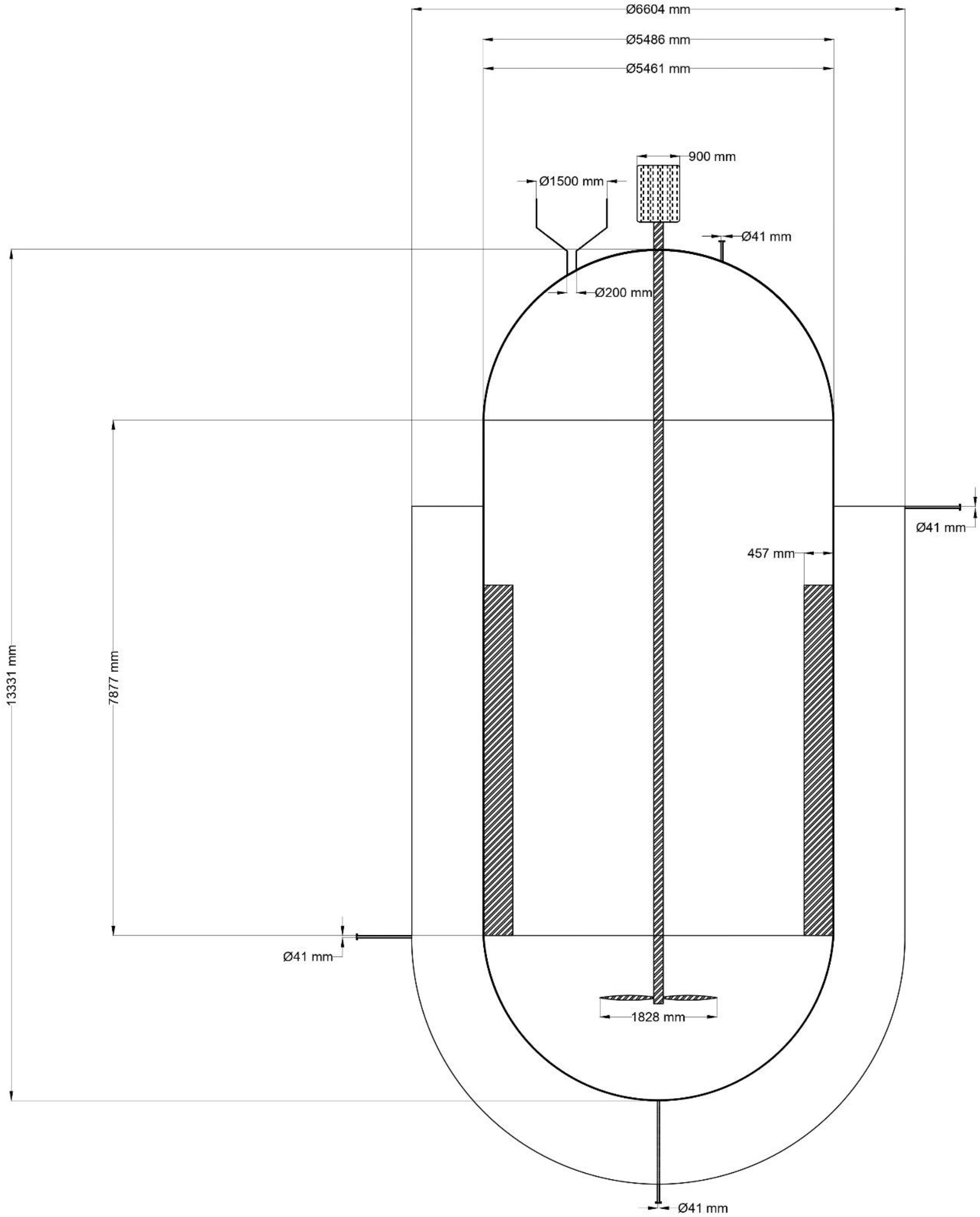
LAMPIRAN 1
REAKTOR (R-01/R-02/R-03/R04)

Nama alat	: Reaktor
Kode alat	: R-01/R-02/R-03/R04
Jenis alat	: Reaktor Alir Tangki Berpengaduk (RATB)
Fungsi	: Tempat mereaksikan bahan baku molases dengan bakteri Lactobcilus dellbruecki
Material	: *SA 135 B Carbon Stell (Apendix D, Brownell Young. Item 1)*
Suhu	: 45 ℃
Tekanan	:1 Atm
Koversi	: 90%
Jumlah Reaktor	: 4 Buah
Jenis Pengaduk	: *marine propeller with 3 blades*

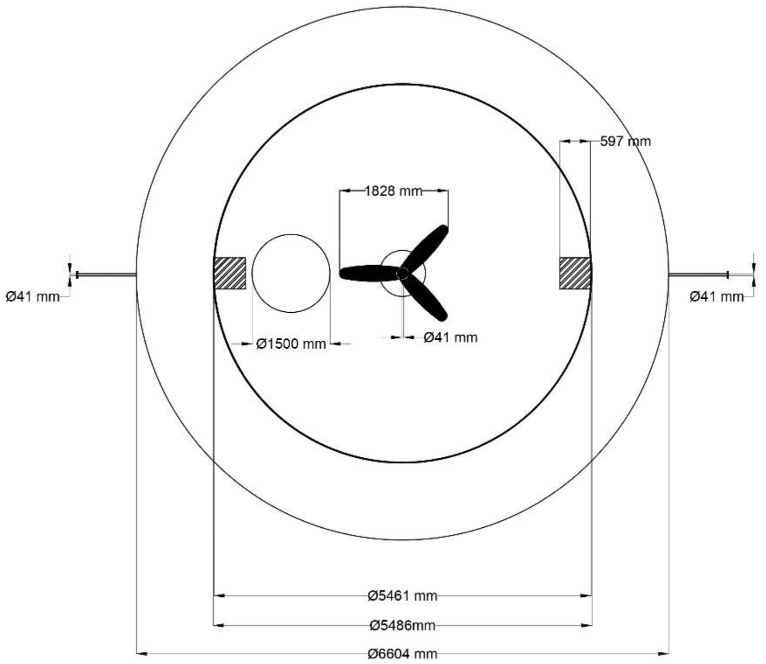
Alasan pemilihan Reaktor RATB :

1. Fase reaksi Padat-Cair dan prosesnya berlasngsung secara kontinue
2. Pada Reaktor Alir Tangki Berpengaduk suhu dan komposisi campuran dalam reaktor selalu seragam. Hal ini memungkinkan melakukan suatu proses isotermal dalam reaktor RATB
3. Pada Reaktor Alir Tangki Berpengaduk karena volume reaktor relatif besar dibandingkan dengan Reaktor Alir Pipa, maka waktu tinggal juga besar, berarti zat pereaksi dapat lebih lama bereaksi di dalam reaktor.

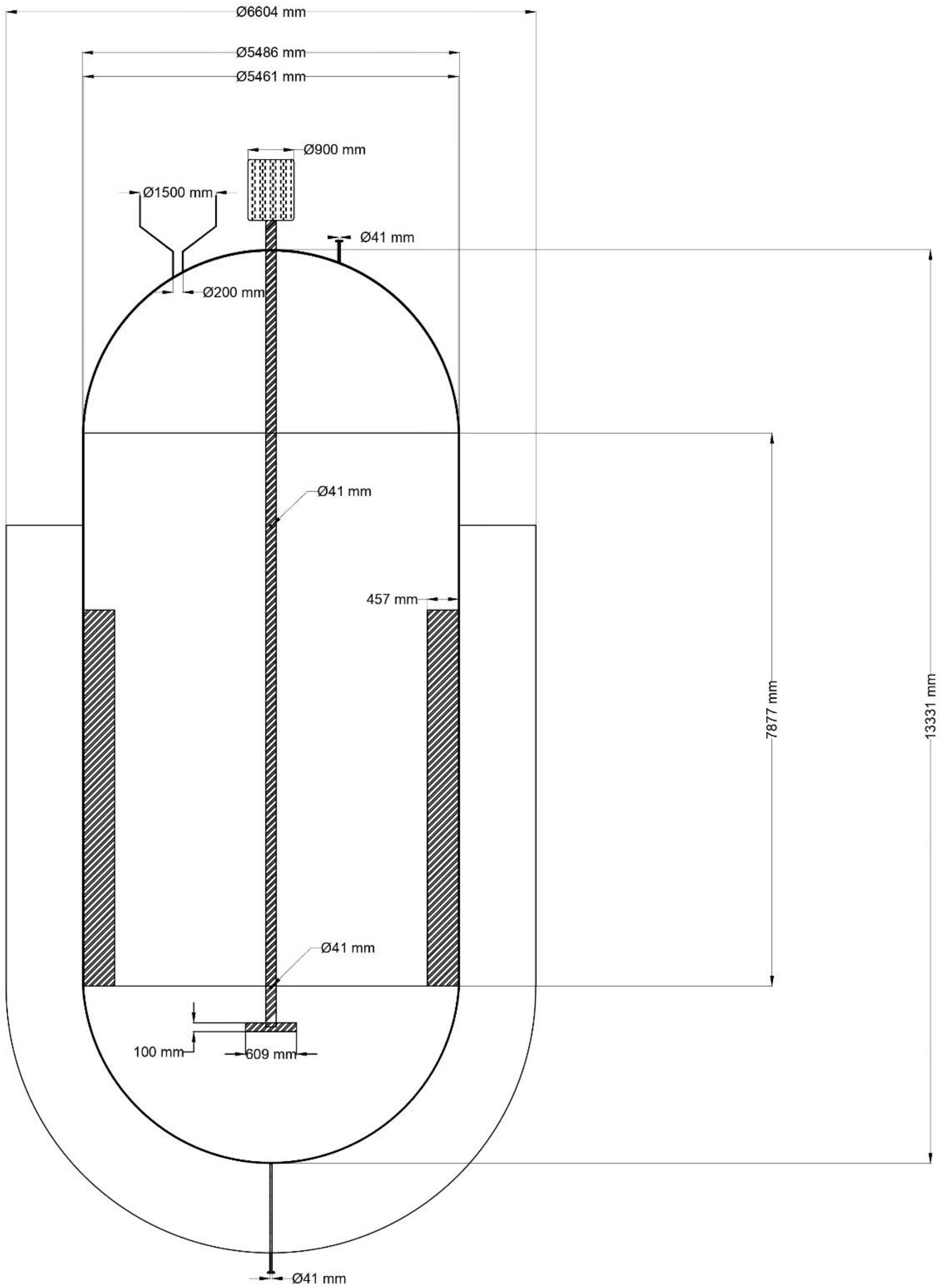
Gambar Tampak Depan Reaktor



Gambar Tampak Atas Reaktor



Gambar Tampak Samping Reaktor



Susunan bahan yang masuk reaktor :

