PRA RANCANGAN PABRIK ETILEN DARI DEHIDRASI ETANOL DENGAN KAPASITAS 85.000 TON/TAHUN

LAPORAN TUGAS AKHIR

Disusun Oleh:

AHMAD DANI (1142020002)

ZIQHRIL HAKIM (1142020033)

Program Studi Teknik Kimia

Institut Teknologi Indonesia

Tangerang Selatan

2024

HALAMAN PERNYATAAN ORISINALITAS

Laporan penulisan ini adalah hasil karya saya sendiri dan semua sumber baik yang dikutip maupun dirujuk telah saya nyatakan dengan benar.

Nama 1	:	**AHMAD DANI**
NRP	:	**1142020002**
Tanda tangan	:	

Nama 2	:	**ZIQHRIL HAKIM**
NRP	:	**1142020033**
Tanda tangan	:	

Tanggal	:	**29 Februari 2024**

HALAMAN PENGESAHAN

Laporan Tugas Akhir diajukan oleh:

Nama : Ahmad Dani (1142020002)
Ziqhril Hakim (1142020033)

Judul : Pra Rancangan Pabrik Etilen dari Dehidrasi Etanol dengan Kapasitas 85.000 Ton/Tahun

Telah berhasil dipertahankan di hadapan Dewan Penguji dan diterima sebagai bagian persyaratan yang diperlukan untuk memperoleh gelar Sarjana Teknik pada Program Studi Teknik Kimia, InstitutTeknologi Indonesia

DEWAN PEMBIMBING

Pembimbing 1 : Dr. Ir. Sidik Marsudi, M.Si., IPM ( )

DEWAN PENGUJI

Penguji 1 : Prof. Dr. Ratnawati, M.Eng.Sc, IPM ( )

Penguji 2 : Dr. Ir. Kudrat Sunandar, S.T., M.T., IPM ( )

Penguji 3 : Dra. Ermizar Tarmizi, M.Si ( )

Ditetapkan di : Tangerang Selatan

Tanggal :

Mengetahui,

Ketua Program Studi Teknik Kimia

PROGRAM STUDI TEKNIK KIMIA * ITI *

(Dr. Ir. Aniek Sri Handayani, M.Si., IPM)

HALAMAN REVISI

HALAMAN PERNYATAAN PERSETUJUAN PUBLIKASI LAPORAN UNTUK KEPENTINGAN AKADEMIS

Sebagai civitas akademik Institut Teknologi Indonesia, saya yang bertandatangan di bawah ini:

Nama	**:**	**1. Ahmad Dani / 1142020002** **2. Ziqhril Hakim / 1142020033**
Program Studi	**:**	**Teknik Kimia**
Jenis Karya	**:**	**Laporan Tugas Akhir**

Demi pengembangan ilmu pengetahuan, menyetujui untuk memberikan kepada Institut Teknologi Indonesia **Hak Bebas Royalti Non-eksklusif (*Non-exclusive Royalty-Free Right*)** atas karya ilmiah saya yang berjudul : **"Pra Rancangan Pabrik *Ethylene* dari Dehidrasi Etanol dengan Kapasitas 85.000 Ton/Tahun"** beserta perangkat yang ada (jika diperlukan). Dengan Hak Bebas Royalti Non-eksklusif ini Institut Teknologi Indonesia berhak menyimpan, mengalih media/formatkan, mengelola dalam bentuk angka dan data (*database*), merawat, dan mempublikasikan laporan saya selama tetap mencantumkan nama saya sebagai penulis / pencipta dan sebagai pemilik Hak Cipta.

Demikian pernyataan ini saya buat dengan sebenarnya

Dibuat di : Tangerang Selatan

Pada tanggal :

Yang menyatakan:

Ahmad Dani | Ziqhril Hakim

ABSTRAK

Nama : **1. Ahmad Dani / 1142020002**
2. Ziqhril Hakim / 1142020033

Nama Pembimbing : **1. Dr. Ir. Sidik Marsudi, M.Si., IPM**

Program Studi : **Teknik Kimia**

Judul : **PRARANCANGAN PABRIK ETILEN DARI DEHIDRASI ETANOL DENGAN KAPASITAS 85.000 TON/TAHUN**

Indonesia merupakan salah satu negara yang saat ini sedang giat melakukan pembangunan di segala bidang diantaranya manufaktur bahan kimia. Industri kimia yang memiliki prospek menjanjikan untuk didirikan diantaranya produksi “Etilen”. Etilen merupakan senyawa hidrokarbon alkena paling sederhana dengan memiliki rumus kimia C_2H_4 dan berat molekul 28.0536 gram/mol. Etilen biasanya diperoleh dari hasil gas bumi dan natural gas, dimana etilen ini merupakan salah satu produk olefin yang paling terkenal karena sebagian besar produksinya digunakan pada industri petrokimia sebagai bahan baku intermediate diantaranya seperti *Ethylene Glikol* (EG), *Ethyl Benzene* (EB), *Ethylene Dichlorida* (EDC), *Ethyl alkohol*, *Vinil Asetat* dan lain sebagainya.

Ada beberapa cara dalam teknologi proses produksi etilen diantaranya dehidrasi etanol, cracking hidrokarbon dan oxidative coupling of methane (OCM). Namun di Indonesia yaitu PT Chandra Asri Pacific Tbk menggunakan cracking hidrokarbon, oleh karena itu kami memilih teknologi proses produksi etilen secara dehidrasi etanol. Pabrik yang akan kami didirikan nantinya dimulai tahun 2023 , dimana tahun konstruksi 2025 dan komersial tahun 2027. Oleh karena itu peluang pasar yang didapatkan dengan menggunakan metode proyeksi tool Forecast Sheet sebesar 2.935 ton/tahun, dikarenakan belum mencukupi TKDN dalam negeri digunakanlah kapasitas ekonomis pabrik yang telah berdiri dengan ketersediaan bahan baku etanol yang nantinya disupply di daerah Lampung. Oleh karena itu, kapasitas pabrik yang akan didirikan sebesar 85.000 ton/tahun di kawasan industri Cilegon (titik ordinat 5°59'25.7"S 105°59'28.1"E) dengan mempertimbangkan faktor-faktor utama dan sekunder.

Teknologi proses etilen dari dehidrasi etanol mengacu ke Patent US 009663414B2 (2017), dimana sehingga didapatkan peluang selektivitas produk sampai 99%. Proses pembentukan etilen dari bahan dasar etanol ini menggunakan reaksi dehidrasi etanol dengan bantuan katalis ZSM-5 pada suhu 470°C dan tekanan 6 atm dalam reaktor fixed bed multitube. Berikut reaksi yang terjadi; $C_2H_5OH(g)$ (katalis) $\rightarrow$ $C_2H_4(g)$ + $H_20(l)$. Dimana dari proses pabrik tersebut dibutuhkan data utilitas diantaranya steam sebesar 38.875 kg/jam, media pendingin 212.971 kg/jam, air 933 kg/jam, listrik 7.271 kW/jam, bahan bakar 211.344 kg/hari.

Manajemen perusahaan kami nantinya berupa perseroan terbatas (PT) dengan berdasarkan RUPS dan dipimpin oleh seorang direksi yang membawahi 146 orang karyawan. Pabrik nantinya beroperasi selama 330 hari per tahun dalam 24 jam secara kontinyu. Dari hasil analisis ekonomi yang telah dilakukan, diperoleh sebagai berikut:

1	Proses pembangunan dan instalasi pabrik dilakukan dalam 1 tahun.	
	Total Modal	= Rp. 11.966.367.511.573
	Modal Sendiri (94,7%)	= Rp. 11.331.367.511.573
	Pinjaman Bank (5,3%)	= Rp. 635.000.000.000
2	Suku bunga per tahun	= 7.95%
3	Jangka waktu pinjaman	= 5 Tahun
4	*Break Even Point* (BEP) tahun pertama	= 65,43 %
5	*Internal rate of Return* (IRR)	= 61,32 %
6	*Minimum Payback Period* (MPP)	= 2 Tahun 6 Bulan

Berdasrkan hasil analisa ekonomi diatas dapat disimpulkan bahwasanya prarancangan pabrik etilen dari dehidrasi etanol dengan kapasitas 85.000 ton/tahun dinyatakan layak didirikan (feasible).

ABSTRACT

Name	:	**1. Ahmad Dani / 1142020002**
		2. Ziqhril Hakim / 1142020033
***Thesis* Advisor**	:	**1. Dr. Ir. Sidik Marsudi, M.Si., IPM**
Department	:	***Chemical Engineering***
Title	:	***PRE-DESIGN OF ETHYLENE PLANT FROM ETHANOL DEHYDRATION WITH CAPACITY OF 85.000 TONS/YEAR***

Indonesia is one of the countries that is currently actively carrying out development in all fields, including chemical manufacturing. Chemical industries that have promising prospects for being established include the production of "ethylene". Ethylene is the simplest alkene hydrocarbon compound with the chemical formula C2H4 and a molecular weight of 28.0536 grams/mol. Ethylene is usually obtained from natural gas and natural gas, where ethylene is one of the most well-known olefin products because most of its production is used in the petrochemical industry as an intermediate raw material such as Ethylene Glycol (EG), Ethyl Benzene (EB), Ethylene Dichloride. (EDC), Ethyl alcohol, Vinyl Acetate and so on.

There are several methods in the ethylene production process technology, including ethanol dehydration, hydrocarbon cracking and oxidative coupling of methane (OCM). However, in Indonesia, PT Chandra Asri Petrochemical uses hydrocarbon cracking, therefore we chose the ethylene production process technology using ethanol dehydration. The factory that we will build will start in 2023, where the construction year will be 2025 and the commercial year will be 2027. Therefore, the market opportunity obtained using the Forecast Sheet tool projection method is 2,935 tons/year, because domestic TKDN is not sufficient, the economic capacity of the factory is used. has been established with the availability of ethanol raw materials which will later be supplied in the Lampung area. Therefore, the factory capacity to be established is 85,000 tons/year in the Cilegon industrial area (ordinate point 5°59'25.7"S 105°59'28.1"E) taking into account primary and secondary factors.

The ethylene process technology from ethanol dehydration refers to US Patent 009663414B2 (2017), which provides an opportunity for product selectivity of up to 99%. The process of forming ethylene from ethanol as a base material uses an ethanol dehydration reaction with the help of a ZSM-5 catalyst at a temperature of 470°C and a pressure of 6 atm in a multitube fixed bed reactor. The following reaction occurs; C2H5OH(g) (catalyst) → C2H4(g) + H2O(l). Where the factory process requires utility data including steam of 38.875 kg/hour, cooling media 212.971 kg/hour, water 933 kg/hour, electricity 7.271 kW/hour, fuel 211.344 kg/day.

Our company management will be a limited liability company (PT) based on a GMS and led by a director who oversees 146 employees. The factory will operate 330 days per year, 24 hours continuously. From the results of the economic analysis that has been carried out, the following are obtained:

1 *The construction and installation of the plant is abot 1 year.*

	Total Capital Investment	= Rp. 11.966.367.511.573
	Self Investment (94,7%)	= Rp. 11.331.367.511.573
	Bank Loan (5,3%)	= Rp. 635.000.000.000
2	*Interest rate per year*	*= 7.90%*
3	*Loan term*	*= 5 Years*
4	*Break even Point (BEP) First year*	= 65,43 %
5	*Internal Rate of Return (IRR)*	= 61,32 %
6	*Minimum Payback Period (MPP)*	*= 2 Year 6 Month*

Based on the results of the economic analysis above, it can be concluded that the preliminary design of an ethylene plant from ethanol dehydration with a capacity of 85,000 tons/year is declared feasible.

KATA PENGANTAR

Puji syukur kehadirat Tuhan Yang Maha Esa, karena berkat rahmat-Nya penulis dapat menyelesaikan laporan tugas akhir yang berjudul "Prarancangan Pabrik Etilen dari Dehidrasi Etanol dengan Kapasitas 85.000 Ton/Tahun"

Dalam penulisan laporan tugas akhir ini penulis menyampaikan ucapan terima kasih dengan tulus kepada;

1. Dr. Ir. Sidik Marsudi, M.Si., IPM selaku pembimbing yang telah membimbing dengan sabar dan penuh perhatian sehingga penulis mampu menyelesaikan tugas akhir dan penyempurnaan laporan ini.
2. Dr. Ir. Aniek Sri Handayani, M.Si., IPM selaku ketua program Teknik Kimia Institut Teknologi Indonesia.
3. Dr. Kudrat Sunandar, S.T., M.T selaku koordinator tugas akhir dan dosen PA program studi Teknik Kimia Institut Teknologi Indonesia.
4. Keluarga besar Zaidun Pauh Limo dan Hasan Basri terkhusus orang tua Ahmad Dani tercinta yang senantiasa mendukung dan mendoakan penulis.
5. Keluarga besar Inyak Jalun terkhusus orang tua Ziqhril Hakim tercinta yang senantiasa mendukung dan mendoakan penulis.
6. Seluruh pegawai PT Indofarma Tbk atas kesediannnya memberikan izin dalam progres penyelesain tugas akhir.
7. Seluruh pegawai PT Landson Pertiwi Agung atas kesediannnya memberikan izin dalam progres penyelesain tugas akhir.
8. Seluruh pegawai Institut Teknologi Indonesia dan teman-teman Teknik Kimia Institut Teknologi Indonesia angkatan 2020 yang telah memberikan dukungan dan semangat dalam penyusunan laporan tugas akhir ini.
9. Terakhir, untuk diri kami sendiri yang mampu berjuang dengan keras dan sejauh ini sehingga mampu menyelesaikan tugas akhir dengan baik.

Penulis menyadari bahwa penyusunan laporan tugas akhir ini masih banyak diperbaiki, untuk itu kritik dan saran yang membangun sanagat diharapkan. Penulis berharap semoga tugas akhir ini dapat bermanfaat bagi pembaca.

Tangerang Selatan, 29 Februari 2024

Penulis

DAFTAR ISI

HALAMAN PERNYATAAN ORISINALITAS i

HALAMAN PENGESAHAN ii

HALAMAN REVISI iii

HALAMAN PERNYATAAN PERSETUJUAN PUBLIKASI LAPORAN UNTUK KEPENTINGAN AKADEMIS iv

ABSTRAK v

ABSTRACT vii

KATA PENGANTAR ix

DAFTAR ISI xi

DAFTAR GAMBAR xix

DAFTAR TABEL xxi

BAB 1 PENDAHULUAN 1

1.1 Latar Belakang 1

1.2 Data Analisis Pasar 2

1.2.1 Data Produksi 2

1.2.2 Data Impor 4

1.2.3 Data Konsumsi 5

1.2.4 Data Ekspor 6

1.3 Penentuan Kapasitas Pabrik 6

1.4 Pemilihan Lokasi Pabrik 9

1.4.1 Faktor Primer Pemilihan Lokasi Pabrik 9

1.4.2 Faktor Sekunder Pemilihan Lokasi Pabrik 11

BAB 2 TEKNOLOGI PROSES 13

2.1 Teknologi yang tersedia 13

2.1.1 Patent US 20130090510A1 (2013) 15
2.1.2 Patent US 009663414B2 (2017) 16
2.2 Seleksi Proses 17
2.2.1 Efisiensi Proses 17
2.2.2 Keamanan Proses 18
2.2.3 Biaya 18
BAB 3 RANCANGAN PROSES 19
3.1 Uraian Proses 19
3.1.1 Deskripsi Proses 19
3.1.2 Diagram Alir Kuantitatif 22
3.1.3 Diagram Alir Kuantitatif Energi 23
3.1.4 Sistem Pengendalian Alat Utama 20
3.1.5 Kebutuhan Utilitas 25
3.2 Tata Letak Alat 35
3.3 Tata Letak Pabrik 38
BAB IV SPESIFIKASI ALAT 41
4.1 Peralatan Proses 41
4.1.1 Tangki Penyimpanan Etanol TP-01 41
4.1.2 Tangki Penyimpanan Air TP-02 41
4.1.3 Tangki Penyimpanan TP-03 42
4.1.4 Mixer M-01 42
4.1.5 Reaktor Fixed Bed Multitube R-01 43
4.1.6 Flash Drum FD-01 44
4.1.7 Kompresor K-01 44
4.1.8 Kompresor K-02 45
4.1.9 Blower BL-01 45

4.1.10 Blower BL-01 45
4.1.11 Vaporizer VP-01 45
4.1.12 Heater H-01 46
4.1.13 Condenser Parsial CD-01 47
4.1.14 Cooler C-01 48
4.1.15 Cooler C-02 49
4.1.16 Pompa P-01 50
4.1.17 Pompa P-02 51
4.1.18 Pompa P-03 51
4.1.19 Pompa P-04 52
4.2 Peralatan Utilitas 52
4.2.1 Screen 52
4.2.2 Reservoir 53
4.2.3 Bak Pengadukan Cepat (BPC) 53
4.2.4 Bak Pengendap I 53
4.2.5 Bak Pengendap II 54
4.2.6 Tangki Filtrasi *(sand filter)* 54
4.2.7 Bak Penampung Air Bersih 55
4.2.8 Tangki Demineralisasi *(ion exchanger)* 55
4.2.9 Bak Umpan Boiler 56
4.2.10 Bak Umpan Cooling Tower 56
4.2.11 Bak Penampung Air Domestik 56
4.2.12 Bak Penampung Limbah 57
4.2.13 Spesifikasi Pompa PU-01 57
4.2.14 Spesifikasi Pompa PU-02 58
4.2.15 Spesifikasi Pompa PU-03 58

