ $(\frac{C\mu}{k})^{\frac{1}{3}} = 1,0225$	k = 68 btu/hr.ft. ºF $\frac{C\mu}{k} = (\frac{0,57\frac{Btu}{lb}.°F\ X\ 0,0363\frac{lb}{ft}.hr}{68\frac{Btu}{hr}.ft.°F})$ $\frac{C\mu}{k} = 0,0297$ $(\frac{C\mu}{k})^{\frac{1}{3}} = 0,3102$
10. Menghitung Ho $ho = jH\ X \frac{k}{De}(\frac{Cp.\mu}{k})^{1/3}(\frac{\mu}{\mu w})^{0,14}$ $ho = 700X\frac{0,0163}{0,0561}X\ 1,0255\ X\ 1$ $ho = 207,9761\ btu/h.ft^2.°F$	***7. Menghitung Hio*** $hi = 190\ X \frac{k}{De}(\frac{Cp.\mu}{k})^{1/3}(\frac{\mu}{\mu w})^{0,14}$ $hi = 100\ X\frac{68}{0,1150}X\ 0,3102\ X\ 1$ $hi = 34864,8\ btu/h.ft^2.°F$ **KONVERSI hi ke hio** $hio = hi\ X\frac{ID}{OD}$ $hio = 34864X\frac{0,1150}{0,1383} = 28984\ btu/h.ft2.°F$
11. ***Clean Overall Coefficient,*** **U_c** $Uc = \frac{h_{io}h_o}{h_{io} + h_o}$ $Uc = \frac{34864\ X\ 207,97}{34864 + 207,97}$ $Uc = 415,9522 btu/h.ft2.°F$	
12. ***Design Overall Coefficient*** **(Ud)** $\frac{1}{UD} = \frac{1}{UC} + Rd$	

$$\frac{1}{Ud} = \frac{1}{415{,}9522} + 0{,}002$$ $$\frac{1}{Ud} = 0{,}0044$$ $$Ud = 227{,}0600 \frac{Btu}{h} ft2.°F$$
13. Panjang Pipa yang dibutuhkan (L) Ud = 227,0600 Btu/jam.ft^{2}F Q = 4208733,3832 Btu/jam a'' = 0,4350 ft^2/ft $$L = \frac{A}{a''} = \frac{105{,}0168\ ft^2}{0{,}4350\ ft^2/ft}$$ $$L = 241{,}4179\ ft$$ Dipilih Panjang *hair pin* = 20 ft Hairpin yang dibutuhkan (n) = $\frac{L}{2\ X\ Panjang\ Pipa}$= 6,0354 Hair pin yang digunakan sebanyak = 6 buah
14. *Required Surface* A actual = 241,4179 X 0,4350 = 105,0168 ft^2 $$UD\ aktual = \frac{Q}{A\ X\ \Delta T\ LMTD}$$ $$UD\ aktual = 227{,}0600\ \text{btu/h.ft2.°F}$$ $$Rd\ = \frac{Uc - Ud}{Uc\ X\ Ud} = 0{,}0020\ \text{hr.ft2.°F/Btu}$$

Pressure Drop : Annulus	***Pressure Drop : Pipe***
1. Rea	**1. Rep**

<table>
<tr>
<td>
De’ = D2 – D1
= 0,1983 – 0,1722
= 0,0261 ft
$$Rea = \frac{Deq\ X\ Ga}{\mu}$$
$$Rea = \frac{0{,}0261\ X\ 630.579{,}66}{0{,}0363}$$
$$Rea = 452.92{,}2265$$
$$f = 0{,}0035 + \frac{0{,}264}{327{,}5478}$$
$$f = 0{,}0043$$
</td>
<td>
Rep = 80.699,02
$$f = 0{,}0035 + \frac{0{,}264}{115{,}05} = 0{,}0058$$
</td>
</tr>
<tr>
<td>
2. Δfa
$$\Delta Fa = \frac{4\ X\ f\ X\ Ga^2 X\ La}{2\ X\ g\ X\ \rho^2 X\ D}$$
$$\Delta Fa = 1{,}0884\ ft$$
</td>
<td>
2. Δfp
$$\Delta Fp = \frac{4\ X\ f\ X\ Gp^2 X\ L}{2\ X\ g\ X\ \rho^2 X\ D}$$
$$\Delta Fp = 33{,}8945\ ft$$
</td>
</tr>
<tr>
<td>
3. Δpa
$$\mathrm{V} = \frac{\mathrm{Ga}}{3600 \times \rho}$$
$$\mathrm{V} = \frac{630.579{,}66}{3600 \times 264} = 0{,}66\ ft/s$$
$$\Delta\mathrm{Ft} = 3\ \mathrm{X}\frac{\mathrm{V}^2}{2\mathrm{g}}$$
$$\Delta\mathrm{Ft} = 3\ \mathrm{X}\frac{0{,}44}{64{,}4} = 0{,}02\ ft$$
$$\Delta\mathrm{pa} = \frac{(\Delta\mathrm{Fa} + \Delta\mathrm{Ft})\rho}{144}$$
$$\Delta\mathrm{pa} = 2{,}03\ psi$$
Memenuhi karena <10 psi
</td>
<td>
4. ΔPp
$$\Delta\mathrm{Pp} = \frac{(\Delta\mathrm{Fp})\ \mathrm{X}\ \rho}{144}$$
$$\Delta\mathrm{Pp} = 6{,}1519\ \mathrm{psi}$$
Memenuhi karena <10 psi
</td>
</tr>
</table>

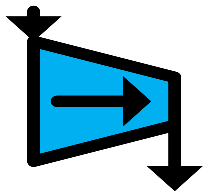
4. **Kompresor (CM-1041)**

Fungsi : Menaikkan tekanan $H2$ sebelum masuk reaktor

Tipe : Centrifugal Compressor

Bahan : Carbon Steel

Gambar :



Data

Laju alir massa (m)	= 19891,30 Kg/jam	= 43852,37 Ib/jam
Densitas Hidrogen (ρ)	= 8,9880 Kg/m^3	= 0,56 Ib/ft^3
Tekanan masuk P1 (atm)	= 1 atm	
Tekanan keluar P2 (atm)	= 30 atm	
Faktor keamanan	= 10 %	

TABLE 6
Factors in equipment scale-up and design (Continued)

Type of equipment	Is pilot plant usually necessary?	Major variables for operational design (other than flow rate)	Major variables characterizing size or capacity	Maximum scale-up ratio based on indicated characterizing variable	Approximate recommended safety or over-design factor, %
Reciprocating compressors	No	Compression ratio	Flow rate Power input Piston displacement	>100:1 >100:1 >100:1	10

(Petter, Tabel 6. Hal 37)

- **Laju alir volumetrik (Qv)**

Qv = m/ ρ

= 43852,37/0,56

= 78151,38 ft^3/jam

= 21,71 ft^3/s

- **Menentukan jenis Compressor**

  Dari laju alir volumetrik dan tekanan keluar dapat dilihat jenis Compressor pada grafik dibawah ini (Branan, 2004)

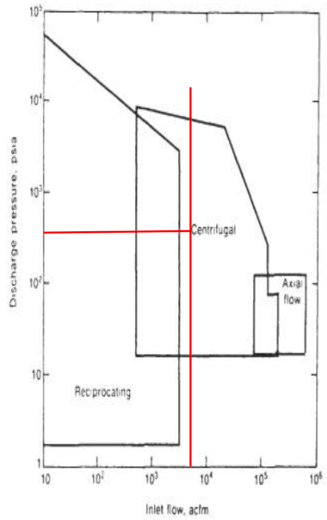


Figure 1. Approximate ranges of application for reciprocating, centrifugal, and axial-flow compressors.

Maka, jenis compressor yang digunakan adalah Centrifugal Compressor

- **Daya Kompresor**

  $P_{gas} = 3{,}03 \times 10^{-5} \times P1 \times Qv \times \text{Ln P2/P1}$

  = 3,5704 lbft/s

  = 0,0065 Hp

  Faktor keamanan 10%

  = 0,0065 Hp / 10%

  = 0,0072 Hp

5. Reaktor (H-2051)

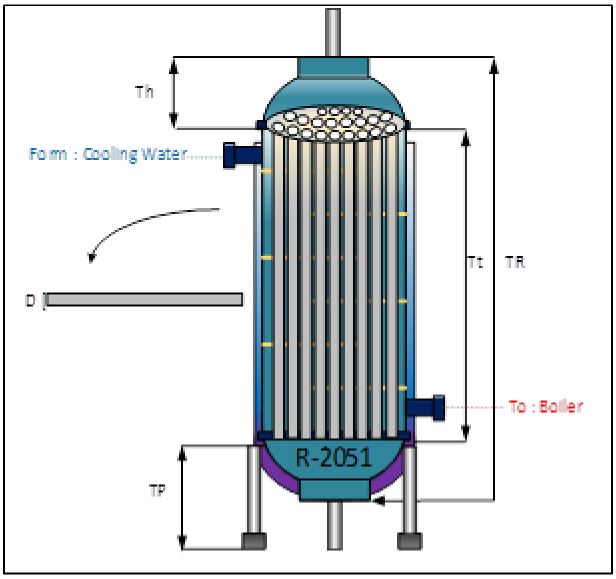
Fungsi : Tempat mereaksikan gliserol dengan hidrogen

Tipe : FIX BAD REACTOR MULTITUBULAR

Bahan : Carbon Steel SA 285 Grade A

Jumlah : 1 Unit

Gambar :



Data :

- Laju alir campuran (m) : 12920,35 kg/jam = 28484,21 lb/jam
- Densitas Campuran (ρ) : 1247,7838 kg/m^3 = 77,90 lb/ft^3
- Temperatur (T) : 150 °C
- Tekanan (P) : 1 atm
- Viskositas (μ) :0,55 cP = 0,00037 lb/ft.s
- Waktu (t) : 80 menit = 1,3 Jam
- π : 3,14
- Faktor Keamanan : 20% (Petter, Tabel 6. Hal 37)

1. Kapasitas Rektor (Vr)

$$Vc = \frac{Massa}{Densitas}$$

$$Vc = 18{,}4083\ m^3$$

Faktor Keamanan = 10%

$$Volume\ Tangki = \frac{Vc}{90\%}$$

$$Volume\ Tangki = 23{,}0103\ m^3 = 812{,}5862\ ft^3$$

2. Dimensi Reaktor (Dr)

- **Volume silinder, V_s**

$$V_s = \frac{\pi}{4}\ x\ D_t^2\ x\ H_s \qquad H_s = 1{,}5\ D_t$$

Maka,

$$V_s = \frac{\pi}{4}\ x\ \ 1{,}5\ D_t^3$$

$$V_s = \frac{3{,}14}{4}\ x\ \ 1{,}5\ D_t^3$$

Vs = $1{,}1775 D_t^3$

- **Volume Elipsoidol**

$$V_e = \frac{\pi}{24}\ x\ \ D_e^3$$

$$V_e = \frac{3{,}14}{24}\ x\ \ D_e^3$$

Ve = 0,131 D_e^3

- **Diameter tangki, D_t**

Vt = Vs + 2Ve
= 1,4392 Dt^2
Dt^3 = 57,2892 m^3
Dt = 3,8550 m

- **Tinggi tangki, H_t**

Tinggi silinder, H_s
Hs = 1,5 Dt
= 5,7825 m
= 227,6569 in
Tinggi Elipsoidal, He

He = 0,25 Dt

= 0,9637 m

= 37,9428 in

Tinggi total tangki, H_t

H_t = H_s + 2 H_e

= 9,74 m

- **Tinggi Cairan, H_c**

$$H_c = \frac{Volume\ Cairan}{Volume\ Tangki}\ x\ H_t$$

Hc = 7,7970 m

- **Tekanan Cairan, P_c**

$P_c = \rho\ .g\ .h_c$

Pc =$947{,}7843 \frac{kg}{m^3}\ x\ 9{,}8 \frac{m}{s} x\ 7{,}7970m = 72420{,}7354 \frac{kg}{m}.s^2$

Pc = 0,7025 atm

= 10,3265 *psi*

- **Tekanan Disain, P_d**

Pd = Pop + Pc

= 13 atm + 0,7025 atm

= 13,70 atm

= 201,4265 psi

- **Tebal dinding tangki, t_d**

$$t_d = \frac{PR}{SE - 0.6P} + C \qquad (Walas, Tabel\ 18.3, hal\ 625)$$

- Tekanan desain, P : 13,70 atm = 201,4265 psi
- Jari-jari tangki, R : 75,88 m
- *Allowable stress*, S : 18.700 psi (Peter, Tabel 4 Hal 538)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)
- Tahun digunakan : 10 tahun

$$\boxed{\frac{PR}{SE - 0.6P} + C}$$

Maka,

Td = 1,0428 in

= 0,0265 m

= 26,4866 mm

- **Tebal tutup ellipsoidal, t_e**

$$\frac{PD}{2SE - 0.2P} + C$$

(Wallas, Tabel 18,3.Hal 649)

Te = 0,98 in

= 0,025 m

- **Kebutuhan Katalis**

Densitas Katalis ρ = 83000 kg/m^3

Volume Reaktor Vr = 82,4486 m^3

Jumlah katalis (m) yang dibutuhkan = $m = \frac{\rho}{Vr}$

m = 1006,6874 kg

- **Menghitung Tinggi Timpukan Katalis, Z**

Diameter Tube, Du = 0,0035 × Dr

= 0,0035 X 3,8550

= 0,0135 m

= 0,5312 in

Dipilih tube ukuran standar (Mc.Cabe Lampiran 6) :

844 PROCESS HEAT TRANSFER

TABLE 11. DIMENSIONS OF STEEL PIPE (IPS)

Nominal pipe size, IPS, in.	OD, in.	Schedule No.	ID, in.	Flow area per pipe, in.²	Surface per lin ft, ft.²/ft.		Weight per lin ft, lb steel
					Outside	Inside	
1/8	0.405	40* 80†	0.269 0.215	0.058 0.036	0.106	0.070 0.056	0.25 0.32
1/4	0.540	40* 80†	0.364 0.302	0.104 0.072	0.141	0.095 0.079	0.43 0.54
3/8	0.675	40* 80†	0.493 0.423	0.192 0.141	0.177	0.129 0.111	0.57 0.74
1/2	0.840	40* 80†	0.622 0.546	0.304 0.235	0.220	0.163 0.143	0.85 1.09
3/4	1.05	40* 80†	0.824 0.742	0.534 0.432	0.275	0.216 0.194	1.13 1.48
1	1.32	40* 80†	1.049 0.957	0.864 0.718	0.344	0.274 0.250	1.68 2.17
1 1/4	1.66	40* 80†	1.380 1.278	1.50 1.28	0.435	0.362 0.335	2.28 3.00
1 1/2	1.90	40* 80†	1.610 1.500	2.04 1.76	0.498	0.422 0.393	2.72 3.64
2	2.38	40* 80†	2.067 1.939	3.35 2.95	0.622	0.542 0.508	3.66 5.03
2 1/2	2.88	40* 80†	2.469 2.323	4.79 4.23	0.753	0.647 0.609	5.80 7.67
3	3.50	40* 80†	3.068 2.900	7.38 6.61	0.917	0.804 0.760	7.58 10.3

IPS = 0,5 in sch 40

OD = 0,8400 in

ID = 0,6220 in

a" = 0,0333 ft^2

Perhitungan tinggi katalis dengan volume tube

$$V_u = \frac{m}{\rho\ katalis}$$

$$Z = \frac{4 \times m}{\pi \times ID^2 \times \rho\ katalis}$$

Dengan :

Z = Tinggi Tumpukan Katalis (m)

V = Volume Katalis dalam Tube (m^3)

m = Massa Katalis (kg)

$$Z = \frac{4 \times 1006{,}6874}{3{,}14 \times 0{,}0002 \times 8300}$$

Z = 619,0089 cm

Tinggi tumpukan katalis keseluruhan = 619,0089 cm

Dipilih tinggi tube 8 ft = 2,4390 m

Tinggi katalis per tube adalah z = 80% dari tinggi tube

Maka, z = 80% × 2,4390 m = 1,9512 m

3. Menghitung jumlah tube, Nt

$$\text{jumlah tube, Nt} = \frac{Tinggi\ katalis\ keseluruhan}{Tinggi\ katalis\ per\ tube}$$

$$Nt = \frac{619{,}0089}{1{,}9512}$$

Nt = 317,2421 buah *tube*

4. Menentukan design tube

Susunan tube = Tiangular Pitch

Bahan = *Stainless Steel*

IPS = 0,5 in sch 40

OD = 0,8400 in

ID = 0,6220 in

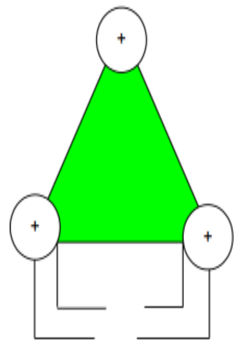
a" = 0,0333 ft²

Tinggi tumpukan Katalis = 1,9512 m

Panjang Pipa = 2,4390m

Susunan pipa yang digunakan adalah segitiga sama sisi dengan tujuan agar memberikan turbulensi yang lebih baik, sehingga akan, memperbesar koofiesien transfer panas dibandingkan susunan *square pitch* (Kern,1983).

Direncanakan *tube* disusun dengan pola *triangular pitch.*



Tebal Pipa = (OD-ID)/2

= (0,8400 - 0,6220)/2

= 0,1090 in

Jarak antar pusat pipa (PT)

PT = 1,25 × OD

PT = 1,25 × 0,8400 = 1,05 in

Jarak antar pipa (*Clearance*)

C' = PT-OD

C' = 1,0500 - 0,8400

C' = 0,21 in

Jarak Buffle, B = Dr × 0,3

= 3,8550 × 0,3

B = 1,1565 m

B = 45,5314 in

Area transfer panas dalam shell, As = ((PT -OD) × ID × B)/PT)

$$\textbf{As} = \frac{0{,}2100 \times 54{,}5770 \times 45{,}5314}{1{,}0500}$$

As = 496,9932 in^2

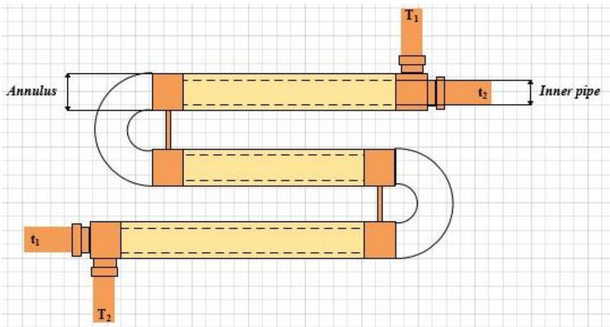
6. *Cooler* (C-2041)

Fungsi : Mendinginkan Produk Reaktor

Tipe : *Double Pipe*

Bahan : *Carbon Steel*

Gambar :



Data :

Fluida Panas **: Produk Reaktor**

- Laju alir massa : 11722 kg/jam = 6633,3 lb/jam
- Temperatur Masuk (T1) : 150°C = 302 F
- Temperatur Keluar (T2) : 60°C = 140 F

Fluida Dingin **: Air Pendingin**

- Laju alir massa : 141,07 kg/jam = 311,0128lb/jam
- Temperatur Masuk (t1) : 25°C = 77 F
- Temperatur Keluar (t2) : 45°C = 113 F

Menghitung ΔL LMTD *(Log Mean Temperature Difference)*

Fluida Panas	Temperatur (^{o}F)	Fluida Dingin	Selisih

T1	302	Temperatur tinggi	t2	113	189
T2	140	Temperatur rendah	t1	77	63

$$LMTD = \frac{\Delta t2 - \Delta t1}{ln\Delta t2 - \Delta t1}$$

LMTD = 114,6901426 F

Faktor Koreksi LMTD

$$R = \frac{T_1 - T_2}{t_2 - t_1} = 4{,}500$$

$$S = \frac{t_2 - t_1}{T - t_1} = 0{,}1600$$

Menentukan Luas Area Perpindahan Panas (A)

Dari tabel 8 D.Q KERN maka didapatkan koefisien perpindahan panas sebesar :

Ud : 450 Btu/jamft^{2}F

Q : 650994,1462 Kj/jam = 617142,4506 Btu/Jam

Sehingga :

$$A = \frac{Q}{U_D \; X \; \Delta t \; LMTD}$$

A = 145,8892 ft^2

Karena nilai A<200 ft^2, maka tipe *Heater* yang digunakan adalah ***Double Pipe*** (D.Q KERN, Hal.103)

Spesifikasi *Double Pipe*

TABLE 6.2. FLOW AREAS AND EQUIVALENT DIAMETERS IN DOUBLE PIPE EXCHANGERS

Exchanger, IPS	Flow area, in.²		Annulus, in.	
	Annulus	Pipe	d_e	d'_e
2 × 1¼	1.19	1.50	0.915	0.40
2½ × 1¼	2.63	1.50	2.02	0.81
3 × 2	2.93	3.35	1.57	0.69
4 × 3	3.14	7.38	1.14	0.53

Dari Tabel 6.2 D.Q Kern Hal.110 diperoleh *flow area* dan De pada *Double Pipe Heat Exchanger*, yaitu :

Exchanger, IPS	**Flow Area, in²**		**Annulus, in**	
	Annulus	**Pipe**	**De**	**De'**
4 X 3	3,14	7,38	1,14	0,53

	Anulus (in)	**(ft)**	**Inner Pipe (in)**	**(ft)**
IPS	4	0,3333	3	0,2499
SC	40		40	
OD (D2)	4,50	0,3749	1,66	0,1382
ID (D1)	4,026	0,3354	3,5	0,2916
a" (ft²/ft)	0,622	Ft²/ft	3,068	Ft²/ft

***Annulus* (Air Pendingin)**	***Inner Pipe* (Produk Reaktor)**
***1. Flow Area,* (aa)** D2 = 0,3749 ft D1 = 0,3354 $aa = \frac{\pi(D2^2 - D1^2)}{4}$ $aa = \frac{3,14\ (0,3749^2 - 0,3354^2)}{4}$	1. ***Flow Area, a_s*** D1 = 0,2916 ft $ap = \frac{\pi(D1^2)}{4}$ $ap = \frac{3,14(0,2916^2)}{4}$ $ap = 0,0667\ ft^2$

<table>
<tr>
<td>

$$aa = 0{,}0220\ ft^2$$

</td>
<td></td>
</tr>
<tr>
<td>

2. *Diameter Ekivalen (De)*

$$De = \frac{(D2^2 - D1^2)}{D1}$$

$$De = \frac{(0{,}3749^2 - 0{,}3354^2)}{0{,}3354}$$

$$De = 0{,}0836\ ft$$

</td>
<td></td>
</tr>
<tr>
<td>

3. *Mass Velocity* (Ga)

$$\text{Ga} = \frac{\text{W}}{aa}$$

$$\text{Ga} = \frac{311{,}0128\ lb/jam}{0{,}0220\ ft^2}$$

$$\text{Ga} = 14128{,}46\ \frac{lb}{h}.ft^2$$

</td>
<td>

2. *Mass Velocity* (Gp)

$$\text{Gp} = \frac{\text{W}}{ap}$$

$$\text{Gp} = \frac{25841{,}5153\ lb/jam}{0{,}0667\ ft^2}$$

$$\text{Gp} = 387277{,}466\frac{lb}{h}.ft^2$$

</td>
</tr>
<tr>
<td>

3. Penentuan nilai <u>μ liquid</u>

$$Rep = \frac{De}{\mu}\ X\ Ga$$

$$Rep = \frac{0{,}0836\ ft\ X 14128{,}46}{0{,}8900}$$

$$Rep = 548{,}510$$

</td>
<td>

4. Penentuan nilai <u>μ liquid</u>

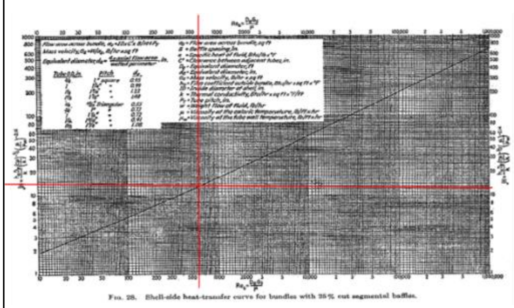
$$Rep = \frac{De}{\mu}\ X\ Gp$$

$$Rep = \frac{0{,}2961\ X\ 387277{,}46}{0{,}0294}$$

$$Rep = 1.584.663$$

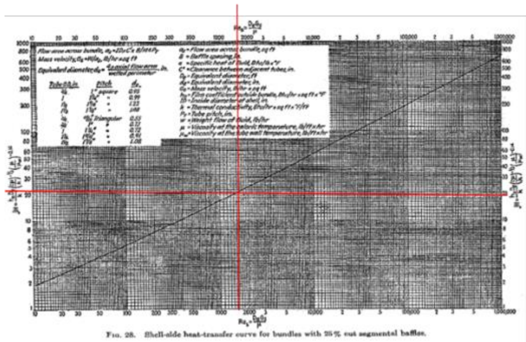
</td>
</tr>
<tr>
<td>

5. Faktor Perpindahan Panas (jH)

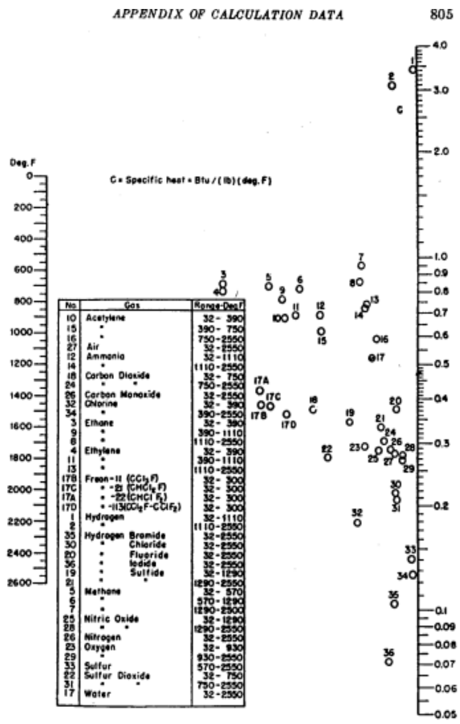
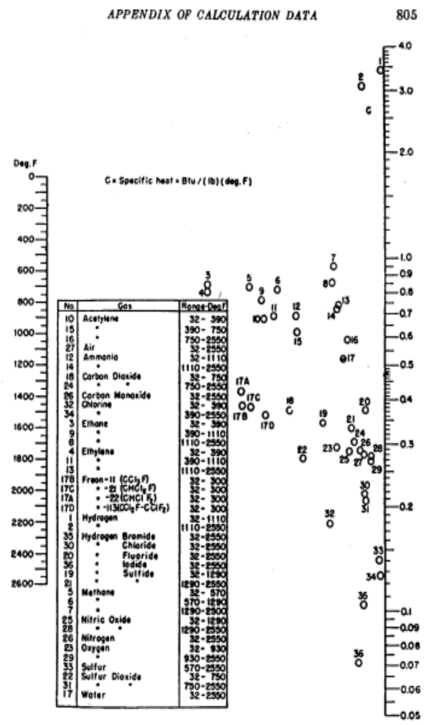