Komponen	Q (kg/Jam)	n (mol)	xi (%)	ρ (kg/m^3)	xi ρ
H_2O	13221,034	734,502	0,963	1000,000	962,695
Sukrosa	2368,900	6,927	0,009	1587,000	14,408
Glukosa	473,780	2,632	0,003	1544,000	5,327
fruktosa	609,146	3,384	0,004	1590,000	7,052
Bakteri	177,017	1,567	0,002	1580,000	3,244
Malt Sprouts	312,382	1,893	0,002	432,112	1,072
$(NH_4)_2HPO_4$	312,382	2,367	0,003	1620,000	5,025
$CaCO_3$	951,000	9,510	0,012	2710,000	33,779
Na_2CO_3	19,408	0,183	0,000	2540,000	0,610
Jumlah	**18445,049**	**762,964**	**1,000**		**1033,212**

Reksi yang terjadi di dalam reactor :

- $C_{12}H_{22}O_{11} + H_2O \rightarrow 4\ C_3H_6O_3$
- $C_6H_{12}O_6 \rightarrow 2C_3H_6O_3$
- $2\ C_3H_6O_3 + CaCO_3 \rightarrow C_6H_{10}CaO_6 + CO_2 + H_2O$

- Data dari perhitungan:

ρ Campuran $= 1366{,}1847$ Kg/m^3

Laju alir massa (F) = 24274,88 Kg/jam = 582597,1752 Kg/24 jam

Konversi (XA) = 90%

Waktu Tinggal (t) = 24 jam (Ghaffar.,2014)

Pada jurnal Ghaffar.,2014 waktu untuk proses fermentasi 24 jam dan waktu yg diperlukan untuk bongkar pasang muat bahan pada reaktor 8 jam sehingga waktu total untuk proses pada reaktor adalah 32 jam. Untuk mempersingkat waktu agar proses produksi tetap berjalan efisien maka digunakan 4 reaktor sehingga masing2 reaktor memiliki flowrate/8 jam.

A. Penentuan Kapasitas Tangki

- **Menghitung volume total tangki**

Bahan Baku Molases = 18445,049 Kg/jam

= 147560,3942 Kg/8 jam

Volume Molases $= \frac{Massa\ Molases}{Densitas\ Molases}$

$= \frac{147560{,}3942\ \frac{\text{Kg}}{\text{8jam}}}{1033{,}21 \frac{\text{Kg}}{\text{m}^3}}$

= 142,8172 m^3

Volume Keamanan $= 20\%\ x\ Volume\ Molases$

$= 0{,}20\ x\ 142{,}8172\ \frac{\text{Kg}}{\text{m3}}$

= 28,5634 m^3

Volume Total Tangki = Volume Molases + Volume Keamanan

= 142, 8172 m^3 + 28, 5634 m^3

= 171,3807 m^3

- **Penentuan Diameter dan Tinggi Tangki**

Volume Tangki Total $= \frac{\pi\ x\ D2\ x\ (1{,}5D)}{4}$

= 1,1775 D^3

$D^3 = \frac{171{,}3807\ m3}{1{,}1775}$

D = $\sqrt[3]{145{,}5462}$ m^3

D = 5,2602 m

D = 207,0936 in

Didapatkan nilai diameter sebesar 206,7690 in, maka :

Tinggi Tangki = 1,5 x D

= 310,6403 in

Tinggi Cairan dalam tangki $= \frac{Volume\ molases}{Volume\ total\ tangki}\ x\ Tinggi\ Tangki$

$= \frac{142{,}8172\ \ m3}{171{,}3807\ m3}\ x\ 310{,}1534\ in$

= 258,86 in

= 6,575 m

- **Menentukan Tekanan pada tangki**

  Data perhitungan:

  - Tekanan Operasi (Po) = 1,0000 atm
  - Percepartan gravitasi (g) = 9,8000 m/s^2
  - Tinggi cairan (hl) = 6,575 m
  - Densitas molasses (r) = 1033,21 kg/m^3

Tekanan Hidrostatis = r x g x hl

= 1033,21 kg/m^3 x 9,8 m/s^2 x 6,575 m

= 66577,2156 kg/m^3.m/s^2.m

= 66577,2156 Pa

= 0,6571 atm

Tekanan Total = Tekanan Operasi + Tekanan hidrostatis

= 1 atm + 0,6571 atm

= 1, 6571 atm

= 24,3589 psi

Tekanan Desain = Tekanan total x Over Desain

= 1,2 x 24,3589 psi

= 29,2306 psi

- **Menentukan Tebal Dinding Tangki**

  Data perhitungan:

  - D = 207,0936 in
  - P = 29,2306 psi
  - f =12750 (Bahan SA 135 B *carbon steel*. Apendix D, Brownell Young)
  - E =80% (Doubel welded, tabel 13,2 Brownell and Young page 254)
  - c = 0,125 in (Peters, ed. 3, hlm 792). Umur alat diperkirakan 10 tahun

Tebal Tangki (ts) $= \frac{P\ x\ D}{2\ x\ f\ x\ E\ -(0,6\ x\ P)} + c$

$= \frac{29{,}2306 \text{ psi x } 207{,}0936 \text{ in}}{2\ x\ 12750 \text{ x } 0{,}8-(0{,}6 \text{ x } 29{,}2306 \text{ psi})} + 0{,}125$

= 0,4220 in

= 0,4375 (ts di standarisasi berdasarkan tabel 5.4. brownell & young hal 87)

- **Menentukan Diameter sesungguhnya**

Diameter luar tangki (OD) = D + (2 x ts)

= 207,0936 in + (2 x 0,43 in)

= 207,9375in

= 216 in (standarisasi berdasarkan tabel 5.7. brownell & young hal 89)

Diameter dalam tangki (ID) = OD – (2 x ts)

= 216 in - (2 x 0,5 in)

= 215 in

- **Menentukan Tebal *Head* Tangki (*torispherical*)**

Tebal *head* tangki $= \frac{0{,}885\ x\ P\ x\ rc}{f\ x\ E-0{,}1\ x\ P} + C$

$= \frac{0{,}885\ x\ 29{,}2306\ \ x\ 216}{12750\ x\ 0{,}8-0{,}1\ x\ 29{,}2306} + 0{,}125$

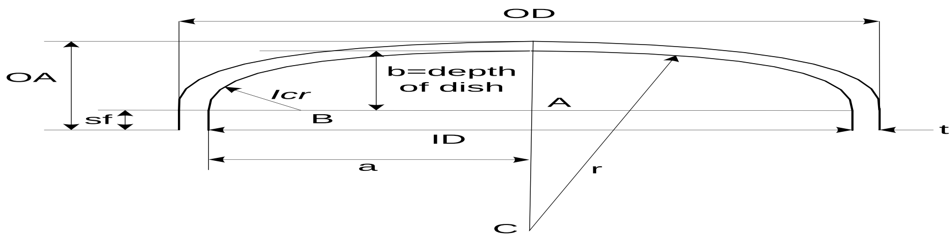
= 0,4257 in

= 0,437 in (standarisasi berdasarkan brownell & young pada tabel 5.4 hal 87

- **Menentukan Tinggi *Head* Tangki (*torispherical*)**

Berdasarkan hal.87 fig.5.8 brownell & young:

- a = ID/2
- AB = a-icr
- BC = r-icr
- AC = $\sqrt{BC^2 - AB^2}$
- b = r - AC



Dimana :

ID = 215 in

ts = 1/2 in

th = 3/4 in

r = 108 in

kr = 6 % (knuckle radius untuk *torispherical* dished *head* adalah 6%) (Brownell & Young hal 88)

icr = r x kr

= 6,48 in

Jadi:

a	=	ID/2	=	107,5 in
AB	=	a-irc	=	101,08 in
BC	=	rc-irc	=	101,52 in
AC	=	√BC2 - AB2	=	9,4148 in
b	=	r - AC	=	98,585 in

Dari tabel 5.6 Brownell hal.88 dengan th 5/8 in didapat sf = 1.5 - 3.5 in

Dipilih nilai sf sebesar: 3 in

Jadi, Tinggi total *Head* :

AO = Sf + b + t *head*

= 102,335 in

= 2,599 m

Didapatkan Tinggi tangki total sebesar:

TT = H +(2 x AO)

= 310,640 in (2 x 102,335 in)

= 515,3107 in

= 13,089 m

- **Menghitung Volume *Head***
  - ➢ Bagian lengkung *torispherical head* (Vh')

    Vh' = 0,000049 x ID^3 (Pers. 5,11 Brownell & Young, hlm.88)

    = 0,000049 x 215^3

    = 487,8303 ft^3

    = 13,813 m^3
  - ➢ Bagian straight flange (Vsf)

    Vsf = 1/4 x π x ID^2 x sf

    = ¼ x 3,14 x 215^2 x 3

    = 108986,49 ft^3

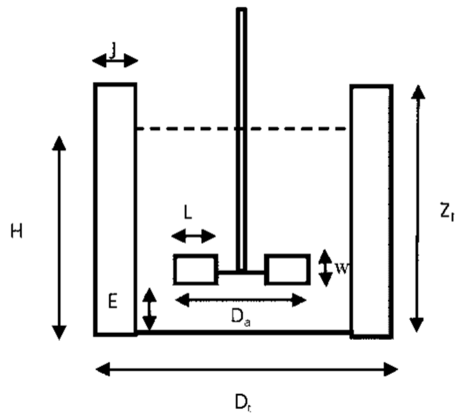
    = 1,7860 m^3
  - ➢ Volume total *head* (Vh)

    Vh = Vh' + Vsf

    = 13,813 m^3 + 1,7860 m^3

    = 15,599 m^3

B. Penentuan desain pengaduk



Nilai viskositas campuran bahan = 0,8607 cp

Dengan viskositas bahan sebesar 0,8607 cp maka di pilih pengadik jenis *marine* propaller with 3 blade sesuai dengan referensi buku brown pada fig.477 dan buku howard rase fig. 8.4.