4.2.16 Spesifikasi Pompa PU-04 58
4.2.17 Spesifikasi Pompa PU-05 59
4.2.18 Spesifikasi Pompa PU-06 59
4.2.19 Spesifikasi Pompa PU-07 60
4.2.20 Spesifikasi Pompa PU-08 60
4.2.21 Spesifikasi Pompa PU-09 60
4.2.22 Spesifikasi Pompa PU-10 61
4.2.23 Spesifikasi Pompa PU-11 61
4.2.24 Spesifikasi Pompa PU-12 61
BAB V ASPEK KESELAMATAN, KESEHATAN KERJA DAN LINGKUNGAN 63
5.1 Deskripsi Singkat 63
5.2 Pertimbangan Aspek Keselamatan Pabrik 63
5.3 Pertimbangan Aspek Keselamatan dan Kesehatan Kerja 67
5.4 Pertimbangan Aspek Lingkungan Pabrik 69
BAB 6 ANALISIS KELAYAKAN PABRIK 72
6.1 Manajemen Perusahaan 72
6.1.1 Diagram Organisasi 73
6.1.2 Perincian Jabatan dan Penggolongan Gaji 80
6.1.3 Penggiliran Tugas 85
6.2 Kelayakan Ekonomi 87
6.2.1 Asumsi dan Parameter 87
6.2.2 Fixed Capital 87
6.2.3 Modal Kerja *(Working Capital Investment)* 89
6.2.4 Biaya Produksi 90
6.2.5 Pengeluaran Umum 91
6.2.6 Penjualan dan Keuntungan 92

6.2.7 *Break Event Point* (BEP) 93
6.2.8 Analisis Ekonomi 94
DAFTAR PUSTAKA 97
LAMPIRAN 1 DATA 86
L1.1 Spesifikasi Bahan Baku dan Produk 86
L1.2 Literatur 89
L1.2.1 Patent US 20130090510A1 (2013) 89
L1.2.2 Patent US 009663414B2 (2017) 117
LAMPIRAN 2 NERACA MASSA DAN ENERGI 154
L2.1 Informasi Umum 154
L2.1.1 Perhitungan Kebutuhan Bahan Baku 155
L2.2 Neraca Mixer (M-01) 156
L2.2.1 Neraca Massa Mixer (M-01) 157
L2.3 Neraca *Vaporizer* (VP-01) 157
L2.3.1 Neraca Massa *Vaporizer* (VP-01) 157
L2.3.2 Neraca Energi *Vaporizer (HE-301)* 158
L2.4 Neraca Heater/Pemanas 161
L2.4.1 Neraca Energi Heater (H-01) 161
L2.5 Neraca Reaktor (R-01) 163
L2.5.1 Neraca Massa Reaktor (R-01) 164
L2.5.2 Neraca Energi Reaktor (R-01) 165
L2.6 Neraca *Condenser Parsial* (CD-01) 168
L2.6.1 Neraca Energi *Condenser Parsial* (CD-01) 168
L2.7 Neraca Flash Drum (FD-01) 172
L2.7.1 Neraca Flash Drum (FD-01) 172
L2.8 Neraca Cooler/Pendingin 174

L2.8.1 Neraca Energi Cooler (C-01) 174
L2.8.2 Neraca Energi Cooler (C-02) 176
LAMPIRAN 3 PERHITUNGAN SPESIFIKASI ALAT 179
L3.1 Perhitungan Spesifikasi Alat Proses 179
L3.1.1 Perhitungan Tangki Penyimpanan TP-01 179
L3.1.2 Perhitungan Tangki Penyimpanan TP-02 184
L3.1.3 Perhitungan Tangki Penyimpanan TP-03 188
L3.1.4 Perhitungan Mixer M-01 191
L3.1.5 Perhitungan Kompresor K-01 202
L3.1.6 Perhitungan Kompresor K-02 204
L3.1.7 Perhitungan Blower BL-01 206
L3.1.8 Perhitungan Blower BL-02 208
L3.1.9 Perhitungan Reaktor Fixed Bed Multitube R-01 210
L3.1.10 Perhitungan Flash Drum FD-01 218
L3.1.11 Perhitungan Vaporizer VP-01 230
L3.1.12 Perhitungan Heater H-01 234
L3.1.13 Perhitungan Condenser Parsial CD-01 238
L3.1.14 Perhitungan Cooler C-01 243
L3.1.14 Perhitungan Cooler C-02 251
L3.2 Pehitungan Spesifikasi Alat Utilitas 255
L3.2.1 Perhitungan Pompa Proses P-01 255
L3.2.2 Perhitungan Pompa Utilitas 251
LAMPIRAN 4 PERHITUNGAN UTILITAS 253
L.4.1 Unit Penyedia Air 253
L.4.1.1 Kebutuhan Air Pemanas (*Steam)* 253
L.4.1.2 Kebutuhan Media Pendingin *(Cooling Tower)* 255

L.4.1.3 Kebutuhan Air untuk *Dowtherm* ... 258
L.4.1.4 Kebutuhan Air Domestik ... 259
L.4.1.5 Unit Pengolahan Air ... 260
L4.2 Perhitungan Daya Pabrik ... 276
L.4.2.1 Daya Proses ... 276
L.4.2.2 Daya Penunjang ... 277
L.4.3 Perhitungan Bahan Bakar ... 278
L.4.3.1 Penyediaan Bahan Bakar ... 278
L.4.3.2 Perancangan Tangki Bahan Bakar ... 279
LAMPIRAN 5 ANALISIS EKONOMI ... 283
L.5.1 Ketetapan Analisa Ekonomi ... 283
L.5.2 Perhitungan Gaji Karyawan ... 287
L.5.3 Harga Alat ... 291
L.5.3.1 Daftar Harga Alat-Alat ... 291
L.5.3.2 Harga Peralatan Utama ... 292
L.5.3.3 Harga Peralatan Penunjang ... 295
L.5.4 Total Modal Investasi ... 301
L.5.5 Struktur Permodalan ... 301
L.5.6 Produk dan Bahan Baku ... 302
L.5.6.1 Hasil Penjualan Produk Per Tahun ... 302
L.5.6.2 Perhitungan Biaya Bahan Baku dan Bahan Penunjang ... 303
L.5.7 Salvage Value (Nilai Aset) dan Depresiasi ... 305
L.5.7.1 Salvage Value (Nilai Aset) ... 305
L.5.7.2 Depresiasi ... 305
L.5.8 Perhitungan Biaya Produksi Total (TPC) ... 307
L.5.9 *Break Even Point* (BEP) ... 318

L.5.10 Laba Rugi dan Pajak 319

L.5.11 *Minimum Payback Period* (MPP) 321

L.5.12 *Internal Rate of Return* 323

DAFTAR GAMBAR

Gambar 1.1 Lokasi Pendirian Pabrik Etilen 12

Gambar 2.1 Diagram Alir Proses Patent US 20130090510A1 (2013) 15

Gambar 2.2 Diagram Alir Proses Patent US 009663414B2 (2017)....................................... 16

Gambar 3.1 Proses Flow Diagram Pabrik *Ethylene* 21

Gambar 3.2 Blok Flow Neraca Masa 22

Gambar 3.3 Blok Flow Neraca Energi 23

Gambar 3.4 Konfigurasi Sistem Pengendalian *feedback* 20

Gambar 3.5 Alat Kontrol pada Reaktor (R-01)... 23

Gambar 3.6 Alat Kontrol pada Tangki (T-01) 24

Gambar 3.7 Alat Kontrol pada Mixer (P-03).. 24

Gambar 3.8 Alat Kontrol pada Vaporizer (VP-01) 25

Gambar 3.9 Alat Kontrol pada Flash Drum (FD-01).. 25

Gambar 3.10 Sistem Pengolahan Air 35

Gambar 3.11 Tata Letak Alat.. 37

Gambar 3.12 Tata Letak Pabrik 40

Gambar 6.1 Diagram Organisasi 75

Gambar L2.1 Diagram Alir Mixer (M-01)... 156

Gambar L2.2 Diagram alir Vaporizer (VP-01) 157

Gambar L2.3 Diagram Alir Reaktor (R-01)... 164

Gambar L2.4 Diagram alir Sub Cooler Condenser (CD-01) 168

Gambar L2.5 Diagram Alir Flash Drum (FD-01) 172

Gambar L3.1 Ukuran berbagai macam Head (Brownell&Young Fig 5.7)............................... 181

Gambar L3.2 Standar Dimensi untuk Head Flanged 182

Gambar L3.3 Hubungan dimensi Flange dan Dished Head 183
Gambar L3.3 Desain basis tangki berpengaduk 196
Gambar L3.5 Konsumsi daya untuk berbagai macam pengaduk 199
Gambar L3.6 Nilai Konstanta kL dan kT 201
Gambar L3.7 Nilai Friction Factor Suction 258
Gambar L3.8 Grafik S_b/S_a vs N_{Re} Suction 259
Gambar L3.9 Nilai Friction Factor Disccharge 261
Gambar L3.10 Grafik S_b/S_a vs N_{Re} Discharge 262
Gambar L3.10 Grafik Efisiensi Pompa 264
Gambar L3.11 Grafik Efisiensi Motor 264
Gambar L4.1 Unit Pengolahan Air 261

DAFTAR TABEL

Tabel 1.1 Data Produksi Etilen dari Tahun 2018 - 2022 3

Tabel 1.2 Proyeksi Produksi Etilen dari Tahun 2023 – 2027.. 3

Tabel 1.3 Data Impor Etilen dari tahun 2018 - 2022 4

Tabel 1.4 Proyeksi Peningkatan Impor Etilen dari Tahun 2023 - 2027.................................... 4

Tabel I.5 Data Konsumsi Etilen dari tahun 2018 - 2022 5

Tabel 1.6 Proyeksi Konsumsi Etilen dari Tahun 2023 - 2027 5

Tabel 1.7 Data Ekspor Etilen dari Tahun 2018 - 2022 6

Tabel 1.8 Proyeksi Perkembangan Ekspor Etilen.. 6

Tabel 1.9 Data Proyeksi Pasar Etilen Tahun 2023 – 2027 7

Tabel 1.10 Selisih antara Penawaran dan Permintaan pada Tahun Pendirian Pabrik 2023 - 2027

Tabel 1.11 Kapasitas Produksi Pabrik Etilen yang Sudah Ada.. 8

Tabel. 1.12 Produsen dan Kapasitas Etanol di Indonesia 10

Tabel 2.1 Parameter Perbandingan Teknologi Proses Produksi Etilen 14

Tabel 3.1 Kebutuham Steam Peralatan Pabrik Etilen 26

Tabel 3.2 Kebutuhan Cooling Tower Pabrik Etilen 30

Tabel 3.3 Kebutuhan Dowtherm Pabrik Etilen 31

Tabel 3.4 Kebutuhan Keseluruhan Air Pabrik Etilen.. 32

Tabel 3.5 Daya Proses Pabrik Ethylene 32

Tabel 5.1 Identifikasi Hazard Berdasarkan MSDS 63

Tabel 5.2 Identifikasih Hazard Peralatan Proses... 65

Tabel 5. 3 Identifikasi Hazard Tata Letak dan Lokasi 66

Tabel 5.4 Identifikasi Potensi Paparan Kimia... 67

Tabel 5.5 Identifikasi Potensi Paparan Fisis 68

Tabel 5.6 Identifikasi Hazard Emisi Gas 69
Tabel 5.7 Identifikasi Hazard Limbah Cair 70
Tabel 5.8 Identifikasi Hazard Limbah Padat 71
Tabel 6.1 Rincian Gaji Karyawan 82
Tabel 6.2 Jadwal Kerja Karyawan Shift 86
Tabel 6.3 Jadwal Kerja Karyawan Non Shift 86
Tabel 6.4 Asumsi & Parameter Analisa Ekonomi 87
Tabel 6.5 Modal tetap langsung 88
Tabel 6.6 Modal tetap tidak langsung 88
Tabel 6.7 Biaya kebutuhan bahan baku 89
Tabel 6.8 Biaya kebutuhan sarana penunjang 89
Tabel 6.9 Biaya produksi tahun ke-1 dan ke-2 90
Tabel 6.10 Biaya pengeluaran umum 91
Tabel 6.11 Penjualan dan Laba 10 tahun 92
Tabel 6.12 Nominal aliran masuk 93
Tabel 6.13 Break Event Point (BEP) 94
Tabel 6.14 MPP & NCFPV 95
Tabel 6.15 Internal Rate of Return 96
Tabel L1.1 Spesifikasi Bahan Baku Etanol 86
Tabel L1.2 Spesifikasi Produk Ethylene 87
Tabel L2.1 Berat Molekul Masing-Masing Komponen 154
Tabel L2.2 Panas Spesifik Fasa Gas Bahan Baku dan Produk 154
Tabel L2.3 Panas Spesifik Fasa Cair Bahan Baku dan Produk 154
Tabel L2.4 Entalpi Steam yang digunakan 117
Tabel L2.5 Entalpi Dowtherm A yang digunakan 154
Tabel L2.6 Neraca Massa Mixer (M-01) 157

Tabel L2.7 Neraca Massa Vaporizer (VP-01) 157
Tabel L2.8 Neraca Energi Aliran 3 Masuk Vaporizer (VP-01) 158
Tabel L2.9 Neraca Energi Fase Liquid Vaporizer (VP-01) 158
Tabel L2.10 Neraca Energi Perubahan Fase Komponen Vaporizer (VP-01) 159
Tabel L2.11 Neraca Energi Fase Gas Vaporizer (VP-01) 159
Tabel L2.12 Neraca Energi Aliran 4 Keluar Vaporizer (VP-01) 160
Tabel L2.13 Neraca Energi Total Vaporizer (VP-01) 160
Tabel L2.14 Neraca Energi Masuk Heater (H-01) 162
Tabel L2.15 Neraca Energi Keluar Heater (H-01) 162
Tabel L2.16 Neraca Energi Total (H-01) 163
Tabel L2.17 Neraca Massa Reaktor (R-01) 164
Tabel L2.18 Neraca Energi Aliran Masuk Reaktor (R-01) 165
Tabel L2.19 Neraca Energi Reaktan Reaktor (R-01) 166
Tabel L2.20 Neraca Energi Produk Reaktor (R-01) 166
Tabel L2.21 Neraca Energi Aliran Keluar Reaktor (R-01) 167
Tabel L2.22 Neraca Energi Total Reaktor (R-01) 168
Tabel L2.23 Neraca Energi Aliran 7 Masuk Condenser Parsial (CD-01) 169
Tabel L2.24 Neraca Energi Fase Gas Condenser Parsial (CD-01) 169
Tabel L2.25 Neraca Energi Perubahan Fase Komponen Condenser Parsial (CD-01) 170
Tabel L2.26 Neraca Energi Fase Liquid Condenser Parsial (CD-01) 170
Tabel L2.27 Neraca Energi Aliran 8 Keluar Condenser Parsial (CD-01) 171
Tabel L2.28 Neraca Energi Total Condenser Parsial (CD-01) 171
Tabel L2.29 Neraca Massa Mixer (FD-01) 172
Tabel L2.30 Neraca Energi Masuk Cooler (C-01) 175
Tabel L2.31 Neraca Energi Keluar Cooler (C-01) 175
Tabel L2.32 Neraca Energi Total (C-01) 176

Tabel L2.33 Neraca Energi Masuk Cooler (C-02) ... 177
Tabel L2.34 Neraca Energi Keluar Cooler (C-02) ... 177
Tabel L2.35 Neraca Energi Total (C-02) ... 178
Tabel L4.1 Kebutuham Steam Peralatan Pabrik Ethylene ... 253
Tabel L4.2 Kebutuhan Cooling Tower Pabrik Ethylene ... 255
Tabel L4.3 Kebutuhan Cooling Tower Pabrik Ethylene ... 256
Tabel L4.4 Kebutuhan Air umtuk *Dowtherm* Pabrik Ethylene ... 259
Tabel L4.5 Kebutuhan Keseluruhan Air Pabrik Ethylene ... 260
Tabel L4.6 Daya Proses Pabrik Ethylene ... 276
Tabel L5.1 Suku Bunga Dasar Kredit Bank Umum Konvensional di Indonesia ... 283
Tabel L5.2 Tabel Cost Index ... 286
Tabel L5.3 Gaji Karyawan ... 288
Tabel L5.4 Kenaikan Gaji Karyawan dalam 10 Tahun ... 290
Tabel L5.5 Total Penjualan Produk dalam 10 Tahun ... 302
Tabel L5.6 Biaya Bahan Baku dalam 10 Tahun ... 303
Tabel L5.7 Biaya Penunjang dalam 10 Tahun ... 304
Tabel L5.8 Depresiasi ... 306
Tabel L5.9 Perhitungan Biaya Produksi Total dalam 10 Tahun ... 308
Tabel L5.10 Nilai *Break Even Point* (BEP) ... 318
Tabel L5.11 Laba Rugi dan Pajak dalam 10 Tahun ... 320
Tabel L5.12 Jumlah Nominal Aliran Masuk ... 321
Tabel L5.13 *Minimum Payback Period* (MPP) ... 322
Tabel L5.14 *Internal Rate of Return* (IRR) ... 323

BAB 1

PENDAHULUAN

1.1 Latar Belakang

Indonesia merupakan salah satu negara yang saat ini sedang giat melakukan pembangunan di segala bidang. Peningkatan standar hidup telah menyebabkan masyarakat menggunakan produk yang lebih praktis dan modern. Perkembangan pesat teknologi dan industri kimia telah meningkatkan kebutuhan akan bahan kimia. Mengingat Indonesia masih mengimpor bahan kimia dasar dari luar negeri, kebutuhan akan bahan kimia dasar menyebabkan Indonesia harus memproduksi bahan kimia yang sangat dibutuhkan konsumen dalam negeri. Perkembangan industri kimia Indonesia diharapkan dapat mengurangi ketergantungan impor bahan kimia dari negara lain. Kementerian Perindustrian (Kemenperin) menyebut Indonesia masih menghadapi permasalahan berupa ketergantungan impor bahan baku kimia. Kementerian Perindustrian meyakini angka tersebut bisa semakin meningkat jika tidak dilakukan upaya pengembangan industri kimia nasional. Salah satu produk kimia yang dibutuhkan di Indonesia adalah etilen.