</td>
<td>

4. Faktor Perpindahan Panas (jH)



</td>
</tr>
</table>

Masukkan nilai Re = 548,510 ke fig. 4 Maka, didapatkan nilai factor perpindahan panas : 15	Masukkan nilai Re = 1.584.663 ke fig. 4 Maka, didapatkan nilai factor perpindahan panas : 22
5. Koefisien Perpindahan Panas (k) Tc = 35 F  **Sumber : DQ Kern. Fig.3, Hal. 805** Dari Fig.3 di atas, maka didapatkan : c = 0,0294 Btu/lb. ^{o}F k = 0,356 Btu/hr.ft. ^{o}F $\frac{C\mu}{k} = (\frac{0,0294\frac{Btu}{lb}.^\circ F\ X 2,1538\frac{lb}{ft}.hr}{0,356\frac{Btu}{hr}.ft.^\circ F})$ $\frac{C\mu}{k} = 0,1781$ $(\frac{C\mu}{k})^{\frac{1}{3}} = 0,5698$	**6. Koefisien Perpindahan Panas (k)** Tc = 221 F  **sumber : D.Q Kern. Fig. 3, Hal.805** dari Fig. 3 di atas, maka didapatkan : c = 0,25 Btu/lb.^{o}F k = 0,0510 btu/hr.ft. ^{o}F $\frac{C\mu}{k} = (\frac{0,25\frac{Btu}{lb}.^\circ F\ X\ 2,1538\frac{lb}{ft}.hr}{0,0510\frac{Btu}{hr}.ft.^\circ F})$ $\frac{C\mu}{k} = 10,5665$ $(\frac{C\mu}{k})^{\frac{1}{3}} = 2,19265$

7. Menghitung Ho

$$ho = jH\ X \frac{k}{De} (\frac{Cp.\mu}{k})^{1/3} (\frac{\mu}{\mu w})^{0,14}$$

$$ho = 15X \frac{0,356}{0,0836} X0,5698X\ 1$$

$$ho = 35,9535\ btu/h.ft^2.°F$$

8. Menghitung Hio

$$hi = 22\ X \frac{k}{De} (\frac{Cp.\mu}{k})^{1/3} (\frac{\mu}{\mu w})^{0,14}$$

$$hi = 22\ X \frac{0,0510}{0,0667} X2,19265X\ 1$$

$$hi = 46,33\ btu/h.ft^2.°F$$

KONVERSI hi ke hio

$$hio = hi\ X \frac{ID}{OD}$$

$$hio = 46,33\ X \frac{0,2916}{0,1382}$$
$$= 97,70\ btu/h.ft2.°F$$

9. *Clean Overall Coefficient,* U_c

$$Uc = \frac{h_{io} h_o}{h_{io} + h_o}$$

$$Uc = \frac{97,70\ X\ 35,9535}{97,70 + 35,9535}$$

$$Uc = 71,90\ btu/h.ft2.°F$$

10. *Design Overall Coefficient* (Ud)

$$\frac{1}{UD} = \frac{1}{UC} + Rd$$

$$\frac{1}{Ud} = \frac{1}{71,90} + 0,002$$

$$\frac{1}{Ud} = 0,016$$

$$Ud = 62,8661 \frac{Btu}{h} ft2.°F$$

11. Panjang Pipa yang dibutuhkan (L)

Ud = 62,8661Btu/jam.ft^{2}F

Q = 650994,1462 Btu/jam

a'' = 3,068 ft^2/ft

$$L = \frac{A}{a''} = \frac{91{,}20089\ ft^2}{3{,}068\ ft^2/ft}$$

$$L = 29{,}7565\ ft$$

Dipilih Panjang *hair pin* = 12 ft

Hairpin yang dibutuhkan (n) = $\frac{L}{2\ X\ Panjang\ Pipa}$ = 1,236

Hair pin yang digunakan sebanyak = 10 buah

12. *Required Surface*

A actual =29,7265 X 3,068

= 91,20089 ft^2

$$UD\ aktual = \frac{Q}{A\ X\ \Delta T\ LMTD}$$

$$UD\ aktual = 62{,}2374\ \text{btu/h. ft2. °F}$$

$$Rd\ = \frac{Uc - Ud}{Uc\ X\ Ud} = 0{,}002\ \text{hr. ft2. °F/Btu}$$

Pressure Drop : Annulus	***Pressure Drop : Pipe***
13. Rea De' = D2 – D1 = 0,3749 – 0,3354 = 0,0395 ft $Rea = \frac{Deq\ X\ Ga}{\mu}$ $Rea = \frac{0{,}0395X\ 14128{,}46}{2{,}1538}$	**3. Rep** Rep = 1.584.663,032 $f\ = 0{,}0009 + \frac{0{,}264}{401{,}7663} = 0{,}0016$

$Rea = 259{,}0078$ $f = 0{,}0021 + \frac{0{,}264}{14{,}14026}$ $f = 0{,}0221$	
4. $\Delta \boldsymbol{fa}$ $\Delta Fa = \frac{4\ X\ f\ X\ Ga^2 X\ La}{2\ X\ g\ X\ \rho^2 X\ D}$ $\Delta Fa = 7{,}00854\ ft$	*5.* $\Delta \boldsymbol{fp}$ $\Delta Fp = \frac{4\ X\ f\ X\ Gp^2 X\ L}{2\ X\ g\ X\ \rho^2 X\ D}$ $\Delta Fp = 0{,}00040 ft$
6. $\Delta \boldsymbol{pa}$ $\mathrm{V} = \frac{\mathrm{Ga}}{3600 \times \rho}$ $\mathrm{V} = \frac{14128{,}46}{3600 \times 476{,}9271} = 0{,}0082 ft/s$ $\Delta \mathrm{Ft} = 3\ \mathrm{X} \frac{\mathrm{V}^2}{2\mathrm{g}}$ $\Delta \mathrm{Ft} = 3\ \mathrm{X} \frac{6{,}77144}{64{,}4} = 3{,}1544\ ft$ $\Delta \mathrm{pa} = \frac{(\Delta \mathrm{Fa} + \Delta \mathrm{Ft})\rho}{144}$ $\Delta \mathrm{pa} = 0{,}00024257\ psi$ **Memenuhi karena <10 psi**	*7.* $\Delta \boldsymbol{Pp}$ $\Delta \mathrm{Pp} = \frac{(\Delta \mathrm{Fp})\ \mathrm{X}\ \rho}{144}$ $\Delta \mathrm{Pp} = 5{,}42802\ \mathrm{psi}$ **Memenuhi karena <10 psi**

7. Storage Tank (ST-2002)

Tangki Penyimpanan Gliserol (ST-1101)

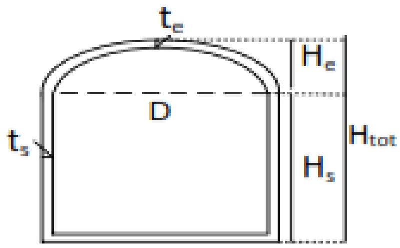
Fungsi : Tempat menyimpan Acetol

Bahan konstruksi : *Stainless Steel*

Jumlah : 1 unit

Lama Penyimpanan : 7 Hari

Gambar :



Parameter	M (kg/jam)	ρ (kg/m3)	V (m3/jam)	%	Xi / p (kg/m3)	densitas mix
$C_3H_8O_3$	225,053	1261	0,178472117	1,92	0,000015	
H_2O	2626,43	997	2,634340729	22,40695652	0,0000	938,7962
$C_3H_6O_2$	8870,04	917	9,672886653	75,67304348	0,0008	
Total	**11721,528**		**12,4856**	**100**	**0,001065**	**938,7962**

Data :

Laju alir = 11721,528kg/jam = 25841,281 lb/jam

Volumetrik = 12,4856 m^3/jam = 0,1224 ft^3/s

Densitas mix = 938,7962 kg/m^3 = 58,6090 lb/ft^3

T = 60^0C

t = 7 hari = 168 jam

P = 1 atm = 14,7 psi

Kapasitas tangki, V_t

$$V_b = \frac{m \times t}{\rho} = \frac{11721{,}528 \frac{kg}{jam} x\ 168\ jam}{938{,}7962 ^{kg}/_{m^3}}$$

= 196.216,722 m^3

= 69.540.919,32 ft^3

= 18.370.763 gal

Faktor keamanan 10%

Maka,

$V_p = 0{,}9\ V_t$

$V_t = \frac{V_p}{0{,}9}$

$= \frac{196.216{,}722\ m^3}{0{,}9}$

$= 2.188.018{,}58\ m^3$

$= 7.726.768{,}13\ ft^3$

$= 578.013.244{,}3\ gal$

Dimensi tangki

- **Volume silinder, V_s**

  $V_s = \frac{\pi}{4} \times D_t^2 \times H_t$ $H_t = 1{,}5D_t$

Maka,

$V_s = \frac{\pi}{4} \times D_t^3$

- **Volume ellipsoidal, V_e**

  $V_e = \frac{\pi}{24} \times D_t^3$ He = ¼ Dt (peter's, Tabel 4)

  $V_e = 0{,}1308 \times D_t^3$

- **Diameter tangki, D_t**

  $V_t = V_s + V_e$

  $= \left(\frac{1{,}5\pi}{4} \times D_t^3\right) + (0{,}1308 \times D_t^3)$

  $V_t = 1{,}3083 \times D_t^3$

  $D_t^3 = \frac{Vt}{1{,}3083}$

  $D_t^3 = \frac{2.188.018{,}58}{1{,}3083}$

  $D_t^3 = 1.672.413{,}498\ m^3$

  $D_t = \sqrt[3]{1.672.413{,}498}$

  $= 118{,}13\ m$

  $= 387{,}47\ ft$

= 4650,91 in

- **Tinggi tangki, H_t**

Tinggi silinder, Hs = 1,5 D_t

= 177,20 m

= 581,21 ft

= 6976,37 in

Tinggi ellipsoidal, H_e = 1/4D_t

= 29,53 m

= 96,86 ft

= 1162,72 in (walas, Tabel 18.5)

Tinggi total, H_t = tinggi silinder + tinggi ellipsoidal

H_t = 177,20 m + 29,53 m

= 206,73 m

= 206733,8 mm

- **Tinggi cairan dalam tangki, H_c**

H_c $= \frac{volume\ bahan}{volume\ tangki} \times H_t$

$= \frac{1969216{,}722\ \ m^3}{206{,}733\ m^3} \times 206\ m$

= 186 m

= 610,2781 ft

= 7325,1987 in

- **Tekanan cairan dalam tangki, P_c**

P_c $= \rho \times g \times H_c$

= 1823,39212 kg/m s^2

= 0,0176 atm

= 2,522 psi

- **Tekanan desain, P_d**

  Pd = Pop + Pc

  = (1 + 0,0176) atm

  = 1,0176 atm

  = 14,95 psi

- **Tebal dinding tangki, t_d**

  $t_d = \frac{PR}{SE-0,6P} + C$ (Walas, Tabel 18.4)

- Tekanan desain, P_d : 2,552 psi
- Jari-jari, R : 2325,45 in
- *Allowable stress*, S : 13700 psi (Walas, Tabel 18.4)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)
- Lama tahun digunakan : 10 tahun

Maka,

$$td = \frac{2,552\ psi \times 2325,45 in}{(13700\ psi \times 0,85) - (0,6 \times 2,552\ \text{psi})} + 0,002 \frac{in}{tahun} \times 10 tahun$$

= 0,53 in

= 0,013 m

= 0,044 ft

= 13,46 mm

- **Tebal dinding ellipsoidal,t_e**

$$\boxed{t_e = \frac{PD_t}{2SE - 0,2P} + C}$$

(*Walas-Chemical Process Equipment*, Tabel 18.4)

$$t_e = \frac{2,552\ psi \times 6976,37 in}{2(13700\ psi \times 0,85) - (0,2 \times 2,552\ \text{psi})} + 0,002 \frac{in}{tahun} x 10 tahun$$

= 0,784 in

= 0,0199 m

= 0,065 ft

= 19,93 mm

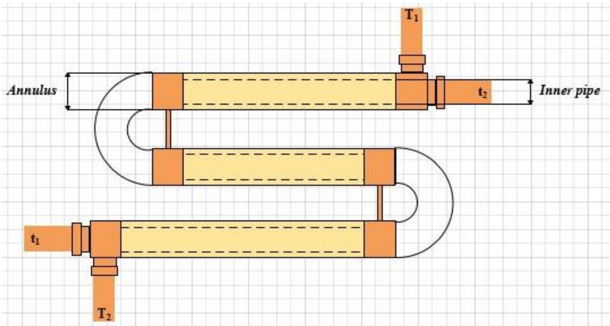
8. ***Heat Exchanger* (HE-2022)**

Fungsi : Memanaskan Acetol sebelum masuk Reaktor

Tipe : *Double Pipe*

Bahan : *Carbon Steel*

Gambar :



Data :

Fluida Panas : ***Steam***

- Laju alir massa : 3008,82 kg/jam = 6633,3 lb/jam
- Temperatur Masuk (T1) : 250°C = 482 F
- Temperatur Keluar (T2) : 250°C = 482 F

Fluida Dingin **: Acetol**

- Laju alir massa : 11721,52 kg/jam = 25841,51529 lb/jam
- Temperatur Masuk (t1) : 60°C = 140 F
- Temperatur Keluar (t2) : 210°C = 410 F

Menghitung ΔL LMTD *(Log Mean Temperature Difference)*

Fluida Panas		**Temperatur (^{o}F)**	**Fluida Dingin**		**Selisih**
T1	482	Temperatur tinggi	t2	410	72
T2	482	Temperatur rendah	t1	140	342

$$LMTD = \frac{\Delta t2 - \Delta t1}{ln\Delta t2 - \Delta t1}$$

LMTD = 173,283 F

Faktor Koreksi LMTD

$$R = \frac{T_1 - T_2}{t_2 - t_1} = 0,000$$

$$S = \frac{t_2 - t_1}{T - t_1} = 0,7895$$

Menentukan Luas Area Perpindahan Panas (A)

Dari tabel 8 D.Q KERN maka didapatkan koefisien perpindahan panas sebesar :

Ud : 150 Btu/jamft^{2}F

Q : 516382,02 Kj/jam = 4895075,78 Btu/Jam

Sehingga :

$$A = \frac{Q}{U_D \ X \ \Delta t \ LMTD}$$

A = 188,33 ft^2

Karena nilai A<200 ft^2, maka tipe *Heater* yang digunakan adalah ***Double Pipe*** (D.Q KERN, Hal.103)

Spesifikasi *Double Pipe*

TABLE 6.2. FLOW AREAS AND EQUIVALENT DIAMETERS IN DOUBLE PIPE EXCHANGERS

Exchanger, IPS	Flow area, in.^{2}		Annulus, in.	
	Annulus	Pipe	d_e	d_e'
2 × 1¼	1.19	1.50	0.915	0.40
2½ × 1¼	2.63	1.50	2.02	0.81
3 × 2	2.93	3.35	1.57	0.69
4 × 3	3.14	7.38	1.14	0.53

Dari Tabel 6.2 D.Q Kern Hal.110 diperoleh *flow area* dan De pada *Double Pipe Heat Exchanger*, yaitu :

Exchanger, IPS	**Flow Area, in^2**		**Annulus, in**	
	Annulus	**Pipe**	**De**	**De'**
2 X 1,25	1,19	1,5	0,915	0,4

	Anulus (in)	**(ft)**	**Inner Pipe (in)**	**(ft)**
IPS	2	0,1666	1,25	0,1041
SC	40	3,3320	40	3,3320
OD (D2)	2,38	0,1983	1,66	0,1383
ID (D1)	2,067	0,1722	1,38	0,1150
a" (ft^2/ft)	0,6220	0,0518	0,435	0,0362

***Annulus* (Steam)**	***Inner Pipe* (Fluida Dingin)**
15. ***Flow Area,* (aa)** D2 = 0,1983 in = 0,1983 ft D1 = 2,067in = 0,1722 ft $$aa = \frac{\pi(D2^2 - D1^2)}{4}$$ $$aa = \frac{3,14\ (0,1983^2 - 0,1722^2)}{4}$$ $$aa = 0,0076\ ft^2$$	**8.** ***Flow Area, a_s*** D1 = 0,1150 ft $$ap = \frac{\pi(D1^2)}{4}$$ $$ap = \frac{3,14(0,1150^2)}{4}$$ $$ap = 0,0104 ft^2$$
16. ***Diameter Ekivalen (De)*** $$De = \frac{(D2^2 - D1^2)}{D1}$$ $$De = \frac{(0,1983^2 - 0,1722^2)}{0,1983}$$	

<table>
<tr><td>

$$De = 0{,}561\ ft$$

</td><td></td></tr>
<tr><td>

17. ***Mass Velocity*** **(Ga)**

$$\text{Ga} = \frac{\text{W}}{aa}$$

$$\text{Ga} = \frac{6633{,}3\ lb/jam}{0{,}0076\ ft^2}$$

$$\text{Ga} = 874900\ \frac{lb}{h}.ft^2$$

</td><td>

9. Mass Velocity **(Gp)**

$$\text{Gp} = \frac{\text{W}}{ap}$$

$$\text{Gp} = \frac{25841{,}5153\ lb/jam}{0{,}0104\ ft^2}$$

$$\text{Gp} = 2491151{,}52\frac{lb}{h}.ft^2$$

</td></tr>
<tr><td>

18. Penentuan nilai μ liquid

$$Rep = \frac{De}{\mu}\ X\ Ga$$

$$Rep = \frac{0{,}561\ ft\ X\ 874900}{0{,}0228}$$

$$Rep = 889.455{,}51$$

</td><td>

19. Penentuan nilai μ liquid

$$Rep = \frac{De}{\mu}\ X\ Gp$$

$$Rep = \frac{0{,}1150\ X\ 2491151}{0{,}5524}$$

$$Rep = 214.212{,}31$$

</td></tr>
<tr><td>

20. Faktor Perpindahan Panas (jH)

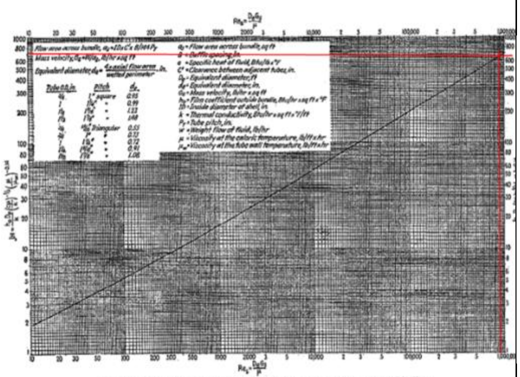


Fig. 28. Shell-side heat-transfer curve for bundles with 25% cut segmental baffles.

Masukkan nilai Re = 889.455,51 ke fig. 4 Maka, didapatkan nilai factor perpindahan panas : 750

</td><td>

21. Faktor Perpindahan Panas (jH)

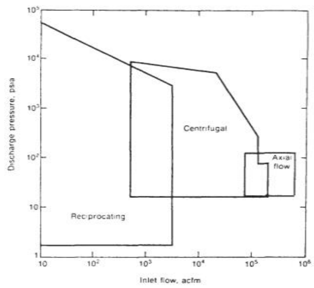


Figure 1. Approximate ranges of application for reciprocating, centrifugal, and axial-flow compressors.

Masukkan nilai Re = 214.212,31ke fig. 4 Maka, didapatkan nilai factor perpindahan panas : 300

</td></tr>
<tr><td>

22. Koefisien Perpindahan Panas (k)

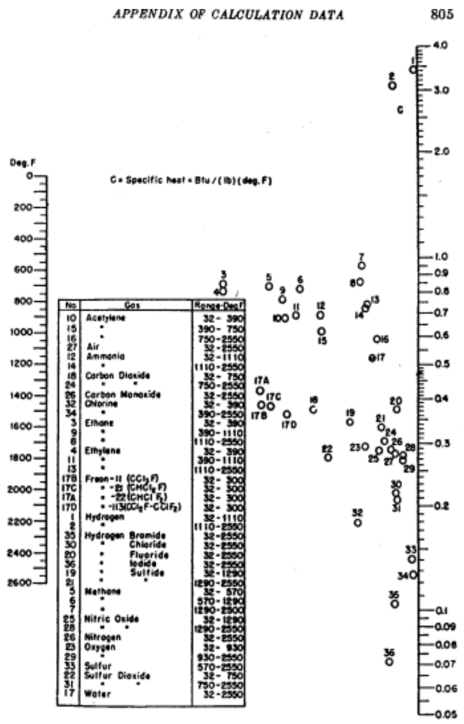
Tc = 120 F

</td><td>

23. Koefisien Perpindahan Panas (k)

Tc = 86 F

</td></tr>
</table>



Sumber : DQ Kern. Fig.3, Hal. 805

Dari Fig.3 di atas, maka didapatkan :

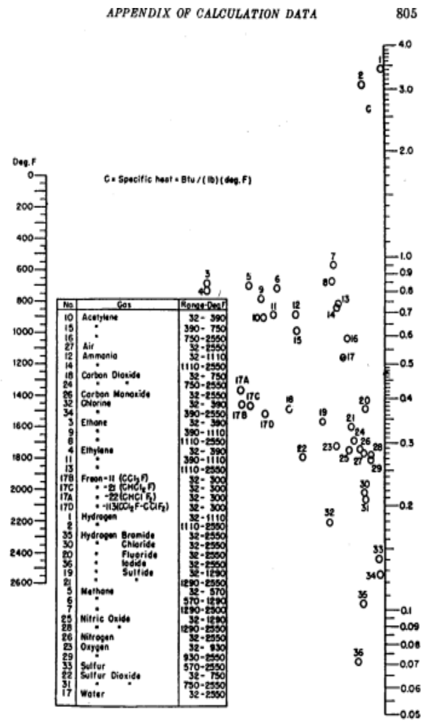
c = 0,48 Btu/lb. °F

k = 0,0163 Btu/hr.ft. °F

$$\frac{C\mu}{k} = (\frac{0{,}48\frac{Btu}{lb}.°F\ X\ 0{,}0552\frac{lb}{ft}.hr}{0{,}0163\frac{Btu}{hr}.ft.°F})$$

$$\frac{C\mu}{k} = 1{,}6248$$

$$(\frac{C\mu}{k})^{\frac{1}{3}} = 1{,}1754$$



sumber : D.Q Kern. Fig. 3, Hal.805

dari Fig. 3 di atas, maka didapatkan :

c = 0,57 Btu/lb.°F

k = 68 btu/hr.ft. °F

$$\frac{C\mu}{k} = (\frac{0{,}57\frac{Btu}{lb}.°F\ X\ 1{,}3368\ \frac{lb}{ft}.hr}{68\frac{Btu}{hr}.ft.°F})$$

$$\frac{C\mu}{k} = 0{,}0112$$

$$(\frac{C\mu}{k})^{\frac{1}{3}} = 0{,}2241$$

24. Menghitung Ho

$$ho = jH\ X \frac{k}{De}(\frac{Cp.\mu}{k})^{1/3}(\frac{\mu}{\mu w})^{0,14}$$

$$ho = 750X\frac{0{,}0163}{0{,}0561}X\ 1{,}1754\ X\ 1$$

$$ho = 256{,}1713\ btu/h.ft^2.°F$$

10. Menghitung Hio

$$hi = 190\ X\frac{k}{De}(\frac{Cp.\mu}{k})^{1/3}(\frac{\mu}{\mu w})^{0,14}$$

$$hi = 300\ X\frac{68}{0{,}1150}X\ 0{,}2241\ X\ 1$$

$$hi = 39.771{,}46\ btu/h.ft^2.°F$$

	KONVERSI hi ke hio $$hio = hi\,X\,\frac{ID}{OD}$$ $$hio = 39771{,}46\,X\,\frac{0{,}1150}{0{,}1383}$$ $$= 33063\ btu/h.ft2.°F$$