- **Menghitung Dimensi Pengaduk**

Diameter pengaduk (Da) $= \frac{1}{3} x\ Dt$

$= \frac{1}{3} x\ 5,48640\ m$

$= 1,82\ m$

$= 5,9999\ ft$

Jarak pengaduk dari dasar (E) $= 1,3\ x\ Da$

$= 1,3\ x\ 1,8287\ m$

$= 2,3773\ m$

Lebar pengaduk (w) $= \frac{1}{3} x\ Da$

$= \frac{1}{3} x\ 1,8287\ m$

$= 0,6095\ m$

Lebar buffle (j) $= \frac{1}{12} x\ Dt$

$= \frac{1}{12} x\ 5,48640\ m$

$= 0,4572\ m$

Panjang pengaduk (L) $= \frac{1}{4} x\ Da$

$= \frac{1}{4} x\ 1,8287\ m$

$= 0,4571\ m$

Tinggi tiang pengaduk (T) $= Dt - E$

$= 5,48640\ m - 1,8287\ m$

$= 3,6576\ m$

- **Menghitung Kecepatan Pengadukan**

Water Equipment Liquid Height $= H \; x \; \frac{\rho \; cairan}{\rho \; air}$

$= 6{,}575 \; x \; \frac{1033{,}21}{1000}$

= 6,7935 m

= 22,2887 ft

Kecepatan Putaran Pengaduk $= \sqrt{\frac{WELH}{2 \; x \; Da}} x \; \frac{600}{\pi \; x \; Da}$

$= \sqrt{\frac{22{,}2887}{2 \; x \; 5{,}999}} x \; \frac{600}{3{,}14 \; x \; 5{,}999}$

= 43,4033rpm

Dipilih kecepatan pengaduk standart yang digunakan sebesar (μ) : 50 rpm atau 0,8333333 rps

- **Menghitung Power Konsumsi**

N_{Re} $= \frac{N \; x \; Da2 \; x \; \rho}{\mu}$

$= \frac{0{,}83333 \; x \; 5{,}9999^2 \; x \; 85{,}3141}{0{,}0008}$

= 3198180,6

Dari (fig 477, brown hal 507) pada kurva no 3 di plotkan dengan Re didapatkan hasil

Po = 1,8

Gc = 32,174 ft /s^2

Sehingga dapat di hitung power konsumsi yang digunakan :

Power Konsumsi $= \frac{Po \; x \; N^3 \; x \; Da^5 \; x \; \rho}{Gc}$

$= \frac{1{,}9 \; x \; 0{,}83^3 rps \;\; x \; 5{,}9999^5 ft \; x \; 85{,}3141 \; lb/ft3}{32{,}174 \;\; ft/s2}$

= 16243,47 ft.lbf/det

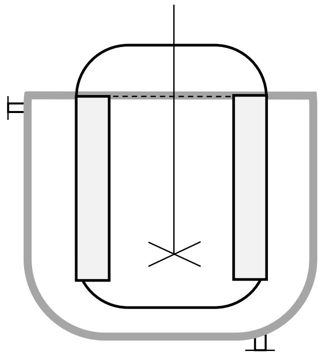
Effisiensi yang di gunakan sebesar 80%, jadi nilai efisiensi sebesar

Effisiensi $= \frac{16243,47 \quad ft.\frac{lbf}{det}}{0,8}$

= ft.lbf/det

= 36,9174 Hp ≈ 50 Hp

C. Penentuan Desain Jaket Pemanas



- **Menghitung Luas Permukaan Transfer Panas**

  Data dari perhitungan:

  – Debit = 2339,4819 Btu/jam.ft2.F

  – Suhu masuk reaktor (T1) = 126,212 °F

  – Suhu keluar reaktor (T2) = 126,68 °F

  – Suhu pemanas masuk (t1) = 293,360 °F

  – Suhu pemanas keluar (t2) = 293 °F

  – Ud yang di pilih = 200 Btu/jam/ft^2.F

  Rumus yang gunakan :

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

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

  $= \frac{(293,360°F-126,68°F)-(293°F-126,212°F)}{ln\frac{(293,360°F-126,68°F)}{(293°F-126,212°F)}}$

  = 166,734 °F

  Luas Tranfer Panas (A) $= \frac{Q}{Ud \ x \ \Delta LMTD}$

  $= \frac{2339,4819 \ Btu/jam.ft2.F}{200 \ Btu/jam.ft2.F x \ 166,734°F}$

  = 0,0701 °F

- Menghitung Diameter Luar Jaket

  Asumsi:

  Jarak antara ID jaket dan OD jaket (DD) = 50 cm

  = 0,5 m

  = 19,685 in

  Maka, Diameter Luar Jaket dapat di hitung dengan :

  OD jaket = OD reaktor + (2 x DD)

  = 216 in + (2 x 19,685 in)

  = 255,3701 in

  = 6,4864 m

  = 21,2806 ft

- Menghitung Tekanan Hidrostatik

  Data dari perhitungan :

  g = 32,15184 ft/s2

  gc = 32,17 ft/s2 (tetapan)

  ODJ = 21,2806 ft

  Maka Tekanan hidrostatik:

  Phidrostatik $= \rho \; x \; \frac{g}{gc} \; x \; ODJ$

  $= 0{,}001 \; x \; \frac{32{,}1518}{32{,}17} \; x \; 21{,}2806$

  = 0,0213 psia

  Tekanan desain yang di pakai dengan over desain sebesar 10% :

  Pdesain = Poperasi + Phidrostatik + (10% x(Poperasi + Phidrostatik))

  = 14,7 psia + 0,0213 psia +(10%x(14,7 psia + 0,0213 psia))

  =16,19 psia

- **Menentukan Tebal Jaket**

  Data dari Brownell and Young 1959 :

  Maximum allowable stress (f) = 12750 psi

  Faktor koreksi (C) = 0,125 in

Jari - Jari dalam tangki (ri) = 108,0 in

Tekanan jaket (P) = 16,19 psi

Efisiensi maksimum (E) = 0,8

Maka Tebal jaket dapat di tentukan dengan:

Tebal Jaket (t) $= \frac{P\ x\ ri}{2\ x\ f\ x\ E - (0{,}6\ x\ P)} + C$

$= \frac{16{,}19 \text{ x } 108{,}0}{2\ x\ 12750\ x\ 0{,}8 - (0{,}6\ x\ 16{,}19)} + 0{,}125$

= 0,2966 in

Jadi nilai tebal jaket yang telah di standarisasi berdasarkan buku brownell & young, tabel 5.4. sebesar : 5/16 in

- **Menentukan Tebal Bottom**

Data dari Brownell and Young 1959 :

Maximum allowable stress (f) = 12750 psi

Faktor koreksi (C) = 0,125 in

Jari - Jari dalam tangki (ri) = 108,0 in

Tekanan jaket (P) = 16,19 psi

Efisiensi maksimum (E) = 0,8

Maka Tebal Bottom dapat ditentukan dengan :

Tebal Bottom (tb) $= \frac{0{,}885\ x\ P\ x\ ri}{f\ x\ E - (0{,}1\ x\ P)} + C$

$= \frac{0{,}885 \text{ x } 16{,}19 \text{ x } 108{,}0}{12750\ x\ 0{,}8 - (0{,}1\ x\ 16{,}19)} + 0{,}125$

= 0,2768 in

Jadi nilai tebal jaket yang telah di standarisasi berdasarkan buku brownell & young, tabel 5.4. sebesar : 5/16 in

- **Menentukan Jarak antara ID jaket dan OD jaket (DD) sesungguhnya**

Nilai OD jaket distandarisasi berdasarkan Buku Brownell & Young, hal.90, didapatkan:

OD jaket = 260 in

icr = $7\frac{4}{5}$in

r = 130 in

Jadi nilai DD dapat di hitung dengan

DD = $\frac{(OD\ jaket - OD\ reaktor)}{2}$

= $\frac{(260\ in - 216\ in)}{2}$

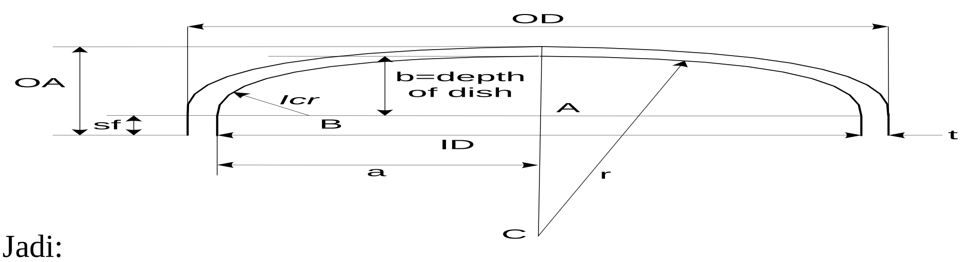
= 22 in

=55,88 cm

- Menentukan Tinggi Bottom dan Tinggi Jaket

Berdasarkan hal.87 fig.5.8 brownell & young:

- a = $\frac{ID}{2}$
- AB = a - icr
- BC = r-icr
- AC = $\sqrt{BC^2} - AB^2$
- b = r - AC



Jadi:

a	=	$\frac{ID}{2}$	=	108	in
AB	=	a-irc	=	100,2	in
BC	=	rc-irc	=	122,2	in
AC	=	$\sqrt{BC^2 - AB^2}$	=	69,948	in
b	=	r - AC	=	60,051	in

Dari tabel 5.6 Brownell hal.88 dengan th 5/8 in didapat sf = 1.5 - 3.5 in

Dipilih nilai sf sebesar: 3 in

Tinggi Bottom dapat ditentukan dengan :

T bottom = tb + sf + b

= 0,3125 in + 3 in + 60,051 in

= 63,3639 in

= 1,6094 m

Jadi, tinggi total jaket adalah:

TT jaket = Tinggi bottom + Tinggi cairan pada *shell*

= 1,6094 m + 6,5649 m

= 8,1744 m

D. Penentuan Desain Hopper Reaktor

Komponen	Q (kg/Jam)	n (mol)	xi (%)	ρ (kg/m^3)	xi ρ
Malt Sprouts	2499,06	15,15	0,08	432,11	33,90
$(NH_4)_2HPO_4$	19992,48	151,46	0,78	1620,00	1270,92
$CaCO_3$	2499,06	24,99	0,13	2710,00	350,80
Na_2CO_3	155,27	1,46	0,01	2540,00	19,27
Jumlah	**8381,95**	**193,06**	**1,00**		**1674,88**

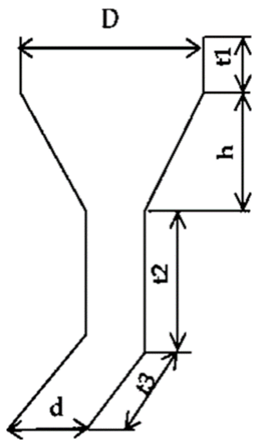
Data Ukuran Hopper untuk kapasitas 900 kg :

- h = 66 cm 0,66 m
- D = 76 cm 0,76 m
- t1 = 30 cm 0,3 m
- t2 = 24 cm 0,24 m

- t3 = 16 cm 0,16 m
- d = 10,16 cm 0,10 m

Dari data di atas didapat ukuran dimensi untuk Hopper Berkapasitas 11.100kg:

- h = 1,8 m
- D = 2,1 m
- t1 = 0,8 m
- t2 = 0,7 m
- t3 = 0,4 m
- d = 0,3 m



- **Menentukan Volume Total Hopper**

V hopper $= \pi \; x \; \frac{h}{12} \; x \; D^2 \; x \; (D + d) \; x \; d^2$

$= 3,14 \; x \; \frac{1,8 \; m}{12} \; x \; 2,1^{\;2} \; x \; (2,1 + 0,3) \; x \; 0,3^2$

= 2,521 m^3

V tabung atas $= \pi \; x \; r^2 \; x \; t1$

$= 3,14 \; x \; 0,75^2 \; x \; 0,8$

= 2,986 m^3

V tabung Ouput = t2 + t3

= 0,7 + 0,4

= 1,12 m^3

Dari hasil perhitugan diatas, nilai volume total hopper dapat di tentukan dengan :

V total = Vhopper + Vtabung atas + Vtabung Output

= 2,521 + 2,986 + 1,12

= 6,63 m^3

Jadi kapasitas hopper desain sebesar:

Kapasitas Hopper = Densitas Campuran x V total

=1674,88 x 6,63

= 11.100 kg

E. Penentuan Pipa Masuk dan Keluar

- **Desain Pipa Feed Masuk Reaktor**

Komponen	kg/ jam	rho, kg/m3	m3/jam	μ,cp
H_2O	8523,411618	1000	8,5234	0,8177
Sukrosa	32,61531236	1587	0,0206	1,51
Glukosa	203,1143367	1544	0,1316	0,778
fruktosa	344,8461	1590	0,2169	0,766
Bakteri	114,1202	1580	0,0722	1,036
Jumlah	**9218,1075**		**8,9646**	

Di Opt $= 260 \; x \; G^{0,52} \; x \; \rho^{-0,37}$

$= 260 \; x \; (\frac{9218,10}{3600})^{0,52} \; x \; (\frac{9218,10}{8,9646})^{-0,37}$

= 32,5709 mm

= 1,2823 in

= 0,0325 m

Dari perhitungan Di Opt dipilih pipa standart berdasarkan data dari buku Unit Operation By G.G. Brown Tabel 23 sebagai berikut:

- NPS = 1,5 in
- Sch. No = 40
- ID = 1,61 in = 0,040894 m
- OD = 1,9 in = 0,048260 m

Data diatas digunakan untuk menghitung kecepatan aliran pada pipa

Kecepatan Aliran $= \frac{Fv}{A}$

$= \frac{(\frac{8,9646}{3600})}{(\frac{3,14}{4}) x \; 0,0408^2}$

=1,8969 m/s

- **Desain Pipa Feed Keluar Reaktor**

Komponen	kg/hr	rho, kg/m3	m3/jam	µ,cp
$C_{12}H_{22}O_{11}$	130,90	1587	0,0825	1,5100
H_2O	8441,27	1000	8,4413	0,8177
bakteri	1141,20	1580	0,7223	1,036
Malt	20,14	260	0,0775	1,0644
$(NH_4)_2HPO_4$	20,14	1620	0,0124	1,032
$C_6H_{10}CaO_6$	2072,67	1490	1,3911	1,0543
$C_6H_{12}O_6$(G)	52,36	1544	0,0339	0,778
$C_6H_{12}O_6$(F)	567,24	1590	0,3568	0,766
Na_2CO_3	256,69	2540	0,1011	1,1501
Jumlah	**12702,61648**		**11,2187**	

Di Opt $= 260 \ x\ G^{0,52}\ x\ \rho^{-0,37}$

$= 260\ x\ (\frac{12702,616}{3600})^{0,52}\ x\ (\frac{12702,616}{11,2187})^{-0,37}$

= 37,1329 mm

= 1,4619 in

= 0,03713 m

Dari perhitungan Di Opt dipilih pipa standart berdasarkan data dari buku Unit Operation By G.G. Brown sebagai berikut:

- NPS = 1,5 in
- Sch. No = 40
- ID = 1,61 in = 0,040894 m
- OD = 1,9 in = 0,048260 m

Data diatas digunakan untuk menghitung kecepatan aliran pada pipa

Kecepatan Aliran $= \frac{Fv}{A}$

$= \frac{(\frac{11,2187}{3600})}{(\frac{3,14}{4})x\ 0,0408^2}$

=2,3738 m/s

- **Desain Pipa Steam Masuk dan Keluar Reaktor**

Komponen	kg/ jam	rho, kg/m3	m3/jam	µ,cp
Steam	673,0534214	1000	0,6731	0,8177
Jumlah	**673,0534**		**0,6731**	

Di Opt $= 260\ x\ G^{0,52}\ x\ \rho^{-0,37}$

$= 260\ x\ (\frac{673,053}{3600})^{0,52}\ x\ (\frac{673,053}{0,6731})^{-0,37}$

= 8,4388 mm

= 0,3322 in

= 0,008438822 m

Dari perhitungan Di Opt dipilih pipa standart berdasarkan data dari buku Unit Operation By G.G. Brown sebagai berikut:

- NPS = 0,375 in
- Sch. No = 40
- ID = 0,493 in = 0,012522 m
- OD = 0,675 in = 0,017145 m

Data diatas digunakan untuk menghitung kecepatan aliran pada pipa

Kecepatan Aliran $= \frac{Fv}{A}$

$= \frac{(\frac{0,6731}{3600})}{(\frac{3,14}{4})x\ 0,0125^2}$

= 1,5189 m/s

F. Tabel Jadwal Proses *Batch* Reaktor

Keterangan:

REAKTOR	PROSES (JAM)															
	4	8	12	16	20	24	28	32	36	40	44	48	52	56	60	64
1																
2																
3																
4																

	Muat
	proses
	Bongkar

LAMPIRAN 2
EVAPORATOR (E-01)