Etilen (C2H4) merupakan salah satu gas yang mudah terbakar, tidak berwarna dan berbau manis serta memiliki banyak manfaat misalnya sebagai komponen utama dari sebagian besar produk plastik sehingga memiliki nilai ekonomi yang besar. Etilen merupakan bahan baku utama untuk pembuatan produk bahan kimia antara seperti etilen glikol (EG), etilbenzena (EB), etilen diklorida (EDC), etilen dibromida, etil alkohol, vinil asetat, etil klorida, etilen klorohidrin dan produk hilir seperti polietilen (PE). Sementara untuk produk akhir yang dibuat dengan etilen termasuk kemasan makanan, mainan, wadah makanan, botol pipa, karpet isolasi dan peralatan rumah tangga.

Reaksi :

$C_2H_5OH \rightarrow C_2H_4 + H_2O$

Etanol Etilen Air

Permintaan global terhadap etilen terus meningkat oleh karena itu untuk memenuhi permintaan ini, industri petrokimia menyiapkan fasilitas produksi etilen di seluruh dunia, termasuk salah satunya di Indonesia. Berdasarkan data Badan Pusat Statistik (2022), kebutuhan pasar etilen di Indonesia yakni sebesar 1.553.725 ton/tahun, sementara produksi etilen di Indonesia melalui PT Chandra Asri Pacific Tbk bahwasanya produksi etilennya sebesar 900.000 ton/tahun, lalu sementara impor yang dilakukan oleh Indonesia untuk memenuhi kebutuhan pasar etilen sebesar 850.633,537 ton/tahun. Oleh karena masih tingginya pendirian pabrik etilen di Indonesia maka pembangunannya memiliki peluang yang cukup besar.

Permintaan etilen di Indonesia selama ini mengalami pertumbuhan yang cukup tinggi dan terus membesar sehingga tidak bisa dipenuhi oleh satu pabrik yang ada yaitu PT Chandra Asri Pacific Tbk. Maka dari itu perlu didirikan lagi pabrik yang memproduksi etilen dengan pertimbangan sebagai berikut:

1. Potensi pasar dalam negeri yang masih tinggi.
2. Potensi pasar ekspor yang tinggi.
3. Dapat menyediakan lapangan pekerjaaan baru.
4. Menempatkan Indonesia pada skala internasional untuk industri petrokimia.
5. Meningkatkan produk hilir produksi plastik di Indonesia.

1.2 Data Analisis Pasar

Analisis pasar dilakukan untuk mengetahui nilai produksi, konsumsi, ekspor-impor, dan pertumbuhan pasar untuk 5 tahun kedepan terhadap produk yang dihasilkan. Guna mengetahui peluang pasar sebelum pabrik etilen didirikan. Analisis pasar dilakukan agar pabrik tidak salah dalam perancangan strategi pembangunan pabrik dan pemasaran produknya ke pasaran. Peluang pasar dihitung berdasarkan data kapasitas konsumsi, produksi, ekspor dan impor yang telah diproyeksikan dengan menggunakan perhitungan persen pertumbuhan. Direncanakan pendirian pabrik etilen ini dapat membantu memenuhi kebutuhan dalam negeri dan juga luar negeri mengingat kebutuhan akan eilen sebagai sumber energi sangat tinggi.

1.2.1 Data Produksi

Berdasarkan data produksi yang didapatkan dari Annual Report PT Chandra Asih tahun 2018 hingga 2022, dapat ditunjukkan dalam Tabel 1.1.

Tabel 1.1 Data Produksi Etilen dari Tahun 2018 – 2022

(PT Chandra Asri, 2018-2022)

Tahun	Produksi (Ton/Tahun)	% Pertumbuhan
2018	900.000	-
2019	900.000	9,7561
2020	900.000	0,0000
2021	900.000	0,0000
2022	900.000	0,0000
Persen rata-rata produksi per tahun (%)		2,4390

Sumber : https://www.chandra-asri.com/investor-relations/reports/annual-reports

Dari data diatas, kita dapat memproyeksikan jumlah produksi untuk tiga tahun kedepan yaitu tahun 2025. Karena di Indonesia hanya terdapat satu pabrik yang memproduksi etilen, sehingga dapat diperkirakan produksi etilen sampai tahun 2025 sama dengan kapasitas pabrik yang ada. Proyeksi produksi etilen dapat dilihat pada tabel I.2 di bawah ini:

Tabel 1.2 Proyeksi Produksi Etilen dari Tahun 2023 – 2025

Tahun	Produksi (Ton/Tahun)
2023	900.000
2024	900.000
2025	900.000
SD	-
Rata-rata	900.000

1.2.2 Data Konsumsi

Berdasarkan data konsumsi yang didapatkan dari Badan Pusat Statistika (BPS), data konsumsi dapat ditunjukan dalam Tabel 1.3 di bawah ini:

Tabel 1.3 Data Konsumsi Etilen dari tahun 2018 - 2022

Tahun	Konsumsi (Ton/Tahun)	% Pertumbuhan
2018	769.952	
2019	952.352	23,69
2020	1.039.390	9,14
2021	1.039.390	7,31
2022	1.106.223	-0,82
Persen rata-rata konsumsi per tahun (%)		9,83

Sumber : https://www.bps.go.id/id/exim

BPS Data Impor 2018 – 2022 Kode HS 2901210000

Dari data Tabel 1.3 terkait konsumsi etilen di Indonesia, menunjukan tingkat kenaikan setiap tahunnya yang disebabkan oleh kebutuhan etilen di indonesia semakin meningkat dengan persen pertumbuhan rata-rata 9,83%. Proyeksi konsumsi etilen dapat dilihat pada tabel 1.4 di bawah ini:

Tabel 1.4 Proyeksi Peningkatan Konsumsi Etilen dari Tahun 2023 - 2025

Tahun	Konsumsi (Ton/Tahun)
2023	1.214.961
2024	1.334.388
2025	1.465.555
SD	125.343
Rata-rata	1.338.301

Hasil analisa menggunakan data sekunder jumlah konsumsi dalam rentang waktu lima tahun sejak 2018 – 2022 menunjukkan persentase pertumbuhan rata-rata sebanyak 9,83% yang artinya seiring berjalannya waktu, angka konsumsi akan etilen semakin meningkat.

1.2.3 Data Impor

Impor merupakan salah satu cara untuk memenuhi kebutuhan etilen. Dengan banyaknya pabrik etilen diharapkan dapat menekan kebutuhan Impor etilen di Indonesia. Berdasarkan data impor yang didapatkan dari Badan Pusat Statistika (BPS), data konsumsi dapat ditunjukan dalam Tabel 1.5 di bawah ini:

Tabel I.5 Data Impor Etilen dari tahun 2018 - 2022

Tahun	Impor (Ton/Tahun)	% Pertumbuhan
2018	261.175	-
2019	121.249	-53,58
2020	8.296	-93,16
2021	2.209	-73,37
2022	7.116	222,06
Persen rata-rata impor per tahun (%)		0,49

Tabel 1.5 menunjukan bahwa adanya peningkatan jumlah impor etilen setiap tahunnya, dengan rata-rata persen pertumbuhan sebesar 0,49%. Berikut adalah proyeksi jumlah impor etilen tahun 2023-2025:

Tabel 1.6 Proyeksi Impor Etilen dari Tahun 2023 - 2025

Tahun	Impor (Ton/Tahun)
2023	7.151
2024	7.186
2025	7.221
SD	35
Rata-rata	7.186

1.2.4 Data Ekspor

Berdasarkan data ekspor yang didapatkan dari Badan Pusat Statistika (BPS), data konsumsi dapat ditunjukan dalam Tabel 1.7 di bawah ini:

Tabel 1.7 Data Ekspor Etilen dari Tahun 2018 - 2022

Tahun	Ekspor (Ton/Tahun)	% Pertumbuhan
2018	131.127	-
2019	173.601	32,39
2020	147.686	-14,93
2021	217.552	47,31
2022	213.339	-1,94
Persen rata-rata ekspor per tahun (%)		15,71

Sumber : https://www.bps.go.id/id/exim

BPS Data Ekspor 2018 – 2022 Kode HS 2901210000

Dari data diatas, kemudian kita bisa memproyeksikan ekspor etilen pada tahun 2023 hingga 2025 adalah sebagai berikut:

Tabel 1.8 Proyeksi Perkembangan Ekspor Etilen

Tahun	Ekspor (Ton/Tahun)
2023	246.851
2024	285.628
2025	330.495
SD	41.859
Rata-rata	287.658

1.3 Penentuan Kapasitas Pabrik

Ada dua parameter yang dijadikan acuan dalam menentukan kapasitas pabrik ;

1. Perbedaan antara nilai supply dan demand di tahun pabrik akan beroperasi.
2. Kapasitas ekonomis terpasang.

Kapasitas produksi antara nilai supply dan demand merupakan hal yang wajib dipertimbangkan dengan baik saat melakukan perancangan pendirian pabrik. Semakin besar kapasitas pabrik dan semakin banyak produk yang dapat dihasilkan maka semakin besar keuntungan yang dapat diperoleh. Karena itu penentuan kapasitas berperan sangat penting dalam perhitungan teknis maupun ekonomis dari pabrik yang akan didirikan.

Tabel 1.9 Data Proyeksi Pasar Etilen Tahun 2023 – 2025

Tahun	Produksi (Ton/Tahun)	Konsumsi (Ton/Tahun)	Impor (Ton/Tahun)	Ekspor (Ton/Tahun)
2023	900.000	1.214.961	7.151	246.851
2024	900.000	1.334.388	7.186	285.628
2025	900.000	1.465.555	7.221	330.495
SD	-	125.343	35	41.859
Rata-rata	900.000	1.338.301	7.186	287.658

Untuk menentukan seberapa besar kinerja produksi yang diharapkan dibutuhkan di pasar, dapat menggunakan perhitungan analisis pasar, atau perhitungan penawaran dan permintaan.

Kapasitas Produksi = (Konsumsi + Ekspor) – (Impor + Produksi)

Pembangunan pabrik etilen dijadwalkan pada tahun 2023, dan pabrik ini diharapkan dapat beroperasi pada tahun 2025. Permintaan dan penawaran disajikan pada Tabel 1.9 berdasarkan data perkiraan produksi, konsumsi, ekspor dan impor.

Tabel 1.10 Selisih antara Penawaran dan Permintaan pada Tahun Pendirian Pabrik 2023 - 2027

Penawaran (ton)		Permintaan (ton)	
Produksi	900.000	Konsumsi	1.338.301
Ekspor	287.658	Impor	7.186

Total	1.187.658	1.345.487
Selisih	157.829	

Pada tahun 2025 diproyeksikan bahwa total penawaran etilen adalah sebesar 1.187.658 ton, sedangkan total permintaan etilen adalah sebesar 1.345.487 ton. Dari Tabel 1.10 diperoleh selisih antara penawaran dan permintaan adalah 157.829 ton. Selain itu, faktor lain dalam penentuan kapasitas produksi etilen adalah melihat kapasitas ekonomis yang sudah tersedia secara global. Berikut adalah data kapasitas ekonomis di dunia dapat dilihat pada Tabel 1.11

Tabel 1.11 Kapasitas Produksi Pabrik Etilen yang Sudah Ada

Perusahaan	Lokasi	Ton/Tahun
Formosa Petrochemical Corporation	Mailiao, Taiwan	2.935.000
Arabian Petrochemical Company	Jubail, Saudi Arabia	2..250.000
ExxonMobil Chemical Company	Baytown, TX, USA	2.197.000
Chevron Philips Chemical Company	Sweeny, TX, USA	1.865.000
Dow Chemical Company	Terneuzen, Netherland	1.800.000
Braskem	Triunfo, Brazil	200.000
Solvay Indupa	Santo Andre, Brazil	60.000
Chandra Asri Petrochemical	Indonesia	860.000

(*Chemview*, 2019)

Jika ditinjau dari data tersebut, kapasitas produksi pabrik etilen yang ditentukan berdasarkan perkiraan data kekosongan pasar terhadap etilen pada tahun 2023 merupakan kapasitas yang termasuk rentang kapasitas produksi dari Produsen etilen yang sudah ada di dunia yaitu antara 60.000 sampai 2.935.000 ton/tahun. Untuk memenuhi permintaan yang semakin meningkat maka kapasitas pabrik yang akan didirikan sebesar 85.000 ton/tahun. Kapasitas pabrik tersebut masih berada diatas kapasitas ekonomis.

1.4 Pemilihan Lokasi Pabrik

Salah satu aspek terpenting dalam perencanaan pabrik adalah pemilihan dan penentuan lokasi pabrik untuk memudahkan kegiatan produksi dan distribusi produk. Hal ini sangat menentukan kemajuan pabrik pada saat produksi dan kedepannya. Oleh karena itu, dalam menentukan lokasi suatu pabrik perlu memperhatikan aspek teknis dan ekonomis, seperti perhitungan biaya produksi dan biaya distribusi minimal. Pertimbangan lainnya adalah aspek sosiologis, yakni berupa mempelajari sifat dan perilaku masyarakat di sekitar lokasi rencana pabrik, dengan mempertimbangkan hambatan sosiologis yang muncul dari luar.
Pemilihan lokasi pabrik nantinya akan menjadi faktor penentu terhadap produksi, distribusi, dan eksistensi dari pabrik tersebut. Faktor-faktor yang perlu diperhatikan dalam melakukan penentuan lokasi pabrik yang akan didirikan yaitu sumber bahan baku, pasar, transportasi, ketersediaan tenaga kerja, iklim, dan juga kebijakan pemerintah pada daerah setempat (Peter dan Timmerhaus, 1991).

1.4.1 Faktor Primer Pemilihan Lokasi Pabrik

1. Pasokan bahan baku

Sumber bahan baku menjadi pertimbangan terpenting dalam memilih lokasi pabrik dengan pembiayaan yang ekonomis dan transportasi yang mudah, sehingga dapat mempengaruhi biaya transportasi. Bahan baku untuk proses produksi etilen diperoleh dari dalam negeri. Indonesia memiliki beberapa pabrik yang memproduksi etanol. Lihat Tabel 1.12

Tabel. 1.12 Produsen dan Kapasitas Etanol di Indonesia

No.	Nama Pabrik	Produksi (Kiloliter/tahun)	Lokasi
1.	PT Indo acidatama	50.000	Solo
2.	PT Molindo Raya	80.000	Jawa Timur - Lampung
3.	Aneka Kimia Nusantara	14.850	Jawa Timur
4.	PG Rajawali II	10.500	Jawa Barat
5.	PT Perkebunan Nasional XI	7.000	Surabaya
6.	PT Perkebunan Nasional X	30.000	Jawa Timur
7.	PT Ethanol Indonesia Industri	50.000	Lampung
8.	PT Lampung Distilerry	50.000	Lampung

(Kemenprin,2018)

Dari table 1.12, kami bisa menyimpulkan bahwasanya untuk menghilangkan langkah distribusi dan transportasi, pabrik harus berlokasi sedekat mungkin dengan sumber bahan mentah dan pasar tempat produk dijual. Namun bahan bakunya banyak ditemukan di pulau Jawa, dan pasar etilen sebagian besar berlokasi di Cilegon, provinsi Banten. Pabrik etilen berbahan dasar etanol dipilih karena lokasinya di Cilegon, Provinsi Banten, dekat dengan pasar karena etilen merupakan gas berbahaya yang mudah terbakar dan harus disimpan dalam fase gas dalam tabung bertekanan tinggi. Oleh karena itu, penyimpanan memerlukan penanganan khusus dan harus memperhatikan keselamatan karyawan dan lingkungan.

2. Lokasi berkenaan dengan pasar

Pabrik yang menggunakan produk etilen sebagai bahan bakunya sebagian besar berlokasi di provinsi Banten dan Jawa Barat, sedangkan beberapa pabrik berlokasi di provinsi Metropolitan dan Jawa Tengah.sebagai bahan baku meliputi etil diklorida, etil oksida, etil eter, etilen oksida, dan etilen glikol, yang merupakan bahan baku utama polimer seperti plastik, resin, serat, dan elastomer, serta pelarut, surfaktan, cat, dan antibeku termasuk pabrik. Indonesia memiliki produsen dan konsumen produk etilen terbesar yaitu PT Chandra Asri Pacific Tbk dan PT Lotte Chemical, dimana etilen dibagi lagi menjadi polietilen atau grade polimer yaitu Low Linear Density Polyethylene (LLPDE) dan High Density Polyethylene (HDPE).

3. Fasilitas transportasi

Dalam mendirikan pabrik, ketersediaan transportasi sangat berguna sehingga memudahkan pendistribusian produk dan bahan baku untuk mengoperasikan pabrik. Pengiriman bahan baku dan produk dapat dilakukan melalui laut maupun darat. Wilayah Cilegon memiliki pilihan transportasi yang sangat baik dan memadai seperti Jalan Raya Merak, Tol Tangerang – Merak dan Tol Trans Jawa.Dekat juga dengan terminal penerimaan kargo PT Indah Kiat Pulp & Paper yaitu Pelabuhan Cikandan, Pelabuhan Merak Mas dan Pelabuhan Cikading 6.

4. Ketersediaan utilitas

Utilitas utama yang dibutuhkan dalam kelancaran proses produksi yaitu air, listrik dan bahan bakar. Kebutuhan air proses dapat dipenuhi dari pengolahan air tanah agar memenuhi standar air industri. Air tanah diperoleh dari PT Peteka Eka Karya dengan nantinya kita juga membangun water treatment plant untuk kebutuhan sehari-hari kita nantinya. Sedangakan sumber listrik dapat dipenuhi dari PLN dimana disamping itu energi listrik juga dapat diproduksi sendiri menggunakan diesel generator jet.

1.4.2 Faktor Sekunder Pemilihan Lokasi Pabrik

1. Ketersediaan tenaga kerja

Lokasi tenaga kerja yang dibutuhkan dapat direkrut dari tenaga ahli dan berpengalaman di lapangan serta tenaga kerja lokal yang berasal dari sekitar pabrik yang tidak jauh dari permukiman masyarakat. Oleh karena itu, maka pabrik dapat membuka lapangan kerja baru dan mampu meningkatkan taraf hidup masyarakat sekitar.