25. *Clean Overall Coefficient,* U_c

$$Uc = \frac{h_{io}h_o}{h_{io} + h_o}$$

$$Uc = \frac{33063\,X\,256{,}1713}{33063 + 256{,}1713}$$

$$Uc = 512{,}34\ btu/h.ft2.°F$$

26. *Design Overall Coefficient* (Ud)

$$\frac{1}{UD} = \frac{1}{UC} + Rd$$

$$\frac{1}{Ud} = \frac{1}{512{,}34} + 0{,}002$$

$$\frac{1}{Ud} = 0{,}0040$$

$$Ud = 253{,}0481\frac{Btu}{h}ft2.°F$$

27. Panjang Pipa yang dibutuhkan (L)

Ud = 253,0481 Btu/jam.ft^2F

Q = 4895075,76 Btu/jam

a’’ = 0,4350 ft^2/ft

$$L = \frac{A}{a''} = \frac{111{,}6350\ ft^2}{0{,}4350\ ft^2/ft}$$

$$L = 256{,}6322\ ft$$

Dipilih Panjang *hair pin* = 20 ft

Hairpin yang dibutuhkan (n) = $\frac{L}{2\ X\ Panjang\ Pipa}$ = 6,4158

Hair pin yang digunakan sebanyak = 6 buah

28. *Required Surface*

A actual = 256,6322 X 0,4350

= 111,6350 ft^2

$$UD\ aktual = \frac{Q}{A\ X\ \Delta T\ LMTD}$$

$$UD\ aktual = 253{,}0481\ \text{btu/h. ft2. °F}$$

$$Rd\ = \frac{Uc - Ud}{Uc\ X\ Ud} = 0{,}0020\ \text{hr. ft2. °F/Btu}$$

Pressure Drop : Annulus	***Pressure Drop : Pipe***
14. Rea De' = D2 – D1 = 0,1983 – 0,1722 = 0,0261 ft $Rea = \frac{Deq\ X\ Ga}{\mu}$ $Rea = \frac{0{,}0261\ X\ 874.900{,}07}{0{,}0552}$ $Rea = 413.425{,}8021$ $f = 0{,}0035 + \frac{0{,}264}{315{,}2334}$ $f = 0{,}0043$	**5. Rep** Rep = 214.212,31 $f = 0{,}0035 + \frac{0{,}264}{173{,}36} = 0{,}0050$
6. Δfa $\Delta Fa = \frac{4\ X\ f\ X\ Ga^2 X\ La}{2\ X\ g\ X\ \rho^2 X\ D}$	***8.*** **Δfp** $\Delta Fp = \frac{4\ X\ f\ X\ Gp^2 X\ L}{2\ X\ g\ X\ \rho^2 X\ D}$

$\Delta Fa = 2{,}2435\ ft$	$\Delta Fp = 60{,}1211 ft$
9. $\boldsymbol{\Delta pa}$ $\mathrm{V} = \dfrac{\mathrm{Ga}}{3600 \times \rho}$ $\mathrm{V} = \dfrac{874.800{,}07}{3600 \times 264} = 0{,}92 ft/s$ $\Delta \mathrm{Ft} = 3\ \mathrm{X} \dfrac{\mathrm{V}^2}{2\mathrm{g}}$ $\Delta \mathrm{Ft} = 3\ \mathrm{X} \dfrac{0{,}85}{64{,}4} = 0{,}04\ ft$ $\Delta \mathrm{pa} = \dfrac{(\Delta \mathrm{Fa} + \Delta \mathrm{Ft})\rho}{144}$ $\Delta \mathrm{pa} = 4{,}19\ psi$ **Memenuhi karena <10 psi**	***10.*** $\boldsymbol{\Delta Pp}$ $\Delta \mathrm{Pp} = \dfrac{(\Delta \mathrm{Fp})\ \mathrm{X}\ \rho}{144}$ $\Delta \mathrm{Pp} = 10{,}9120\ \mathrm{psi}$ **Memenuhi karena <10 psi**

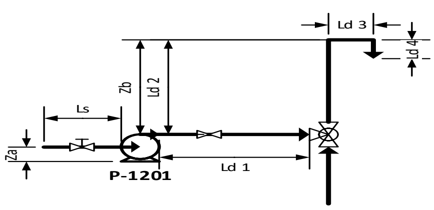
9. Pompa (P-2012)

Fungsi : Mengalirkan Acetol ke Reaktor

Tipe : *Centrifugal Pumps*

Bahan : *Carbon Steel*

Gambar :



Data :

- Laju alir massa, m : 11721,5281 kg/jam = 7,1781 lb/dt
- Densitas campuran, ρ : 939 kg/m^3 = 58,6090 lb/ft^3
- Viskositas Campuran, μ : 0,5524 cP = 0,00037 lb/ft.s

- Tinggi pompa terhadap cairan masuk, Za : 0 m = 0 ft
- Tinggi pompa terhadap cairan keluar, Zb : 8 m = 26,24 ft
- Panjang pipa hisap, Ls : 5 m = 16,4 ft
- Panjang pipa buang, Ld : 10 m = 32,8 ft
- Faktor keamanan 10%

Pemilihan faktor keamanan pada pompa *centrifugal pump* dapat dilihat pada Gambar 10.1

TABLE 6
Factors in equipment scale-up and design

Type of equipment	Is pilot plant usually neces-sary?	Major variables for operational design (other than flow rate)	Major variables characterizing size or capacity	Maximum scale-up ratio based on indi-cated char-acterizing variable	Approxi-mate recom-mended safety or over-design factor, %
Agitated batch crystallizers	Yes	Solubility-temperature relationship	Flow rate Heat transfer area	>100:1	20
Batch reactors	Yes	Reaction rate Equilibrium state	Volume Residence time	>100:1	20
Centrifugal pumps	No	Discharge head	Flow rate Power input Impeller diameter	>100:1 >100:1 10:1	10

Sumber : (Peter, Pers 14.15 Hal 496)

Laju alir volumetrik, Q_v

$$Q_p = \frac{m}{0,9}$$

$$= \frac{7,1781\ lb/s}{0,9} = 7,9757\ \text{lb/s}$$

$$Q_v = \frac{Q_p}{\rho}$$

$$= \frac{7,9757\ lb/s}{58,6090\ lb/ft^3}$$

$$= 0,1360\ \text{ft}^3/\text{s}$$

$$= 61,082\ \text{gal/min}$$

Diameter optimum, D_{opt}

Asumsi aliran turbulen

$D_{opt} = 3,9*Q_v^{0,45}*\rho^{0,13}$ (Peter, Pers 14.15 Hal 496)

Untuk mendapatkan rumus diameter optimum pada aliran turbulen dapat dilihat pada Gambar 10.2

For turbulent flow ($N_{Re} > 2100$) in steel pipes

$$D_{i,opt} = 3.9 q_f^{0.45} \rho^{0.13} \quad (15)$$

For viscous flow ($N_{Re} < 2100$) in steel pipes

$$D_{i,opt} = 3.0 q_f^{0.36} \mu_c^{0.18} \quad (16)$$

MATERIALS TRANSFER, HANDLING, AND TREATMENT EQUIPMENT 497

$D_{i\ opt}$ = optimum inside pipe diameter, in.
q_f = fluid flow rate, ft^3/s
ρ = fluid density, lb/ft^3
μ_c = fluid viscosity, centipoises

Sumber : (Peter, Pers 14.15 Hal 496)

$D_{opt} = 3,9*Q_v^{0,45}*\rho^{0,13}$

$= 3,9*(0,1360)^{0,45}*(58,6090)^{0,13}$

$= 2,6984$ in

Berdasarkan Tabel 11 Kern, diperoleh pipa baja dengan ukuran sebagai berikut,

	Suction (a)				Discharge (b)			
IPS	3 in sch 40							
ID	3,0680	in	0,2557	ft	3,0680	in	0,2557	ft
OD	3,500	In	0,2917	ft	3,500	in	0,2917	ft
a"	0,0513				ft^2			

Kecepatan aliran, V

Va = Vb, karena ukuran pipa hisap dan pipa buang sama

$V = \frac{Q_v}{a"}$

$$= \frac{0{,}1361\, ft^3/dt}{0{,}0513\ ft^2} = 2{,}6553 \text{ ft/s}$$

$$\frac{V^2}{2g_c} = \frac{(2{,}6553)^2\ ft/s}{2 \times 32{,}17\ ftlbm/lbfs^2} = 0{,}1096 \text{ ft-lb}_f\text{/lb.m}$$

Bilangan Reynolds, N_{Re}

$$N_{Re} = \frac{\rho \times V \times D}{\mu} \qquad \text{(Mc Cabe, pers 3.8)}$$

$$N_{Re} = \frac{\rho \times V \times D}{\mu}$$

$$= \frac{58{,}6091\ \frac{lb}{ft^3} \times 2{,}6553 \frac{ft}{dt} \times 0{,}2557 ft}{0{,}00037\ \frac{lb}{ft.hr}} = 107.180{,}44$$

- **Rugi Gesek pada Pipa hisap (*suction*)**

  Pada pipa hisap, rugi gesek timbul akibat gesekan dengan kulit pipa

The hydraulic radius is a useful parameter for generalizing fluid-flow phenomena in turbulent flow. Equation (5.7) can be so generalized by substituting $4r_H$ for D or $2r_H$ for r_w:

$$h_{fs} = \frac{\tau_w}{\rho r_H} \Delta L = \frac{\Delta p_s}{\rho} = f \frac{\Delta L}{r_H} \frac{\bar{V}^2}{2g_c} \qquad (5.56)$$

$$N_{Re} = \frac{4r_H \bar{V} \rho}{\mu} \qquad (5.57)$$

The simple hydraulic-radius rule does not apply to laminar flow through noncircular sections. For laminar flow through an annulus, for example, f and N_{Re} are related by the equation[3]

$$f = \frac{16}{N_{Re}} \phi_a \qquad (5.58)$$

$$r_H = \frac{S}{L_p} \qquad (5.54)$$

where S = cross-sectional area of channel
L_p = perimeter of channel in contact with fluid

Thus, for the special case of a circular tube, the hydraulic radius is

$$r_H = \frac{\pi D^2/4}{\pi D} = \frac{D}{4}$$

The equivalent diameter is $4r_H$, or simply, D.
An important special case is the annulus between two concentric pipes. Here the hydraulic radius is

$$r_H = \frac{\pi D_o^2/4 - \pi D_i^2/4}{\pi D_i + \pi D_o} = \frac{D_o - D_i}{4} \qquad (5.55)$$

$$h_{fsa} = f \frac{\Delta L}{r_H} \frac{V^2}{2g_c} \qquad \text{(Mc Cabe, Pers 5.56)}$$

$$r_H = \frac{ID}{4} \qquad \text{(Mc Cabe, Hal 103)}$$

$= \frac{0,2557 ft}{4}$

$= 0,0639$ ft

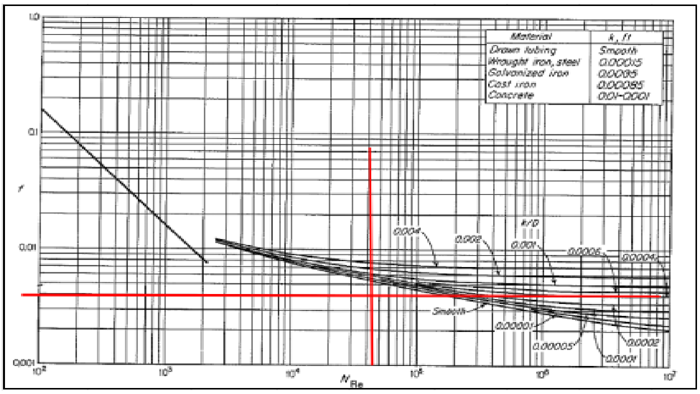
N_{Re} $= 107.180,44$

Material yang digunakan untuk konstruksi pipa adalah *carbon steel pipe*, dimana

K $= 0,00015$ ft (Mc Cabe, Fig. 5.9)

$\frac{k}{ID}$ $= \frac{0,00015\ ft}{0,2557\ ft} = 0,0006$

f $= 0,004$ (Mc Cabe, Fig. 5.9)



Maka,

h_{fsa} $= 0,004 \times \frac{16,4000 ft}{0,0639\ ft} \times 0,1096\ ft.lb_f/lb$

$= 0,1125$ ft-lb_f/lb

- Rugi gesek akibat *fitting* dan *valve*

h_{ff} $= K_f \frac{V^2}{2g_c}$ (Mc Cabe, Pers 5.67)

EFFECT OF FITTINGS AND VALVES. Fittings and valves disturb the normal flow lines and cause friction. In short lines with many fittings, the friction loss from the fittings may be greater than that from the straight pipe. The friction loss h_{ff} from fittings is found from an equation similar to Eqs. (5.59) and (5.65):

$$h_{ff} = K_f \frac{\bar{V}_a^2}{2g_c} \qquad (5.67)$$

where K_f = loss factor for fitting
$\bar{V}_a$ = average velocity in pipe leading to fitting

Factor K_f is found by experiment and differs for each type of connection. A short list of factors is given in Table 5.1.

K_f (*gate valve*) = 0,2 = 1 unit (Mc Cabe, Tabel 5.1)

TABLE 5.1
Loss coefficients for standard threaded pipe fittings†

Fitting	K_f
Globe valve, wide open	10.0
Angle valve, wide open	5.0
Gate valve	
Wide open	0.2
Half open	5.6
Return bend	2.2
Tee	1.8
Elbow	
90°	0.9
45°	0.4

† From J. K. Vennard, in V. L. Streeter (ed.), *Handbook of Fluid Dynamics*, McGraw-Hill Book Company, New York, 1961, p. 3-23.

Total K_f = (1 x 0,2) = 0,2

Maka,

h_{ff} $= 0{,}2 \times 0{,}1096\ ft.lb_f/lb$ = 0,0219 ft-lb$_f$/lb

- **Rugi Gesek pada Pipa buang (*discharge*)**

Pada pipa buang, rugi gesek timbul akibat gesekan dengan kulit pipa, pengaruh *fitting* dan *valve*.

- Pada pipa hisap, rugi gesek timbul akibat gesekan dengan kulit pipa

h_{fsb} $= f\,\dfrac{\Delta L}{r_H}\,\dfrac{V^2}{2g_c}$ (Mc Cabe, Pers 5.56)

r_H $= \dfrac{ID}{4}$ (Mc Cabe, Hal 103)

$= \dfrac{0{,}2557\ ft}{4}$ = 0,0639 ft

N_{Re} = 107.180,44

Material yang digunakan untuk konstruksi pipa adalah *carbon steel pipe*, dimana

k = 0,00015 ft (Mc Cabe, Fig. 5.9)

$\dfrac{k}{ID}$ $= \dfrac{0{,}00015 ft}{0{,}2557 ft}$ = 0,0006

f = 0,004 (Mc Cabe, Fig. 5.9)

Maka,

h_{fsb} $= 0{,}004 \times \frac{32{,}8000\ ft}{0{,}0639\ ft} \times 0{,}1096 ft.lb_f/lb$

=0,2249 ft-lb$_f$/lb

- Rugi gesek akibat *fitting* dan *valve*

h_{ffb} $= K_f \frac{V^2}{2g_c}$ (Mc Cabe, Pers 5.67)

EFFECT OF FITTINGS AND VALVES. Fittings and valves disturb the normal flow lines and cause friction. In short lines with many fittings, the friction loss from the fittings may be greater than that from the straight pipe. The friction loss h_{ff} from fittings is found from an equation similar to Eqs. (5.59) and (5.65):

$$h_{ff} = K_f \frac{\bar{V}_a^2}{2g_c} \quad (5.67)$$

where K_f = loss factor for fitting
$\bar{V}_a$ = average velocity in pipe leading to fitting

Factor K_f is found by experiment and differs for each type of connection. A short list of factors is given in Table 5.1.

K_f (*elbow* 90°) = 0,9 = 2 unit (Mc Cabe, Tabel 5.1)

K_f (*globe valve*) = 10 = 1 unit (Mc Cabe, Tabel 5.1)

TABLE 5.1
Loss coefficients for standard threaded pipe fittings†

Fitting	K_f
Globe valve, wide open	10.0
Angle valve, wide open	5.0
Gate valve	
Wide open	0.2
Half open	5.6
Return bend	2.2
Tee	1.8
Elbow	
90°	0.9
45°	0.4

† From J. K. Vennard, in V. L. Streeter (ed.), *Handbook of Fluid Dynamics*, McGraw-Hill Book Company, New York, 1961, p. 3-23.

Total K_f = (2x0,9) + (1x10) = 11,8

Maka,

h_{ffb} $= 11{,}8 \times 0{,}1096\ ft.lb_f/lb$ = 1,2931 ft-lb$_f$/lb

Sehingga, total rugi gesek adalah

$= h_{fsa} + h_{ff} + h_{fsb} + h_{ff}$

h_f = 1,6524 ft-lb_f/lb

Daya pompa (BHP)

Daya pompa dihitung menggunakan Persamaan Bernoulli :

(Mc.Cabe, pers 4.32)

$$\frac{P_a}{\rho} + \frac{gZ_a}{g_c} + \frac{\alpha_a V_a^2}{2g_c} + \eta W_p = \frac{P_b}{\rho} + \frac{gZ_b}{g_c} + \frac{\alpha_b V_b^2}{2g_c} + h_f$$

fluid is $W_p - h_{fp}$. In practice, in place of h_{fp} a pump efficiency denoted by η is used, defined by the equation

$$W_p - h_{fp} \equiv \eta W_p$$

or

$$\eta = \frac{W_p - h_{fp}}{W_p} \qquad (4.31)$$

The mechanical energy delivered to the fluid is, then, ηW_p, where $\eta < 1$. Equation (4.29) corrected for pump work is

$$\frac{p_a}{\rho} + \frac{gZ_a}{g_c} + \frac{\alpha_a \bar{V}_a^2}{2g_c} + \eta W_p = \frac{p_b}{\rho} + \frac{gZ_b}{g_c} + \frac{\alpha_b \bar{V}_b^2}{2g_c} + h_f \qquad (4.32)$$

Equation (4.32) is a final working equation for problems on the flow of incompressible fluids.

Atau

$$\eta W_p = \left(\frac{P_b}{\rho} + \frac{gZ_b}{g_c} + \frac{\alpha_b V_b^2}{2g_c}\right) - \left(\frac{P_a}{\rho} + \frac{gZ_a}{g_c} + \frac{\alpha_a V_a^2}{2g_c}\right) + h_f$$

Dimana

$P_a = P_b$

$V_a = Vb$

$\rho_a = \rho_b$

$g/g_c = 1$

$\alpha_a = \alpha_b$

$\eta = \frac{58+8}{2} = 43$ % (Peters, Fig. 14.37)

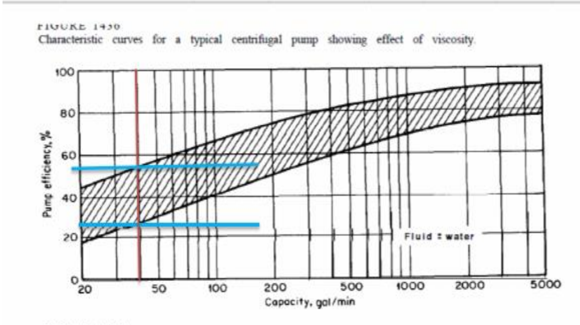


FIGURE 1437
Efficiencies of centrifugal pumps.

Sehingga persamaan di atas dapat disederhanakan menjadi :

$\eta W_p = (Z_b - Z_a) + h_f$

$0{,}43\ W_p = (26{,}24 - 0)\text{ft} + 1{,}6524\ \text{ft-lb}_f/\text{lb}$

$W_p = 64{,}8660\ \text{ft-lb}_f/\text{lb}$

$BHP = \frac{W_p \times m}{550}$

$= \frac{64{,}8660\ ^{ft.lb_f}/_{lb} \times 7{,}1781\ ^{lb}/_{dt}}{550}$

$= 0{,}8466$ Hp

Daya motor (MHP)

$MPH = \frac{BHP}{\eta}$

$\eta = 80\ \%$ (Peters, Fig 14.38)

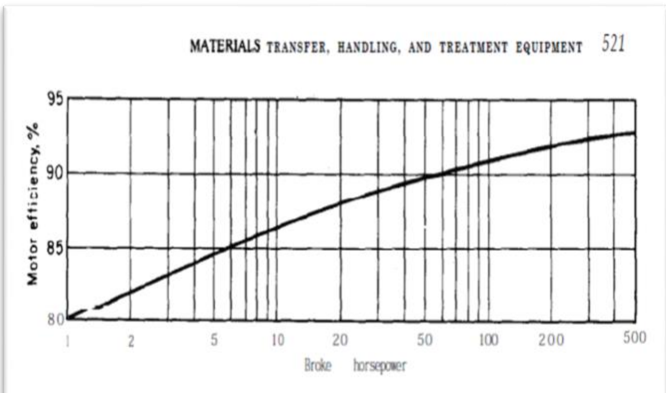


FIGURE 1438
Efficiencies of three-phase motors.

MPH $= \frac{0,8466\ hp}{0,8}$

$= 1,0582$ Hp

10. Reaktor (R-2071)

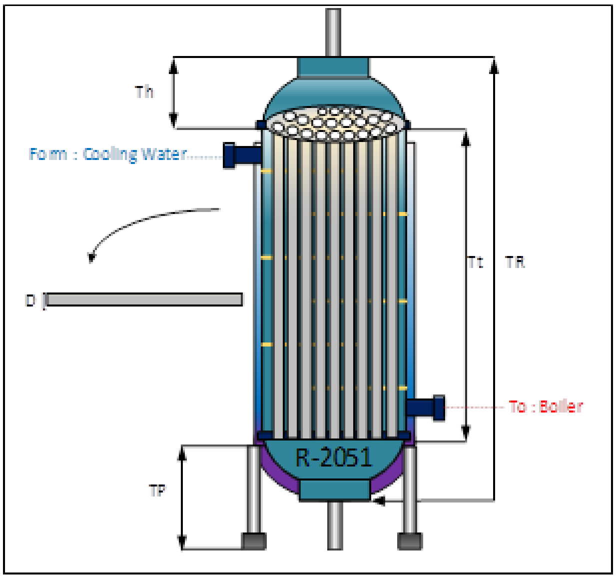
Fungsi : Tempat mereaksikan gliserol dengan hidrogen

Tipe : CSTR

Bahan : Carbon Steel SA 285 Grade A

Jumlah : 1 Unit

Gambar :



Data :

- Laju alir campuran (m) : 12920,35 kg/jam = 28484,21 lb/jam
- Densitas Campuran (ρ) : 947,7843 kg/m^3 = 59,17 lb/ft^3
- Temperatur (T) : 210 °C
- Tekanan (P) : 13 atm
- Viskositas (μ) :0,55 cP = 0,00037 lb/ft.s
- Waktu (t) : 80 menit = 1,3 Jam
- π : 3,14
- Faktor Keamanan : 20% (Petter, Tabel 6. Hal 37)

5. Kapasitas Rektor (Vr)

$$Vc = \frac{Massa}{Densitas}$$

$$Vc = 65{,}9589\ m^3$$

Faktor Keamanan = 10%

$$Volume\ Tangki = \frac{Vc}{90\%}$$

$$Volume\ Tangki = 82{,}4486\ m^3 = 2911{,}59\ ft^3$$

6. Dimensi Reaktor (Dr)

- **Volume silinder, V_s**

  $V_s = \frac{\pi}{4}\ x\ D_t^2\ x\ H_s$ $\qquad H_s = 1{,}5\ D_t$

Maka,

$$V_s = \frac{\pi}{4}\ x\ \ 1{,}5\ D_t^3$$

$$V_s = \frac{3{,}14}{4}\ x\ \ 1{,}5\ D_t^3$$

Vs = $1{,}1775 D_t^3$

- **Volume Elipsoidol**

$$V_e = \frac{\pi}{24}\ x\ \ D_e^3$$

$$V_e = \frac{3{,}14}{24}\ x\ \ D_e^3$$

Ve = 0,131 D_e^3

- **Diameter tangki, D_t**

  Vt = Vs + 2Ve

  = 1,4392 Dt^2

  Dt^3 = 57,2892 m^3

  Dt = 3,8550 m

- **Tinggi tangki, H_t**

  Tinggi silinder, H_s

  Hs = 1,5 Dt

  = 5,7825 m

= 227,6569 in

Tinggi Elipsoidal, He

He = 0,25 Dt

= 0,9637 m

= 37,9428 in

Tinggi total tangki, H_t

H_t = H_s + 2 H_e

= 9,74 m

- **Tinggi Cairan, H_c**

$$H_c = \frac{Volume\ Cairan}{Volume\ Tangki}\ x\ H_t$$

Hc = 7,7970 m

- **Tekanan Cairan, P_c**

$P_c = \rho \,.\, g \,.\, h_c$

Pc $= 947{,}7843\frac{kg}{m^3}\ x\ 9{,}8\frac{m}{s}\ x\ 7{,}7970m = 72420{,}7354\frac{kg}{m}.s^2$

Pc = 0,7025 atm

= 10,3265 *psi*

- **Tekanan Disain, P_d**

Pd = Pop + Pc

= 13 atm + 0,7025 atm

= 13,70 atm

= 201,4265 psi

- **Tebal dinding tangki, t_d**

$$t_d = \frac{PR}{SE - 0.6P} + C \qquad (Walas, Tabel\ 18.3, hal\ 625)$$

- Tekanan desain, P : 13,70 atm = 201,4265 psi
- Jari-jari tangki, R : 75,88 m
- *Allowable stress*, S : 18.700 psi (Peter, Tabel 4 Hal 538)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)
- Tahun digunakan : 10 tahun

$$\frac{PR}{SE - 0.6P} + C$$

Maka,

Td = 1,0428 in

= 0,0265 m

= 26,4866 mm

- **Tebal tutup ellipsoidal, t_e**

$$\frac{PD}{2SE - 0.2P} + C$$ (Wallas, Tabel 18,3.Hal 649)