Nama alat : Evaporator

Kode alat : E-01

Jenis alat : *Long Tube Vertical Evaporator*

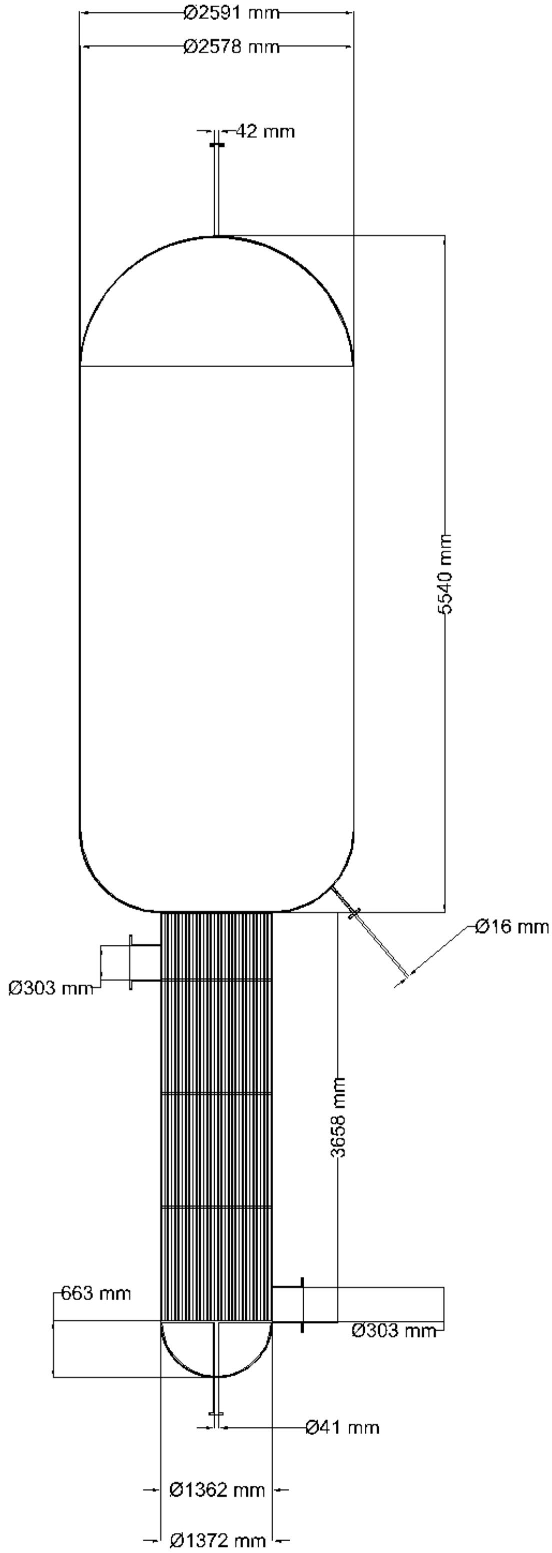
Fungsi : Memekatkan $C_3H_6O_3$ dari konsentrasi 12% menjadi 75%

Material : *Stainess Stell SA 167 grade 10 type 310*

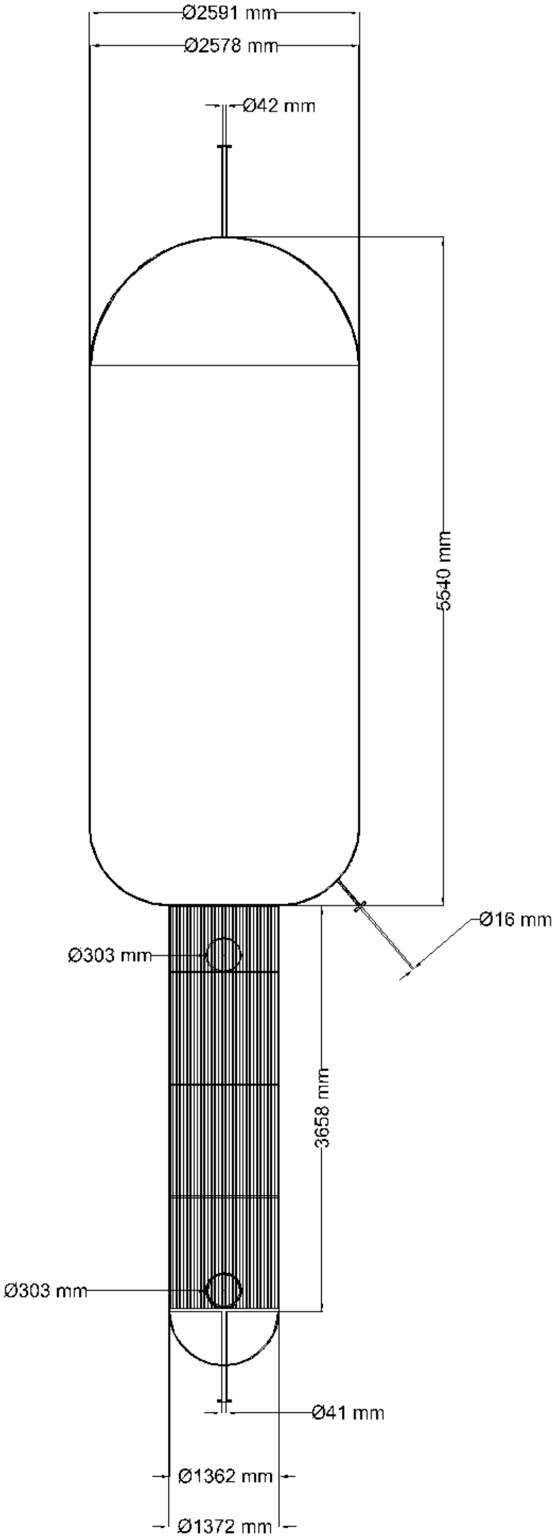
Alasan pemilihan *Long Tube Vertical Evaporator* :

1. *Long Tube Vertical Evaporator* umum digunakan di industri karena relatif murah serta lebih mudah dalam pengoperasian dan pembersihannya (Perry's, 1989:11-109)
2. *Long Tube Vertical Evaporator* memiliki ukuran *tube* lebih panjang (12-24 ft) dibandingkan dengan ukuran *tube* evaporator lainnya sehingga dapat memperbesar serta mempercepat sirkulasi cairan agar proses perpindahan panas lebih besar, sehingga baik digunakan untuk perbedaan temperatur yang rendah atau tinggi (Faputri, A.,2018)
3. *Long Tube Vertical Evaporator* cocok digunakan untuk bahan yang berbusa

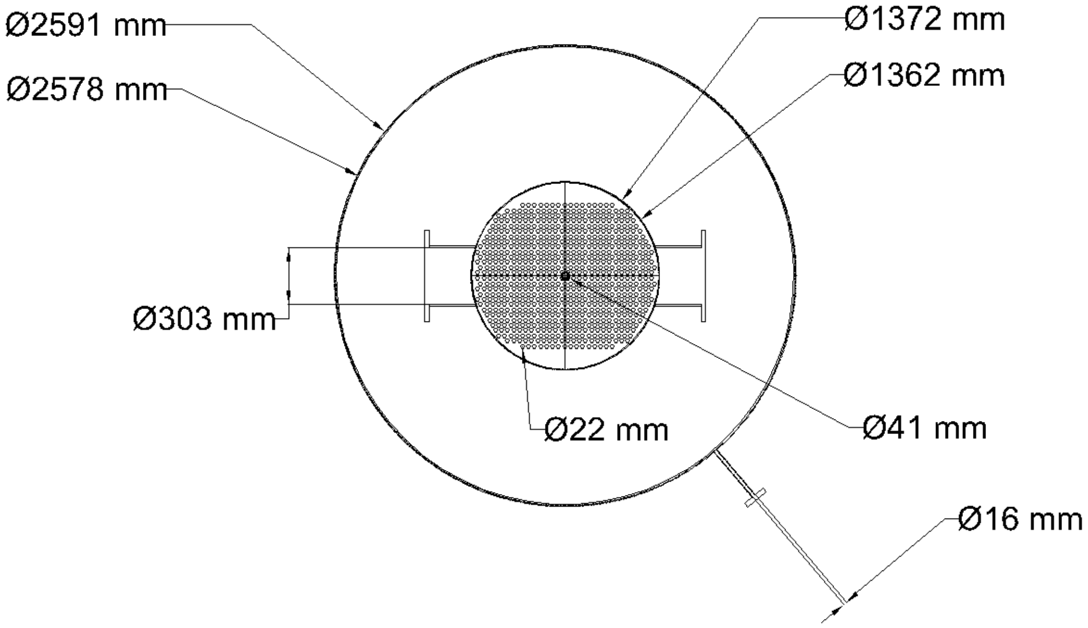
Gambar tampak depan evaporator



Gambar tampak samping evaporator



Gambar tampak atas evaporator



➢ Neraca Massa

- Laju alir larutan masuk :

Komponen	Massa, (kg/jam)
H_2O	12.185,610
$C_{12}H_{22}O_{11}$	220,308
$C_3H_6O_3$	2.111,128
$C_6H_{12}O_6$(G)	44,062
$C_6H_{12}O_6$(F)	182,744
Na_2CO_3	18,050
Total :	**14.761,901**

- H_2O yang diuapkan :

  V = 11.941,898 kg/jam

- Larutan pekat hasil :

Komponen	Massa, (kg/jam)
H_2O	243,712
$C_3H_6O_3$	2.111,128
$C_{12}H_{22}O_{11}$	220,308
$C_6H_{12}O_6$(G)	44,062
$C_6H_{12}O_6$(F)	182,744
Na_2CO_3	18,050
Total :	**2.820,002**

sehingga, fraksi massa solute :

x_f : 0,17452294
x_3 : 0,91357731

➢ Perancangan Suhu Dan Tekanan

Dirancang : T_{Steam} = 145°C, P_{Steam} = 4,10 atm

P_0 =	122,765	InHg
=	4,102	atm
P_1=	119,183	InHg
=	3,983	atm

Pressure drop :

$$\Delta P = \frac{(P_0 - P_1)}{1}$$

ΔP = 3,58 inHg

= 0,1197 atm

Sehingga spesifikasi steam adalah sebagai berikut :

Steam	Tekanan			Suhu didih air murni		
	Simbol	InHg	atm	Simbol	°C	°K
ke efek 1	P0	122,77	4,10	T0	145,0000	418,0000
uap efek 1	P1	119,18	3,983	T'1	143,9393	416,9393

- **Perhitungan Laju Alir Steam**

Data – data :

- Persamaan Boiling Point Rise untuk Larutan $C_3H_6O_3$:

  BPR (°C) = 164,05x^5 – 233,12x^4 + 146,48x^3 – 16,13x^2 + 3,8695x

  Dimana :

  x = fraksi massa solute dalam larutan

- Kapasitas panas solute

Komponen	Massa, kg/jam	mr	kj/kmol	xi	mr campuran	cp campuran, kj/kmol
H_2O	12.185,610	18	75,327	0,874	15,734	65,844
$C_{12}H_{22}O_{11}$	220,308	342	226,62	0,00027	0,092	0,061
$C_3H_6O_3$	2.111,128	90	142,95	0,1238	11,142	17,697
$C_6H_{12}O_6$(G)	44,062	180	207,108	0,00010	0,019	0,022
$C_6H_{12}O_6$(F)	182,744	180	203,869	0,0011	0,210	0,238
Na_2CO_3	18,050	106	112,3	0,0005	0,0562	0,0595
Jumlah	**14.761,901**	**916**			**27,255**	**83,924**

Hasil Cp campuran sebesar 83,9244 kj/kmol atau 91,620 j/kg.