2. Ketersediaan tanah yang cocok

Kondisi lahan yang masih relatif luas dan datar sangat produktif. Cilegon juga merupakan salah satu kawasan industri di Indonesia, sehingga terdapat peraturan ketat dan tindakan perbaikan terkait dampak lingkungan.

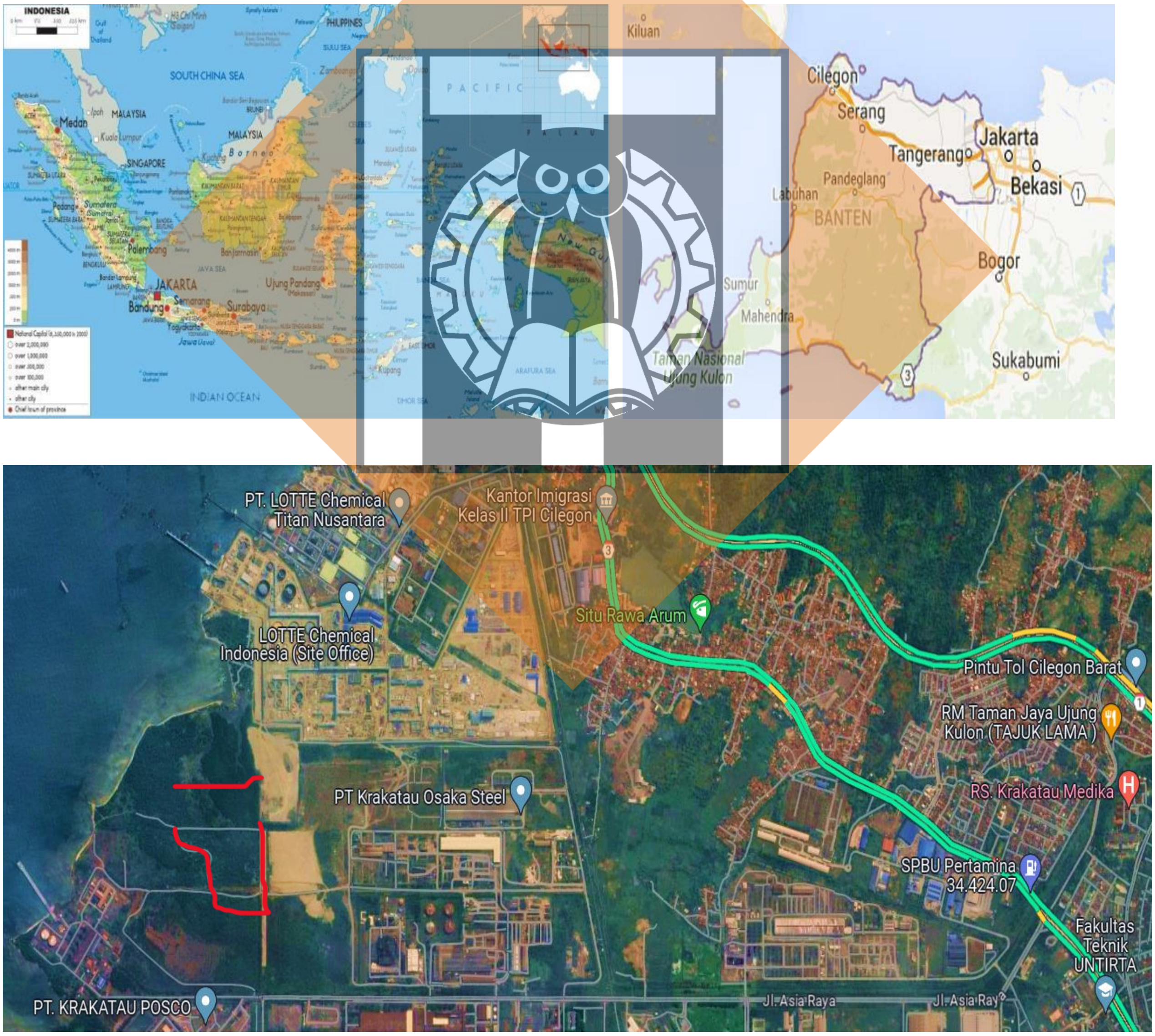
3. Dampak lingkungan

Pembangunan pabrik disuatu daerah akan membawa dampak baik secara positif maupun secara negatif. Bagi kehidupan sosial, industri cenderung membawa dampak positif, tapi bagi lingkungan hidup industri membawa banyak dampak negatif seperti pencemaran udara, polusi, air dan lain sebagainya. Selain yang telah disebutkan tadi, dalam lingkungan sosial industri biasanya mendapat tuntutan sosial, oleh karena itu perlu dilakukan monitoring terhadap pengolahan limbah sehingga meminimalisir pencemaran lingkungan di sekitar pabrik.

4. Iklim

Indonesia terletak di daerah tropis dan hanya memiliki dua iklim; musim hujan dan kemarau yang menjadikan pembangunan pabrik dan kelancaran proses produksi dan penjualan menguntungkan dan mudah.

Berdasarkan dari faktor-faktor tersebut pendirian pabrik pembuatan Etilen direncanakan akan berlokasi di Jalan Amerika Kelurahan Warnasari, Kecamatan Citangkil, Kota Cilegon, Provinsi Banten (titik ordinat 5°59'25.7"S 105°59'28.1"E) yang dapat dilihat pada Gambar 1.1



GAMBAR 1.1 LOKASI PENDIRIAN PABRIK ETILEN

BAB 2

TEKNOLOGI PROSES

2.1 Teknologi yang tersedia

Menurut Mc. Ketta (1984), ada beberapa cara untuk menghasilkan etilen:

A. Dehidrasi Etanol

Diharapkan bahwa sumber alternatif produksi etilen akan berasal dari reaksi dehidrasi etanol menjadi etilen. Etanol dan nitrogen kemudian dihilangkan dari pengotor sulful dan kemudian dikirim ke vaporizer untuk diuapkan dan dipanaskan di heater sebelum masuk ke reaktor. Reaktor fixed bed multitube digunakan, dan katalis yang digunakan adalah *zeolit ZSM-5*. Persamaan reaksinya adalah sebagai berikut: $C_2H_5OH \xrightarrow{\text{Zeolit ZSM-5}} C_2H_4 + H_2O$. Reaksi dehidrasi etanol membentuk etilen dengan nilai konversi tergantung ke standarnya dengan pada tekanan sesuai >1 atm.

B. Cracking Hidrokarbon

Pirolisis hidrokarbon dapat menghasilkan etilen dalam jumlah besar; bahan baku biasanya adalah etana, propana, butana, dan gas minyak. Bahan baku hidrokarbon dimasukkan ke dalam reaktor pirolisis, juga dikenal sebagai furnace. Di dalam furnace, reaksi tidak isotermal tetapi endotermis. Proses berlangsung secara konsisten pada suhu 650–950 °C dan perbandingan steam 0,3– 0,7 untuk bahan baku. Untuk mencegah proses polimerisasi, querching tower mendinginkan gas yang dihasilkan dari reaksi secara cepat. Selanjutnya, pemurnian dilakukan dengan menggunakan kolom fraksinasi untuk mernisahakan bagian hidrokarbon yang terbentuk dari reaksi (Mackenzie et al., 1983).

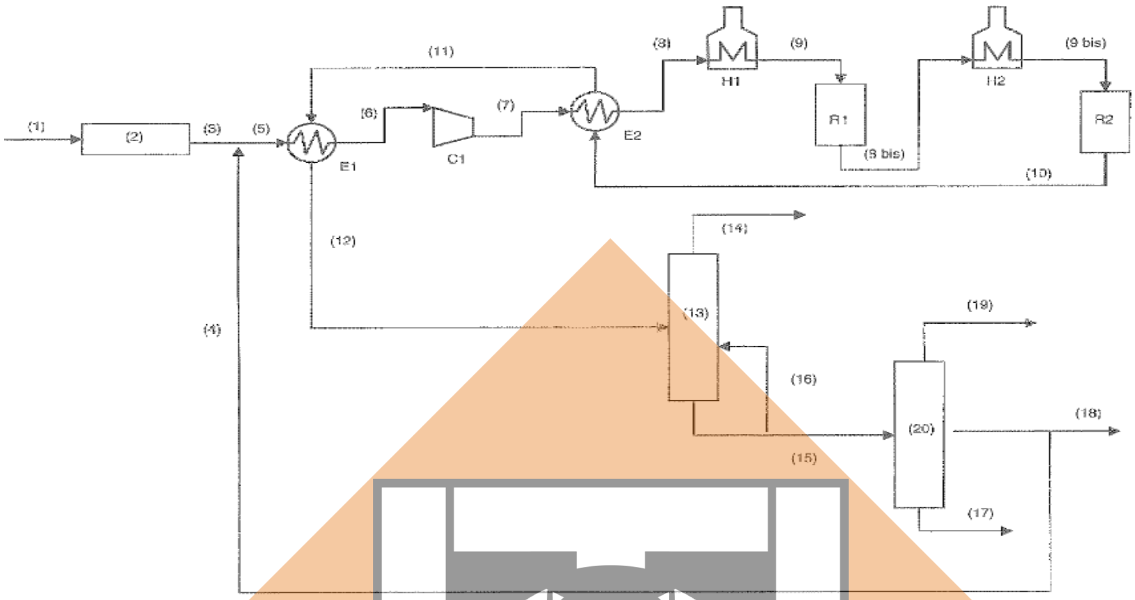
C. Oxidative Coupling of Methane (OCM)

Gas metana digunakan dalam metode Oxidative Coupling of Methane (OCM). Gas ini dapat berasal dari gas alam atau dari gasifikasi batu bara. Gas metana ini berinteraksi dengan reaktor OCM dengan bantuan oksigen dalam reaksi berikut: 2CH4 (g) + O2 (g) → C2H4 (g) + 2H2O (g). Kelemahan reaksi OCM konvensional termasuk oksigen murni yang telah dipisahkan dari nitrogen saat berinteraksi dengan gas metana, tingkat etilen yang rendah karena oksidasi gas metana yang terlalu lama, dan kemungkinan ledakan karena gas metana bercampur dengan oksigen murni. Salah satu keuntungan menggunakan reaktor ini dalam reaksi OCM adalah bahwa ia menggantikan oksigen murni dengan udara sebagai reaktan, menghilangkan proses pemurnian oksigen, meningkatkan produksi etilen, dan mengurangi kemungkinan ledakan. Reaktor tubular membran SOFC sebanding dengan reaktor OCM lainnya. Dengan katalis Mn-Ce-Na2WO4/SiO2gel, konversi gas metana mencapai 60,7%, selektifitas C2+ sebesar 41,6%, perbandingan etilen dengan gas metana yang terbentuk sebesar 5,8, dan produksi etilen sebesar 19,4%.

Tabel 2.1 Parameter Perbandingan Teknologi Proses Produksi Etilen

Parameter	Dehidrasi Etanol	Cracking Hidrokarbon	Oxidative Coupling of Methane (OCM)
Bahan Baku	Etanol	Nafta	Gas Metana
Tekanan (atm)	6 atm	± 3,5 atm	-
Suhu (˚C)	350 – 500	650 – 900	-
Katalis	Zeolit	-	Mn-Ce-Na2Wod/SiO2gel
Fase Reaksi	Gas-Padat	Gas – Gas	Gas – Gas
Reaktor	Fixed Bed Multitube	RAP Multibular Pirolisis dengan Box-Type Furnace	Reaktor Tubular
Konversi Reaktor	99%	92%	60,7%

2.1.1 Patent US 20130090510A1 (2013)



GAMBAR 2.1 DIAGRAM ALIR PROSES PATENT US 20130090510A1 (2013)

Metode ini menjelaskan mengenai teknologi pembuatan etilen dari proses dehidrasi etanol encer dengan konsumsi rendah tanpa daur ulang air.

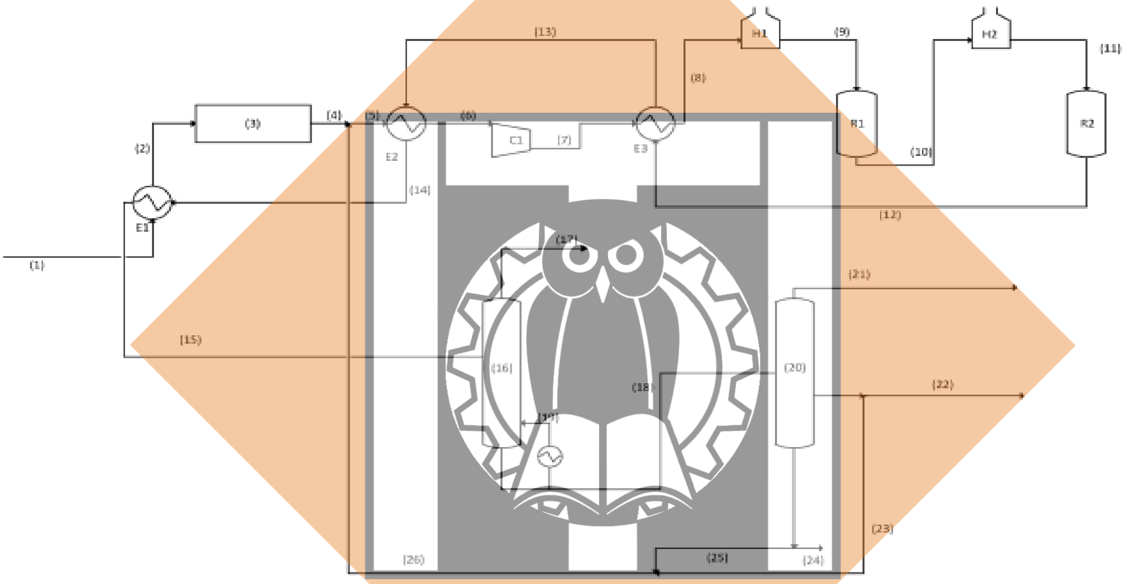
Proses pembuatan etilen dengan tahapan proses sebagai berikut :

a. Penguapan etanol encer dilakukan penguapan bahan baku etanol encer dengan persen massa antara 2% hingga 55% yang diuapkan dalam suatu *exchanger*, bahan baku etanol tersebut dimasukkan ke dalam tahap penguapan tersebut pada tekanan antara 0,1 MPa hingga 1,4 MPa
b. Proses kompresi bahan baku tersebut yang akan diuapkan akan dikompresikan di dalam kompresor, dimana tekanan pada akhir tahap kompresi adalah antara 0,3 Mpa hingga 1,8 MPa.
c. Dilanjutkan dengan proses pemasukan bahan baku yang telah diuapkan dan dikompresi, pada suhu masuk antara 350°C hingga 550°C. Dan pada tekanan masuk antara 0,3 MPa hingga 1,8 MPa ke dalam reaktor isothermal yaitu *multitube fixed-bed reactors* yang ditandai dengan aliran ke bawah dan ke atas dengan menggunakan katalis dehidrasi *amorphous acid catalyst* atau *zeolitic acid catalyst* dan di sinilah

reaksi dehidrasi berlangsung.

d. Proses pemisahan efluen yang diperoleh dari reaktor isothermal yang terdiri dari efluen yang terdiri dari etilen dan efluen yang terdiri dari air dengan menggunakan separator.
e. Proses produksi diakhiri dengan pemurnian sebagian dari efluen yang terdiri dari air yang diperoleh dari tahap sebelumnya dan pemisahan sebagian aliran etanol yang tidak bereaksi didaur ulang dan dicampur dengan bahan baku etanol untuk diuapkan tanpa mendaur ulang aliran air murni tersebut.

2.1.2 Patent US 009663414B2 (2017)



GAMBAR 2.2 DIAGRAM ALIR PROSES PATENT US 009663414B2 (2017)

Metode ini menjelaskan mengenai teknologi pembuatan etilen dari proses dehidrasi etanol encer dengan konsumsi rendah.

Proses pembuatan Etilen dengan tahapan proses sebagai berikut :

a. Pemanasan bahan baku etanol dengan persentase massa 35% dan 99,9% berat, langkah pemanasan awal bahan baku etanol tersebut antara suhu 100°C hingga 130°C dengan *vaporizer*.
b. Selanjutnya bahan baku etanol dengan resin penukar ion pada padatan asam yaitu resin TA801, dimana mempunyai kapasitas pertukaran untuk menangkap spesies basa dan kationik paling sedikit 0,1 mmol H ekuivalen per gram yang beroperasi pada suhu antara

100 dan 130°C. Untuk menghasilkan bahan baku etanol yang telah diolah sebelumnya.

c. Tahap penguapan bahan baku yang diuapkan terdiri dari bahan baku etanol yang telah diolah sebelumnya dan sebagian dari aliran air yang telah didaur ulang melalui *vaporizer* dengan efluen dari reaktor, pada tekanan antara 0,1 MPa dan 1,4 MPa sehingga menghasilkan bahan baku yang diuapkan.
d. Proses kompresi bahan baku tersebut yang akan diuapkan akan dikompresikan di dalam kompresor, dimana tekanan pada akhir tahap kompresi adalah antara 0,3 Mpa hingga 1,8 MPa dengan menggunakn *exchanger* dengan tipe *single phase gas type.*
e. Tahap dehidrasi bahan baku yang telah terkompresi dengan menggunakan reaktor isothermal yang mengandung katalis dehidrasi yaitu *amorphous acid catalyst* atau *zeolitic acid catalyst* dan di mana reaksi dehidrasi yang beroperasi pada suhu masuk antara 350°C hingga 550°C dan pada tekanan masuk antara 0,3 MPa hingga 1,8 MPa.
f. Selanjutnya tahap pemisahan efluen dari reaktor isothermal yaitu reactor *fixed bed multitube* dimana efluen yang terdiri dari etilen pada tekanan di bawah 1,6 MPa dan efluen yang terdiri dari air.
g. Proses produksi diakhiri dengan pemurnian sebagian dari efluen yang terdiri dari air yang diperoleh dari tahap sebelumnya dan pemisahan sebagian aliran etanol yang tidak bereaksi didaur ulang dan dicampur dengan bahan baku etanol untuk diuapkan dan mendaur ulang aliran air murni tersebut.