Te = 0,98 in

= 0,025 m

- **Kebutuhan Katalis**

Densitas Katalis ρ = 83000 kg/m^3

Volume Reaktor Vr = 82,4486 m^3

Jumlah katalis (m) yang dibutuhkan = $m = \frac{\rho}{Vr}$

m = 1006,6874 kg

- **Menghitung Tinggi Timpukan Katalis, Z**

Diameter Tube, Du = 0,0035 × Dr

= 0,0035 X 3,8550

= 0,0135 m

= 0,5312 in

Dipilih tube ukuran standar (Mc.Cabe Lampiran 6) :

844 *PROCESS HEAT TRANSFER*

TABLE 11. DIMENSIONS OF STEEL PIPE (IPS)

Nominal pipe size, IPS, in.	OD, in.	Schedule No.	ID, in.	Flow area per pipe, in.^{2}	Surface per lin ft, ft.^{2}/ft. Outside	Surface per lin ft, ft.^{2}/ft. Inside	Weight per lin ft, lb steel
1/8	0.405	40*	0.269	0.058	0.106	0.070	0.25
		80†	0.215	0.036		0.056	0.32
1/4	0.540	40*	0.364	0.104	0.141	0.095	0.43
		80†	0.302	0.072		0.079	0.54
3/8	0.675	40*	0.493	0.192	0.177	0.129	0.57
		80†	0.423	0.141		0.111	0.74
1/2	0.840	40*	0.622	0.304	0.220	0.163	0.85
		80†	0.546	0.235		0.143	1.09
3/4	1.05	40*	0.824	0.534	0.275	0.216	1.13
		80†	0.742	0.432		0.194	1.48
1	1.32	40*	1.049	0.864	0.344	0.274	1.68
		80†	0.957	0.718		0.250	2.17
1 1/4	1.66	40*	1.380	1.50	0.435	0.362	2.28
		80†	1.278	1.28		0.335	3.00
1 1/2	1.90	40*	1.610	2.04	0.498	0.422	2.72
		80†	1.500	1.76		0.393	3.64
2	2.38	40*	2.067	3.35	0.622	0.542	3.66
		80†	1.939	2.95		0.508	5.03
2 1/2	2.88	40*	2.469	4.79	0.753	0.647	5.80
		80†	2.323	4.23		0.609	7.67
3	3.50	40*	3.068	7.38	0.917	0.804	7.58
		80†	2.900	6.61		0.760	10.3

IPS = 0,5 in sch 40

OD = 0,8400 in

ID = 0,6220 in

a" = 0,0333 ft²

Perhitungan tinggi katalis dengan volume tube

$$V_u = \frac{m}{\rho\ katalis}$$

$$Z = \frac{4 \times m}{\pi \times ID^2 \times \rho\ katalis}$$

Dengan :

Z = Tinggi Tumpukan Katalis (m)

V = Volume Katalis dalam Tube (m^3)

m = Massa Katalis (kg)

$$Z = \frac{4 \times 1006{,}6874}{3{,}14 \times 0{,}0002 \times 8300}$$

Z = 619,0089 cm

Tinggi tumpukan katalis keseluruhan = 619,0089 cm

Dipilih tinggi tube 8 ft = 2,4390 m

Tinggi katalis per tube adalah z = 80% dari tinggi tube

Maka, z = 80% × 2,4390 m = 1,9512 m

7. Menghitung jumlah tube, Nt

$$\text{jumlah tube, Nt} = \frac{Tinggi\ katalis\ keseluruhan}{Tinggi\ katalis\ per\ tube}$$

$$Nt = \frac{619{,}0089}{1{,}9512}$$

Nt = 317,2421 buah *tube*

8. Menentukan design tube

Susunan tube = Tiangular Pitch

Bahan = *Stainless Steel*

IPS = 0,5 in sch 40

OD = 0,8400 in

ID = 0,6220 in

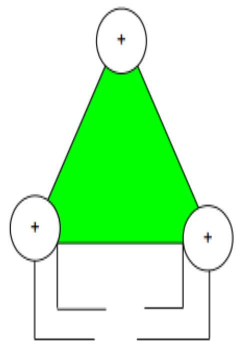
a" = 0,0333 ft²

Tinggi tumpukan Katalis = 1,9512 m

Panjang Pipa = 2,4390m

Susunan pipa yang digunakan adalah segitiga sama sisi dengan tujuan agar memberikan turbulensi yang lebih baik, sehingga akan, memperbesar koofiesien transfer panas dibandingkan susunan *square pitch* (Kern,1983).

Direncanakan *tube* disusun dengan pola *triangular pitch*.



Tebal Pipa = (OD-ID)/2

= (0,8400 - 0,6220)/2

= 0,1090 in

Jarak antar pusat pipa (PT)

PT = 1,25 × OD

PT = 1,25 × 0,8400 = 1,05 in

Jarak antar pipa (*Clearance*)

C' = PT-OD

C' = 1,0500 - 0,8400

C' = 0,21 in

Jarak Buffle, B = Dr × 0,3

= 3,8550 × 0,3

B = 1,1565 m

B = 45,5314 in

Area transfer panas dalam shell, As = ((PT -OD) × ID × B)/PT**)**

$$\textbf{As} = \frac{0{,}2100 \times 54{,}5770 \times 45{,}5314}{1{,}0500}$$

As = 496,9932 in^2

11. ***Storage Tank*** **(ST-2003)**

Tangki Penyimpanan (ST-2003)

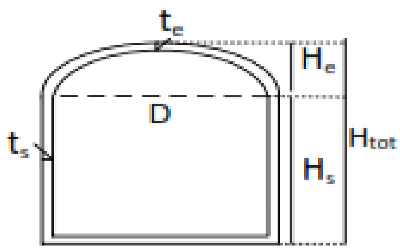
Fungsi : Tempat menyimpan Propilen Glikol

Bahan konstruksi : *Stainless Steel*

Jumlah : 1 unit

Lama Penyimpanan : 7 Hari

Gambar :



Parameter	**M (kg/jam)**	**ρ (kg/m3)**	**V (m3/jam)**	**%**	**Xi / p (kg/m3)**	densitas mix
$C_3H_8O_3$	225,053	1261	0,178472117	1,88227335	0,000015	1031,576737
H_2O	2626,43	997	2,634340729	21,96667558	2,203E-04	
$C_3H_6O_2$	8870,04	917	9,672886653	1,483722427	0,0000	
$C_3H_8O_2$	8927,57	1040	8,584204265	74,66732864	0,000718	
Total	**11956,4642**		**11,5904**	**100**	**0,000969**	**1031,576737**

Data :

Laju alir = 11956,4622 kg/jam = 26359,22 lb/jam

Volumetrik = 11,5904 m^3/jam = 0,11369619 ft^3/s

Densitas mix = 1031,576737 kg/m^3 = 64,4013lb/ft^3

T = 114^{0}C

t = 7 hari = 168 jam

P = 1 atm = 14,7 psi

Kapasitas tangki, V_t

$$V_b = \frac{m \times t}{\rho} = \frac{11956{,}4622 \frac{kg}{jam} x\ 168\ jam}{1031{,}576737 \, {}^{kg}/_{m^3}}$$

= 2.008.685,99 m^3

= 70.934.737,04 ft^3

= 18.738.971.352 gal

Faktor keamanan 10%

Maka,

$$V_p = 0{,}9\ V_t$$

$$V_t = \frac{V_p}{0{,}9}$$

$$= \frac{2008685{,}99\ m^3}{0{,}9}$$

= 2.231.873,32 m^3

= 78.816.374,48 ft^3

= 589598439 gal

Dimensi tangki

- **Volume silinder, V_s**

  $$V_s = \frac{\pi}{4} \times D_t^2 \times H_t \qquad H_t = 1{,}5D_t$$

Maka,

$$V_s = \frac{\pi}{4} \times D_t^3$$

- **Volume ellipsoidal, V_e**

$V_e \quad = \frac{\pi}{24} \times D_t^3$ He = ¼ Dt (peter's, Tabel 4)

$V_e \quad = 0{,}1308 \times D_t^3$

- **Diameter tangki, D_t**

$V_t \quad = V_s + V_e$

$= \left(\frac{1{,}5\pi}{4} \times D_t^3\right) + (\,0{,}1308 \times D_t^3)$

$V_t \quad = 1{,}3083 \times D_t^3$

$D_t^3 \quad = \frac{Vt}{1{,}3083}$

$D_t^3 \quad = \frac{2.231.873{,}32}{1{,}3083}$

$D_t^3 \quad = 1.705.933{,}9 \text{ m}^3$

$D_t \quad = \sqrt[3]{1.705.933{,}9}$

= 118,91 m

= 390,04 ft

= 4681,75 in

- **Tinggi tangki, H_t**

Tinggi silinder, Hs = 1,5 D_t

= 178,3752 m

= 585,07 ft

= 7022,63 in

Tinggi ellipsoidal, H_e = 1/4D_t

= 29,7292 m

= 97,511 ft

= 1170,43 in (walas, Tabel 18.5)

Tinggi total, H_t = tinggi silinder + tinggi ellipsoidal

H_t = 178,37 m + 29,72m

= 208,10 m

= 208104,49 mm

- **Tinggi cairan dalam tangki, H_c**

H_c $= \frac{volume\ bahan}{volume\ tangki} \times H_t$

$= \frac{2008685{,}1\ m^3}{2231873{,}3\ m^3} \times 208\ m$

= 187,3 m

= 614,3244 ft

= 7373,76 in

- **Tekanan cairan dalam tangki, P_c**

P_c $= \rho \times g \times H_c$

= 1835,4816 kg/m s^2

= 0,0178 atm

= 2,57 psi

- **Tekanan desain, P_d**

Pd = Pop + Pc

= (1 + 0,0178) atm

= 1,0178 atm

= 14,96 psi

- **Tebal dinding tangki, t_d**

$t_d = \frac{PR}{SE-0{,}6P} + C$ (Walas, Tabel 18.4)

- Tekanan desain, P_d : 2,5696 psi
- Jari-jari, R : 2340,87 in
- *Allowable stress*, S : 13700 psi (Walas, Tabel 18.4)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)
- Lama tahun digunakan : 10 tahun

Maka,

td $= \frac{2{,}5696\,psi \times 2340{,}87\,in}{(13700\,psi \times 0{,}85) - (0{,}6 \times 2{,}5696\text{ psi})} + 0{,}002\frac{in}{tahun} \times 10tahun$

= 0,5366 in

= 0,0136 m

= 0,044 ft

= 13,63 mm

- **Tebal dinding ellipsoidal,t_e**

$$t_e = \frac{PD_t}{2SE - 0{,}2P} + C$$

(Walas-Chemical Process Equipment, Tabel 18.4)

t_e $= \frac{2{,}5696psi \times 7022{,}63in}{2(13700\,psi \times 0{,}85) - (0{,}2 \times 2{,}5696\text{ psi})} + 0{,}002\frac{in}{tahun}x10tahun$

= 0,7948 in

= 0,020 m

= 0,066 ft

= 20,18 mm

12. Menara Distilasi (D-3081)

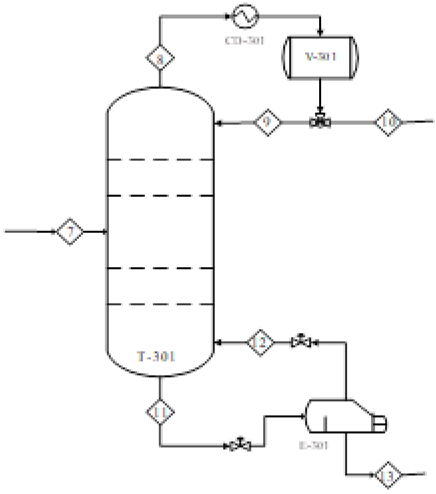
Fungsi : Untuk

Tipe :

Bahan :

Jumlah : 1 unit

Gambar :



Data Neraca Massa Distilasi

Tabel LC-1. Komposisi Aliran Feed Distilasi

Komponen	BM (kg/kmol)	Laju alir Molar	Fraksi Mol	Laju alir Massa (Fi)	Fraksi Massa	xF
		kmol/h		kg/h		
$C_3H_6O_2$	74	2,397	0,009	177,401	0,015	0,009
H_2O	18	145,913	0,544	2626,438	0,220	0,544
$C_3H_8O_2$	76	117,468	0,438	8927,572	0,747	0,438
$C_3H_8O_3$	92	2,446	0,009	225,053	0,019	0,009
Total (kg/h)		**268,22**	**1,00**	**11956,46**	**1,00**	**1,00**

Tabel LC-2. Komposisi Aliran Distilat (Top)

Komponen	BM (kg/kmol)	Laju alir Molar	Fraksi Mol	Laju alir Massa (Fi)	Fraksi Massa	xD
		kmol/h		kg/h		
$C_3H_6O_2$	74	2,396	0,018	177,322	0,067	0,018
H_2O	18	132,648	0,974	2387,671	0,900	0,974
$C_3H_8O_2$	76	1,175	0,009	89,276	0,034	0,009
$C_3H_8O_3$	92	0,000	0,000	0,000	0,000	0,000

Total (kg/h)		**136,219**	**1,000**	**2654,268**	**1,000**	**1,000**

Tabel LC-3. Komposisi Alirn Reboiler (Bottom)

Komponen	**BM (kg/kmol)**	**Laju alir Molar**	**Fraksi Mol**	**Laju alir Massa (Fi)**
		kmol/h		**kg/h**
$C_3H_6O_2$	74	0,00	0,0000	0
H_2O	18	13,2648	0,1005	238,7671
$C_3H_8O_2$	76	116,2934	0,8810	8838,2967
$C_3H_8O_3$	92	2,4462	0,0185	225,0533
Total (kg/h)		**132,0055**	**1,0000**	**9302,1958**

Kondisi Operasi Menara Distilasi

a. Menentukan Kondisi Operasi Umpan

Bubble Point (Kondisi Umpan Cair Jenuh)

T = 121°C = 394,39 K

P = 1 atm = 1,5 bar

b. Menentukan Kondisi Operasi Distilat

T = 112,67°C = 385,67 K

P = 1 atm = 1,5 bar

c. Menentukan Kondisi Operasi Bottom

T = 206,26°C = 479,26 K

P = 1 atm = 1,5 bar

Menentukan *Stage Minimum* (Nm)

Nm = 2,905 stages = 8

= 2,905 – 1

= 1,905 = 2

Menghitung Refluk Minimum

Rm = 0,03

Rop = 0,04

Menghitung Jumlah Tahap (*tray*) Teoritis dan Aktual

N actual = 8,301

Menetukan Lokasi Umpan Masuk (*Feed Stage*)

N_R = 2 (*Rectifying section*/*Stage* atas)

N_s = 6 (*Stripping secton*/*stage* bawah)

N_R/N_S = 0,2861

Maka, *Feed Stage* berada di stage :

N_S = 6

Perhitungan Komposisi Komponen di Distilasi dan Reboiler

$$V_{n+1} = Ln + D$$

Dimana :

R = Ln/D

Ln = R X D

Ln = 0,04 X 136,22

= 5,27 kmol/jam

Vn = 5,27 + 136,22

= 141,49 kmol/jam

$$L_m = Ln + (q\ X\ F)$$

L_m = 273,50 kmol/jam

$$V_n = Vm + ((1 - q)\ X\ F)$$

V_m = V_n – ((1-q) X F)

= 141,49 – 0

= 141,49 kmol/jam

$$B = Lm - Vm$$

B = 273,50 – 141,49

= 132,01 kmol/jam

Menghitung Tinggi Distilasi

Berdasarkan *plate spacing towler,* pada umumnya jarak antar *plate* yang digunakan ialah 0,15 m – 1 m (Towler, 2020. Hal.693)

Digunakan jarak antar *plate* = 0,15 m

= 15 X 0,15 m

= 2,25 m

Tinggi Menara Distilasi (H)

H = Jumlah *tray actual* X *Plate spacing*

= 8,301 X 0,15 m

= 1,2451

Menghitung Berat Molekul Campuran

Komponen	BM	$Z_{i,f}$	BMf kg/kmol	$X_{i,D}$	BMd kg/kmol	$X_{i,B}$	BMb kg/kmol
$C_3H_6O_2$	74,000	0,009	0,661	0,018	1,302	0,000	0,001
H_2O	18,000	0,544	9,792	0,974	17,528	0,100	1,809
$C_3H_8O_2$	76,000	0,438	33,284	0,009	0,655	0,881	66,954
$C_3H_8O_3$	92,000	0,009	0,839	0,000	0,000	0,019	1,705
Total	260,000	1,000	44,576	1,000	19,485	1,000	70,468

Menghitung Densitas Campuran Liquid dan Vapor

Data Properties Densitas Cairan

Komponen	A	B	Tc (K)	n
$C_3H_6O_2$	0,272	0,272	512,580	0,233
H_2O	0,266	0,264	516,250	0,237
$C_3H_8O_2$	0,277	0,272	536,710	0,249
$C_3H_8O_3$	0,347	0,274	647,130	0,286

a. Menghitung Densitas Liquid Umpan

Komponen	ρ (kg/m^3)	xF	ρ_c (kg/m^3)
$C_3H_6O_2$	685,9138698	0,009	6,130478296
H_2O	684,5852372	0,544	372,4115941
$C_3H_8O_2$	705,1922191	0,438	308,8363181
$C_3H_8O_3$	933,5414499	0,009	8,513973703
Total	**3009,232776**	**1**	**695,8923643**

b. Menghitung Densitas Top Distilasi

Komponen	ρ (kg/m^3)	xD	ρ_c (kg/m^3)
$C_3H_6O_2$	696,6028539	0,018	12,2539956
H_2O	695,3456372	0,974	677,1174768
$C_3H_8O_2$	715,1042004	0,009	6,166666957
$C_3H_8O_3$	942,5737722	0,000	8,52923E-07
Total	**3049,626464**	**1**	**695,5381403**

c. Menghitung Densitas Campuran Bottom

Komponen	ρ (kg/m^3)	xB	ρ_c (kg/m^3)
$C_3H_6O_2$	541,4924526	0,000	0,004360458
H_2O	542,451424	0,100	54,50931323
$C_3H_8O_2$	583,5566858	0,881	514,0980734
$C_3H_8O_3$	836,9734916	0,019	15,51019501

Total	2504,474054	1	584,1219421

d. Menghitung Densitas Campuran Vapor

$$\rho\ gas = \frac{P\ X\ BM}{R\ X\ T} X\ Fraksi\ Mol$$

Komponen	BM (kg/kmol)	Distilat		Bottom	
		xD	ρ kg/m³	xB	ρ kg/m³
$C_3H_6O_2$	74,000	0,018	0,041	0,000	0,000
H_2O	18,000	0,974	0,554	0,100	0,046
$C_3H_8O_2$	76,000	0,009	0,021	0,881	1,704
$C_3H_8O_3$	92,000	0,000	0,000	0,019	0,043
Total		**1,0**	0,041	**1,0**	**1,8**

Menghitung *Mixture Surface Tension*

Berdasarkan buku ***Yaws, 1999.*** **Tabel 9.1** *Surface Tension* ditentukan oleh

$$\sigma = A\ X\ (\frac{1-T}{Tc})^n$$

Mixture Surface Tension pada Distilat

Komponen	A	Tc	n	σ (dyne/cm)	XD	σm (N/m)
$C_3H_6O_2$	68,329	512,58	1,2222	57,60212279	0,018	0,001013283
H_2O	67,036	516,25	1,2222	56,60347432	0,974	0,055119641
$C_3H_8O_2$	65,93	508,31	1,2222	55,47337397	0,009	0,000478372
$C_3H_8O_3$	132,674	647,13	0,955	107,6831773	0,000	9,74411E-11
Total				277,3621484	1	0,056611296

Mixture Surface Tension pada Bottom

Komponen	A	Tc	n	σ (dyne/cm)	XB	σm (N/m)

$C_3H_6O_2$	68,329	512,58	1,2222	45,86886319	0,000	3,69367E-07
H_2O	67,036	516,25	1,2222	45,19218194	0,100	0,004541227
$C_3H_8O_2$	65,93	508,31	1,2222	44,03572016	0,881	0,03879431
$C_3H_8O_3$	132,674	647,13	0,955	88,15399072	0,019	0,001633607
Total				223,250756	1	0,044969513

Menghitung Diameter Menara Distilasi

1. Menentukan *Liquid Vapor Flow*

$$F_{LV} = \frac{L_w}{V_w}\sqrt{\frac{\rho_v}{\rho_L}}$$

Distilat :
Ln = 0,02854 kg/s
Vn = 0,7658 kg/s
ρv = 0,041 kg/m^3
ρl = 695,5 kg/m^3
Flv = 0,000
Bottom :
Lm =5,35358kg/s
Vn = 2,7696 kg/s
ρv = 1,79 kg/m^3
ρl = 2504,47 kg/m^3
Flv = 0,05172

2. Menentukan Nilai K1

Konstanta K1 ditentukan melalui kurva *flooding velocity*, sieve plates (Towler, 2020. Hal.720)

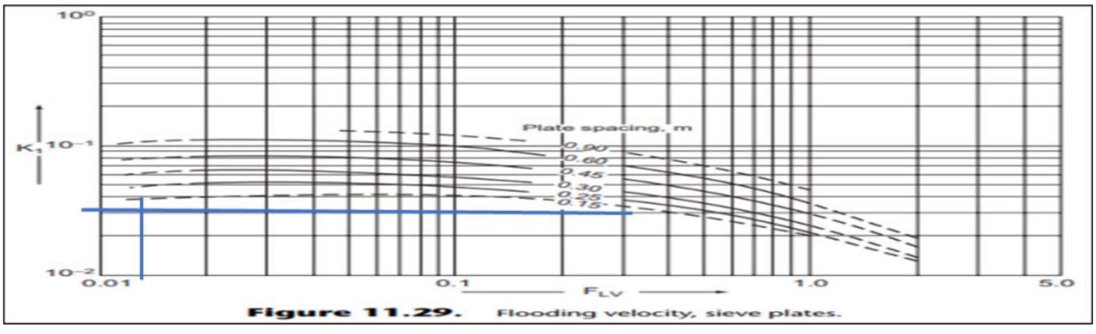


Figure 11.29. Flooding velocity, sieve plates.

Plate spacing digunakan = 0,15 m

K1 top = 0,02

K1 bottom = 0,02

$$K_1^* = K1\ X\ [\frac{\sigma}{0,02}]^2$$

Maka diperoleh :

K1 top = 0,0245

K1 bottom = 0,0234

3. Menentukan *Flooding Vapor Velocity* (Uf)

$$u_f = K_1 \sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$$

Uf top = 0,647474503 m/s

Uf bottom = 1,17814 m/s

Didapatkan *flooding* pada max flowrate design (un)

Un top = 0,5179 m/s

Un bottom = 0,9393 m/s

4. Menentukan Net Area

Net Area = Max. Vol Rate/ Max Rate Flooding

Top = 35,91 m^2

Bottom = 1,64 m^2

5. Menghitung *Downcorner Area* (Ad)

Ad = Net Area/ (1-0,16)

Top = 42,76 m^2

Bottom = 1,95 m^2

6. Menghitung Diameter Kolom (Dc)

$$D_c = \sqrt{\frac{A_d \times 4}{\pi}}$$

Diameter top = 7,369 m

Diameter bottom = 1,576 m

Perancangan *Plate*

Diameter kolom (Dc) = 1,5767 m (Brownell)

Luas kolom (Ac) = 1,9515 m^2

Luas Downsorner (Ad) = 16% Ac = 0,31222 m^2

Luas net area (An) = Ac-Ad = 1,6393 m^2

Luas *active area* (Aa) = Ac-2Ad = 1,3270 m^2

Luas *hole area* (Ah) = 10% Aa = 0,13270 m^2

Panjang Weir (Lw) = (Ad/Ac) X100%

= 16%

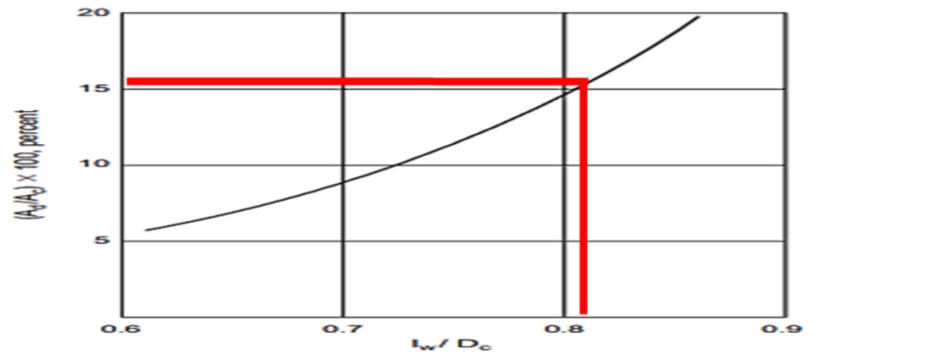


Figure 11.33. Relation between downcomer area and weir length.