- Persamaan Kapasitas Panas Air Fasa Cair (Cpl)

  Koefisien persamaan :

  A = 15341,104 D = -0,00078

  B = -116,019 E = 5,201E-07

C = 0,451

Cpl (J/kg.^{o}C) = A + BT + CT^2 + DT^3 + ET^4

Cpl = 4296,7 J/kg.^{o}C

- Persamaan Kapasitas Panas Air Fasa Uap (Cpv)

  Koefisien persamaan :

  A = 1859,400 D = 4,472E-07

  B = -0,16048 E = -7,08E-10

  C = 0,000648 D = 2,135E-13

Cpv (J/kg.^{o}C) = A + BT + CT^2 + DT^3 + ET^4 + FT^5

Cpv = 1918,9 J/kg.^{o}C

- Persamaan Panas Laten Penguapan Air (λw)

  Tc = 150 ^{o}C

  = 423 K

  A = 2889425,47 C = -0,011767

  B = 0,017757 D = 0,01432

λw (J/kg) = A (1 - Tr) (B + CT + DT^2)

λw = 2650991,72 J/kg

- T_{ref} = 25 ^{o}C,

  = 298 K

- Trial jumlah steam

Steam	Laju Alir	
	Simbol	kg/jam
Masuk	V0	9270,965397
Keluar	v1	8229,9011

- Perhitungan BPR dan suhu didih larutan

Larutan	Laju Alir		x	BPR	Suhu didih larutan		
	Simbol	kg/jam		°C	Simbol	°C	°K
Masuk	L1	6531,999	0,0753	0,2555	T1	144,19	417,19

- Perhitungan entalpi larutan umpan

  $Hf = (xf \times Cps)+((1\text{-}xf) \times Cpl) \times (T_f\text{-}T_{ref})$

  Hf = 34203,55 J/kg

- Perhitungan entalpi larutan meninggalkan evaporator

  $Hl = (x \times Cps)+((1\text{-}x) \times Cpl) \times (T_{didih\ larutan}\text{-}T_{ref})$

  Hl = 474386,61 J/kg

- Perhitungan entalpi uap meninggalkan evaporator

  $Hv = (Cpl \times (T_{suhu\ out} - 25)) + \lambda w + (Cpv \times (T_{didih\ larutan}\text{-}T_{ref})$

  Hv = 3390779 J/kg

- Jumlah H_2O yang teruapkan menurut perhitungan

  $$V1 = \frac{\text{Q1 – (massa total molases masuk } \times \text{ Hf)}}{Hv - Hl}$$

  V_1 = 8227,74 kg/jam

➢ **Perhitungan Luas Perpindahan Panas**

- Overall Coefficient (Ud) = 700 Btu/jam.ft^2.^{o}F

  = 3976 W/m^2.K

  ***Asumsi :* karena long *tube* vertical evaporator, natural siculation memiliki nilai Ud 200-700 Btu/jam.ft^2.^{o}F

- $\Delta T = 10°C$
- Luas Perpindahan Panas (A)

  $$A = \frac{Q}{\Delta T \times Ud}$$

  A = 171,167 m^2

➢ **Spesifikasi *Tube***

OD = 1 in t = 0,065 in

	=	25,4 mm		=	0,00165 m
	=	0,0254 m	Lt	=	12 ft
BWG	=	16		=	3,657 m
ID	=	0,87 in	tts	=	25 mm
	=	0,022 m			0,025 m

- **Perhitungan Jumlah *Tube***

Panjang *tube* efektif — Lt eff = lt – 2tts = 3,607 m

Luas permukaan 1 buah *tube* — As1 = □.Lt.OD = 0,2877 m^2

Jumlah *tube* yang diperlukan — Nt = A/As1= 594,9

Dipilih : — Nt standar = 664 buah

- **Penentuan A koreksi**

A,kor = Nt.As1

A,kor = 191,04 m^2

(karena nilai A perhitungan lebih kecil dari A koreksi maka jenis evaporator yang digunakan adalah *single effect*)

- **Perhitungan *Tube Bundle* Diameter**

Dipilih :

Tata letak *tube* : *Triangular pitch*

Jumlah lewatan : 1 *shell*, 2 *tube*

Dari Tabel 12.4 C & R ED. 4 (hal 649) diperoleh :

K_1 = 0,249

n_1 = 2,207

sehingga :

$$Db = OD \times \left(\frac{Nt}{K_1}\right)^{\frac{1}{n_1}}$$

Db = 0,906 m

A. Section Bawah

Diketahui :

H = Panjang *tube*

H = 3,657 m

= 143,99 in

H = 3D

D = $\frac{3,657}{3}\ m$

D = 1,219 m

= 47,99 in

Volume total *shell* = ¼ × π × D × H

= 4,267 m^3

Volume total *tube* = 0,93 m^3

Volume *shell* tanpa *tube* = 3,3369 m^3

H steam dalam *shell* = $\frac{Volume\ shell\ tanpa\ tube}{Volume\ total\ shell} \times Tinggi\ shell$

= 2,86 m

- **Menentukan Tebal *Shell***

Dimana :

Tekanan Operasi (Po) = 1,0000 atm

Percepartan gravitasi (g) = 9,8000 m/s^2

tinggi cairan (hl) = 2,8598 m

Densitas molasses (r) = 1025,099 kg/m^3

Ph = Tekanan hidrostatik (atm)

Ph = r x g x hl

= 28728,9654 kg/m3.m/s2.m

= 28728,9654 Pa

= 0,2835 atm

P total = P o + Ph

= 1,2835 atm

= 18,8679 psi

P desain = 22,6415 psi

Data yang digunakan sebagai berikut :

D = 47,9999 in

R = 24,0000

P = 22,6415 psi

f = 18.750 (*SA-167 Stainless steel*)

E = 0,8

c = 0,125 in

t = 0,1613 in

t *shell* (ts) = *0,1875* in = *0,004*8 m

(t standard berdasarkan tabel 5.4. *brownell & young hal 87*)

- **Diameter tangki sesungguhnya**

Diameter luar tangki = ID + (2*t)

= 48,3749 in

OD = 54,0 in = 1,372 m

ID = 53,63 in = 1,362 m

r = 27,00 in = 0,6858 m

- ***Tube* Sheet**

material *tube* = SA-167

f = 18.750 psi

$$T = \frac{FG}{2}\left(\frac{P}{S}\right)^{1/2}$$

T = 0,6776 in

digunakan tebal standar 1 in

- **Tebal bottom Tangki (*torispherical*)**

$$t_h = \frac{0{,}885.P.r_c}{f.E - 0{,}1P} + c$$

Dimana :

th = tebal *head* tangk, in

P = Tekanan desain, psi *(Brownell & Young 1959 Eq. (13.10) hal 256)*

rc = OD = Diameter luar tangki, in *(Brownell & Young hal 88)*

f = tekanan maksimum yang diizinkan pada bahan, psi

E = Efisiensi maksimum, % *(Tabel 13.2 hal 254)*

c = Faktor koreksi, in

tb = 0,1971 in

tb= 0.1971 maka berdasarkan *brownell & young pada tabel 5.4 hal 87* didapat tebal standar sebesar = 0,25 in

Berdasarkan *hal.87 fig.5.8 brownell & young* :

a = ID/2

AB = a-icr

BC = r-icr

AC = √BC2 - AB2

b = r - AC

dimana :

ID = diameter dalam = 54 in

ts = tebal *shell* = 3/16 in

th = tebal tutup = 1/4 in

r = 27,00 in

kr = *knuckle radius* = 0.06 rc = 1,62

icr = 1,62 in *(knuckle radius untuk torispherical dished head adalah 6%) (Brownell & Young hal 88)*

a = ID/2 = 26,8125 in

AB = a-irc = 25,19 in

BC = rc-irc= 25,38 in

AC = √BC2 - AB2 = 3,0793 in

b = r – AC = 23,9207 in

Dari tabel 5.6 *Brownell hal.88* dengan th 3/16 in didapat sf = 1.5 – 2,5 in perancangan digunakan sf = 2,5 in

Jadi tinggi bottom total (AO) = Sf + b + t bottom

= 26,671 in

= 0,6774 m

Jadi tinggi tangki = H + AO

= 170,6704 in

= 4,335 m

= 14,2232 ft

B. Section Atas

H/D = 3

D atas = 2. D bawah

= 2,438399922 m

D = 95,99980493 in

H = 2D

= 4,876799844 m

= 191,9996099 in

Densitas cairan = 1025,098525 kg/m^3

Densitas uap = 0,70450431 kg/m^3

Laju Volumetrik Uap = $\frac{\text{massa uap}}{\text{densitas uap}}$

= 11566,1705 m^3/jam

Kec uap max, u = $0{,}035\sqrt{\frac{\rho l}{\rho v}}$

= 1,335085322 m/s

= 4806,307159 m/jam

vol uap = 11566,1705 m3/jam

volume 20% keamanan = 2313,2341 m^3/jam

volume total tangki (vt) = 13879,4046 m^3/jam

h uap dalam *shell* = 4,06399987 m

= 159,9996749 in

- **Tebal *shell***

Menentukan tekanan pada tangki

Dimana :

Tekanan Operasi (Po) = 1 atm

Percepartan gravitasi (g) = 9,8 m/s^2

tinggi cairan (hl) = 4,06399987 m

Densitas uap (r) = 0,70450431 kg/m^3

Ph = Tekanan hidrostatik (atm)