2.2 Seleksi Proses

Pemilihan teknologi untuk suatu pabrik kimia diawali dengan mengevaluasi beberapa alternatif teknologi yang ada menggunakan tiga parameter pembanding, yaitu:

2.2.1 Efisiensi Proses

Efisiensi proses berdasarkan teknologi, bahan baku, tingkat kompleksitas teknologi, limbah yang dihasilkan, pemakaian energi dan pemakaian katalis. Hasil dari paten kedua menunjukkan bahwa proses produksi etilen lebih kompleks dan lebih efisien dibandingkan pada paten pertama, lalu untuk kedua paten sama-sama adanya proses pre-treatment dimana pre-treatment ini bertujuan untuk menghilangkan pengotor sulfur dan nitrogen sehingga dapat memperpanjang umur katalis dan paten pertama menyatakan bahwa konversi 99,4% dibandingkan paten kedua 99% namun untuk dari selektivitas etilen yang dihasilkan paten pertama 97% sementara pada paten kedua 99% lalu paten satu tidak adanya recyle produk sementara paten kedua adanya recyle produk sehingga nantinya menghemat efesiensi proses salah satunya tidak memerlukan banyaknya steam.

2.2.2 Keamanan Proses

Dalam hal keamanan, kondisi operasi tertentu memerlukan pengendalian faktor keamanan. Salah satunya saat proses pembentukan produk etilen dimana berlangsung pada suhu yang cukup tinggi. Dalam paten pertama maupun paten kedua, katalis *zeolit ZSM-5* mengubah etanol menjadi etilen lalu pada paten pertama dengan reaktor isothermal yang digunakan pada suhu reaktor 440˚C dan tekanan 6 atm sementara pada paten kedua dengan reaktor isothermal yang digunakan suhu reaktor 470˚C sesuai suhu dehidrasi etanol dan tekanan 6 atm.

2.2.3 Biaya

Biaya secara kualitatif dengan membandingkan jumlah peralatan yang dibutuhkan untuk tiap teknologi, katalis digunakan serta bahan baku didapatkan secara impor atau tidak yang nantinya akan diterapkan dalam pembangunan pabrik. Biaya produksi harus dipertimbangkan secara menyeluruh dan optimal untuk mendapatkan keuntungan terbaik bagi perusahaan. Kedua proses produksi etilen yang telah dijelaskan sebelumnya menunjukkan bahwa paten kedua lebih kompleks secara teknologi dibandingkan dengan paten pertama. Namun, karena menggunakan suhu tinggi di reaktor, paten kedua menghemat energi karena menggunakan panas yang dihasilkan dari reaktor untuk digunakan kembali untuk memanaskan dan menguapkan bahan baku.

Tabel 2. 4 Perbandingan Proses Pembuatan Etilen

Proses		
Status Teknologi	Dehidrasi Etanol (Paten US 20130090510A1 2013)	Dehidrasi Etanol (Paten US 009663414B2 2017)
Bahan Baku	Etanol	Etanol
Katalis	*ZSM-5 (Zeolit)*	*ZSM-5 (Zeolit)*
Senyawa	C_2H_50H, C_2H_4, H_20	C_2H_50H, C_2H_4, H_20
Recyle/Reuse	Tidak	Ya
Selektivitas	97,4%	99%

Vessel / Reaktor Bertekanan Tinggi	Ya dengan 6 atm dan suhu 440˚C	Ya dengan 6 atm dan suhu 470˚C

BAB 3

RANCANGAN PROSES

3.1 Uraian Proses

Pada sub bab uraian proses akan dijelaskan tahapan proses produksi etilen dari dehidrasi etanol, diagram alir kuantitaif massa yang menunjukkan ringkasan laju alir massa untuk setiap aliran proses (tidak termasuk utilitas), dan diagram alir kuantitatif energi yang menunjukkan ringkasan alIran energi untuk setiap unit yang memiliki proses perpindahan panas.

Proses pembuatan etilen dari dehidrasi etanol melalui beberapa tahap yaitu:

1. Tahap persiapan bahan baku

2. Tahap pembentukan Etilen

3. Tahap pemurnian etilen

3.1.1 Deskripsi Proses

Produksi etilen dilakukan dengan melalui beberapa tahapan, yaitu proses penguapan etanol, proses dehidrasi etanol dan proses pemurnian (purification).

a. Perlakuan Awal Bahan Baku

Dalam pembuatan etilen digunakan bahan baku berupa etanol yang disimpan pada tangki penyimpanan etanol (TP-01) dengan suhu awal 30ºC dan tekanannya 1 atm. Kandungan etanol yang digunakan dalam proses ini adalah etanol 99%. Sebelum dilakukan proses dehidrasi, etanol diencerkan sampai konsentasi rendah dengan cara penambahan air dari (TP-02) sebagai medium pengencer, lalu dicampur menggunakan mixer (M-01) sampai konsentrasi 42%. Bahan baku etanol diperoleh dari daerah Lampung diantaranya PT Molindo Raya Plant Lampung.

b. Proses Utama

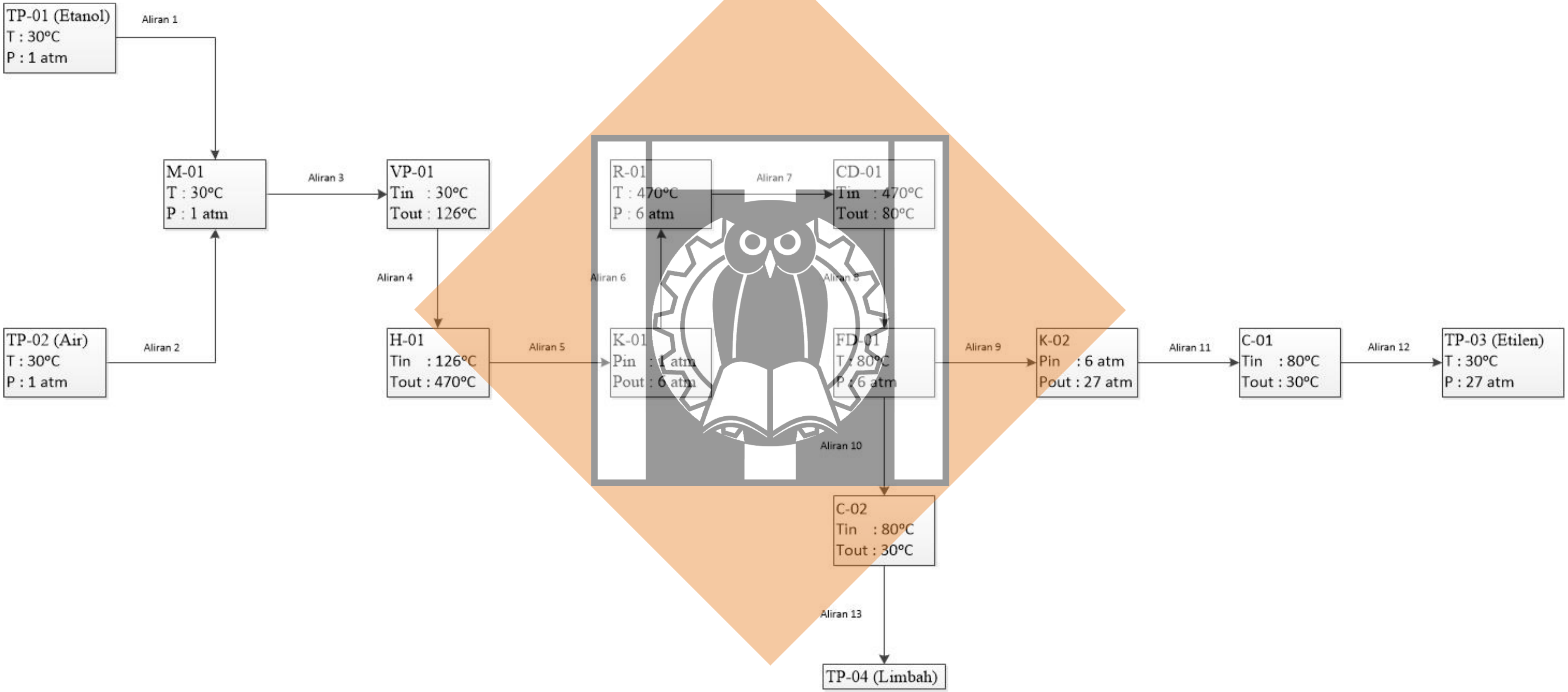
Larutan etanol encer dengan fraksi murni dialirkan ke vaporizer (VP-01) dan terjadi penguapan etanol dengan suhu masuk 30^oC dan tekanannya 1 atm sehingga suhu keluaran naik menjadi 126^oC dalam bentuk fase gas lalu dikompresi ke kompressor (K-01) agar tekanan uapnya naik pada tekanan 6 atm. Setelah fasenya sudah menjadi uap dan tekanan uap naik aliran dialirkan lagi menuju heater (H-01) dan dipanaskan suhunya menjadi 470^oC sehingga pada saat masuk ke reaktor adiabatik (R-01) sehingga terjadi reaksi pada suhu tersebut yaitu 470^oC yaitu suhu yang kompatibel dengan reaksi dehidrasi dan tekanannya 6 atm. Dimana reaksi yang terjadi berlangsung pada fase gas dengan persamaan reaksi sebagai berikut :

C_2H_5OH (*ZSM-5*) $\longrightarrow$ $C_2H_4 + H_2O$

Selektivitas produk hasil dehidrasi etanol ini adalah *etilen* 99% dimana nilai konversi *etanol* adalah 99%.

c. Pemurnian Produk

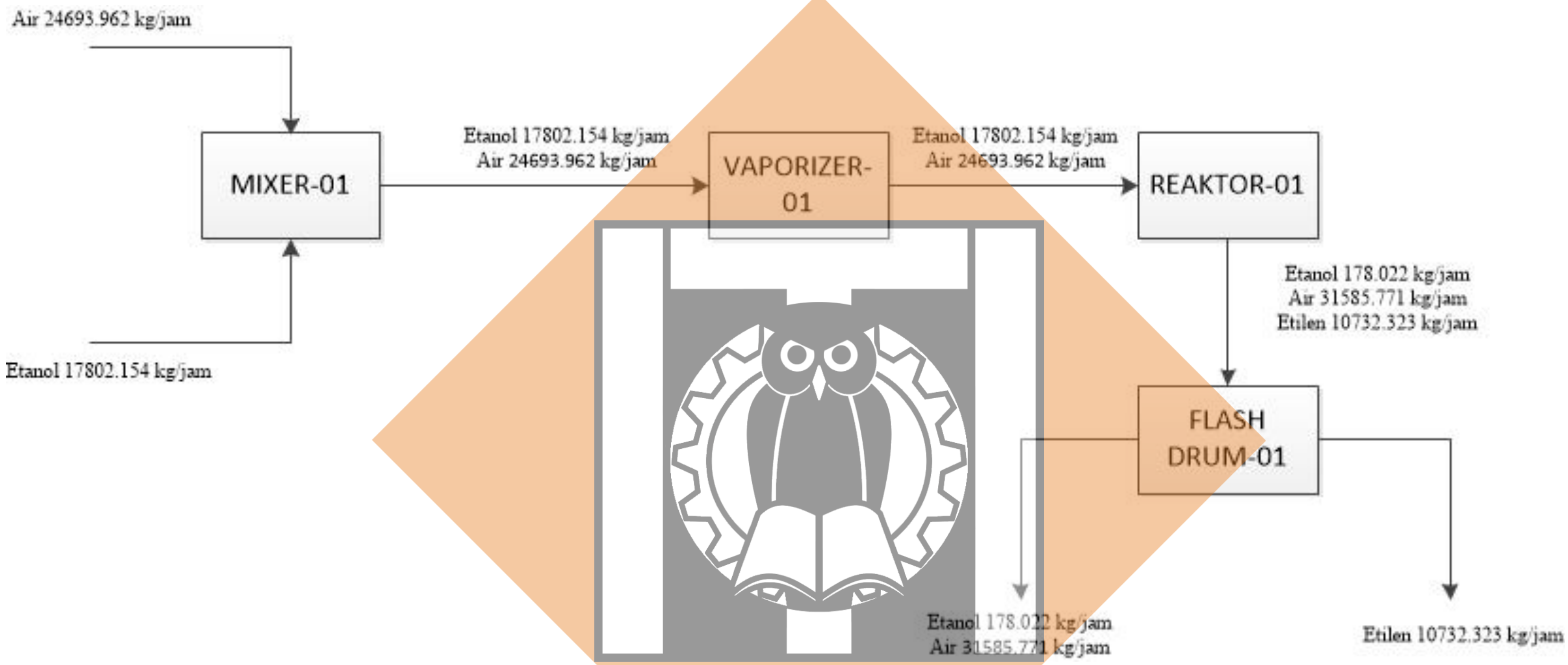
Efluen yang berasal dari reaktor (R-01) dialirkan ke kondensor parsial (CD-01) pada suhu 80^oC untuk merubah fasenya dimana etilen tetap menjadi gas, sementara etanol dan air akan berubah menjadi fase cair sebelum masuk ke dalam kondisi operasi dari unit flash drum (FD-01) yang beroperasi pada suhu 80^oC dan tekanan 6 atm, yang mana dalam proses pemisahan tersebut didapatkan 2 fasa yaitu fasa cair yang terdiri dar air dan etanol serta fasa gas etilen yang kemudian ditampung dalam tangki penampung etilen (TP-03) dengan tekanan 27 atm dan suhu 30^oC.



GAMBAR 3.1 PROSES FLOW DIAGRAM PABRIK *ETHYLENE*

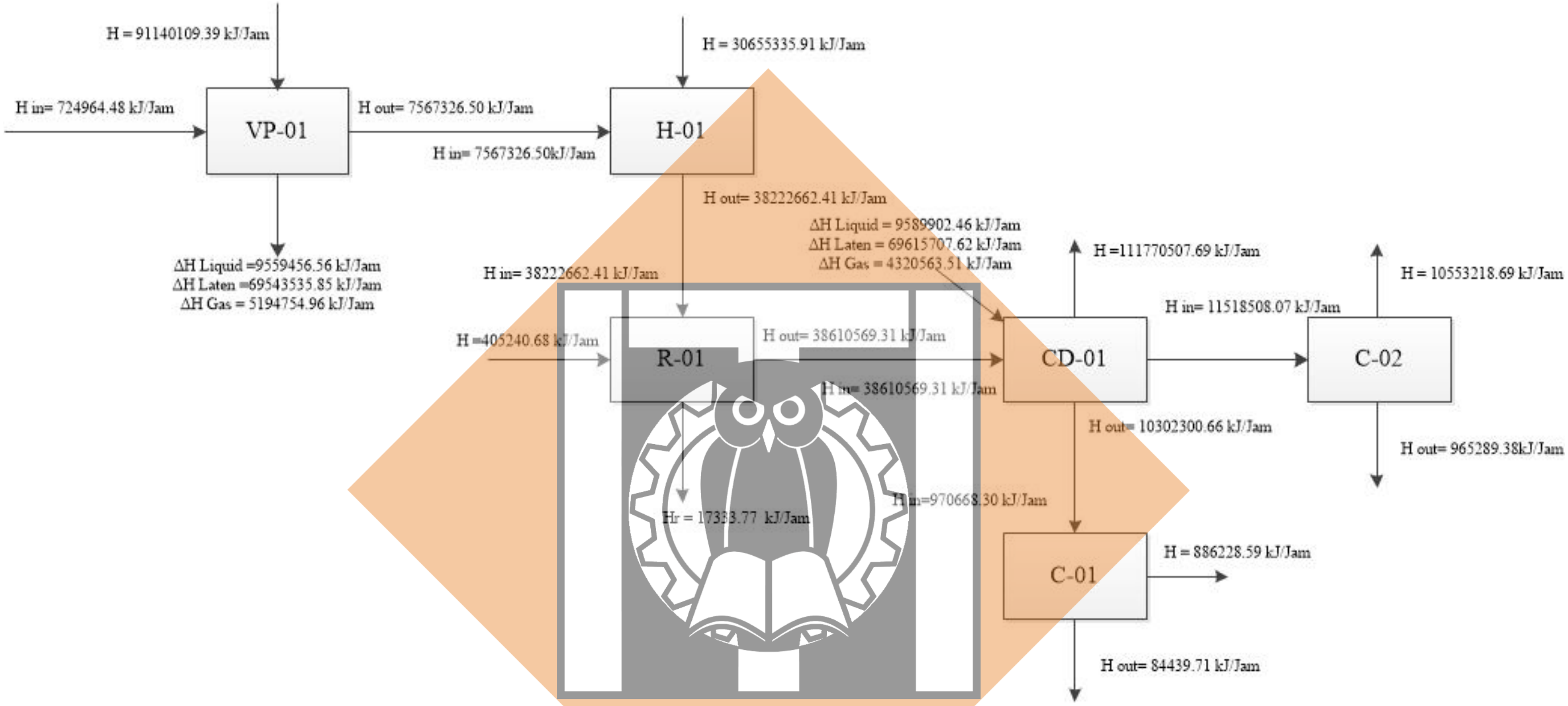
3.1.2 Diagram Alir Kuantitatif

Berikut merupakan diagram alir kuantitatif massa untuk setiap proses (tidak termasuk utilitas).