Maka diperoleh :

Lw/Dc = 0,81

Lw = 0,81 X 1,5767

= 1,277 m

Menghitung Jumlah Lubang *Plate*

$$luas\ Satu\ Lubang = \frac{\pi\ dh^2}{4}$$

Luas satu lubang = 0,0001656 m^2

Jumlah Lubang = Ah/ Luas satu lubang

= 0,121/0,0001656

= 727,7383 lubang

Menentukan Tebal *Shell*

Tensile strength = 75.000 psi

Allowable stress = 17.000 psi

Fakor korosi = 0,1 in

Diameter = 1,58 m

R = 0,7883 m = 31,03 in

Ptotal = Poperasi + Phidrostatis

= 15,5373 psi

P desain = 25 psia + Ptotal

= 40,5373 psi

$$ts = \frac{P\ X\ R}{f\ X\ E - 0.6\ P} + C$$

Ts = 0,1926 in

Menentukan Tebal *Head*

ID = 62,08 in

OD = 62,58 in

Icr = 2,5

r = 36

$$th = \frac{Prc\ W}{2\ F\ X\ E - 0.2\ P} + C$$

Th = 0,144 in

Tebal head standar = 0,250 in

= 0,0064 in

Menentukan Tinggi *Head*

th = 0,250 in

sf = 2 in

a = ID/2 = 18,750 in

AB = a-icr = 16,250 in

BC = rc-icr = 33,5 in

AC = $(BC^2-AB^2)^{0,5}$ = 29,295 in

b = rc-Ac = 6,705 in

Tinggi head (th) = sf + b + tebal head

= 8,955 in

= 0,227 m

Menghitung Tinggi Total Menara

Tinggi Menara(h) = 1,2451 m

Tinggi head atas = 0,227 m

Tinggi head bawah = 0,227 m

Tinggi total kolom distilasi :

Hd = H + Tinggi head atas + tinggi head bawah

= 1,700 m

S

B. SPESIFIKASI PERALATAN UTILITAS

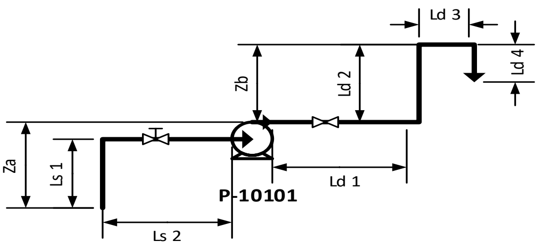
1 Pompa

Fungsi : Mengalirkan air dari waduk ke bak penampungan

Tipe : *Centriugal pump*

Bahan : *Commercial steel pipe*

Gambar :



Data :

- Laju alir massa, m	: 8302,28 kg/jam
	: 18303,20 lb/jam
- Densitas air, ρ	: 1000 kg/m^3 = 62,43 lb/ft^3
- Viskositas air, μ	: 1 cp
	: 0,00067 lb/ft.s
- Tinggi pompa terhadap cairan masuk, Za	: 1 m = 3,28 ft
- Tinggi pompa terhadap cairan keluar, Zb	: 5 m = 16,40 ft
- Panjang pipa hisap, Ls	: 12 m = 39,37 ft
- Panjang pipa buang, Lb	: 50 m = 164,04 ft
- Faktor keamanan	: 10 %

Laju alir volumetrik, Qv

Qp = m/0.9

= 20336,892 lb/jam /0,9 = 5,6491 lb/s

Qv = Qp/ρ

= (5,6491 lb/s)/(62,43 lb/ft^3)

= 0,0905 ft^3/s

Diameter optimum, Dopt

Asumsi aliran turbulen

Dopt = $3{,}9 \times Qv^{0,45} \times \rho^{0,13}$ (Peter, Pers 14.15)

= $3{,}9 \times (0{,}0905)^{0,45} \times (62{,}43)^{0,13}$

= 2,2663 in

Dari Appendix 5 Mc. Cabe , hal 1107), diperoleh data;

	Suction (a)				**Discharge (b)**			
IPS	2,5 in sch 40							
ID	2,4690	In	0,2058	ft	2,4690	in	0,2058	ft
OD	2,8800	In	0,2400	ft	2,8800	in	0,2400	ft
a"	0,0333	ft^2			0,0333	ft^2		

Sumber : (Mc Cabe, Lampiran 6)

Kecepatan aliran, V

Va = Vb, karena ukuran pipa hisap dan pipa buang sama

$$V = \frac{Q_v}{a''}$$

$$= \frac{0{,}0905\ ft^3/s}{0{,}0333\ ft^2} = 2{,}7203 \text{ ft/s}$$

$$\frac{V^2}{2g_c} = \frac{7{,}400}{2 \times 32{,}17} = 0{,}1150 \text{ ft-lb}_f/\text{lb}$$

Bilangan Reynolds, N_{Re}

$$N_{Re} = \frac{\rho \times V \times D}{\mu}$$

$$= \frac{62{,}43 \frac{lb}{ft3} x\ 2{,}7203\ X\ 0{,}2058\ ft}{0{,}00067 \frac{lb}{ft.s}} = 51997{,}1619$$

Rugi Gesek

- **Pipa hisap (*suction*)**

Pada pipa hisap, rugi gesek timbul akibat gesekan dengan kulit pipa, serta pengaruh *fitting* dan *valve*.

- Rugi gesek akibat kulit

h_{fsa} $= f \frac{\Delta L}{r_H} \frac{V^2}{2g_c}$ (Mc Cabe, Pers 5.56)

r_H $= \frac{ID}{4} = \frac{0{,}2058\, ft}{4}$ = 0,0514 ft

N_{Re} = 51997,1620

Material yang digunakan untuk konstruksi pipa adalah *commercial steel pipe*, dimana

K = 0,00015 ft (Mc Cabe, Fig. 5.9)

$\frac{k}{ID}$ $= \frac{0{,}00015\, ft}{0{,}2058}$

= 0,00072ft

F = 0,005 (Mc Cabe, Fig. 5.9)

Maka,

h_{fsa} = 0,005 x $\frac{39{,}37\, ft}{0{,}0514\, ft}$ x 0,1155 ft.lbf/lb

= 0,44010 ft-lb_f/lb

- Rugi gesek akibat *fitting* dan *valve*

h_{ffa} $= K_f \frac{V^2}{2g_c}$ (Mc Cabe, Pers 5.67)

K_f (*elbow* 90°) = 0,9 (Mc Cabe, Tabel 5.1)

K_f *gate valve* = 0,2

Maka,

h_{ffa} $= (0{,}2 + 0{,}9) \times 0{,}4401\ \text{ft.} \frac{lb_f}{lb}$ = 0,4632 ft-lb_f/lb

- **Pipa buang (*discharge*)**

Pada pipa buang, rugi gesek timbul akibat gesekan dengan kulit pipa, serta pengaruh *fitting* dan *valve*.

- Rugi gesek akibat kulit

h_{fsb} $= f \frac{\Delta L}{r_H} \frac{V^2}{2g_c}$ (Mc Cabe, Pers 5.56)

r_H $= \frac{ID}{4}$

$= \frac{0,2058\ ft}{4}$

= 0,0514 ft

N_{Re} = 51997,1620

Material yang digunakan untuk konstruksi pipa adalah *commercial steel pipe*, dimana

k = 0,00015 ft (Mc Cabe, Fig. 5.9)

$\frac{k}{ID} = \frac{0,00015\ ft}{0,2058 ft} = 0,0007290$

f = 0,005 (Mc Cabe, Fig. 5.9)

Maka,

$h_{fsb} = 0,005 \times \frac{164,04\ ft}{0,0514 ft} \times 0,1155$ ft.lbf/lb

= 1,8339 ft-lb_f/lb

- Rugi gesek akibat *fitting* dan *valve*

$h_{ffb} = K_f \frac{V^2}{2g_c}$ (Mc Cabe, Pers 5.67)

K_f (*elbow* 90°) = 0,9 (Mc Cabe, Tabel 5.1)

K_f (*sambungan lurus*) = 0,2 (Mc Cabe, Tabel 5.1)

K_f *(globe valve)* = 10

Total K_f = (6 x 0,9) + (1 x 10) +(0,2 x 2) = 15,8

Maka,

$h_{ffb} = 15,8 \times 0,1150 ft.lb_f/lb$

= 1,77121 ft-lb_f/lb

Sehingga, total rugi gesek adalah :

$= h_{fsa} + h_{fsb} + h_{ffa} + h_{ffb}$

= 3,605 ft-lbf/lb

Daya pompa (BHP)

Daya pompa dapat dihitung dengan menggunakan Persamaan *Bernoulli* :

$$\frac{P_a}{\rho}+\frac{gZ_a}{g_c}+\frac{\alpha_a V_a^2}{2g_c}+\eta \mathrm{W_p}=\frac{P_b}{\rho}+\frac{gZ_b}{g_c}+\frac{\alpha_b V_b^2}{2g_c}+h_f$$

Atau

$$\eta \mathrm{W_p}=\left(\frac{P_b}{\rho}+\frac{gZ_b}{g_c}+\frac{\alpha_b V_b^2}{2g_c}\right)-\left(\frac{P_a}{\rho}+\frac{gZ_a}{g_c}+\frac{\alpha_a V_a^2}{2g_c}\right)+h_f$$

Dimana

$P_a = P_b$

$V_a = V_b$

$\rho_a = \rho_b$

$g/g_c = 1$

$\alpha_a = \alpha_b$

η = 33,5% (Peters, Fig. 14.37)

Sehingga persamaan di atas dapat disederhanakan menjadi :

$$\eta \mathrm{W_p} = (\mathrm{Z_b}-\mathrm{Z_a})+h_f$$

0,335 W_p = (16,40-3,28) ft-lb_f/lb + 4,0855 ft-lbf/lb

Wp = 51,3996 ft-lbf/lb

BHP $= \frac{W_p \times m}{550}$

$= \frac{51{,}3186\ {}^{ft.lb_f}/_{lb} \times 5{,}6491\ {}^{lb}/_{dt}}{550}$

= 0,5271 HP

Daya motor (MHP)

MPH $= \frac{BHP}{\eta}$

η = 75 % (Peters, Fig 14.38)

MPH $= \frac{0{,}5271\ 0HP}{0{,}75}$

= 0,7028 HP

Dengan cara yang sama, maka diperoleh daya pada masing-masing pompa untuk peralatan utilitas seperti pada Tabel C.1

Tabel C.1 Daya Pompa Pada Peralatan Utilitas

Kode Alat	**Keterangan**	**Daya (hp)**
P-10101	Pompa dari telaga menuju Bak Sedimentasi	0,53
P-10102	Pompa dari Bak Sedimentasi Menuju *Clarifier*	0,31
P-10103	Pompa dari tangki pelarutan tawas Menuju *Clarifier*	0,26
P-10104	Pompa dari tangki soda *Ash* Menuju *Clarifier*	0,15
P-10105	Pompa dari Bak Penampung Menuju *Sand Filter*	0,14
P-10106	Pompa dari *Sand Filter* menuju tangki menara	0,55
P-10107	Pompa dari tangki Kaporit menuju Air sanitasi	0,13
P-20108	Pompa dari *Kation Anion Exchanger* Menuju *Feed water tank*	0,12
P-20109	Pompa dari *Feed water tank* menuju *Daerator, Cooling tower* dan Air Proses	0,13
P-20110	Pompa *Recycle* Kondensat menuju *Daerator*	0,22
P-30111	Pompa umpan Boiler	0,21
P-30112	Pompa dari *Cooling Tower* Menuju Unit Proses	0,44

2. Bak Sedimentasi

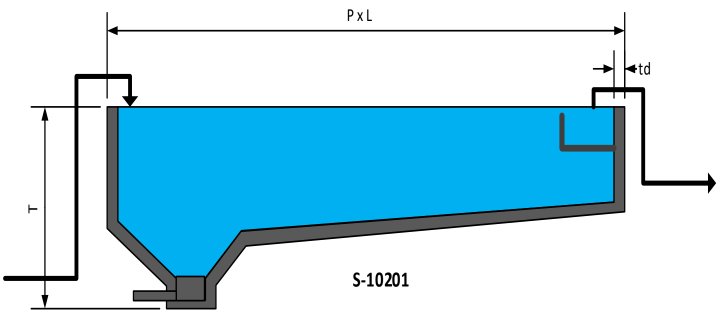
Fungsi : Mengendapkan lumpur/ padatan berdasarkan gaya gravitasi

Jenis : Bak berbentuk persegi panjang

Jumlah : 1 buah

Konstruksi : Beton bertulang

Gambar :



Data :

Laju alir massa = 8302,2785 kg/jam

Densitas = 1000 kg/m^{3}[19]

Waktu tinggal = 24 jam

Laju alir volumetrik, Q

$$Q = \frac{m}{\rho} = \frac{8302{,}2785 \frac{kg}{jam}}{1000\ kg/m^3} = 8{,}3023\ m^3/jam$$

Dimensi bak

V = 8,3023 m^3/jam x 24 jam

= 199,2547 m^3

Faktor keamanan 10%

$$\text{Volume bak} = \frac{199{,}2547\ m^3}{0{,}9} = 221{,}3941\ m^3$$

Perbandingan dimensi bak penampung yaitu P : L : T = 3 : 2 : 1

Volume bak = panjang x lebar x tinggi

199,2547 = 3T x 2T x T

6T^3 = 199,2547 m^3

T = 3,205 m

Sehingga diperoleh dimensi bak :

Panjang = 3T = 9,6316 m

Lebar = 2T = 6,4211 m

Tinggi = 1T = 3,2105 m

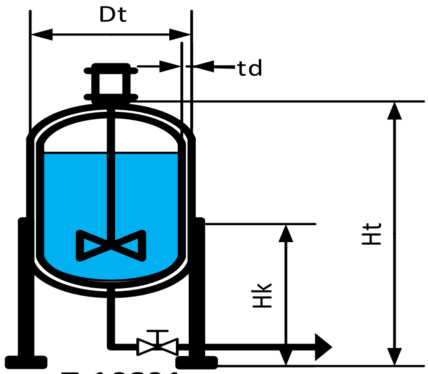
3 Tangki Pelarutan Alum

Fungsi : Tempat melarutkan alum ($Al_2(SO_4)_3$)

Jenis : Silinder vertikal dengan alas dan tutup *ellipsoidal*

Konstruksi : *Carbon Steel*

Gambar :



Data :

Laju alir massa Air = 8302,28 kg/jam = 18298,22 lb/jam

Laju alir volumetrik = 4,81 m^3/jam = 4812,92 liter/jam

= 394,31 ft^3/s

Densitas = 1.725 kg/m^3 = 107,69 lb/ft^3

Waktu tinggal = 8,00 Jam (Setiap Pergantian Shift)

Viskositas Campuran = 1,00 cp = 0,00067 lb/ft h

Kebutuhan alum

Diasumsi kekeruhan air sungai ciasem adalah sebesar <100 NTU

Berdasarkan Jurnal Amina I (3),2020. Untuk Turbidity 72 NTU Memakai dosis tawas 4 ml dalam 1 Liter Air

Konsentrasi Alum = 1% (10 gr/L)

Larutan Alum yang dibutuhkan = 4 mL dalam 1 liter Air

= 0,004 liter Alum x 4823,02 liter/jam

= 19,252 liter/jam = 0,02 m^3/jam Alum

Massa Alum yang dibutuhkan = 19,292 liter / 0,004 liter = 4823,02 liter

= 4812,92 liter / 10 gr =481,29 gr

= 0,48 Kg/jam

Kapasitas tangki

Kebutuhan alum direncanakan untuk pemakaian selama 8 jam

Volume Kapasitas = 0,02 m^3 x 8 jam

= 0,15 m^3

Faktor Keamanan = 10%

Volume Tangki = 0,15 m^3/ 0,9

= 0,1711 m^3

Dimensi tangki

- **Volume silinder, V_s**

  $V_s = \frac{\pi}{4} \times D_t^2 \times H_s$ $H_s = D_t$

Maka,

  $V_s = \frac{\pi}{4} \times D_t^3$

- **Volume ellipsoidal, V_e**

  $V_e = \frac{\pi}{24} \times D_t^3$ He = ¼ Dt (peter's, Tabel 4)

  V_e = 0,13 x D_t^3

- **Diameter tangki, D_t**

  $V_t = V_s + V_e$

  $= (0{,}785 \times D_t^3) + (0{,}13 \times D_t^3)$

  V_t = 0,915 x D_t^3

  $D_t^3 = \frac{Vt}{0{,}915}$

  $D_t^3 = \frac{0{,}1715}{0{,}915}$

  D_t^3 = 0,1874 m^3

  $D_t = \sqrt[3]{0{,}1874}$

  = 0,5722 m

  = 22,5272 ft

  = 1,8768 in

- **Tinggi tangki, H_t**

Tinggi Kaki Tangki = 0,5 m

Tinggi silinder, Hs = D_t

= 0,5726 m

Tinggi ellipsoidal, H_e = 1/4D_t

= 0,1431 m

(walas, Tabel 18.5)

Tinggi total, H_t = tinggi silinder + 2tinggi ellipsoidal + tinggi kaki

H_t = 1,3583 m

- **Tinggi cairan dalam tangki, H_c**

H_c $= \frac{volume\ bahan}{volume\ tangki} \times H_t - Hk$

= 0,7725 m

= 2,5337 ft

- **Tekanan cairan dalam tangki, P_c**

P_c $= \rho \times g \times H_c$

= 1725 kg/m^3 x 10 m/dt^2 x 0,7730 m

= 13324,894 /m s^2

= 1,9326 psi

Pop = 1 atm

= 14,7 psi

- **Tekanan desain, P_d**

Pd = Pop + Pc

= 16,6326 Psi

- **Tebal dinding tangki, t_d**

t_d $= \frac{PR}{SE-0,6P} + C$ (Walas, Tabel 18.4)

- Tekanan desain, P_d : 16,6 psi

- Jari-jari, R : 11,2636 in
- *Allowable stress*, S : 13700 psi (Walas, Tabel 18.4)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)
- Lama tahun digunakan : 10 tahun

Maka,

$$td = \frac{16{,}3\ psi \times 11{,}2714\ in}{(13700\ psi \times 0{,}85) - (0{,}6 \times 16{,}3\ psi)} + 0{,}002 \frac{in}{tahun} \times 10 tahun$$

= 0,0361 in

= 0,0009 m

= 0,917 mm

- **Tebal dinding ellipsoidal,t_e**

$t_e = \frac{PD_t}{2SE - 0{,}2P} + C$ *(Walas-Chemical Process Equipment*, Tabel 18.4)

$$t_e = \frac{17{,}26\ psi \times 9{,}37\ in}{2(13700\ psi \times 0{,}85) - (0{,}2 \times 9{,}37\ psi)} + 0{,}002 \frac{in}{tahun} x 10 tahun$$

= 0,033 in

= 0,0011 m

- **Design Pengaduk**

Untuk umpan dengan viskositas ≤ 4.000 cP, maka dipilih pengaduk jenis propeller berdaun tiga (Walas, hal 288)

Untuk mencegah vortex, maka pada tangki dipasang baffle

Dt : diameter tangki

Da : diameter impeller

E : tinggi turbin dari dasar tangki

L : panjang blade pada turbin

W : lebar blade pada turbin

j : lebar baffle

Dt : 0,5726 m

Diameter Pengaduk (d)

D $= \frac{Dt}{3}$

$= 0{,}1907$m

$= 0{,}6256$ ft

Panjang daun pengaduk (L)

L $= \frac{D}{4}$

$= 0{,}0477$ m

$= 0{,}1564$ ft

Lebar daun pengaduk (W)

W $= \frac{D}{5}$

$= 0{,}0381$ m

$= 0{,}1251$ ft

Tinggi pengaduk dari dasar tangki (E)

E $= \frac{Dt}{3}$

$= 0{,}1907$ m

$= 0{,}6256$ ft

Lebar baffle (j)

J $= \frac{Dt}{12}$

$= 0{,}0477$ m

$= 0{,}1564$ ft

Kecepatan putar pengaduk (N) (Treybal, Pers 6.18)

$$\frac{N \times d}{\left(\frac{\sigma\ g_c}{\rho}\right)^{0.25}} = 1{,}22 + 1{,}25\left(\frac{D_B}{d}\right)$$

Σ = 0,0000054 lb/ft

gc = 32,17 ft/dt^2

N = 1,94505rps

Daya pengadukan

$$N_{Re} = \frac{\rho \times N \times d^2}{\mu}$$

Nre = 122331

Karena NRe > 10000, maka

$$P = \frac{K_T N^3 d^5 \rho}{g_c}$$

(Mc Cabe, Pers 9.24)

Kt = 0,32

P = 0,7617 lb/ft s

= 0,001385 HP

Efisiensi motor penggerak 80%

Daya Motor = 0,001385 /0,8

= 0,001731 HP

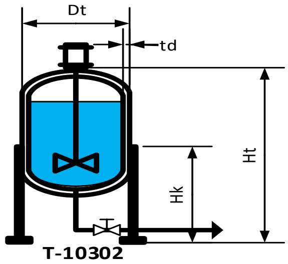
4. Tangki Soda Ash

Fungsi : Tempat melarutkan Soda Ash

Jenis : Silinder vertikal dengan alas dan tutup *ellipsoidal*

Konstruksi : *Carbon Steel*

Gambar :



Data :

Laju alir massa Air	= 8302,28 kg/jam	= 18298,22lb/jam
Laju alir volumetrik	= 3,27 m^3/jam	= 3268,61liter/jam
Densitas	= 2540 kg/m^3	= 158,57 lb/ft^3
Waktu tinggal	= 1 Hari	= 24 jam
Viskositas Campuran	= 0,45 cp	= 0,00014 lb/ft h

Kebutuhan Soda Ash

Diasumsi kekeruhan air telaga bidadari adalah sebesar <100 NTU

Berdasarkan Jurnal Amina I (3),2020. Untuk Turbidity 72 NTU Memakai dosis Soda Ash 2 ml dalam 1 Liter Air

Konsentrasi Soda Ash = 1% (10 gr/L)

Larutan Soda Ash yang dibutuhkan = 2 mL dalam 1 liter Air

= 0,002 liter Soda Ash x 6,537 liter/jam

= 6,537 liter/jam = 0,007 m^3/jam Soda Ash

Massa Soda Ash yang dibutuhkan = 6,537 liter / 0,002 liter = 3268,61liter

= 3268,61 liter / 10 gr = 326,86gr

= 0,33 Kg/jam

Kapasitas tangki

Kebutuhan alum direncanakan untuk pemakaian selama 8 jam

Volume Kapasitas = 0,007 m^3/jam x 24 jam

= 0,16 m^3

Faktor Keamanan = 10%

Volume Tangki = 0,16 m^3 / 0,9 = 0,1743 m^3

Dimensi tangki

- **Volume silinder, V_s**

  $V_s = \frac{\pi}{4} \times D_t^2 \times H_t$ $H_t = D_t$

Maka,

  $V_s = \frac{\pi}{4} \times D_t^3$

- **Volume ellipsoidal, V_e**

  $V_e = \frac{\pi}{24} \times D_t^3$ He = ¼ Dt (peter's, Tabel 4)

  V_e = 0,13 x D_t^3

- **Diameter tangki, D_t**

  $V_t = V_s + V_e$

  $= (0{,}785 \times D_t^3) + (0{,}13 \times D_t^3)$

  V_t = 0,915 x D_t^3

  $D_t^3 = \frac{Vt}{0{,}915}$

  $D_t^3 = \frac{0{,}1743}{0{,}915}$

  D_t^3 = 0,1905 m^3

  $D_t = \sqrt[3]{0{,}1905}$

  = 0,5757 m

  = 1,8884 ft

  = 22,6666 in

- **Tinggi tangki, H_t**

  Tinggi Kaki Tangki = 0,5 m

  Tinggi silinder, Hs = D_t

  = 0,5757 m

  Tinggi ellipsoidal, H_e = ¼ D_t

= 0,1439 m

(walas, Tabel 18.5)

Tinggi total, H_t = tinggi silinder + 2tinggi ellipsoidal + tinggi kaki

H_t = 1,3636 m

- **Tinggi cairan dalam tangki, H_c**

H_c $= \frac{volume\ bahan}{volume\ tangki} \times H_t - Hk$

= 0,7772 m

= 2,5493 ft

- **Tekanan cairan dalam tangki, P_c**

P_c $= \rho \times g \times H_c$

= 2540 kg/m^3 x 10 m/dt^2 x 0,7772 m

= 19741,853 kg/m s^2

= 2,8633 psi

Pop = 1 atm

= 14,7 psi

- **Tekanan desain, P_d**

Pd = Pop + Pc

= 17,5633 Psi

- **Tebal dinding tangki, t_d**

t_d $= \frac{PR}{SE - 0{,}6P} + C$ (Walas, Tabel 18.4)

- Tekanan desain, P_d : 17,6 psi
- Jari-jari, R : 11,3333 in
- *Allowable stress*, S : 13700 psi (Walas, Tabel 18.4)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)

- Lama tahun digunakan : 10 tahun

Maka,

td $= \frac{17,6\ psi \times 11,3333\ in}{(13700\ psi \times 0,85) - (0,6 \times 17,6\ psi)} + 0,002 \frac{in}{tahun} \times 10 tahun$

= 0,0371 in

= 0,0009 m

- **Tebal dinding ellipsoidal,t_e**

$$t_e = \frac{PD_t}{2SE - 0,2P} + C$$

(Walas-Chemical Process Equipment, Tabel 18.4)

t_e $= \frac{17,6\ psi \times 22,6666\ in}{2(13700\ psi \times 0,85) - (0,2 \times 17,6\ psi)} + 0,002 \frac{in}{tahun} x 10 tahun$

= 0,0371 in = 0,0009 m

- **Design Pengaduk**

Untuk umpan dengan viskositas ≤ 4.000 cP, maka dipilih pengaduk jenis propeller berdaun tiga (Walas, hal 288)