Ph = r x g x hl

= 28,05843317 kg/m^3.m/s^2.m

= 28,05843317 Pa

= 0,000276915 atm

P total = P o + Ph

= 1,000276915 atm

= 14,70407065 psi

P desain = 17,64488478 psi ~ (over desain =1,2)

$$t = \frac{PD}{2\,fE - (0{,}6xP)} + c$$

D = 95,99980493 in

r = 47,99990246 in

P = 17,64488478 psi

f = 18.750 SA-167 grade 10 (*Stainless steel*)

E = 0,8

c = 0,125 in

t = 0,181503396 in

t *shell* (ts) = ¼ in = 0,00635 m

(t standard berdasarkan tabel 5.4. *brownell & young hal 87*)

- Diameter tangki sesungguhnya

Diameter luar tangki = ID + (2*t)

= 96,36281172 in

OD = 102,0 in = 2,5908 m

ID =101,50 in = 2,5781 m

r = 51 in = 1,2954 m

➢ **Tebal *head* Tangki (*torispherical*)**

$$t_h = \frac{0{,}885.P.r_c}{f.E - 0{,}1P} + c$$ *(Brownell & Young 1959 Eq. (13.10) hal 256)*

Dimana :

th = tebal *head* tangk, in

P = Tekanan desain, psi *(Brownell & Young 1959 Eq. (13.10) hal 256)*

rc = OD = Diameter luar tangki, in *(Brownell & Young hal 88)*

f = tekanan maksimum yang diizinkan pada bahan, psi

E = Efisiensi maksimum, % *(Tabel 13.2 hal 254)*

c = Faktor koreksi, in

th = 0,178099705

th = 0.178099705 maka berdasarkan *brownell & young* pada tabel 5.4 hal 87 didapat tebal standar sebesar = 0,25 in = 0,00635 m

Berdasarkan hal.87 fig.5.8 brownell & young

a = ID/2

AB = a-icr

BC = r-icr

AC = √BC2 - AB2

b = r - AC

dimana :

ID = diameter dalam = 101,50 in

ts = tebal *shell* = 0,25 in

th = tebal tutup = 0,25 in

r = 51 in

kr = *knuckle radius* = 0.06 rc = 3,06 in

icr = 3,06 in *(knuckle radius untuk torispherical* dished *head adalah 6%) (Brownell & Young hal 88)*

a = ID/2 = 50,75 in

AB = a-irc = 47,69 in

BC = rc-irc = 47,94 in

AC = √BC2 - AB2 = 4,88952963 in

b = r – AC = 46,11047037 in

Dari tabel *5.6 Brownell hal.88* dengan th 1/4 in didapat sf = 1.5 - 2.5 in. perancangan digunakan sf = 2 in

Jadi tinggi *head* total (AO) = Sf + b + t *head*

= 48,36 in

= 1,228355947 m

Jadi tinggi tangki = H + AO

= 240,36 in

= 6,11 m

= 20,03098415 ft

C. Perhitungan Pipa Masuk dan Keluar

1. Umpan larutan asam laktat

Diameter pipa optimum untuk *stainless steel* :

$D_{i,\,optimum} = 260G^{0,5}\rho^{-0,35}$ (Coulson Vol.6, 1983, pers 5.15 hal.161)

P = 1,0 atm,

T = 307,50 K = 34,50°C

Komponen	**kg/jam**	**rho, kg/m^3**	**m^3/jam**
H_2O	7868,061909	997	7,8917
$C_{12}H_{22}O_{11}$	2,434786988	1590	0,0015
$C_3H_6O_3$	1114,3818	1209	0,9217
$C_6H_{12}O_6$(G)	0,973914795	1560	0,0006
$C_6H_{12}O_6$(F)	10,55074362	1690	0,0062
Na_2CO_3	4,774506122	2540	0,0019
Jumlah	9001,1777		8,8238

rho = 1020,1075 kg/m^3

G = $\frac{massa\ total}{3600}$

= 2,5003 kg/s

Di,opt = 36,3870 mm

= 1,4326 in

= 0,036386953 m

Dipilih pipa standar :

NPS = 1,5 in

Sch. No = 40 (G.G Grown table 23 hal 123)

ID = 1,61 in

OD = 1,9 in

2. Umpan Steam masuk *shell*

Diameter pipa optimum untuk *stainless steel* :

$D_{i,\,optimum} = 260 G^{0,5} \rho^{-0,35}$ (*Coulson Vol.6, 1983, pers 5.15 hal.161*)

P = 4,1 atm,

T = 418 K = 145°C

massa steam = 9270,965397 kg/jam

densitas steam = 2,255 kg/m^3

laju alir steam = 4111,292859 m^3/jam

rho = 2,255 kg/m^3

G = $\frac{massa\ total}{3600}$

= 2,5752 kg/s

Di,opt = 272,84 mm

= 10,74 in

= 0,2728 m

Dipilih pipa standar :

NPS = 12 in

Sch. No = 40 (*G.G Grown table 23 hal 123*)

ID = 11,938 in

OD = 12,75 in

3. Umpan Produk Larutan Asam Laktat

Diameter pipa optimum untuk *stainless steel* :

$D_{i,\,optimum} = 260 G^{0,5} \rho^{-0,35}$ (*Coulson Vol.6, 1983, pers 5.15 hal.161*)

P = 1,0 atm,

T = 378 K = 105°C

Komponen	kg/hr	rho, kg/m3	m3/jam
H_2O	157,3612382	997	0,1578
$C_3H_6O_3$	1114,3818	1209	0,7009
$C_{12}H_{22}O_{11}$	2,434786988	1590	0,0020
$C_6H_{12}O_6$(G)	0,973914795	1560	0,0006
$C_6H_{12}O_6$(F)	10,55074362	1690	0,0062
Na_2CO_3	4,774506122	2540	0,0019
Jumlah	1290,4770		0,8695

rho = 1484,220006 kg/m^3

G = $\frac{massa\ total}{3600}$

= 0,35846583 kg/s

Di,opt = 10,50286838 mm

= 0,413498755 in

= 0,010502868 m

Dipilih pipa standar :

NPS = 0,5 in

Sch. No = 40 (*G.G Grown table 23 hal 123*)

ID = 0,622 in

OD = 0,84 in

4. Produk uap

Diameter pipa optimum untuk *stainless steel* :

$D_{i,\ optimum} = 260G^{0,5}\rho^{-0,35}$(*Coulson Vol.6, 1983, pers 5.15 hal.161*)

massa uap = 7705,5 kg/jam

densitas steam = 954,65 kg/m^3

laju alir steam = 8,071544545 m^3/jam

rho = 954,65 kg/m^3

G = $\frac{massa\ total}{3600}$

= 2,140416667 kg/s

Di,opt = 29,95105601 mm

= 1,179175433 in

= 0,029951056 m

Dipilih pipa standar :

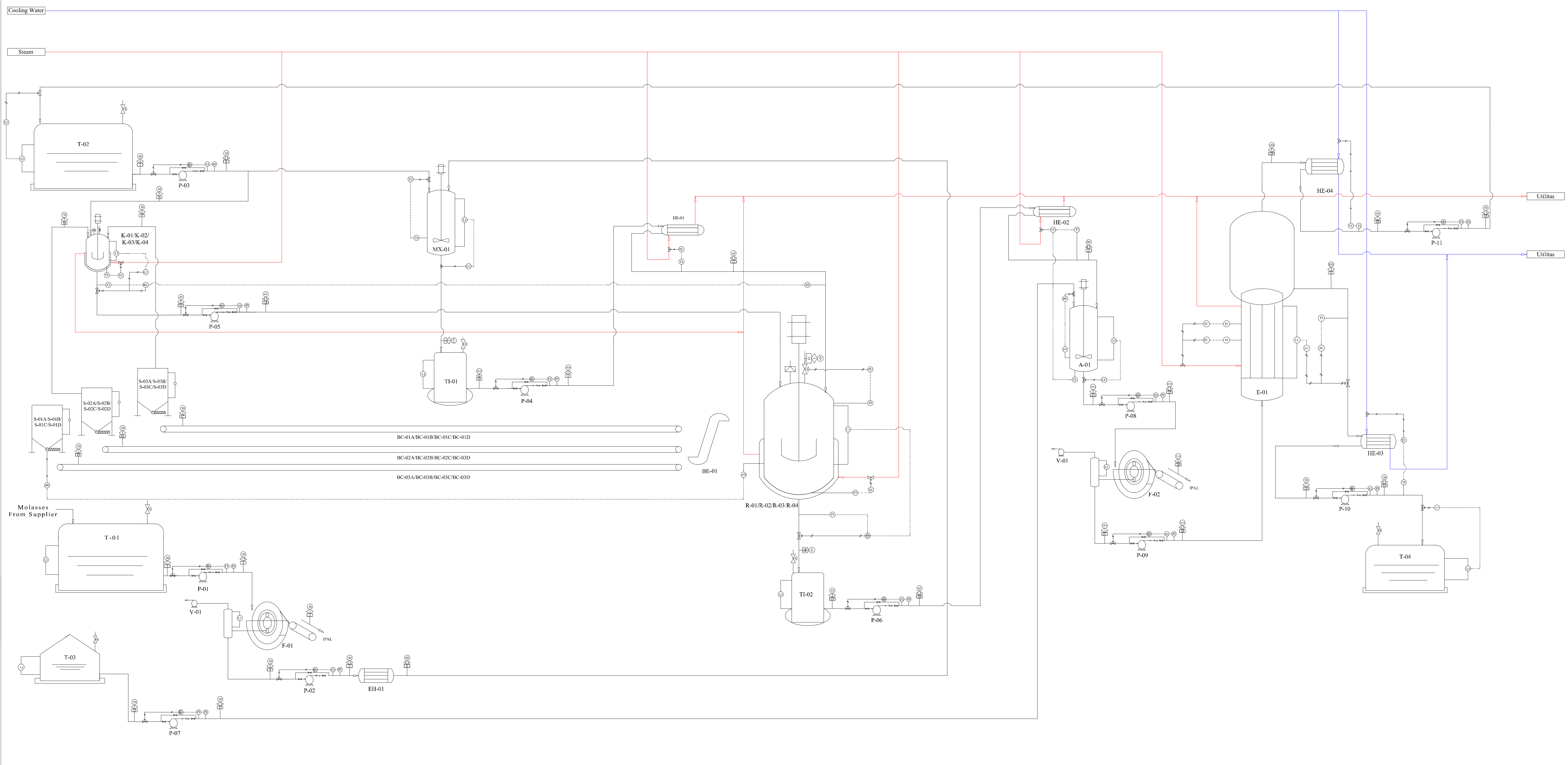
NPS = 1,25 in

Sch. No = 40 (*G.G Grown table 23 hal 123*)