GAMBAR 3.2 BLOK FLOW NERACA MASA

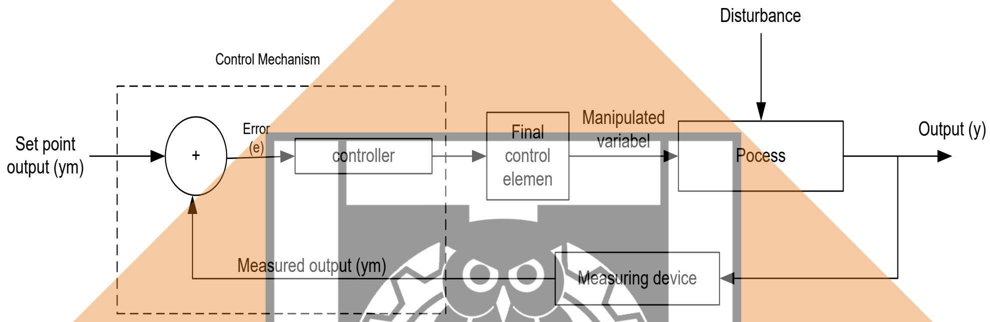
3.1.3 Diagram Alir Kuantitatif Energi



GAMBAR 3.3 BLOK FLOW NERACA ENERGI

3.1.4 Sistem Pengendalian Alat Utama

Tujuan pengendalian adalah untuk mempertahankan variabel yang dikendalikan pada harga yang diinginkan (setpoint). Pengendali yang digunakan pada rancangan awal pabrik etilen ini adalah pengendali yang menggunakan konfigurasi kendali umpan balik. Pengendalian dengan sistem ini mempunyai beberapa keunggulan: pengaturan sistem kontrol berbiaya rendah dan mudah. Selain itu, sistem ini dapat secara langsung mengukur variabel kontrol untuk menentukan harga variabel yang dimanipulasi Gambar 3.4 menunjukkan konfigurasi sistem feedback.



GAMBAR 3.4 KONFIGURASI SISTEM PENGENDALIAN *FEEDBACK*

Dalam pengoperasian pabrik propionaldehida ini banyak gangguan dari luar yang bisa dialami. Ada lima persyaratan yang perlu dipenuhi dalam pengoperasian pabrik yaitu:

1. Terjaminnya keamanan.
2. Terpenuhinya spesifikasi produk.
3. Peraturan lingkungan.
4. Kendala-kendala operasional pada masing-masing alat pemroses.
5. Keekonomisan untuk memperoleh keuntungan maksimum.

Untuk memenuhi semua persyaratan tersebut diperlukan adanya pengawasan dan intervensi dari luar. Sistem pengendalian diharapkan dapat memenuhi tiga kelompok kebutuhan berikut:

1. Menekan pengaruh gangguan eksternal.
2. Memastikan kestabilan suatu proses.
3. Optimasi kinerja suatu proses.

Variabel-variabel yang terlibat dalam proses operasi pabrik adalah F (laju alir), T (temperatur), P (tekanan) dan L (level). Variabel-variabel tersebut dapat dikategorikan menjadi dua kelompok, yaitu variabel input dan variabel output.

1. Variabel input

Variabel input adalah variabel yang menandai efek lingkungan pada proses kimia yang dituju. Variabel ini juga diklasifikasikan dalam dua kategori, yaitu:

- *Manipulated (adjustable) variable*, jika harga variabel tersebut dapat diatur dengan bebas oleh operator atau mekanisme pengendalian.
- *Disturbance variable*, jika harga tidak dapat diatur oleh operator atau sistem pengendali, tetapi merupakan gangguan.

2. Variabel output

Variabel output adalah variabel yang menandakan efek proses kimia terhadap lingkungan yang diklasifikasikan dalam dua kelompok:

- *Measured output variables*, jika variabel dapat diketahui dengan pengukuran langsung.
- *Unmeasured output variables*, jika variabel tidak dapat diketahui dengan pengukuran langsung.

Berikut merupakan alat-alat utama yang variabelnya sangat di perhatikan dalam prarancangan ini.

1. Reaktor (R-01)

Reaksi dehidrasi etanol adalah sebagai berikut:

$C_2H_5OH \rightarrow C_2H_4 + H_20$

Etanol → *Etilen + Air*

Reaksi *fixed bed multitube* dijalankan pada *temperature* 470°C dan tekanan 6 atm. Etanol diumpankan berlebih melaui aliran 5 setelah pemanasan untuk memastikan reaksi dehidrasi etanol terjadi sempurna dan menghindari terjadinya reaksi samping. Reaksi dehidrasi etanol berjalan selama 7 jam dan dengan *yield* diharapkan mencapai produk etilen terhadap etanol sebesar ±99%.

Reaksi tersebut bersifat eksotermis sehingga selama terjadinya reaksi *temperature* reaktor dan tekanan reaktor harus dijaga agar mencapai hasil produk akhir sesuai keinginan. Variabel-variabel yang mempengaruhi kondisi reaktor adalah *laju alir*, *temperature* dan tekanan reaktor. Untuk mengendalikan variable tersebut dipasang *controller* untuk mengetahui dan mengendalikan variable tersebut. *Controller* yang digunakan pada reaktor antara lain adalah:

a. *Temperature Indicator Controller* (TIC)

Fungsi : - Untuk mengetahui kondisi temperatur di dalam reaktor.

- Mencegah terjadinya perubahan temperatur di dalam reaktor.

Jenis kontrol yang digunakan adalah *fail open*. TIC dipasang di dalam reaktor. Besar kecilnya temperatur akan diterima berbentuk sinyal elektrik oleh TIC, dan akan diteruskan menuju *transduser* untuk diubah menjadi sinyal *pneumatic* yang akan menggerakkan valve steam.

Tindakan : Jika kondisi temperatur di dalam reaktor dibawah 470C, maka TIC akan meneruskan sinyal menuju *transduser* dan akan diubah menjadi sinyal *pneumatic* yang akan memperbesar bukaan *control valve* pada aliran steam pemanas. Sedangkan jika kondisi temperatur di dalam reaktor diatas 470°C, maka bukaan control valve pada aliran steam akan diperkecil.

b. *Flow Indicator Controller* (FIC)

Fungsi : Sebagai alat pengontrol aliran di dalam reaktor.

Pada sistem ini valve yang digunakan yaitu jenis *Air to open*, dimana gas akan menekan valve sehingga memberikan bukaan lebih besar. FIC dipasang pada aliran *offgas*. Besar kecilnya *aliran* akan diterima berbentuk sinyal elektrik oleh FIC, dan akan diubah menjadi sinyal digital yang akan diteruskan ke ruang kontrol, kemudian ruang kontrol akan mengirim sinyal secara digital ke FIC untuk diubah menjadi sinyal *pneumatic* yang akan menggerakkan valve pada aliran *offgas*.

Tindakan : Jika kondisi aliran membuat tekanan operasi di dalam rektor lebih besar dari 6 atm maka bukaan *valve offgas* akan diperbesar. Sedangkan jika kondisi operasi tekanan di dalam reaktor lebih kecil dari 6 atm maka valve pada *offgas* akan ditutup.

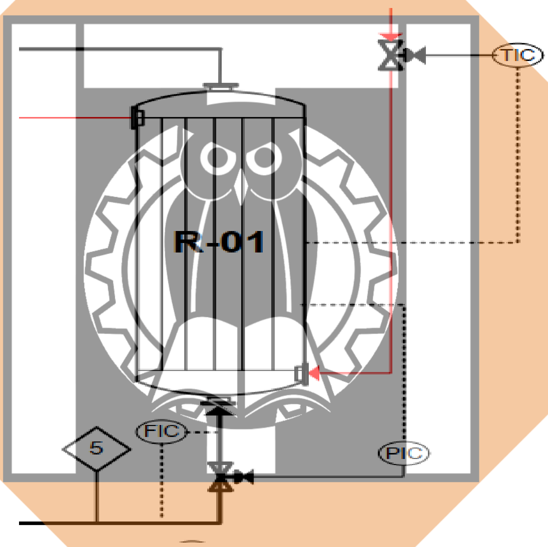
c. Pressure Indicator Controller (PIC)

Fungsi : - Untuk mengetahui kondisi tekanan di dalam reaktor.

- Mencegah terjadinya perubahan temperatur di dalam reaktor.

Jenis kontrol yang digunakan adalah pelampung dan lengan gaya. Prinsipnya adalah perubahan gaya apung yang dialami pelampung akibat perubahan level cairan. PIC dipasang pada aliran produk. Jenis kontrol yang digunakan adalah fail open. PIC dipasang di dalam reaktor. Besar kecilnya tekanan akan diterima berbentuk sinyal elektrik oleh PIC, dan akan diteruskan menuju transduser untuk diubah menjadi sinyal pneumatic yang akan menggerakkan valve steam .

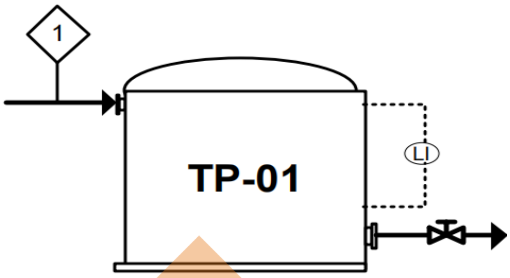
Tindakan : Jika kondisi tekanan di dalam reaktor dibawah 6 atm, maka PIC akan meneruskan sinyal menuju transduser dan akan diubah menjadi sinyal pneumatic yang akan memperbesar bukaan control valve. Sedangkan jika kondisi tekanan di dalam reaktor diatas 6 atm, maka bukaan control valve akan diperkecil..



GAMBAR 3.5 ALAT KONTROL PADA REAKTOR (R-01)

2. Tangki (TP-01,02,03)

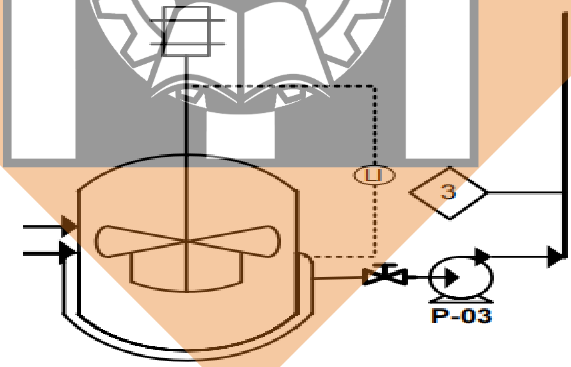
TP-01 dan TP-02 berfungsi sebagai tempat penyimpanan bahan baku. Oleh karena itu, Ketinggian cairan dalam TP-01 & TP 02 dijaga dengan bantuan Level Indicator Controller (LIC). Sedangkan TP-03 sebagai tempat penyimpanan produk utama, bekerja pada tekanan 27 atm. Oleh sebab itu, pressure dalam TP-03 dijaga tetap pada 27 atm dengan bantuan Pressure Indicator Controller (PIC) yang terhubung pada aliran utilitas steam. PIC akan mengirimkan sinyal pada control valve apabila tekanan dalam TP-03 kurang atau berlebih dan membuat valve membuka lebih lebar/lebih kecil sesuai dengan kondisi yang dibutuhkan TP-03.



GAMBAR 3.6 ALAT KONTROL PADA TANGKI (TP-01)

3. *Mixer* (M-01)

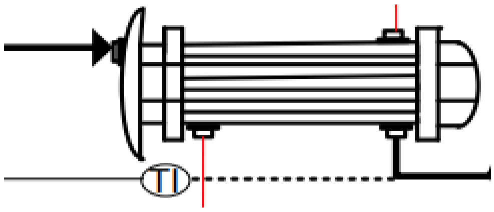
Pada *mixer* utama ini terjadi pencampuran antara aliran bahan baku utama, dan aliran air. Kedua aliran inlet ini *temperature*nya telah di jaga melalui pendinginan agar saat masuk ke dalam M-01 *temperature* operasinya tetap pada 30 ºC. Ketinggian cairan dalam M-01 dijaga dengan bantuan *Level Indicator Controller* (LIC) yang terhubung pada aliran keluar *mixer*.



GAMBAR 3.7 ALAT KONTROL PADA MIXER (M-01)

4. Vaporizer (VP-01)

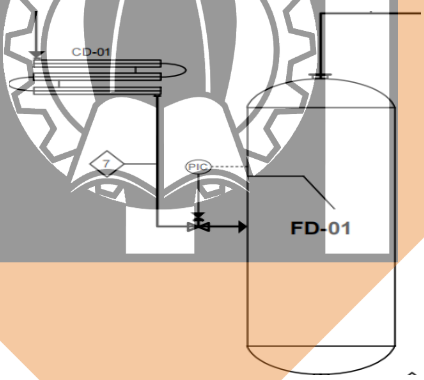
Vaporizer bekerja pada temperature 126˚C, dimana, vaporizer berfungsi sebagai tempat penguapan bahan baku setelah dari proses mixer sebelum dialirkan ke proses berikutnya. Oleh karena itu, *temperature* dalam vaporizer dijaga tetap pada 126˚C dengan bantuan *Temperature Indicator Controller* (TIC). TIC akan mengirimkan sinyal pada *control valve* apabila temperature dalam tangki kurang atau berlebih dan membuat *valve* membuka lebih lebar/lebih kecil sesuai dengan kondisi yang dibutuhkan vaporizer.



GAMBAR 3.8 ALAT KONTROL PADA VAPORIZER (VP-01)

5. *Flash Drum* (FD-01)

Flash drum bekerja pada *tekanan* 6 atm. Flash drum bekerja memisahkan produk utama dari air berdasarkan prinsip kesetimbangan uap air. Oleh sebab itu, *pressure* dalam FD-01 dijaga tetap pada 6 atm dengan bantuan *Pressure Indicator Controller* (PIC) yang terhubung pada aliran utilitas *steam.* PIC akan mengirimkan sinyal pada *control valve* apabila tekanan dalam FD-01 kurang atau berlebih dan membuat *valve* membuka lebih lebar/lebih kecil sesuai dengan kondisi yang dibutuhkan FD-01.



GAMBAR 3.9 ALAT KONTROL PADA FLASH DRUM (FD-01)

3.1.5 Kebutuhan Utilitas

Utilitas merupakan bagian penting dalam suatu kegiatan operasional pabrik yang bertujuan untuk membantu kelancaran proses unit produksi. Utilitas yang dibutuhkan dalam pabrik etilen berupa listrik, air, dan bahan bakar.

1. Unit Penyedia Air

Kebutuhan air parbik meliputi kebutuhan air domestik, air untuk *steam,* air untuk pendinginan dan lain-lain. Pemasok kebutuhan air berasal dari PT Krakatau Tirta. Dengan beberapa dasar perhitungan sebagai berikut.

a. Kebutuhan Air Pemanas (*Steam)*

Steam dihasilkan dalam sebuah *boiler fire tube* yang menggunakan bahan bakar solar, dengan jenis *steam* yang dihasilkan adalah *superheated steam.* Spesifikasi *steam* yang digunakan dalam kegiatan proses pabrik etilen ini adalah sebagai berikut (TLV, 2024):

Temperature = 480.00ºC

Tekanan (P*steam*) = 600 kPa

Entalpi *steam* = 3457.79 Kj/Kg

- Kebutuhan *Steam* untuk peralatan pabrik adalah sebagai berikut:

Tabel 3.1 Kebutuham Steam Peralatan Pabrik Etilen

No	Alat	Kebutuhan (Kg/h)
1	*Vaporizer* (VP-01)	26357.9076
2	*Heater* (H-01)	8865.5864
3	*Reaktor* (R-01)	117.1964
	Total Kebutuhan Steam	35340.6904

Dengan asumsi faktor keamanan dan kehilangan panas dimasing-masing alat, setiap *steam* dialirkan ke unit proses di lebihkan 10% sehingga :

$$Total + FK\ 10\% = 1.1\ \times 35340.6904 = 38874.7595\,\frac{Kg}{jam}$$

Atau setara dengan 85703.29474 lb/jam.

- Boiler Untuk Mendapatkan *Steam*

Boiler untuk mendapatkan steam berjenis *Boiler fire tube* dengan jumlah 1 unit. Penentuan daya *boiler* adalah sebagai berikut (Severn, 2004):

$$BHP = \frac{Ms\ \times (Hv - Hf)}{Cf\ \times (34.5)}$$

Keterangan :

Ms = Massa *Steam* = 85703.2947 lb/jam

Hv = Entalpi *vapor* pada *temperature* 480 ºC = 1478.8836 Btu/lb

Hf = Entalpi *steam* pada suhu 30 ºC = 54.0602 Btu/lb

Cf = Panas laten penguapan air pada 100 ºC = 970.0999 Btu/lb

Konversi= $34.5 \frac{\frac{\text{Lb}}{\text{h}}}{\text{HP}}$

$$BHP = \frac{85703.2947 \times (1478.8836 - 54.0602)}{970.0999 \times (34.5)} = 3648.5725\ HP$$

Heating surface untuk boiler tiap 1 HP adalah 10 ft³/HP (Severn, 2004). Maka,

$$Heating\ Surface\ Boiler = 10 \times 3648.5725 = 36485.7254 ft^3$$

- Kebutuhan Air Untuk Menghasilkan *Steam*

$$\text{Kebutuhan Air} = \frac{\text{MS} \times (\text{Hv} - \text{Hf})}{\text{Cf}}$$

Keterangan :

Ms = Massa *Steam* = 85703.2947 lb/jam

Hv = Entalpi *vapor* pada *temperature* 480 ºC = 1478.8836 Btu/lb

Hf = Entalpi *steam* pada suhu 30 ºC = 54.0602 Btu/lb

Cf = Panas laten penguapan air pada 10 0ºC = 970.0999 Btu/lb

$$Kebutuhan\ Air = \frac{85703.2947 \times (1478.8836 - 54.0602)}{970.0999}$$

$$= 125875.7528\ lb/jam$$

Atau setara dengan 57096.8669 kg/jam.

Densitas air pada 30 ºC adalah = 1022.8753 kg/m³.