Untuk mencegah vortex, maka pada tangki dipasang baffle

Dt : diameter tangki

Da : diameter impeller

E : tinggi turbin dari dasar tangki

L : panjang blade pada turbin

W : lebar blade pada turbin

j : lebar baffle

Dt : 0,5757 m

Diameter Pengaduk (d)

D $= \frac{Dt}{3}$

= 0,1919 m = 0,6295 ft

Panjang daun pengaduk (L)

L $= \frac{D}{4}$

$= 0,0480$ m $= 0,1574$ ft

Lebar daun pengaduk (W)

W $= \frac{D}{5}$

$= 0,0384$ m $= 0,1259$ ft

Tinggi pengaduk dari dasar tangki (E)

E $= \frac{Dt}{3}$

$= 0,1919$ m $= 0,6295$ ft

Lebar baffle (j)

J $= \frac{Dt}{12}$

$= 0,0480$ m $= 0,1574$ ft

Kecepatan putar pengaduk (N) (Treybal, Pers 6.18)

$$\frac{N \times d}{\left(\frac{\sigma\, g_c}{\rho}\right)^{0.25}} = 1{,}22 + 1{,}25\left(\frac{D_B}{d}\right)$$

Σ $= 0,0000054$ lb/ft

gc $= 32,17$ ft/dt^2

N $= 1,22$ rps

Daya pengadukan

$$N_{Re} = \frac{\rho \times N \times d^2}{\mu}$$

Nre $= 567939$

Karena NRe > 10000, maka

$$P = \frac{K_T N^3 d^5 \rho}{g_c}$$ (Mc Cabe, Pers 9.24)

Kt = 0,32

P = 0,2833 lb/ft dt

= 0,000515 HP

Efisiensi motor penggerak 80%

Daya Motor = 0,000515 /0,8

= 0,000644 HP

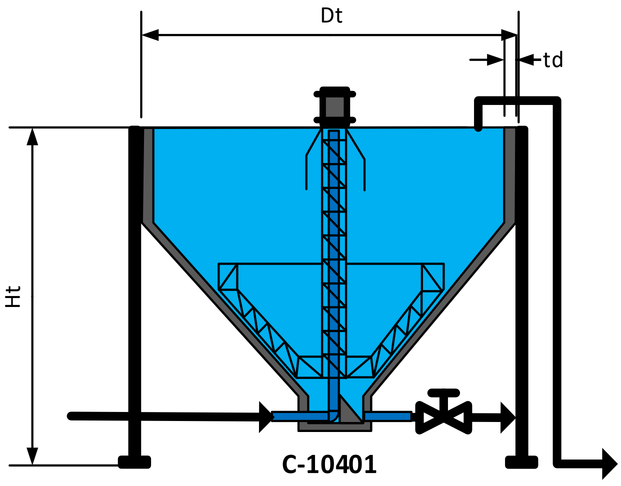
5. Clarifier (C-10401)

Fungsi = Tempat pencampuran, pembentukan dan pengendapan flok-flok yang terkandung dalam air

Dimensi = Silinder dengan alas kerucut

Bahan = Carbon Steel

Gambar =



Data Operasi :

Laju alir massa = 8302,28 kg/jam = 18298,22lb/jam

Laju alir volumetrik = 8,30 m^3/jam = 8302,28liter/jam

= 680,18 ft^3/s

Densitas = 1000,00 kg/m^3 = 62,43 lb/ft^3

Waktu tinggal = 1,00 Jam

Viskositas Campuran = 1,00 cp = 0,00067 lb/ft h

Q = 8,3022785 m^3/Jam

Kapasitas tangki, V_t

$V_b = \frac{m \times t}{\rho}$ $= \frac{8302,28\ \frac{kg}{jam} x\ 1\ jam}{1000 ^{kg}/_{m^3}}$

= 8,30 m^3

= 2193,23gal

Faktor keamanan 10%

Maka,

V_b = 0,9 V_t

$V_t = \frac{V_p}{0,9}$

$= \frac{8,30\ m^3}{0,9}$

= 9,2248 m^3 = 325,73ft^3

Dimensi tangki

- **Volume silinder, V_s**

  $V_s = \frac{\pi}{4} \times D_t^2 \times H_s$ $H_s = D_t$

Maka,

$V_s = \frac{\pi}{4} \times D_t^3$

- **Volume Kerucut, V_k**

  $V_K = \frac{1}{3}\pi \times D_t^2\ x\ He$ He = 1/2 Dt (peter's, Tabel 4)

  V_K = 0,523 x D_t^3

  V_K = 0,915 m

- **Diameter tangki, D_t**

D_t^3 = $\frac{Vt}{1,308}$

D_t^3 = $\frac{9,2248}{1,308}$

D_t^3 = 7,0526 m^3

D_t = $\sqrt[3]{7,0526}$

= 1,9165 m

= 75,4510 in

= 6,2860 ft

- **Tinggi tangki, H_t**

Tinggi silinder, Hs = D_t

= 1,9165 m

Tinggi Kerucut, H_k = ½ D_t

= 0,9582 m

(walas, Tabel 18.5)

Tinggi total, H_t = Hs + Hk

H_t = 2,8747 m

- **Tinggi cairan dalam tangki, H_c**

H_c = $\frac{volume\ bahan}{volume\ tangki} x\ Hk$

= 2,5872 m

= 8,4861 ft

- **Tekanan cairan dalam tangki, P_c**

P_c = $\rho \times g \times H_c$

= 1000 kg/m^3 x 10 m/dt^2 x 2,5872 m

= 25872,198 kg/m s^2

= 3,7525 psi

Pop = 1 atm

= 14,7 psi

- **Tekanan desain, P_d**

  Pd = Pop + Pc

  = (14,7 + 3,7551)

  = 18,4525 Psi

- **Tebal dinding tangki, t_d**

  $t_d = \frac{PR}{SE - 0,6P} + C$ (Walas, Tabel 18.4)

- Tekanan desain, P_d : 18,5 psi
- Jari-jari, R : 37,7255 in
- *Allowable stress*, S : 13700 psi (Walas, Tabel 18.4)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)
- Lama tahun digunakan : 10 tahun

Maka,

$$td = \frac{18,5\ psi \times 37,7255\ in}{(13700\ psi \times 0,85) - (0,6 \times 18,5\ psi)} + 0,002 \frac{in}{tahun} \times 10 tahun$$

= 0,0798 in

= 0,0020 m

- **Design Pengaduk**

  Untuk umpan dengan viskositas ≤ 4.000 cP, maka dipilih pengaduk jenis propeller berdaun tiga (Walas, hal 288)

  Untuk mencegah vortex, maka pada tangki dipasang baffle

  Dt : diameter tangki

  Da : diameter impeller

  E : tinggi turbin dari dasar tangki

  L : panjang blade pada turbin

  W : lebar blade pada turbin

  j : lebar baffle

Dt : 1,9165 m

Diameter Pengaduk (d)

D $= \frac{Dt}{3}$

$= 0{,}6388$ m

$= 2{,}0953$ ft

Panjang daun pengaduk (L)

L $= \frac{D}{4}$

$= 0{,}1597$ m

$= 0{,}5238$ ft

Lebar daun pengaduk (W)

W $= \frac{D}{5}$

$= 0{,}1278$ m

$= 0{,}4191$ ft

Tinggi pengaduk dari dasar tangki (E)

E $= \frac{Dt}{3}$

$= 0{,}6388$ m

$= 2{,}0953$ ft

Lebar baffle (j)

J $= \frac{Dt}{12}$

$= 0{,}1597$ m

$= 0{,}5238$ ft

Kecepatan putar pengaduk (N) (Treybal, Pers 6.18)

$$\frac{N \times d}{\left(\frac{\sigma\, g_c}{\rho}\right)^{0.25}} = 1{,}22 + 1{,}25\left(\frac{D_B}{d}\right)$$

Σ $= 0{,}0000054$ lb/ft

gc = 32,17 ft/dt^2

N = 1,22 rps

Daya pengadukan

$$N_{Re} = \frac{\rho \times N \times d^2}{\mu}$$

Nre = 497654

Karena NRe > 10000, maka

$$P = \frac{K_T N^3 d^5 \rho}{g_c}$$

(Mc Cabe, Pers 9.24)

Kt = 0,32

P = 1,1279 lb/ft dt

= 0,0021 HP

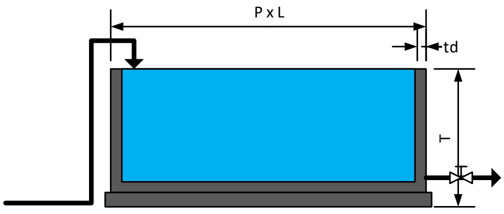
Efisiensi motor penggerak 80%

Daya Motor = 0,0021 /0,8

= 0,0026 HP

6. Bak Penampung Air Sungai

Fungsi : Menampung air dari Clarifier
Jenis : Persegi Panjang
Jumlah : 1 buah
Konstruksi : Beton bertulang
Gambar :



Data :

Laju alir massa = 8302,28 kg/jam

Densitas = 1000 kg/m^3

Waktu tinggal = 1 jam

Laju alir volumetrik, Q

$$Q = \frac{m}{\rho} = \frac{8302{,}28\ \frac{kg}{jam}}{1000\ kg/m^3} = 8{,}3023\ m^3/jam$$

Dimensi bak

$$V = 8{,}3023 m^3/jam \ \text{x}\ 1\ jam$$

$$= 8{,}3023\ m^3$$

Faktor keamanan 10%

$$\text{Volume bak} = \frac{8{,}3023\ m^3}{0{,}9} = 9{,}2248\ m^3$$

Perbandingan dimensi bak penampung yaitu P : L : T = 3 : 2 : 1

Volume bak = panjang x lebar x tinggi

9,2248 = 3T x 2T x T

$6T^3 = 9{,}2248\ m^3$
T = 1,1540 m

Sehingga diperoleh dimensi bak :

Panjang = 3T = 3,4620 m
Lebar = 2T = 2,3080 m
Tinggi = 1T = 1,1540 m

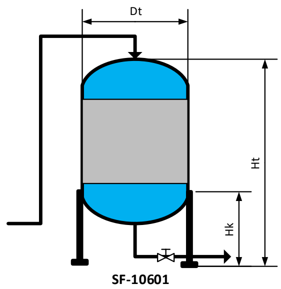
7. Tangki Sand Filter

Fungsi : Menyaring sisa flok dalam air dari Clarifier

Jenis : Silinder vertikal dengan alas dan tutup *ellipsoidal*

Konstruksi : *Carbon Steel*

Gambar :



Data :

Laju alir massa = 8302,28 kg/jam

Densitas = 997 kg/m^3 = 62,24 lb/ft^3

Waktu tinggal = 5 menit = 0,08 jam

Viskositas = 1 cp = 0,0007 lb/ft h

Q = 8,3273 m^3/jam

Vc = 0,6939 m^3

Vt = 0,7710 m^3

Kondisi Filter :

Porositas unggun, Ɛ = 0,4

Air yang terisi dalam unggun 80% dari air masuk

Volume ruang kosong = Volume yang terisi air

Volume unggun = V air yang mengisi unggun + V partikel

Air yang mengisi unggun = 0,617 m^3

Volume partikel = 1,542 m^3

Volume unggun = 2,159 m^3 = 76,2341 ft^3

volume air yang tidak mengisi unggun = 20% x Volume unggun

= 0,432 m^3

= 15,2463 ft^3

Sehingga, Volume Tangki = V unggun + V air yang tidak mengisi unggun
= 2,159 + 0,432

= 2,591 m^3 = 91,478 ft^3

Direncanakan akan digunakan 3 unit sand filter sehingga kapasitas masing-masing bak menjadi :

Vt = 0,96 m^3

Dimensi tangki

- **Volume silinder, V_s**

$V_s = \frac{\pi}{4} \times D_t^2 \times H_t$ $H_t = ½ D_t$

Maka,

$V_s = \frac{\pi}{8} \times D_t^3$

- **Volume ellipsoidal, V_e**

$V_e = \frac{\pi}{24} \times D_t^3$ He = ¼ Dt (peter's, Tabel 4)

$V_e = 0,042 \times D_t^3$

- **Diameter tangki, D_t**

$V_t = V_s + V_e$

$= \left(\frac{\pi}{8} \times D_t^3\right) + (\ 0,042 \times D_t^3)$

$V_t = 0,4345 \times D_t^3$

$D_t^3 = \frac{Vt}{0,4345}$

$D_t^3 = \frac{0,9595}{0,4345}$

$D_t^3 = 2,2083\ m^3$

$D_t = \sqrt[3]{2,2083}$

= 1,3019 m

= 51,2553 in

= 4,2702 ft

- **Tinggi tangki, H_t**

Tinggi Kaki Tangki = 0,5 m

Tinggi silinder, Hs = 1/2 D_t

= 0,6509 m

Tinggi ellipsoidal, H_e = ¼ D_t

= 0,3255 m (walas, Tabel 18.5)

Tinggi total, H_t = tinggi silinder + 2tinggi ellipsoidal + tinggi kaki

H_t = 1,8019 m

- **Tinggi cairan dalam tangki, H_c**

H_c $= \frac{volume\ bahan}{volume\ tangki} \times H_t - Hk$

= 1,1717 m

= 644,434 ft

- **Tekanan cairan dalam tangki, P_c**

P_c $= \rho \times g \times H_c$

= 997 kg/m^3 x 10 m/dt^2 x 1,1717 m

= 11681,834 kg/m s^2

= 1,6943 psi

Pop = 1 atm

= 14,7 psi

- **Tekanan desain, P_d**

Pd = Pop + Pc

= (14,7 +1,6943) psi

= 16,3943 Psi

- **Tebal dinding tangki, t_d**

t_d $= \frac{PR}{SE - 0{,}6P} + C$ (Walas, Tabel 18.4)

- Tekanan desain, P_d : 16,4 psi
- Jari-jari, R : 25,6276in
- *Allowable stress*, S : 13700 psi (Walas, Tabel 18.4)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)

- Lama tahun digunakan : 10 tahun

Maka,

td $= \frac{16{,}4\ psi \times 25{,}6276\ in}{(13700\ psi \times 0{,}85) - (0{,}6 \times 16{,}4\ psi)} + 0{,}002\frac{in}{tahun} \times 10 tahun$

= 0,0561 in

= 0,0014 m

- **Tebal dinding ellipsoidal,t_e**

$$t_e = \frac{PD_t}{2SE - 0{,}2P} + C$$ *(Walas-Chemical Process Equipment*, Tabel 18.4)

t_e $= \frac{16{,}4\ psi \times 1{,}08\ in}{2(13700\ psi \times 0{,}85) - (0{,}2 \times 16{,}4\ psi)} + 0{,}002\frac{in}{tahun} x 10 tahun$

= 0,0561 in

= 0,0014 m

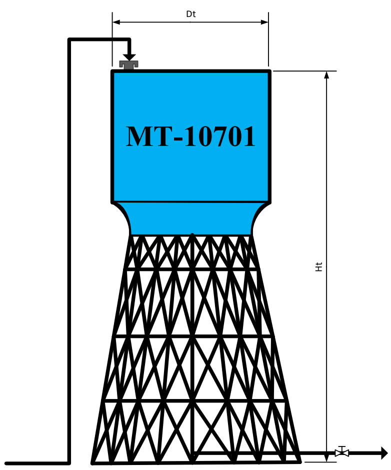
8. Tangki Menara

Fungsi : Menampung Air Pengolahan *Water treatment*

Jenis : Silinder dengan alas dan tutup Datar

Konstruksi : *Carbon Steel*

Gambar :



Data :

Laju alir massa = 8302,28 kg/jam

Densitas = 997 kg/m^3 = 62,24 lb/ft^3

Viskositas = 1 cp = 0,0007 lb/ft h

Q = 8,32726 m^3/jam

Penggunaan besi yang dipakai untuk menara :

Jenis = Besi Siku UNP 100 dan UNP 80

Tebal = 10 mm dan 8 mm

Tinggi Kaki Menara = 5 m

Besi Yang Terpakai = UNP 100

= Jumlah Kaki x Tinggi Menara

= 50 m

= UNP 80 (Asumsi row persilangan 2 m dengan sekat 1 m)

= 100 m

Kapasitas tangki, V_t

$$V_c = \frac{m \times t}{\rho} = \frac{8302{,}28\ \frac{kg}{jam} x\ 1\ jam}{997\ {kg}/{m^3}}$$

= 8,33 m^3

= 2199,8290 gal

Faktor keamanan 10%

Maka,

V_c = 0,9 V_t

$$V_t = \frac{V_p}{0{,}9}$$

$$= \frac{8{,}33\ m^3}{0{,}9}$$

= 9,2525 m^3 = 326,7062 ft^3

Dimensi tangki

- **Volume silinder, V_s**

  $V_s = \frac{\pi}{4} \times D_t^2 \times H_t$ $H_s = D_t$

Maka,

$V_s = \frac{\pi}{4} \times D_t^3$

$V_s = 0{,}785\ D_t^3$

$V_s = 0{,}067\ m^3$

- **Volume Tangki, V_t**

  $V_t = V_s$ (peter's, Tabel 4)

  $V_t = 0{,}785 \times D_t^3$

- **Diameter tangki, D_t**

  $D_t^3 = \frac{Vt}{9{,}2525}$

  $D_t^3 = \frac{0{,}785}{9{,}2525}$

  $D_t^3 = 0{,}0848\ m^3$

  $D_t = \sqrt[3]{0{,}0848}$

  $= 0{,}43983$ m

  $= 17{,}3138$ in

  $= 1{,}4425$ ft

- **Tinggi tangki, H_t**

  Tinggi Kaki Tangki,Hk = 5 m

  Tinggi silinder, Hs $= D_t$

  $= 0{,}4398$ m

  Tinggi total, H_t = Hs + Hk

  H_t = 5,4398 m

- **Tinggi cairan dalam tangki, H_c**

  $H_c = \frac{volume\ bahan}{volume\ tangki} \times H_t - Hk$

= 0,3958 m = 1,2982 ft

- **Tekanan cairan dalam tangki, P_c**

P_c = $\rho \times g \times H_c$

= 997 kg/m^3 x 10 m/dt^2 x 0,3958 m

= 3946,07 kg/m s^2

= 0,5723 psi

Pop = 1 atm = 14,7 psi

- **Tekanan desain, P_d**

Pd = Pop + Pc

= (14,7 + 0,5723) psi

= 15,2723 Psi

- **Tebal dinding tangki, t_d**

$t_d = \frac{PR}{SE-0,6P} + C$ (Walas, Tabel 18.4)

- Tekanan desain, P_d : 15,3 psi
- Jari-jari, R : 8,6569 in
- *Allowable stress*, S : 13700 psi (Walas, Tabel 18.4)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)
- Lama tahun digunakan : 10 tahun

Maka,

$$td = \frac{15,3\ psi \times 8,6569\ in}{(13700\ psi \times 0,85) - (0,6 \times 15,3\ psi)} + 0,002 \frac{in}{tahun} \times 10 tahun$$

= 0,0314 in

= 0,0008 m

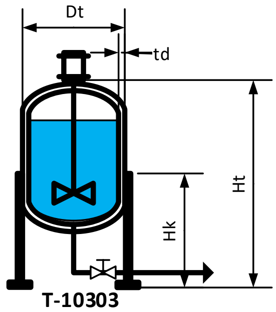
9. Tangki Kaporit

Fungsi : Tempat melarutkan Kaporit

Jenis : Silinder vertikal dengan alas dan tutup *ellipsoidal*

Konstruksi : *Carbon Steel*

Gambar :



Data :

Laju alir massa Air	= 8302,28 kg/jam	= 18298,22 lb/jam
Laju alir volumetrik	= 8,33 m^3/jam	= 8327,26 liter/jam
Densitas	= 997 kg/m^3	= 62,24 lb/ft^3
Waktu tinggal	= 30 Hari	= 720 jam
Viskositas Campuran	= 1,042 cp	= 0,00070 lb/ft h
Densitas Kaporit	= 2350 kg/m^3	= 146,7105 lb/ft^3

Kebutuhan Kaporit :

Dosis Kaporit (Sumber : Jurnal Skala Husada Vol 8 no 2 hal 191-195)

= Kaporit yang digunakan mengandung 60% Chlor

= 3,00 mg/l = 0,000003 kg/l

= 0,000003 kg/l x 8327,26 liter/jam

= 0,02 kg/jam

Volume Kaporit = 0,02 kg/jam / 2350 kg/m^3

= 0,00001 m^3/jam

Kapasitas tangki, V_t

V_c $= \frac{m \times t}{\rho}$ = 0,00001 m^3/jam x 720 jam

= 0,00765 m^3

= 2,02 gal

Faktor keamanan 10%

Maka,

V_c = 0,9 V_t

$V_t = \frac{V_c}{0,9}$

$= \frac{0,00765\ m^3}{0,9}$

= 0,008504 m^3

= 0,30 ft^3

Dimensi tangki

- **Volume silinder, V_s**

$V_s = \frac{\pi}{4} \times D_t^2 \times H_t$ $H_s = D_t$

Maka,

$V_s = \frac{\pi}{4} \times D_t^3$

V_s = 0,785 Dt^3

- **Volume ellipsoidal, V_e**

$V_e = \frac{\pi}{24} \times D_t^3$ He = ¼ Dt (peter's, Tabel 4)

V_e = 0,13 x D_t^3

- **Diameter tangki, D_t**

$V_t = V_s + V_e$

$= (0,785 \times D_t^3) + (0,13 \times D_t^3)$

V_t = 0,915 x D_t^3

$D_t^3 \quad = \frac{Vt}{0,915}$

$D_t^3 \quad = \frac{0,0085}{0,915}$

D_t^3 = 0,0093 m^3

$D_t \quad = \sqrt[3]{0,0093}$

= 0,2106 m

= 0,6907 ft = 8,2906 in

- **Tinggi tangki, H_t**

Tinggi Kaki Tangki = 0,5 m

Tinggi silinder, Hs = D_t

= 0,2106 m

Tinggi ellipsoidal, H_e = ¼ D_t

= 0,0526 m

(walas, Tabel 18.5)

Tinggi total, H_t = tinggi silinder + 2tinggi ellipsoidal + tinggi kaki

H_t = 0,8159 m

- **Tinggi cairan dalam tangki, H_c**

$H_c \quad = \frac{volume\ bahan}{volume\ tangki} \times H_t - Hk$

= 0,2843 m

= 0,9324 ft

- **Tekanan cairan dalam tangki, P_c**

$P_c \quad = \rho \times g \times H_c$

= 2350 kg/m^3 x 10 m/dt^2 x 0,2843 m

= 6680,6753 kg/m s^2

= 0,9690 psi

Pop = 1 atm = 14,7 psi

- **Tekanan desain, P_d**

  Pd = Pop + Pc

  = (14,7 + 0,9690) psi

  = 15,6690 Psi

- **Tebal dinding tangki, t_d**

  $t_d = \frac{PR}{SE - 0{,}6P} + C$ (Walas, Tabel 18.4)

- Tekanan desain, P_d : 15,7 psi
- Jari-jari, R : 4,1453 in
- *Allowable stress*, S : 13700 psi (Walas, Tabel 18.4)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)
- Lama tahun digunakan : 10 tahun

Maka,

$$td = \frac{15{,}7 psi \times 4{,}1453\ in}{(13700\ psi \times 0{,}85) - (0{,}6 \times 15{,}7\ psi)} + 0{,}002 \frac{in}{tahun} \times 10 tahun$$

= 0,0256 in

= 0,0006 m

- **Tebal dinding ellipsoidal,t_e**

  $$t_e = \frac{PD_t}{2SE - 0{,}2P} + C$$ (*Walas-Chemical Process Equipment*, Tabel 18.4)

$$t_e = \frac{15{,}7\ psi \times 0{,}2106\ \ in}{2(13700\ psi\ \times 0{,}85) - (0{,}2 \times 15{,}7\ psi)} + 0{,}002 \frac{in}{tahun} x10 tahun$$

= 0,0256 in

= 0,0006 m

- **Design Pengaduk**

  Untuk umpan dengan viskositas ≤ 4.000 cP, maka dipilih pengaduk jenis propeller berdaun tiga (Walas, hal 288)

  Untuk mencegah vortex, maka pada tangki dipasang baffle

  Dt : diameter tangki

Da : diameter impeller

E : tinggi turbin dari dasar tangki

L : panjang blade pada turbin

W : lebar blade pada turbin

j : lebar baffle

Dt : 0,2106 m

Diameter Pengaduk (d)

D $= \frac{Dt}{3}$

$= 0,0702$ m

$= 0,2302$ ft

Panjang daun pengaduk (L)

L $= \frac{D}{4}$

$= 0,0175$ m

$= 0,0576$ ft

Lebar daun pengaduk (W)

W $= \frac{D}{5}$

$= 0,0140$ m $= 0,0460$ ft

Tinggi pengaduk dari dasar tangki (E)

E $= \frac{Dt}{3}$

$= 0,0702$ m $= 0,2302$ ft

Lebar baffle (j)

J $= \frac{Dt}{12}$

$= 0,0175$ m

$= 0,0576$ ft

Kecepatan putar pengaduk (N) (Treybal, Pers 6.18)

$$\frac{N \times d}{\left(\frac{\sigma\, g_c}{\rho}\right)^{0.25}} = 1{,}22 + 1{,}25\left(\frac{D_B}{d}\right)$$

Σ = 0,0000054 lb/ft

gc = 32,17 ft/dt^2

N = 1,2210 rps

Daya pengadukan

$$N_{Re} = \frac{\rho \times N \times d^2}{\mu}$$

Nre = 575,5038

Karena NRe > 10000, maka

$$P = \frac{K_T N^3 d^5 \rho}{g_c}$$

(Mc Cabe, Pers 9.24)