ID = 1,38 in

OD = 1,66 in

PROCESS ENGINEERING FLOW DIAGRAM
PABRIK ASAM LAKTAT DARI MOLASES
KAPASITAS : 10.000 TON ASAM LAKTAT/TAHUN



DRAW BY :			
1	MEGA TRI UMAMININGRUM	TK 19	190606002
2	BAGAS AJI PRATAMA	TK 19	190606005

TEKNIK KIMIA
UNIVERSITAS MUHAMMADIYAH
GRESIK

PROCESS ENGINEERING FLOW DIAGRAM
LACTIC ACID PLANT

SUPRERVISOR:
BENNY ARIF PAMBUDIARTO S.T., M,Eng
Alviani Hesty P. S.T., M.Sc

Date	: 11 . 07 . 2023		
Drawing With	: 11 . 07 . 2023		
Sheet	1/2	Size	A3

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Fluid	Molase	Molase	Cake (Slurry)	Molase	Molase	Molase	Water	Water	Molase 12%	Molase 12%	Molase 12%	Molase 12%	Lactobacillus Delbrueckii	Lactobacillus Delbrueckii	Calcium Carbonat	Malt	Diammonium Phospat	Calcium Lactate Slurry	Calcium Lactate Slurry
Source	Molase Storage	Pump 01	RDFV 01	RDFV 01	Pump 02	Electric Heater	Water Storage	Pump 03	Mixer 01	Intermediate Tank	Pump 04	Heat Exchanger	Culture Tank	Pump 05	Silo 01	Silo 02	Silo 03	Reactor	Intermediate Tank 02
Destination	Pump 01	RDFV 01	IPAL	Pump 02	Electric Heater	Mixer 01	Pump 03	Mixer 01	Intermediate Tank 01	Pump 04	Heat Exchanger	Reactor	Pump 05	Reactor	Reactor	Reactor	Reactor	Intermediate Tank 02	Pump 06
Type	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Batch	Batch	Batch	Continues	Continues	Continues	Continues	Continues	Batch	Continues
Cycle Time	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Total Mass Flowrate (kg/cycle time)	-	-	-	-	-	-	-	-	-	87225,2193	87225,2193	87225,2193	-	-	-	-	-	798592,5974	-
Total Mass Flowrate (kg/hr)	10412,7481	10412,7481	5471,8991	4940,8490	4940,8490	4940,8490	11700,7723	11700,7723	16641,6213	133132,9701	193776,8149	678927,5729	208,2550	208,2550	970,4082	312,3824	312,3824	152459,7707	19057,4713
Individual Component Flowrates:																			
*(kg/h) -continuous; *(kg/cycle time) -batch																			
Sucrose	3644,4618	3644,4618	728,8924	2368,9002	2368,9002	2368,9002	-	-	2368,9002	18951,2015	18951,2015	18951,2015	-	-	-	-	-	1895,1201	236,8900
Glukose	728,8924	728,8924	2082,5496	473,7800	473,7800	473,7800	-	-	13189,7953	105518,3623	105518,3623	105518,3623	-	-	-	-	-	379,0240	47,3780
Fructose	2290,8046	2290,8046	1275,5616	609,1458	609,1458	609,1458	-	-	473,7800	3790,2403	30321,9224	242575,3790	-	-	-	-	-	4873,1661	609,1458
Water	728,8924	728,8924	255,1123	1489,0230	1489,0230	1489,0230	11700,7723	11700,7723	609,1458	4873,1661	38985,3288	311882,6302	177,0167	177,0167	-	-	-	104620,6738	13077,5842
Abu	2082,5496	2082,5496	328,0016	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Non Sugar	937,1473	937,1473	801,7816	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Lactobacillus Delbrueckii	-	-	-	-	-	-	-	-	-	-	-	-	31,2382	31,2382	-	-	-	14161,3374	1770,1672
Malt	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	312,3824	-	249,9060	31,2382
Diammonium Phospat	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	312,3824	249,9060	31,2382
Calcium Carbonat	-	-	-	-	-	-	-	-	-	-	-	-	-	-	951,0000	-	-	-	-
Sodium Carbonat	-	-	-	-	-	-	-	-	-	-	-	-	-	-	19,4082	-	-	155,2653	19,4082
Calcium Laktate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	25875,3720	3234,4215
Sulfuric Acid	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Calcium Sulfate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Lactic Acid	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Carbon dioxide	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

Stream Number	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34	35	36	37	38
Fluid	Calcium Lactate Slurry	Calcium Lactate Slurry	Sulfuric Acid	Sulfuric Acid	Lactic Acid Slurry	Lactic Acid Slurry	Cake (Slurry)	Lactic Acid	Lactic Acid 12%	Water Vapor	Water	Water	Lactic Acid 86%	Lactic Acid 86%	Lactic Acid 86%	Carbon dioxide	Water	Diammonium Phospat	Malt
Source	Pump 06	Heat Exchanger 02	Sulfuric Acid Storage	Pump 07	Acidifier	Pump 08	RDFV 02	RDFV 02	Pump 09	Evaporator	Heat Exchanger 04	Pump 11	Evaporator	Heat Exchanger 03	Pump 10	Reactor	Water Storage	Silo - 03	Silo - 02
Destination	Heat Exchanger 02	Acidifier	Pump 07	Acidifier	Pump 08	RDFV 02	IPAL	Pump 09	Evaporator	Heat Exchanger 04	Pump 11	Water Storage	Heat Exchanger 03	Pump 10	Lactic Acid Storage	Atmosphere	Culture Tank	Culture Tank	Culture Tank
Type	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Continues	Batch	Continues	Continues	Continues
Cycle Time	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Total Mass Flowrate (kg/cycle time)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	5534,32	-	-	-
Total Mass Flowrate (kg/hr)	19057,4713	19057,4713	1261,1277	1261,1277	18285,4592	18285,4592	5556,6984	10005,8964	10005,8964	11941,8981	11941,8981	11941,8981	2820,0025	2820,0025	2820,0025	3347,5200	31,2382	53,1050	53,1050
Individual Component Flowrates:																			
*(kg/h) -continuous; *(kg/cycle time) -batch																			
Sucrose	236,8900	236,8900	-	-	236,8900	236,8900	16,5823	2,434786988	2,434786988	-	-	-	220,3077	220,3077	220,3077	-	-	-	-
Glukose	47,3780	47,3780	-	-	47,3780	47,3780	3,3165	0,973914795	0,973914795	-	-	-	44,0615	44,0615	44,0615	-	-	-	-
Fructose	609,1458	609,1458	-	-	609,1458	609,1458	426,4020	10,55074362	10,55074362	-	-	-	182,7437	182,7437	182,7437	-	-	-	-
Water	13077,5842	13077,5842	25,2226	25,2226	13102,8068	13102,8068	917,1965	7862,759687	7862,759687	11941,89809	11941,89809	11941,89809	243,7122	243,7122	243,7122	-	31,23824421	-	-
Abu	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Non Sugar	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Lactobacillus Delbrueckii	1770,1672	1770,1672	-	-	1770,1672	1770,1672	1770,1672	-	-	-	-	-	-	-	-	-	-	-	-
Malt	31,2382	31,2382	-	-	31,2382	31,2382	31,2382	-	-	-	-	-	-	-	-	-	-	-	53,10501515
Diammonium Phospat	31,2382	31,2382	-	-	31,2382	31,2382	31,2382	-	-	-	-	-	-	-	-	-	-	53,10501515	-
Calcium Carbonat	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sodium Carbonat	19,4082	19,4082	-	-	19,4082	19,4082	1,3586	18,04959184	18,04959184	-	-	-	18,0496	18,0496	18,0496	-	-	-	-
Calcium Laktate	3234,4215	3234,4215	-	-	485,1632	485,1632	485,1632			-	-	-	-	-	-	-	-	-	-
Sulfuric Acid	-	-	1235,9051	1235,9051	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Calcium Sulfate	-	-	-	-	1715,1336	1715,1336	1715,1336			-	-	-	-	-	-	-	-	-	-
Lactic Acid	-	-	-	-	236,8900	236,8900	158,9021	2111,127689	2111,127689	-	-	-	2111,1277	2111,1277	2111,1277	-	-	-	-
Carbon dioxide	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	3347,52	-	-	-

DRAW BY :			
1	MEGA TRI UMAMININGRUM	TK 19	190606002
2	BAGAS AJI PRATAMA	TK 19	190606005

TEKNIK KIMIA
UNIVERSITAS MUHAMMADIYAH
GRESIK

PROCESS ENGINEERING FLOW DIAGRAM
LACTIC ACID PLANT

SUPRERVISOR:
BENNY ARIF PAMBUDIARTO S.T., M,Eng
Alviani Hesty P. S.T., M.Sc

Date	: 11 . 07 . 2023
Drawing With	: 11 . 07 . 2023
Sheet 2/2	Size A3