Sehingga debit air (Qa) yang dibutuhkan adalah,

$$Qa = \frac{57096.8669}{1022.8753} = 55.8200\ m^3/jam$$

- Kebutuhan Air Untuk *Make Up Boiler*

Kehilangan air dikarenakan *blowdown* dan hilang akibat penguapan diasumsikan sebesar 10%, maka jumlah air yang harus diumpankan sebagai *make up boiler* adalah = 10% x massa air.

$$Volume\ yang\ dibutuhkan = \frac{10\% \times 57096.8669}{1022.8753} = 5.5820\ m^3/jam$$

- Kebutuhan Bahan Bakar Untuk *Boiler*

Bahan bakar yang digunakan untuk *Boiler* adalah solar. Dimana,

$$Solar\ yang\ dibutuhkan = \frac{m\ steam \times (hv - hf)}{\eta \times Hv}$$

Keterangan

m steam = massa steam = 85703.2947 kg/jam = 188941.4836 lb/jam

Hv = *Heating value* solar = 19200 btu/lb

hv = Entalpi vapor pada suhu 480ºC = 1478.8836 btu/lb

hf = Entalpi steam pada suhu 30ºC = 54.0602 btu/lb

η = efisiensi pembakaran = 85%

ρ bio solar = densitas solar = 850 kg/m³

$$Massa\ solar = \frac{188941.4836 \times (19200 - 54.0602)}{85\% \ \times 19200} = 16495.6029\ lb/jam$$

Atau setara dengan 7482.3564 kg/jam.

- **Perancangan *Cooling Tower*.**

Fungsi : Mendinginkan kembali air yang sudah digunakan sebagai media pendingin

Jenis : *Counter Flow Induced Draft*

Cara Kerja : Air yang sudah digunakan sebagai media pendingin dialirkan kedalam *Cooling tower***.** Disini air akan mengalami pendinginan karena adanya panas yang hilang akibat penguapan dan adanya tiupan angin dari kipas (*fan*). Panas yang diperlukan untuk penguapan itu diambil dari bahan air itu sendiri sehingga sebagian sisanya yang tidak teruapkan (tidak menguap) menjadi dingin dan dapat disirkulasi kembali sebagai media pendingin.

Data :

Bahan masuk = 30017.3114 kg/jam

Densitas air (pada 30 ºC) = 1022.8753 kg/m^3

T air masuk = 50ºC = 122 ºF

T air keluar = 30ºC = 86 ºF

Laju alir, Wc = 29.3460 m^3/jam

= (29.3460 m^3/jam) × (264.172 gal/1 m^3) × (1 jam/60 menit)

$= 107.4946$ gpm

Dari *psychometric chart* (fig.12-2 , Perrys) dengan temperatur rata-rata sekitar pabrik 30°C (86°F) dan kelembaban relatif sebesar 65 % diperoleh temperatur bola basah, $T_w = 75.2$ °F

Dari *Perry's* hal 12-15 dipilih *Cooling Tower* jenis *Induced Draft Cooling Tower* dengan pola aliran *counter current*.

Cooling range $= T_1 - T_2 = 36$ °F

Temp. approach $= T_2 - T_w = 10.8$ °F

Diinginkan tinggi menara = 8 m

Dari fig 12-14 *Perry's* pada temperatur air panas, $T_1 = 122$ °F Vs Temperatur air dingin, $T_2 = 86$ °F dengan temperatur bola basah, Tw = 75.2 °F, diperoleh kandungan air = 2.7 gal/min.ft²

- Perhitungan Luas Menara

$$\text{Luas menara} = \frac{\text{Wc}}{\text{Kandungan air}} = 39.8128 \text{ ft}^2$$

Diperkirakan effesiensi menara = 90 %

Maka luas menara aktual =39.8128 ft² / 0,9 = 44.2364 ft²

- Menghitung daya *fan*.

  Dari fig. 12–15 *Perry's* untuk effisiensi kerja *cooling tower* 90% diperoleh daya *fan* = 0.03 Hp/ft².

Daya *fan* aktual = 0.03 Hp/ft² × 44.2364ft² = 1.3271 Hp ≈ 3 Hp

- Menghitung jumlah air *make up*.

  Make up water, $W_m = W_e + W_b + W_d$ (Perry's, pers. 12-9)

  Menghitung jumlah air yang menguap, We : (Perry's, Pers 12-10)

  $W_e = 0.00085 \times Wc \times (T_1 - T_2)$

  Maka $W_e = 32.8933$ gpm

Menghitung jumlah *blow down* dalam air, Wb : (Perry's,pers.12-12)

$W_b = \frac{W_e}{\text{Siklus - 1}}$; siklus = 3 – 5 (diambil siklus = 3)

= 1.0000 gpm

Menghitung *drift loss*, Wd

W_d = 0.1 s/d 0.2 % Wc, (diambil 0.2 % Wc)

= 0.2 × 32.8933 gpm = 6.5787 gpm

maka, $W_m = W_e + W_b + W_d$

= 32.8933 + 1.0000 + 6.5787 = 40.4720 gpm

= 9.3086 m^3/jam

= 9521.4959 kg/jam

= 381267.3500 lb/jam

b. Kebutuhan Air Pendingin *(Cooling Tower)*

Air pendingin yang digunakan memiliki kondisi sebagai berikut :

Temperature masuk = 10°C

Temperature keluar = 25°C

Tabel 3.2 Kebutuhan Cooling Tower Pabrik Etilen

No	Alat	Kebutuhan Cooling Tower (Kg/h)
1	Cooler (C-01)	1921.8838
2	Cooler (C-02)	22885.8115
Total Kebutuhan Cooling Tower		24807.6953

Dengan asumsi factor keamanan dan penyerapan panas dimasing-masing alat, setiap *cooling tower* yang dialirkan ke unit proses dilebihkan 10%, sehingga :

$$Total + FK\ 10\% = 1.1\ \times 24807.6953 = 27288.4649\,\frac{Kg}{jam}$$

c. Kebutuhan Air untuk *Dowtherm*

Air pendingin yang digunakan memiliki kondisi sebagai berikut :

Temperature masuk = 10°C

Temperature keluar = 25°C

Tabel 3.3 Kebutuhan Dowtherm Pabrik Etilen

No	Alat	Kebutuhan Dowtherm (Kg/h)
1	Kondenser Parsial (CS-01)	212971.8203
Total Kebutuhan Dowtherm		212971.8203

Dengan asumsi factor keamanan dan penyerapan panas dimasing-masing alat, setiap *dowtherm* yang dialirkan ke unit proses dilebihkan 10%, sehingga :

$$Total + FK\ 10\% = 1.1\ \times 212971.8203 = 234269.0023\,\frac{Kg}{jam}$$

d. Kebutuhan Air Domestik

- Air Sanitasi

  Kebutuhan sir per orang = 120 L/hari

  Jumlah karyawan = 146

  Total = 17520 L/hari

- Air Laboratorium

  Kebutuhan air lab = 95 -200 L/hari/staff

  Permisalan = 95 L/hari/staff

  Jumlah staff RnD & QC = 20 staff

  Total = 1900 L/hari

- *Hydrant*

  Kebutuhan = 20 L/jam

  Total = 480 L/hari

$$\text{Total Air Domestik} = 17520 + 1900 + 480 = 19900\ \text{L/hari}$$

Atau setara dengan 19.7800 m³/hari atau 0.8242 m³/jam.

Kondisi Air Domestik yang digunakan :

Temperature = 30ºC

ρ = 1022.8753 Kg/m³

Massa air domestik = 848.1341 Kg/jam

Asumsi factor keamanan sebesar 10%, maka jumlah air domestik yang dibutuhkan adalah:

$$Total + FK\ 10\% = 1.1 \times 848.1341 = 932.9475\ Kg/jam$$

Maka kebutuhan Air Keseluruhan adalah ;

Tabel 3.4 Kebutuhan Keseluruhan Air Pabrik Etilen

No	Kebuthan Air	Star Up (Kg/jam)	Kontinyu (Kg/jam)
1	Air Umpan boiler	57096.8669	-
2	*Make Up boiler*	-	5709.6867
3	Air Pendingin	27288.4649	-
4	*Make Up cooling tower*	-	9521.4959
5	Air Pendingin (*Dowtherm*)	234269.0023	-
6	*Make up* air pendingin	-	234269.0023
7	Air Domestik	932.9475	932.9475
Total		319587.2816	250433.1324

Atau setara dengan 269.3158 m³/jam atau 275476.446 kg/jam.

2. Unit Penyedia Listrik

Penyediaan sarana listrik dibagi menjadi 2 golongan yaitu daya proses dan daya penunjang.

a. Daya Proses

Berikut rincian detail daya proses:

Tabel 3.5 Daya Proses Pabrik Etilen

Nama Alat/Kode Alat	Fungsi	Daya (HP)
Pompa Utilitas-01 (PU-01)	Memompa air dari badan sungai ke screen	2.95
Pompa Utilitas-02 (PU-02)	Memompa air dari screen ke reservoir	2.95

Nama Alat/Kode Alat	Fungsi	Daya (HP)
Pompa Utilitas-03 (PU-03)	Memompa air dari reservoir ke bak pengadukan cepat (BPC)	2.95
Pompa Utilitas-04 (PU-04)	Memompa air dari bak pengaduk cepat ke bak pengendap I	2.95
Pompa Utilitas-05 (PU-05)	Memompa air dari bak pengendap I ke bak pengendap II	2.95
Pompa Utilitas-06 (PU-06)	Memompa air dari bak pengendap II ke tangki filtrasi	2.95
Pompa Utilitas-07 (PU-07)	Memompa air dari tangki filtrasi ke bak air bersih	2.95
Pompa Utilitas-08 (PU-08)	Memompa air dari bak air bersih ke tangki demineralisasi	2.95
Pompa Utilitas-09 (PU-09)	Memompa air dari bak bersih ke bak penumpang cooling tower	2.01
Pompa Utilitas-10 (PU-10)	Memompa air dari tangki demineralisasi ke bak umpan boiler	2.01
Pompa Utilitas-11 (PU-11)	Memompa air dari bak air bersih ke bak domestik	0.54
Pompa Utilitas-12 (PU-12)	Memompa air dari bak air bersih ke bak limbah	2.01
Fan cooling tower	Mendinginkan fluida panas menjadi fluida dingin dengan prinsip "dehumineralisasi"	1.33
Total		31,4971

b. Daya Penunjang

- Peralatan bengkel

  Dalam suatu pabrik diperlukan fasilitas pemeliharaan dan perbaikan pabrik.

  Estimasi daya listrik yang dibutuhkan = 50 Kw/hari

- Instrumentasi

  Alat instrumentasi yang digunakan berupa alat control dan alat pendeteksi.

  Estimasi daya listrik yang diburuhkan = 15 Kw/hari

- Penerangan lampu jalan, pendingin ruangan dan perkantoran

  Alat penerangan yang dibutuhkan untuk pabrik, kantor dan lingkungan sekitar pabrik.

  Estimasi daya listrik yang dibutuhkan = 45 Kw/hari

- Alat kantor seperti computer, intercom, pengeras suara dan lainnya

  Estimasi daya listrik yang dibutuhkan = 30 Kw/hari

  $Total\ Daya\ Penunjang = 50 + 15 + 45 + 30 = 140\ KW/hari$

Atau setara dengan 7.8226 HP/jam.

Kebutuhan Listrik Keseluruhan = Daya Penunjang + Daya Proses + Daya Utilitas

= 7.8226 + 8086.9810 + 31.4971

= 8126.3007 HP/jam

Asumsi factor keamanan sebesar 10%, maka jumlah air domestik yang dibutuhkan adalah:

$$Total + FK\ 10\% = 1.1\ \times 8126.3007 = 9751.5609\frac{HP}{jam}$$

Atau setara dengan 7271.7389 KWH.

3. Unit Penyedia Bahan Bakar

Penyediaan bahan bakar untuk generator, boiler dan kendaraan

Bahan Bakar = Solar

Heating value = 19200 Btu/lb

ρ = 850 kg/m³

a. Penyediaan Solar untuk Generator

Diasumsikan:

- Efisiensi pembakaran solar pada generator = 85 %
- Pemadaman listrik = 1 jam/hari
  - Generator yang digunakan = 7271.7389 KWH ~ 24812195.0051 Btu/jam

$$Massa\ Solar = \frac{Kebutuhan\ Listrik}{h\ \times Hv}$$

Keterangan:

Hv = *Heating value* bio solar

h = Efisiensi pembakaran

$$Massa\ Solar = \frac{24812195.0051}{(85\%\ \times 19200)} = 1520.3551\frac{lb}{jam} = 689.5034\frac{kg}{jam}$$

Jika per hari perkiraan pemadaman 1 jam maka, 15858.5791 kg/hari. Oleh karena itu kebutuhan solar untuk generator per hari sebesar 19030.2950 kg/hari ~ 22388.5823 l/hari.

b. Penyediaan Solar untuk Boiler

Kebutuhan solar untuk boiler = 7482.3564 kg/jam

Kebutuhan solar per hari = 179576.5539 kg/hari

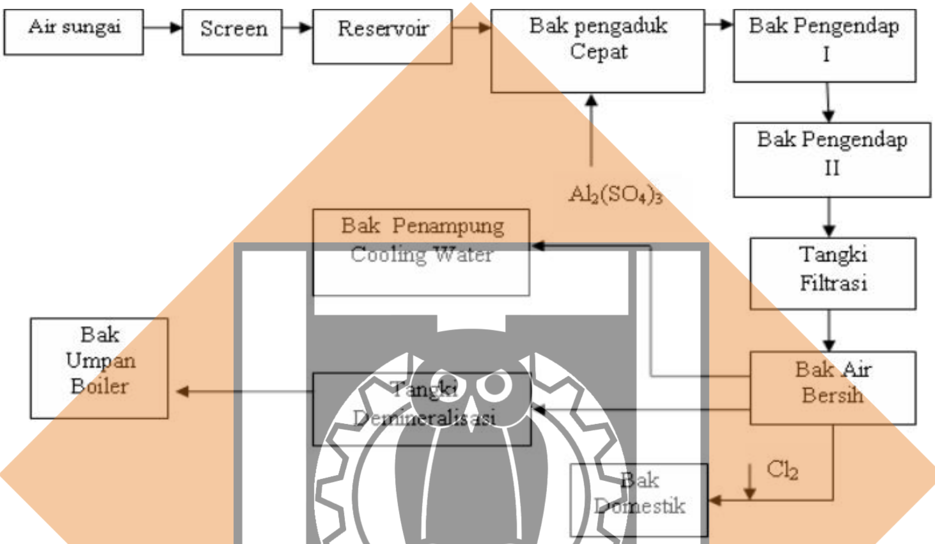
Total kebutuhan bahan bakar solar = 195435.1331 kg/hari ~ 229923.6860 l/hari.

c. Penyediaan Solar untuk Kendaraan

Kebutuhan solar untuk kendaraan = 60 l/hari

$$Total\ kebutuhan\ Solar = Kebutuhan\ untuk\ generator + boiler + kendaraan$$

$$Total\ kebutuhan\ Solar = 22388.5823 + 229923.6860 + 60.0000 = 252372.2683\ l/hari \sim 211344.7122 \text{ kg/hari.}$$



GAMBAR 3.10 SISTEM PENGOLAHAN AIR

3.2 Tata Letak Alat

Penyusunan tata letak alat sangat penting dan memberikan dampak pada jalannya operasi pabrik yang efiien dan meminimalkan biaya pembangunan. Tata letak alat ini sangat berhubungan erat dengan perancangan bangunan pabrik yang bertujuan untuk:

1. Alur proses produksi berjalan lancar dan efisien
2. Agar pekerja dapat bekerja dengan leluasa, aman, selamat dan nyaman.

Terdapat dua macam penyusunan tata letak alat proses, yaitu:

1. Tata Letak Produk atau Garis (*Product Lay Out/Line Lay Out)*

   Dalam tata letak ini susunan mesin/alat berdasarkan urutan proses produksi. Umumnya digunakan pada pabrik yang memproduksi suatu produk dalam skala besar dan kontinyu dimana bahan baku disuplai di satu ujung jalur dan bergerak dari satu operasi ke operasi berikutnya dengan cukup cepat dengan pekerjaan minimum dalam proses, penyimpanan dan penanganan bahan.

2. Tata Letak Proses atau Fungsional *(Process/Functional Lay Out)*

   Dalam tata letak ini peralatan diletakan berdasarkan proses atau fungsi yang sama, peralatan tidak diatur sesuai dengan urutan operasi tetapi diukur dengan sifat atau jenis operasi pada ruang tertentu. Tata letak ini umumnya cocok untuk pekerjaan yang tidak berulang.

Pada saat merancang pabrik etilen, tata letak alat proses yang digunakan menjadi tata letak produk atau lini (product layout/line layout). Tata letak ini dipilih dengan mempertimbangkan fakta bahwa terdapat tiga tahap produksi yang berurutan (tahap persiapan - tahap produksi etilen - pemurnian etilen) dan fokus hanya pada satu produk utama (etilen) dimana artinya tidak perlu mengelompokkan tata letak pabrik tergantung fungsinya.

Pembangunan ekonomi dan pengoperasian pabrik pengolahan yang efisien sangatlah penting. Faktor-faktor berikut dipertimbangkan ketika merancang penyusunan tata letak alat proses:

1. Pertimbangan ekonomis

   Biaya konstruksi dapat dioptimalkan dengan menempatkan peralatan yang membuat sistem perpipaan antar peralatan proses sesingkat mungkin, sehingga mengurangi daya tekan alat terhadap bahan/campuran dimana nantinya akan mengurangi biaya variable.

2. Kemudahan operasional

   Letak tiap alat diusahakan agar dapat memberikan keleluasan bergerak pada para pekerja dalam melaksanakan aktifitas produksi.