Kt = 0,32

P = 0,0007 lb/ft dt

= 0,00000133 HP

Efisiensi motor penggerak 80%

Daya Motor = 0,00000133 /0,8

= 0,00000165705 HP

10. *Feed Water Tank*

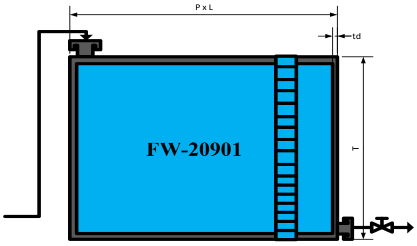
Fungsi : Menampung Air dari *Kation Anion Exchanger* sebagai air umpan boiler dan Air *Cooling Tower*

Jenis : Persegi Panjang

Jumlah : 1 buah

Konstruksi : Beton bertulang

Gambar :



Data :

Laju alir massa = 8302,28 kg/jam

Densitas = 997 kg/m^3

Waktu tinggal = 1 jam

Laju alir volumetrik, Q

$$Q = \frac{m}{\rho} = \frac{8302{,}28\ \frac{kg}{jam}}{997\ kg/m^3} = 8{,}3273\ m^3/jam$$

Dimensi bak

V = 8,3273 m^3/jam x 1 jam

= 8,3273 m^3

Faktor keamanan 10%

$$\text{Volume bak} = \frac{8{,}3273\ m^3}{0{,}9} = 9{,}2525\ m^3$$

Perbandingan dimensi bak penampung yaitu P : L : T = 3 : 2 : 1

Volume bak = panjang x lebar x tinggi

9,2525 = 3T x 2T x T

6T^3 9,2525 m^3

T = 1,1552 m

Sehingga diperoleh dimensi bak :

Panjang = 3T = 3,4655 m

Lebar = 2T = 2,3103 m

Tinggi = 1T = 1,1552 m

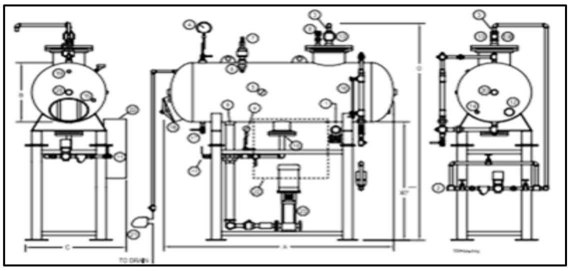
11. *Daerator*

Fungsi : Menghilangkan kadar oksigen terlarut dalam air umpan boiler

Tipe : *Parker Spray Daerator* Model no DAS2436

Jumlah : 1 unit

Gambar :



Data :

Air umpan boiler = 5177,476 kg/jam

= 11411,16 lb/jam

Laju alir volumetrik, Q

$$Q = \frac{m}{\rho} = \frac{5177,476 \frac{kg}{jam}}{997\ kg/m^3}$$

= 5,1931 m^3/jam

= 5193,0547 liter/jam

= 0,0509 ft^3/s

Kebutuhan Hidrazin :

Untuk 1,00 mg/L Oksigen = 1,00 mg/L Hidrazin

jumlah Hidrazin = 5,00 mg/liter

= 0,000005 kg/liter

= 0,0260 kg/jam

= 0,62 kg/hari

= 1,37 lb/hari

Hidrazin yang digunakan berupa larutan Hidrazin dengan konsentrasi 25% berat

Massa Hidrazin = 2,49 kg/hari

Volume Hidrazin = 0,00244 m^3/hari

Densitas =1,02 g/cm^3 = 1020 kg/m^3

Direncanakan akan didesain parker spray deaerator model DAS2436 yang mampu mengolah : 500,6484 lb/jam Air umpan boiler

Berdasarkan data tersebut, diperoleh kapasitas alat sebagai berikut :

Tipe = PARKER SPRAY model no. DAS2436

Length = 54 in = 1,37 m

Diameter = 24 in = 0,61 m

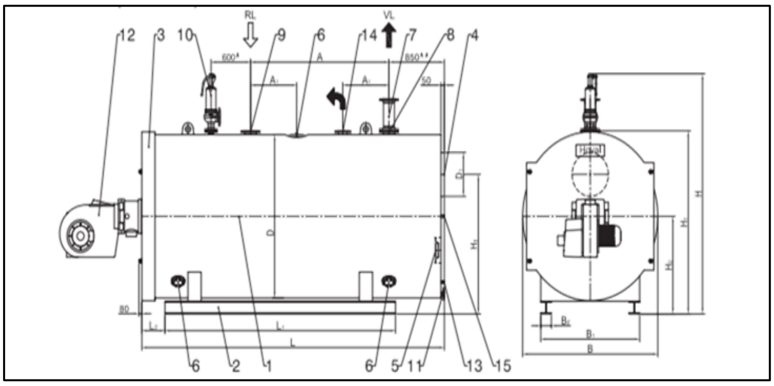
Height = 104 in = 2,64 m

Water inlet = 0,75 in = 0,02 m

Steam inlet = 0,75 in = 0,02 m

12. Boiler

Fungsi : untuk menghasilkan steam
Tipe : *THW-1 THE (10/05)*
Jumlah : 1 buah
Gambar :



Kebutuhan air umpan boiler = 5177,4755 kg/jam

Jika kondensat yang dapat diregenerasi = dan asumsi 89% yang dapat disirkulasikan kembali

Sehingga jumlah kondensat yang disirkulasikan kembali adalah :

= 4607,9532 kg/jam

Maka air yang di make-up yang dibutuhkan oleh boiler adalah :
= 569,5223 kg/jam

Data operasi

Steam yang dibutuhkan : 5177,4755 kg/jam = 11414,2626 lb/jam

Kondensat yang diregenrasi : 4607,9532 kg/jam = 10158,6937 lb/jam

Air make-up : 569,5223 kg/jam = 1255,5689 lb/jam

Jumlah steam yang dihasilkan : 6730,7182 kg/jam = 14838,5414 lb/jam

= 6,7510 m^3/jam

= 6750,9711 L/jam

Berdasarkan data jumlah steam yang dihasilkan, maka dipilih boiler tipe THW-I THE (17/10) dengan spesifikasi sebagai berikut :

THW-I HTE (10/05-34/25)

Technical data

Type		(10/05)	(13/08)	(17/10)	(22/15)	(27/20)	(34/25)
• Nominal output (oil firing)	kW	900/500	1280/800	1550/1000	2200/1500	2700/2000	3400/2500
• Nominal output (gas firing)	kW	900/500	1290/800	1560/1000	2200/1500	2700/2000	3400/2500
• Operating temperature max. (SBT) [1)]				depending on net pressure			
• Temperature level flow/return				depending on net pressure			
• Safety valve pressure	bar	10	10	10	10	10	10
	bar	13	13	13	13	13	13
	bar	16	16	16	16	16	16
• Boiler efficiency at 120 °C (natural gas) *	%	87.8/89.3	88.0/90.0	89.0/90.7	88.3/90.2	88.7/90.2	88.7/90.2
• Boiler efficiency at 120 °C (diesel oil) *	%	88.8/90.2	88.9/90.8	89.8/91.4	89.2/90.9	89.6/90.9	89.6/91.0
• Flue gas resistance	mbar	7.7	10.3	9.7	11.0	11.0	13.0
at max. boiler load of	kW	900	1290	1560	2200	2700	3400
• Water content	l	1700	1900	2100	2800	3500	4500
• Water flow resistance **	mbar	100	100	150	150	200	150
	z-value ***	0.04873	0.02883	0.02523	0.01506	0.01335	0.00631
• Flue gas temperature after boiler (natural gas)	°C	272/235	269/225	249/210	265/224	257/225	258/225
• Flue gas temperature after boiler (diesel oil)	°C	262/227	260/217	241/203	255/216	248/218	249/218

[1)] Country and equipment specific
* efficiency for boiler middle temperature
** for boiler max. load and ΔT = 20 K
***for other flow rates use "z-value" for water side pressure loss calculation: Δp (mbar) = asked flow rate $(m^3/h)^2$ * z

Boiler type	Main dimensions					Boiler foundation					Transport dim.		F/R nozzle			Flue gas con.		SV	BS
	B Width	L Length	H	H_1	H_2	D	L_1	L_2	B_1	B_2	B_{min}	H^4_{min}	A	A_1	DN[1,3]	H_3	D_1	DN[1]	DN[1]
(10/05)	1570	2530	2150	1760	900	1500	1650	230	1050	60	1750	1960	850	300	100	1200	300	25	50
(13/08)	1620	2830	2250	1810	925	1550	2000	230	1100	60	1800	2010	1000	350	100	1250	350	32	50
(17/10)	1670	3030	2400	1860	975	1600	2150	230	1200	60	1900	2060	1000	350	125	1300	400	32	50
(22/15)	1770	3430	2500	1960	1000	1700	2650	230	1250	60	1950	2160	1600	600	150	1400	450	50	65
(27/20)	1870	3930	2650	2060	1050	1800	3000	230	1350	60	2050	2260	1800	600	150	1500	500	50	80
(34/25)	1970	4280	2750	2160	1100	1900	3500	230	1400	60	2150	2360	2100	700	150	1550	500	65	80
(39/30)	2020	4580	2800	2210	1125	1950	3500	230	1450	60	2200	2410	2100	700	200	1600	550	65	80
(43/35)	2070	4730	2980	2260	1150	2000	3500	230	1500	60	2250	2460	2100	700	200	1650	600	65	80
(48/40)	2170	5330	3130	2410	1250	2100	4000	350	1550	160	2350	2610	2500	800	200	1750	600	65	100
(54/45)	2220	5380	3180	2460	1325	2150	4000	350	1600	160	2400	2660	2500	800	200	1800	650	65	100
(59/50)	2270	5430	3280	2560	1350	2200	4500	350	1650	160	2450	2760	2500	800	250	1850	650	65	100
(68/60)	2370	5630	3470	2660	1400	2300	4500	350	1700	160	2550	2860	2500	800	250	1900	700	65	125
(78/70)	2470	5930	3570	2760	1450	2400	5000	350	1800	160	2650	2960	3000	900	250	2050	750	80	125
(89/80)	2570	6230	3670	2860	1500	2500	5000	350	1850	160	2750	3060	3000	900	250	2100	750	80	150
(99/90)	2670	6530	3770	2960	1550	2600	5500	350	1950	160	2850	3160	3000	900	250	2200	800	80	150
(115/100)	2770	6630	3980	3060	1600	2700	5500	350	2000	160	2950	3260	3000	900	300	2300	850	100	150
(130/120)	2870	6980	4130	3210	1700	2800	6000	400	2050	200	3050	3410	3500	1000	250	2400	900	100	150
(150/140)	3070	7180	4330	3410	1800	3000	6000	400	2200	200	3250	3610	3500	1000	250	2700	900	100	150
(170/160)	3270	7380	4500	3610	1900	3200	6000	400	2300	200	3450	3810	4000	1200	300	2650	1050	100	200
(190/180)	3470	7615	4900	3810	2000	3400	6000	400	2500	200	3550	4010	4000	1200	300	2750	1100	125	200
(210/200)	3670	7915	5200	4110	2200	3600	6000	400	2700	200	3700	4310	4000	1200	300	2950	1100	125	200

[1] DN/...PN 16/PN 40
[3] Diameter for standard ΔT = 20 K (from THW-I 130/120 HTE upwards ΔT = 30 K), other dimensions on request
[4] without armature tube

Daya operasi = 900 kw

Efisiensi = 89%

Temperatur flue gas = 180°C

Tekanan operasi	= 10 bar		
Panjang	= 3030 mm	= 119,29 in	= 3,03 m
Lebar	= 1670 mm	= 65,74 in	= 1,67 m
Tinggi	= 2400 mm	= 94,48 in	= 2,4 m

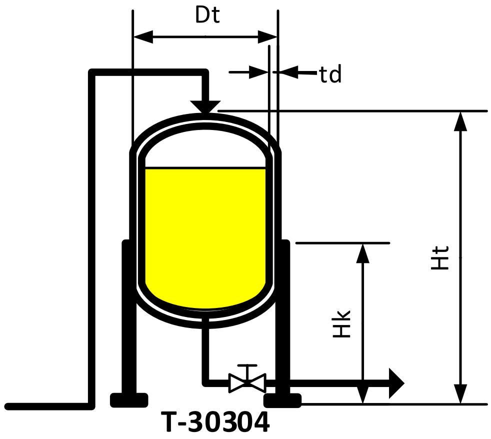
13. Tangki Solar

Fungsi : Tempat Penyimpanan Bahan Bakar

Jenis : Silinder vertikal dengan alas dan tutup *ellipsoidal*

Konstruksi : *Carbon Steel*

Gambar :



Data :

Kebutuhan Listrik = 621,17 kW

Laju alir = 120,51 kg/jam

Densitas = 0,83 kg/m^3 = 0,05 lb/ft^3

Waktu penggunaan = 1 jam

Kebutuhan Solar :

Maka,

Untuk generator berpenggerak motor diesel : **194 x 10 x 24 = 46.560 g/hari = 46,56 kg/hari**

Jika berat jenis bahan bakar solar adalah 0.832 kg/l, maka dalam satuan konsumsi.

Bahan bakar dalam satuan liter adalah :

46,56 / 0.832 = 55,96 liter/hari

Berat jenis bahan bakar

Berat jenis suatu bahan bakar bergantung pada temperatur dan kandungannya, secara umum dapat diambil harga rata-rata dari Wikipedia sebagaimana dibawah ini:

- *Bensin 0,745 kg/l*
- *Solar 0,832 kg/l*

(Sumber : http://gudanggenset.com/berita-dan-artikel/cara-menghitung-pemakaian-bahan-bakar-pada-genset/)

Kapasitas tangki, V_t

V_b $= \frac{m \times t}{\rho}$ $= \frac{120{,}51\ \frac{kg}{jam} x\ 1\ jam}{0{,}83\ {kg}/{m^3}}$

$= 48{,}28\ m^3$

Jumlah 3 unit tangki $= 143{,}45\ m^3 / 3\ unit = 47{,}82\ m^3$

Faktor keamanan 10%

Maka,

$V_b = 0{,}9\ V_t$

$V_t = \frac{V_p}{0{,}9}$

$= \frac{48{,}28\ m^3}{0{,}9}$

$= 53{,}6443\ m^3$

$= 1894{,}18\ ft^3$

Dimensi tangki

- **Volume silinder, V_s**

$V_s = \frac{\pi}{4} \times D_t^2 \times H_t$ $\qquad H_s = D_t$

Maka,

V_s $= \frac{\pi}{4} \times D_t^3$

V_s = 0,785 Dt^3

- **Volume ellipsoidal, V_e**

  V_e $= \frac{\pi}{24} \times D_t^3$ He = ¼ Dt (peter's, Tabel 4)

  V_e = 0,13 x D_t^3

- **Diameter tangki, D_t**

  V_t = $V_s + V_e$

  $= (0{,}785 \times D_t^3)$ $+ (\ 0{,}13 \times D_t^3)$

  V_t = 0,915 x D_t^3

  D_t^3 $= \frac{Vt}{0{,}915}$

  D_t^3 $= \frac{53{,}6443}{0{,}915}$

  D_t^3 = 58,6277 m^3

  D_t $= \sqrt[3]{58{,}6277}$

  = 3,8795 m

  = 152,7368 in = 12,7248 ft

- **Tinggi tangki, H_t**

  Tinggi Kaki Tangki = 0,5 m

  Tinggi silinder, Hs = D_t

  = 3,8795 m

  Tinggi ellipsoidal, H_e = 1/4D_t

  = 0,9699 m (walas, Tabel 18.5)

  Tinggi total, H_t = tinggi silinder + tinggi ellipsoidal + tinggi kaki

  H_t = 5,3494 m

- **Tinggi cairan dalam tangki, H_c**

  H_c $= \frac{volume\ bahan}{volume\ tangki} \times H_t - Hk$

= 4,3645 m

= 14,3154 ft

- **Tekanan cairan dalam tangki, P_c**

P_c $= \rho \times g \times H_c$

= 1 kg/m^3 x 10 m/dt^2 x 4,3645 m

= 36,3123 kg/m s^2

= 0,0053 psi

Pop = 1 atm = 14,7 psi

- **Tekanan desain, P_d**

Pd = Pop + Pc

= (14,7 + 0,0053) psi

= 14,7053 Psi

- **Tebal dinding tangki, t_d**

$t_d = \frac{PR}{SE-0,6P} + C$ (Walas, Tabel 18.4)

- Tekanan desain, P_d : 14,7 psi
- Jari-jari, R : 76,3684 in
- *Allowable stress*, S : 13700 psi (Walas, Tabel 18.4)
- Efisiensi pengelasan, E : 0,85 (Peter, Tabel 4 Hal 538)
- Faktor korosi yang diizinkan : 0,002 in/thn (Perry's Tabel 23-2)
- Lama tahun digunakan : 10 tahun

Maka,

$$td = \frac{14,7\ psi \times 76,3684\ in}{(13700\ psi \times 0,85) - (0,6 \times 14,7\ psi)} + 0,002\frac{in}{tahun} \times 10 tahun$$

= 0,1165 in

= 0,0030 m

= 2,9594 mm

- **Tebal dinding ellipsoidal,t_e**

$$t_e = \frac{PD_t}{2SE - 0,2P} + C$$

(*Walas-Chemical Process Equipment*, Tabel 18.4)

$$t_e = \frac{14{,}7\,psi \times 3{,}8795\,in}{2(13700\,psi \times 0{,}85)-(0{,}2 \times 14{,}7\,psi)} + 0{,}002\frac{in}{tahun}x10tahun$$

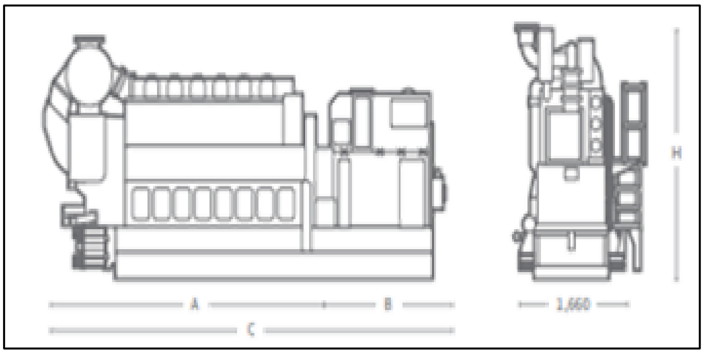
= 0,1164 in

= 0,0030 m

= 2,9578 mm

12. Generator

Fungsi : Sumber Arus Listrik
Tipe : 6H27DF Hyundai
Jumlah : 1 buah
Gambar :



Main Data / **Dimensions**

Speed	900rpm		1,000rpm		Dimension(mm)				Dry Mass(ton)	
Frequency	60Hz		50Hz							
	Eng.(kw)	Gen.(kw)	Eng.(kw)	Gen.(kw)	A	B	C	H	Engine	GenSet
6H27DF	1,710	1,624	1,860	1,767	4,414	2,262	6,676	3,103	23.5	33.7
7H27DF	1,995	1,895	2,170	2,061	4,797	2,262	7,059	3,241	27.7	37.7
8H27DF	2,280	2,177	2,480	2,368	5,311	2,340	7,651	3,371	34.0	44.8
9H27DF	2,565	2,462	2,790	2,678	5,691	2,490	8,181	3,371	36.2	47.2

Based on alternator efficiency of 95~96%.

Data Operasi
Speed = 900 rpm
Frequency = 60 Hz
Daya Engine = 1710 KW
Daya Generator = 1524 KW
Dimensi
Panjang = 6,68 m
Lebar = 1,66 m
Tinggi = 3,10 m

Berat Engine = 23,5 Ton

Berat Genset = 33,7 Ton

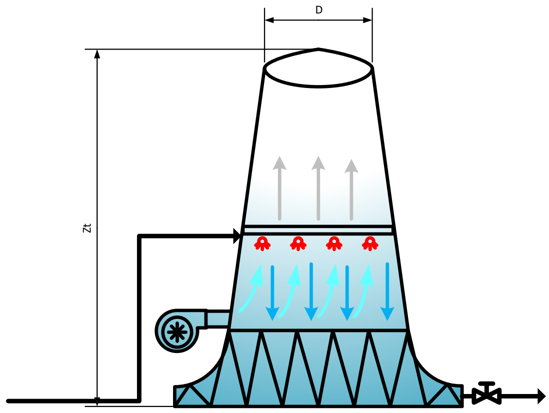
12. Cooling Tower

Fungsi : Mendinginkan air untuk kebutuhan air pendingin pada proses

Tipe : *Induced draft cooling tower*

Jumlah : 1 buah

Gambar :



Data :

Laju alir	= 8302,28 kg/jam	= 18303,2032 lb/jam
Densitas	= 997 kg/m^3	= 62,24 lb/ft^3
Viskositas	= 1 cP	= 0,00067 lb/ft.s
T in	= 40°C	= 104 F
T out	= 27°C	= 80,6 F
h udara	= 36,5 Btu/lb udara kering	
h air	= 105 Btu/lb udara kering	

Fig 12.2 dan fig 12.3 Perry's :

Temperatur bola basah = 72,5 F

Temperatur bola kering = 75 F

Tav = 23,4 F

Laju alir volumetrik, Wc

Wc = 8,3272 m^3/jam

= 36,6638 gal/menit

Tipe cooling tower yang digunakan adalah *Induced draft cooling tower* dengan aliran counter current

Cooling range = 23,4 F

Luas tower, A

Kandungan air, Ca = 1,5 gal/menit.ft^2 Fig 12.14 Perry's

Luas menara = 24,4425 ft^2

Faktor keamanan = 0,1

maka,

A = 27,1584 ft^2

Daya yang dibutuhkan fan

Performa standar menara 97%

maka daya yang diperoleh adalah 0,037 HP/ft^2 Fig 12.15 Perry's

sehingga,

Daya fan = 27,1584 x 0,037

= 1,0049 HP

Dimensi tower : Pers 12.15 Perry's

$$Dt = \frac{A \ x \ \sqrt{Z_t}}{C_t \ x \ \sqrt{C_t}}$$

Hal 12.21 Perry's

Dengan, Dt = Koefisien bahan menara

A = Luas menara

Zt = Tinggi menara

Ct = Koefisien performa menara

Untuk memperoleh nilai , Dt

$$\frac{W_L}{D_t} = 90.85\ x \left(\frac{\Delta h}{\Delta T}\right) x \sqrt{\Delta t + (0.3124\ \ \Delta h)}$$

Dengan, Δh = Perubahan panas

= ha - hu

= 83 Btu/lb

ΔT = Perubahan temperatur melalui menara

= 23,4 F

WL = Beban air pada menara

= 8302,28 kg/jam

= 18303,2032 lb/jam

Δt = T keluar - T bola kering

= 80,6 – 75

= 5,6 F

maka, a = 18303,20 lb/jam beban air pada menara

b = 90,85

c = Δh / ΔT

= 83 Btu/lb / 23,4 F

= 3,55 Btu/lb F

d = 11,00

e = 0,31

f = Δh x e

= 83 Btu/lb x 0,3124

= 25,93

g = d + f

= 36,93

h = $g^{0,5}$

= 6,08

i = b x c x h

= 1958,27 ft

j = a / i

= 9,35 ft

= 2,8496 m

Direncanakan Zt = 1.5 D

= 4,2744 m

= 14,0199 ft

LAMPIRAN D
PERHITUNGAN ANALISA EKONOMI

Analisa ekonomi dihitung untuk menentukan jumlah modal yang dibutuhkan untuk mendirikan dan mengoperasikan pabrik serta tinjauan kelayakan suatu pabrik.

1. Perhitungan Jumlah Modal

Pra rancangan pabrik Propilen glikol dari gliserol dan hidrogen dengan kapasitas produksi 70.000 ton/tahun ini mengolah bahan cair. Dalam hal ini, untuk menentukan jumlah modal yang dibutuhkan untuk mendirikan dan mengoperasikan pabrik diperoleh dari hasil perkiraan dengan metoda *percentage delivered equipment cost* untuk *solid-liquid processing plant* (Peters, 1991).

1.1 Perhitungan Harga Alat

Untuk menghitung harga peralatan pada tahun 2030 ditentukan dengan persamaan :

$$\text{Harga 2030} = \text{Harga tahun 2024} \times \left(\frac{\text{Indeks harga 2028}}{\text{Indeks harga 2022}}\right) \qquad \text{(Peter's, 1991)}$$

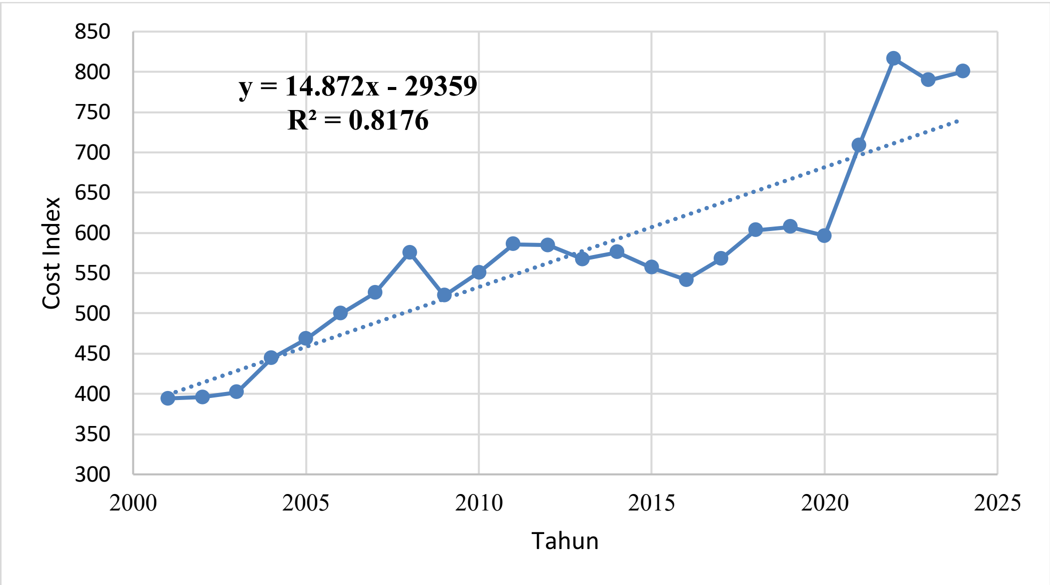
Daftar indeks harga rata-rata tahunan menurut *Engineering Plant Cost* dapat dilihat pada Tabel D.1 dan Gambar D.1 di bawah ini.

Tabel D.1 Daftar Indeks Harga Rata-Rata Tahunan

Tahun	Cost Index
2001	394,3
2002	395,6
2003	402,0
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	603,1
2019	607,5
2020	596,2
2021	708,8
2022	816,0
2023	789,6
2024	800,3
2025	756,8
2026	771,7
2027	786,5
2028	801,4
2029	816,3
2030	831,2

(Sumber : Chemical Engineering Plant Cost index,http://www.chemengonline.com/pci-home)



Gambar D.1 Grafik Hubungan Harga Indeks Terhadap Tahun

Persamaan yang diperoleh sesuai Gambar D.1 adalah :

$$y = 14,872x - 29359$$

Dengan menggunakan persamaan di atas dapat dicari harga indeks pada tahun perhitungan dan perancangan pabrik yaitu tahun 2024 dan perancangan pabrik di tahun 2030 yaitu :

x = 2024

y = 713

- *Cost index* tahun 2028

  y = 14,5 (2028) – 28606

  y = 800

Contoh perhitungan harga peralatan :

Harga tangki penyimpanan gliserol dengan kapasitas 1339,0678 m^3 pada tahun 2014 adalah US$ 22.900

Nilai indeks harga tahun 2014: 576,1

Nilai indeks harga tahun 2030: 831,2

Harga satu buah tangki penyimpanan gliserol tahun 2030 adalah :

Harga 2030 $= 22.900 \times \left(\frac{831,1}{576,1}\right)$

= US$ 33.039

= Rp 1.077.552.138

Diketahui : 1 Dollar = Rp. 16.307,45 (September 2025)

Dengan cara yang sama, diperoleh perkiraan harga peralatan utama dan utilitas seperti yang terlihat pada Tabel D.2 dan Tabel D.3

Tabel D.2 Perkiraan Harga Peralatan Utama

No	Nama Alat	Kode Alat	Unit	Harga per Unit tahun 2014 (Ny)		Harga per Unit tahun 2030 (Nx)	
				USD	Rp	USD	Rp
1	Tangki Penyimpanan C3H8O3 dan C3H6O2		2	22.900	746.881.210	33.038,6	1.077.552.137,7
2	Tangki Penyimpanan		1	19.700	321.256.765	28.421,9	463.488.583,2
3	POMPA1		1	4900	79.906.505	7.069,4	115.283.962,3
4	pompa 2		5	6300	513.684.675	9.089,2	741.111.186,4
5	Heater		1	18000	293.534.100	25.969,2	423.492.106,5
6	Flash Drum		1	6900	112.521.405	9.954,9	162.338.640,8
7	Heater 2		1	18900	308210805	28.000,0	456.608.600,0
8	Coller		2	17800	580.545.220	25.680,7	837.573.277,3
9	Reboiler		1	17100	278.857.395	24.670,8	402.317.501,2

10	Reaktor Fix bed multitubular		2	255100	8.320.060.990	368.041,9	12.003.648.485,4
11	Kompresor		1	1300	21.199.685	1.000,0	16.307.450,0
12	Kondensor		1	25100	409.316.995	36.212,7	590.536.215,2
13	Destilasi		1	19800	322.887.510	28.566,2	465.841.317,2
TOTAL				433.800	12.308.863	625.715,5	17.756.099.463,1

Total Harga Alat Proses

- Harga (A) = US$ 496.157 = Rp 7.602.887.865
- Biaya T dan Asuransi (12,5%A) = US$ 62.020 = Rp 950.360.983
- Pajak bea cukai (10%) = US$ 49.616 = Rp 760.288.787+
- Total = US$ 607.792 = Rp 9.313.537.635

Tabel D.3 Perkiraan Harga Peralatan Utilitas

No	Alat	Jumlah	Harga/Unit US$	2021		2027	
				US$	Rp	US$	Rp
1	Centrifugal Pump (3,5 in)	7	7.500	52.500	804.486.375,00	71.583	1.096.899.926,34
2	Centrifugal Pump (1 1/4 in)	1	3.200	3.200	49.035.360,00	4.363	66.858.662,18
3	Centrifugal Pump (2/3 in)	2	3.200	6.400	98.070.720,00	8.726	133.717.324,35
4	Centrifugal Pump (1/4 in)	2	3.200	6.400	99.070.720,00	8.726	133.717.324,35
5	Bak Sedimentasi	1	287.000	287.000	4.397.858.850,00	391.318	5.996.386.263,97
6	Tangki Pelarutan Tawas	1	14.500	14.500	222.191.475,00	19.770	302.953.312,99
7	Tangki Pelarutan Soda Ash	1	14.500	14.500	222.191.475,00	19.770	302.953.312,99
8	Tangki Pelarutan Kaporit	1	14.500	14.500	222.191.475,00	19.770	302.953.312,99
9	Clarifier	1	435.000	435.000	6.665.744.250,00	593.113	9.088.599.389,65
10	Bak Penampung	1	35.700	35.700	547.050.735,00	48.676	745.891.949,91
11	Sand filter	3	47.500	142.500	2.183.605.875,00	194.296	2.977.299.800,06
12	Menara Tangki	1	319.500	319.500	4.895.874.225,00	435.631	6.675.419.551,71

13	Kation Anion Exchange	2	36.650	73.300	1.123.216.215,00	99.943	1.531.481.230,49
14	Feed Water Tank	1	35.600	35.600	545.518.380,00	48.540	743.802.616,72
15	Dearator	1	397.000	397.000	6.083.449.350,00	541.301	8.294.652.776,30
16	Boiler	1	255.200	255.200	3.910.569.960,00	347.960	5.331.978.308,59
17	Generator	1	492.700	492.700	7.59.913.085,00	671.786	10.294.144.642,02
18	Tangki Solar	3	25.900	77.700	1.190.639.835,00	105.942	1.623.411.890,98
19	Cooling tower	1	407.800	407.800	6.248.943.690,00	556.027	8.520.300.761,14
Total		**32**		**3.071.000**	**47.058.622.050**	**4.187.243**	**64.163.422.358**

Total Harga Alat Utilitas

- Harga (B) = US$ 4.187.243,00 = Rp 64.163.422.358
- Biaya T dan A (12,5%B) = US$ 523.405,00 = Rp 78.020.427.795
- Pajak bea cukai (10%) = US$ 418.724,27 = Rp 6.416.342.236+
- Total = US$ 5.129.372,00 = Rp 78.600.192.388

Total Harga Peralatan

- Total Harga = Harga Peralatan Proses + Harga Peralatan Utilitas
  = US$ 607.792 + 5.129.372
  = US$ 5.737.165
  = Rp 87.913.730.023

1.2 Perhitungan Komponen-komponen Investasi

Perkiraan investasi dihitung dengan menggunakan faktor rasio berdasarkan metode *delivered equipment cost* untuk *fluid-fluid processing plant*
seperti yang dapat dilihat pada Tabel D.4

Tabel D.4 Perhitungan *Capital Investment* Pabrik Propilen glikol

Komponen	%	Biaya (US$)	Biaya (Rp)
Direct Cost			
Biaya peralatan (*Equipment*)	**100**	4.958.834	80.865.942.203
pemasangan alat (47 % *Equipment*)	**47**	1.933.945	31.537.717.459
instrumentasi dan alat kontrol (18 % *Equipment*)	**18**	644.648	10.512.572.486
pemasangan pipa (66 % *Equipment*)	**66**	1.537.239	25.068.442.083
Pemasangan instalasi listrik (11 % *Equipment*)	**11**	495.883	8.086.594.220
Bangunan (18 % *Equipment*)	**18**	1.438.062	23.451.123.239
Pengembangan area (10 % *Equipment*)	**10**	495.883	8.086.594.220

Fasilitas pelayanan (70 % *Equipment*)	**70**	2.727.359	44.476.268.212
Lahan (6 % *Equipment*)	**6**	297.530	4.851.956.532
Total Direct Cost		**14.529.384**	**236.937.210.656**
Indirect Cost			
Engineering and supervision (33 % Direct Cost)	**33**	4.649.403	75.819.907.410
Biaya konstruksi (41 % *Direct Cost*)	**41**	4.939.991	80.558.651623
Total Indirect Cost		**9.589.394**	156.378.559.033
Total DC dan IC		**24.118.778**	**393.315.769.689**
Biaya kontraktor (21 % *Direct Cost and Indirect Cost*)	**21**	4.341.380	70.796.838.544
Biaya tidak terduga (42 % *Direct Cost and Indirect Cost*)	**42**	8.682.760	141.593.677.088
Fixed Capital Investment		**37.142.918**	**605.706.285.322**
Work Capital Investment	**15TCI**	**6.554.633**	**106.889.344.469**
Total Capital Iinvestment		**43.697.551**	**712.595.629.790**

(Sumber : Peters,Tabel 17 Hal 183)

Modal Kerja (*working capital investment*, WCI)

WCI = 15% TCI

TCI = FCI + WCI

TCI = FCI + 0,15 TCI

0,85 TCI = FCI

TCI = US$ 43.697.551 = Rp 712.595.629.790

WCI = US$ 6.554.633 = Rp 106.889.344.469

2. Sumber Investasi

Sumber investasi atau permodalan berasal dari modal sendiri dan modal pinjaman bank dengan persentase 50% : 50%

Modal sendiri = 50% x US$ 43.697.551 = US$ 21.848.776

= Rp 356.297.814.895

Pinjaman bank = 50% x US$43.697.551 = US$ 21.848.776

= Rp 356.297.814.895

Total = US$ 43.697.551 = Rp 712.595.629.790

3. Biaya Produksi Total (*Total Production Cost*)

3.1 Biaya Bahan Baku

Daftar biaya bahan baku prarancangan pabrik Propilen glikol dariPropilen oksida dan Air dapat dilihat pada Tabel D.5

Tabel D.5 Biaya bahan baku dan Bahan Kimia Utilitas

No.	Bahan	Kebutuhan		Harga Pasaran	Harga tahun sebelumnya 2025 (Ny)		Harga tahun dicari 2030 (Nx)	
		kg/jam	kg/tahun	USD	USD	Rp	USD	Rp
1	C3H8O3	11.721,68	84.396.121,64	0,536	45.236.321	737.689.046.107	49.681.053	810.171.283.777
2	H2	1.198,67	8.630.420,79	0,469	4.047.667	66.007.132.908	4.445.374	72.492.717.479
3	Tawas	0,4813	3.465,30	0,225	780	12.714.792	856	13.964.095
4	Soda Ash	0,3269	2.353,4018	0,21	494,2144	8.059.376,1922	543	8.851.257
5	Kaporit	0,0250	179,8688222	0,2	35,97376443	586640,3648	40	644.281
6	Solar	6,03499	43451,90211	0,59	25636,62224	418067935,4	28155,56943	459145540,7
7	Asam Sulfat	0,7209	5190,48	0,287	1489,66776	24292682,51	1636,036278	26679579,81
TOTAL					49.312.425		54.157.657	

3.2 Gaji Karyawan

Sistem gaji karyawan di pabrik Propilen glikol dari gliserol dan hidrogen berdasarkan gaji upah minimum Kota (UMK) provinsi Riau barat tahun 2030 dengan nilai UMR sebesar Rp 4.986.297 Daftar gaji karyawan pra rancangan pabrik Propilen glikol dari gliserol dan hidrogen dapat dilihat pada Tabel D.6

Tabel D.6 Daftar Gaji Karyawan

Jabatan	**Jumlah**	**Sistem Gaji**	**Total/Bulan (US$)**	**Total/tahun (US$)**	**Total/tahun (Rp)**
Dewan Komisaris	1	5 x UMR	1.529	18.346	299.177.809
Direktur Utama	1	4 x UMR	1.223	14.677	239.342.247
Sekretaris	2	3 x UMR	1.835	22.015	359.013.371
Staf Ahli	4	3 x UMR	3.669	44.031	718.026.741
Manager Teknik dan Produksi	1	3 x UMR	917	11.008	179.506.685
Manager Administrasi dan Umum	1	3 x UMR	917	11.008	179.506.685
Kepala Bagian					
-Kabag Produksi	1	2 x UMR	612	7.338	119.671.124
-Kabag Teknik	1	2 x UMR	612	7.338	119.671.124
-Kabag Utilitas	1	2 x UMR	612	7.338	119.671.124
-Kabag Umum dan SDM	1	2 x UMR	612	7.338	119.671.124

-Kabag Keuangan dan Pemasaran	1	2 x UMR	612	7.338	119.671.124
-Kabag Litbang	1	2 x UMR	612	7.338	119.671.124
Kepala Seksi					
-Kasi Pemeliharaan dan Bengkel	1	1,5 x UMR	459	5.504	89.753.343
-Kasi Produksi	4	1,5 x UMR	1.835	22.015	359.013.371
-Kasi Utilitas	1	1,5 x UMR	459	5.504	89.753.343
-Kasi Laboratorium dan Pengendalian Mutu	1	1,5 x UMR	459	5.504	89.753.343
-Kasi Pemasaran	1	1,5 x UMR	459	5.504	89.753.343
-Kasi Keuangan	1	1,5 x UMR	459	5.504	89.753.343
-Kasi K3	1	1,5 x UMR	459	5.504	89.753.343
-Kas Humas	1	1,5 x UMR	459	5.504	89.753.343
-Kasi Keamanan	1	1,5 x UMR	459	5.504	89.753.343
-Kasi Administrasi dan Oprasional	1	1,5 x UMR	459	5.504	89.753.343
Karyawan Keuangan	3	1,2 x UMR	1.101	13.209	215.408.022
Karyawan Pemasaran	3	1,2 x UMR	1.101	13.209	215.408.022
Karyawan K3	3	1,2 x UMR	1.101	13.209	215.408.022
Karyawan Humas	4	1,2 x UMR	1.468	17.612	287.210.696
Karyawan Pembelian dan Pemasaran	3	1,2 x UMR	1.101	13.209	215.408.022
Karyawan Teknik	8	1,2 x UMR	2.935	35.224	574.421.393
Karyawan Produksi	36	1,2 x UMR	13.209	158.510	2.584.896.268
Karyawan Utilitas	28	1,2 x UMR	10.274	123.286	2.010.474.875
Karyawan Laboratorium	8	1,2 x UMR	2.935	35.224	574.421.393
Karyawan Gudang	8	1,2 x UMR	2.935	35.224	574.421.393

Dokter	2	1,5 x UMR	917	11.008	179.506.685
Perawat	8	1,2 x UMR	2935,37306	35224,47672	574421392,8
Petugas Keamanan	8	1 x UMR	2.446	29.354	478.684.494
Petugas Kebersihan	5	1 x UMR	1.529	18.346	299.177.809
Sopir	10	1 x UMR	3.058	36.692	598.355.618
Pesuruh	10	1 x UMR	3.058	36.692	598.355.618
TOTAL	176		71.825	IDR 861.899	IDR 14.055.373.456,11

Maka, gaji total karyawan selama 1 tahun = Rp. 14.055.373.456

= US$ 861.899

3.3 Perhitungan Komponen Biaya Produksi Total

Perhitungan komponen biaya produksi total dapat dilihat pada Tabel D.7

Tabel D.7 Perhitungan komponen biaya produksi (sebelum Perhitungan TPC)

Parameter	Fixed Cost (US$)	Variable Cost (US$)
Direct Production Cost (DPC)		
Raw Materials (10-50% TPC)		54.157.657
Operating Labor (10-20% TPC)	861.899	
Direct Supervisory (10-20% OL)		129.285
Utilities (10-20% TPC)		0,10 TPC
Maintenance and Repairs (2-10% FCI)	2.228.575	
Operating Supplies (0,5-1% FCI)	278.572	
Laboratory Charges (10-20% OL)		129.285
Patents and Royalties (0-6% TPC)	0,01 TPC	
Total DPC	**3.369.046**	**54.416.227**
Fixed Charge		
Depreciation(10% FCI)	3.714.292	
Local Taxes (1-4% FCI)	928.573	

Insurance (0,4-1% FCI)	260.000	
Total FC	**4.902.865**	
Plant Overhead Cost		**0,1 TPC**
General Expenses		
Administrative cost (2-6% TPC)	0,02 TPC	
Distribution Cost (2-20% TPC)	0,02 TPC	
Research and Development (5% TPC)	0,05 TPC	
Financing (0-10% TCI)	2.184.878	
Total General Expenses	**2.184.878**	
Total Production Cost	**10.456.789**	**54.416.227**

Total Production Cost = *Manufacturing Cost* + *General Expenses*

= (*Fixed Cost* + *Variable Cost*)

TPC = 10.456.789 + 54.416.227 + 0,3 TPC

TPC = 64.873.016 + 0,3 TPC

0,7 TPC = US$ 64.873.016

= US$ 92.675.736

= Rp 1.511.304.938.446

Sehingga setelah dilakukan perhitungan TPC, Biaya Produksi total dapat dilihat pada Tabel D.8

Tabel D.8 Perhitungan komponen biaya produksi (setelah Perhitungan TPC)

Parameter	Fixed Cost (US$)	Variable Cost (US$)
Direct Production Cost (DPC)		
Raw Materials (10-50% TPC)		54.157.657
Operating Labor (10-20% TPC)	861.899	
Direct Supervisory (10-20% OL)		129.285
Utilities (10-20% TPC)		6.487.301,55
Maintenance and Repairs (2-10% FCI)	2.228.575	

Operating Supplies (0,5-1% FCI)	278.572	
Laboratory Charges (10-20% OL)		129.285
Patents and Royalties (0-6% TPC)	648.730,16	
Total DPC	**4.017.776**	**60.903.528**
Fixed Charge		
Depreciation(10% FCI)	3.714.292	
Local Taxes (1-4% FCI)	928.573	
Insurance (0,4-1% FCI)	260.000	
Total FC	**4.902.865**	
Plant Overhead Cost		**6.487.302**
General Expenses		
Administrative cost (2-6% TPC)	2.594.920,62	
Distribution Cost (2-20% TPC)		7.136.031,71
Research and Development (5% TPC)	3.243.650,78	
Financing (0-10% TCI)	3.243.650,78	
Total General Expenses	**9.082.222**	
Total Production Cost	**18.002.863**	**74.526.862**

Sehingga diperoleh :

Direct Production Cost = US$ 64.921.304 = Rp. 1.058.700.926.021

Fixed Charge = US$ 4.902.865 = Rp. 79.953.229.662

Plant overhead cost = US$ 6.487.302 = Rp. 105.791.345.691

General Expense = US$ 9.082.222 = Rp. 148.107.883.968

Fixed Cost = US$ 18.002.863 = Rp. 293.580.795.915

Variabel cost = US$ 74.526.862 = Rp. 1.215.343.069.688

4. Harga Penjualan Produk *(Total Sales)*

Produksi = 8838 kg/jam

= 70000000 kg/tahun

Harga jual Pabrik = US$ 1,8/kg

= Rp 29.000

Total Penjualan (TS) = US$ 124.482.600

= Rp 2.030.000.000.000

Harga penjualan produk pertahun dapat dilihat pada Tabel D.8

Tabel D.8 Perhitungan komponen Penjualan produk

Komponen	Produksi (kg/jam)	Produksi (kg/tahun)	Harga/kg ($)	Total Harga ($)	Total Harga (Rp)
Propylene Glycol	8838	70.000.000	1,78	124.482.600	2.030.000.000.000
Total	8838	70.000.000	**1,78**	124.482.600	2.030.000.000.000

Berdasarkan Tabel D.8 diperoleh harga penjualan (TS) adalah sebesar US$ 124.482.600 atau Rp 2.030.000.000.000

5. Analisa Kelayakan Investasi

5.1 Laba

Total Capital Investment (TCI) = US$ 43.697.551 = Rp 712.597.814.668

Depresiasi (10% FCI) = US$ 3.714.292 = Rp 60.570.814.247

Total Penjualan (TS) = US$ 124.482.600 = Rp 2.030.000.000.000

Total Production Cost (TPC) = US$ 92.675.736 = Rp 1.511.309.572.233

Laba Sebelum Pajak (Laba Kotor) = Total Penjualan (TS) – Biaya Produksi (TPC)

= US$ 124.482.600 - US$ 92.675.736

= US$ 31.806.864

= Rp 518.690.427.767

Pajak 25,0% (Dirjen Pajak)

Laba Bersih = Laba Kotor – (Laba kotor x Pajak)

= US$ 31.806.864 - (US$ 31.806.864 x 25,0%)

= US$ 23.855.148

= Rp 389.017.820.825

5.2 Laju Pengembalian Modal (*Rate Of Return*)

$$ROR = \frac{laba\ bersih}{TCI}\ x\ 100\ \% \qquad = \frac{\text{US\$ 43.697.551}}{\text{US\$ 23.855.148}}\ x\ 100\%$$

$$= 54{,}59\ \%$$

5.3 Waktu Pengembalian Modal (*Pay Out Time*)

Masa *start up* : 2 tahun

Umur pabrik :10 tahun

Kapasitas produk pabrik selama beroperasi :

Tahun I	: 70%	= 70% x TS	= US$ 87.137.820
Tahun II	: 90%	= 90% x TS	= US$ 112.034.340
Tahun III dan seterusnya	: 100%	= 100% x TS	= US$ 124.482.600
Fixed Cost	= US$ 18.002.863		
Variable Cost	= US$ 74.526.862		

Keuntungan masing-masing kapasitas setelah ditambah depresiasi

1. Kapasitas 70%
   = total penjualan 70% - [{*fixed cost* + (*variable cost* x 70%)}+ depresiasi]
   = US$ 13.251.861
   = Rp 216.104.732.474
2. Kapasitas 90%
   = total penjualan 90% - [{*fixed cost* + (*variable cost* x 90%)}+ depresiasi]
   = US$ 23.243.009
   = Rp 379.035.373.267
3. Kapasitas 100%
   = total penjualan 100%-[{*fixed cost* + (*variable cost* x 100%)}+ depresiasi]
   = US$ 28.238.583
   = Rp 460.500.693.664

Jumlah keuntungan selama *start up* = US$ 13.251.861 + US$ 23.243.009
= US$ 36.494.871
= Rp 595.140.105.741

Waktu Pengembalian Modal (POT) sebelum pajak

$$POT = Massa\ start\ up + \frac{TCI - jumlah\ keuntungan\ selama\ start\ up}{keuntungan\ saat\ kapasitas\ 100\%}$$

$$= 2 + \frac{\text{US\$ } 43.697.551 \; - \; \text{US\$ } 36.494.871}{\text{US\$ } 28.238.583}$$

$$= 2{,}26 \; tahun$$

Maka diperoleh POT < 5 tahun, maka pabrik masih layak

5.4 Titik Impas (*Break Even Point*)

Total Sales = US$ 348.000.000

Fixed Cost = US$ 53.403.458

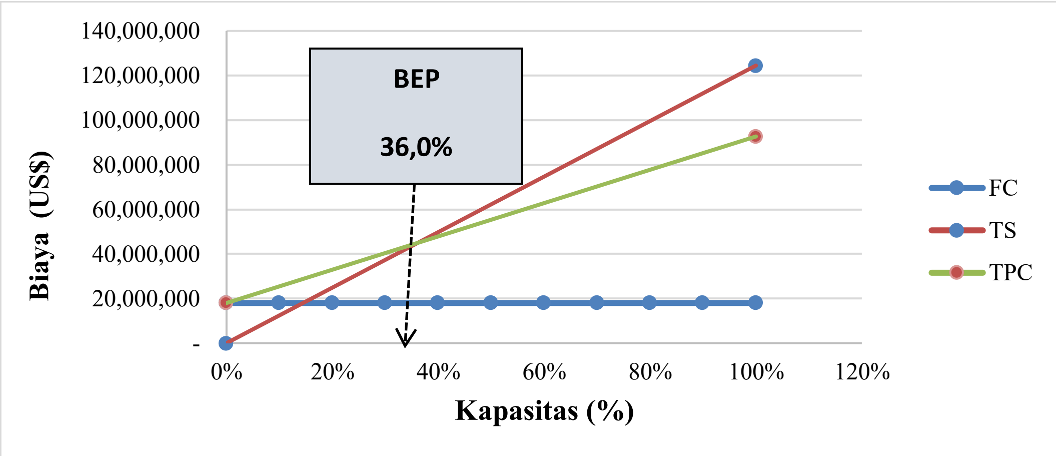
Variabel Cost = US$ 245.942.532

Maka,

$$Break\ Even\ Point\ (BEP) = \frac{Fixed\ Cost}{Total\ Sales - Variabel\ Cost} x\ 100\%$$

$$Break\ Even\ Point\ (BEP) = \frac{18.002.863}{124.482.600 - 74.526.862} x\ 100\%$$

$$Break\ Even\ Point\ (BEP) = 36{,}0\ \%$$



Gambar D.2 Kurva BEP