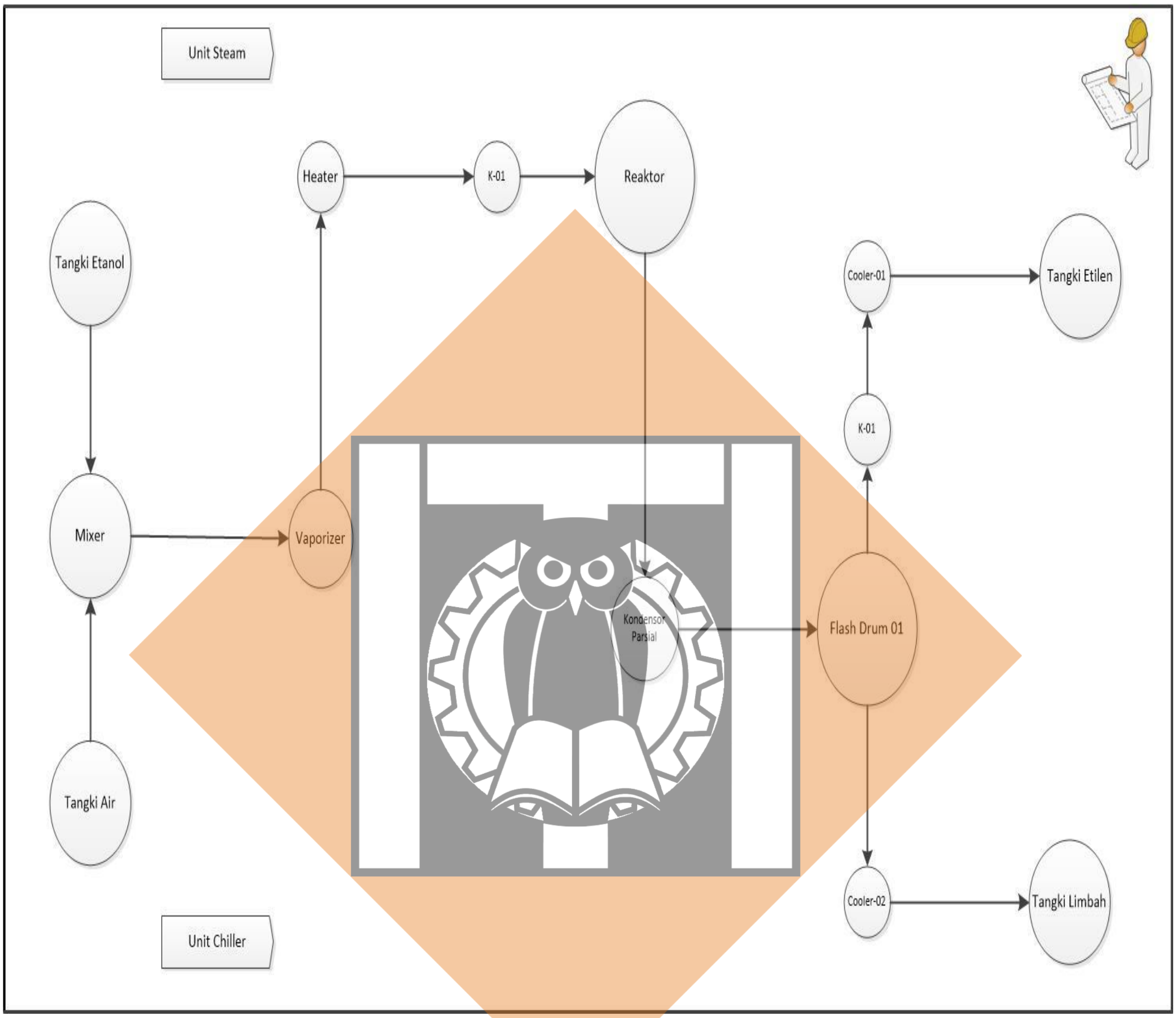
3. Kemudahan pemeliharaan

   Kemudahan pemeliharaan merupakan pertimbangan penting ketika menemukan peralatan proses. Pemeliharaan peralatan proses memastikan bahwa peralatan berfungsi dengan baik dan memiliki masa pakai yang lama. Saat mencari peralatan proses, penting untuk memastikan ruang yang cukup untuk pemeliharaan dan perbaikan.

4. Keamanan operasional

   Keselamatan pekerja dan operasional pabrik merupakan faktor yang sangat penting. Peralatan bersuhu tinggi diisolasi dengan insulasi untuk melindungi pekerja dari kerusakan.

Terdapat pintu keluar darurat sehingga pekerja dapat dengan mudah melindungi diri jika terjadi situasi yang tidak terduga.



GAMBAR 3.11 TATA LETAK ALAT

3.3 Tata Letak Pabrik

Tata letak pabrik merupakan bagian dari perancangan pendirian pabrik yang perlu diperhatikan. Tata letak pabrik mengatur susunan letak bangunan untuk area proses, area perlengkapan, area kantor, area gudang, area utilitas dan area fasilitas lain demi kelancaran proses produksi dengan baik dan efisien serta menjaga keselamatan pekerja dan menjaga keamanan dari pabrik tersebut. Aliran proses dan aktifitas menjadi dasar pertimbangan pengaturan bangunan pabrik, sehingga proses dapat berjalan dengan efektif, aman dan kontinyu.

Berikut merupakan factor-faktor yang menjadi perhatian dan pertimbangan dalam menentukan tata letak pabrik *(Plant Lay Out)* antara lain :

a. Kemudahan dalam operai danproses yang disesuaikan dengan kemudahan memelihara peralatan derta kemudahan control hasil produksi.
b. Distribusi utilitas yang tepat dan ekonomis.
c. Kebebasan bergerak yang cukup di antara alat proses dan alat yang menyimpan bahan berbahaya dan atau alat penunjang lainnya.
d. Antisipasi adanya kemungkinan perluasan pabrik.
e. Penggunaan ruang yang efektif dan ekonomis.

Berdasarkan faktor pertimbangan di atas, pengaturan tata letak pabrik etilen, untuk penempatan bangunan dalam kawasan pabrik dirancang sebagai berikut:

1. Area proses

   Area tempat berlangsungnya proses produksi etilen ditempatkan pada area yang memudahkan distribusi suplai bahan baku dari tangki penyimpanan dan pengiriman produk ke tangki penyimpanan produk serta tempat yang memudahkan pemantauan dan perbaikan alat-alat proses.
2. Area penyimpanan

   Area tempat penyimpanan bahan baku dan produk yang dihasilkan diletakkan pada tempat yang mudah dijangkau alat pengangkut.
3. Area pemeliharaan dan perawatan pabrik

Area perbengkelan untuk melakukan perawatan dan perbaikan sesuati kebutuhan pabrik.

4. Area utilitas/sarana Penunjang

   Area alat-alat penunjang produksi berupa tempat penyediaan air, tenaga listrik, pemanas dan sarana pengolahan limbah.

5. Area administrasi dan perkantoran

   Area pusat kegiatan administrasi pabrik untuk urusan-urusan dengan pihak - pihak luar maupun dalam.

6. Area laboratorium

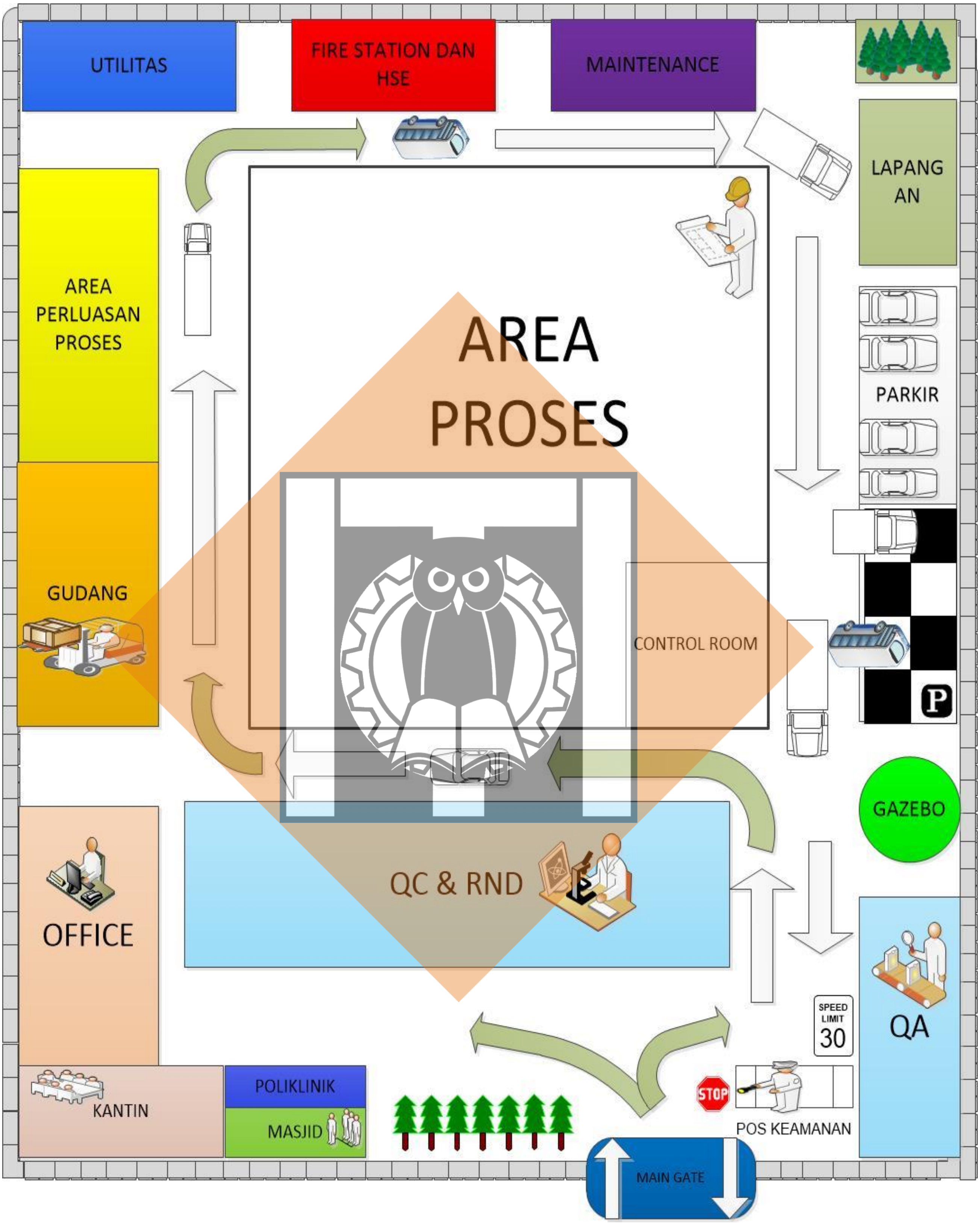
   Area *Quality Control* (QC) terhadap produk maupun bahan baku serta tempat penelitian dan pengembangan (R&D).

7. Fasilitas umum

   Area yang terdiri dari kantin, klinik pengobatan, lapangan parkir serta tempat ibadah. Penempatan fasilitas umum diletakkan sedemikian rupa agar seluruh pekerja dapat memanfaatkannya.

8. Area Perluasan

   Area yang dipersiapkan untuk perluasan pabrik dimasa yang akan dating dikarenakan peningkatan kapasitas produksi akibat adanya peningkatan produk.



GAMBAR 3.12 TATA LETAK PABRIK

BAB IV
SPESIFIKASI ALAT

4.1 Peralatan Proses

Peralatan proses yang terdapat pada pabrik etilen terdiri dari tangki penyimpanan bahan baku cair maupun gas, reaktor fixed bed multitube, mixer pencampuran, vaporizer untuk menguapkan dan kompresor untuk menaikan tekanan. Alat penukar panas yang digunakan berupa *heater dan cooler*. Semua peralatan dirancang sesuai kebutuhan.

4.1.1 Tangki Penyimpanan Etanol TP-01

a. Fungsi : Tempat penyimpanan bahan baku Etanol
b. Kapasitas : 17802.15 Kg/jam
c. Kondisi Operasi
   Temperature : 30 ^oC
   Tekanan : 1 atm
d. Dimensi
   Volume (Vt 20%) : 334.23 m^3
   Diameter
   - ID tangki : 5.04 m
   - OD tangki : 6.10 m

   Tinggi : 13.33 m
e. Material Konstruksi : *Stainless steel SA 167 type 304*
f. Harga Satuan : Rp 6,100,113,028
g. Jumlah Alat : 6 unit

4.1.2 Tangki Penyimpanan Air TP-02

a. Fungsi : Tempat penyimpanan bahan baku air
b. Kapasitas : 24693.96 Kg/jam
c. Kondisi Operasi
   Temerature : 30^oC
   Tekanan : 1 atm
d. Dimensi

Volume (Vt 20%) : 231.76 m^3

Diameter

- ID tangki : 5.44 m
- OD tangki : 5.49 m

Tinggi : 11.78 m

e. Material Konstruksi : *Stainless steel SA 167 Type 304*

f. Harga Satuan : Rp 4,885,049,864

g. Jumlah Alat : 3 unit

4.1.3 Tangki Penyimpanan TP-03

a. Fungsi : Tempat penyimpanan produk Etilen

b. Kapasitas : 10732.32 kg/jam

c. Kondisi Operasi

Temperature : 30^oC

Tekanan : 27 atm

d. Dimensi

Volume gas : 51.35 m^3

Diameter :

- ID tangki : 4.61 m
- OD tangki : 4.69 m

e. Material Konstruksi : *Stainless steel SA-204 grade A*

f. Harga Satuan : Rp 47,527,981

g. Jumlah Alat : 1 unit

4.1.4 Mixer M-01

a. Fungsi : Tempat pencampuran bahan baku Etanol dan Air

b. Kapasitas : 42496.12 Kg/jam

c. Kondisi Operasi

Temperature : 30^oC

Tekanan : 1 atm

d. Dimensi

Volume : 52.98 m^3

Diameter :

- ID tangki : 3.30 m
- OD tangki : 3.28 m

Jenis pengaduk : Propeller turbin dengan 6 *blades*

Diameter pengaduk (Di) : 1.09 m

Lebar Pengaduk (W) : 0.11 m

Panjang poros pengaduk : 6.91 m

Diameter poros : 1.09 m

e. Tinggi : 7.73 m
f. Daya : 7.61 HP
g. Material Konstruksi : *Stainless steel SA-167 grade 11 M type 316*
h. Harga Satuan : Rp 49,594,415
i. Jumlah Alat : 1 unit

4.1.5 Reaktor Fixed Bed Multitube R-01

a. Fungsi : Tempat terjadinya reaksi dehidrasi etanol dan pembentukan produk etilen
b. Fasa : Gas
c. Kondisi Operasi
   Temperature : 470°C
   Tekanan : 6 atm
d. Material Konstruksi : *Stainless steel SA-167 Grade 3 Type 304*
e. Jumlah Alat : 1 unit
f. Dimensi
   - Bagian Tube
     NPS : 3 inch
     Panjang : 6.10 m
     ID : 0.94 m
     OD : 1.07 m
     Jumlah : 350 buah

- Bagian Alat

OD : 1.98 m

ID : 1.94 m

- Bagian Nozzle

Tutup Atas

OD : 0.22 m

ID : 0.20 m

Silinder

OD : 0.005 m

ID : 0.013 m

Tutup Bawah

OD : 0.22 m

ID : 0.20 m

g. Tinggi Reaktor : 9.60 m

h. Harga Satuan : Rp 48,974,484,673

4.1.6 Flash Drum FD-01

a. Fungsi : Memisahkan Etilen dengan Etanol dan air

b. Fasa : Cair-Gas

c. Kondisi Operasi

Temperature : 80°C

Tekanan : 6 atm

d. Material Konstruksi : *Stainless steel SA-240 Grade M Type 316*

e. Jumlah Alat : 1 unit

f. Volume : 187.96 m^3

g. Dimensi

OD : 4.57 m

ID : 4.53 m

h. Tinggi Total :14.74 m

i. Harga Satuan : Rp 34,724,356,137

4.1.7 Kompresor K-01

a. Fungsi : Menaikan tekanan dari 1 atm menjadi 6 atm

b. Jenis : *Reciprocate*
c. Bahan Konstruksi : *Carbon steel, SA-285*
d. Daya : 7300 HP
e. Jumlah Alat : 1 unit
f. Harga Satuan : Rp 92,299,338,920

4.1.8 Kompresor K-02

a. Fungsi : Menaikan tekanan dari 6 atm menjadi 27 atm
b. Jenis : *Reciprocate*
c. Material Konstruksi : *Carbon steel, SA-285*
d. Daya : 730 HP
e. Jumlah Alat : 1 unit
f. Harga Satuan : Rp 12,090,705,056

4.1.9 Blower BL-01

a. Fungsi : Mengalirkan gas keluaran Vaporizer (VP-01)
b. Jenis : *Centrifugal Multiblade Backward Curved Blower*
c. Material Konstruksi : *Carbon steel, SA-285*
d. Daya : 1 HP
e. Jumlah Alat : 1 unit
f. Harga Satuan : Rp 1,105,542,165

4.1.10 Blower BL-01

a. Fungsi : Mengalirakan produk atas (gas) keluaran Flash Drum (FD-01)
b. Jenis : *Centrifugal Multiblade Backward Curved Blower*
c. Material Konstruksi : *Carbon steel, SA-285*
d. Daya : 1 HP
e. Jumlah Alat : 1 unit
f. Harga Satuan : Rp 1,105,542,165

4.1.11 Vaporizer VP-01

a. Fungsi : Memanaskan serta menguapkan bahan baku menjadi bahan baku etanol 42% (fase gas) dari suhu 30˚C menjadi 126˚C
b. Jenis : *Shell and Tube heat exchanger*

c. Kondisi Operasi

Temperature : 126ºC

Tekanan : 1 atm

d. Material Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

e. Jumlah Alat : 1 unit

f. Dimensi

Shell Side : Fluida dingin (Etanol - Air)

- Laju alir massa : 93687.7869 lb/jam
- T in : 86 ºF
- T out : 258 ºF
- ID : 32.64 inch
- Jarak *baffle* : 8.16 inch
- Jumlah *pass* : 2 *passes*

Tube Side : Fluida panas (Steam)

- Laju alir massa : 58109.1703lb/jam
- T in : 896 ºF
- T out : 878 ºF
- OD : 1.00 inch
- ID : 0.67 inch
- BWG : 8
- Panjang : 6 ft
- Jumlah : 520 *tube*
- Pt : 1.25 in
- Bentuk *pitch* : *square*
- Jumlah *pass* : 2 *passes*

g. Harga Satuan : Rp 1,450,636,635

4.1.12 Heater H-01

a. Fungsi : Memanaskan bahan baku etanol 42% (fase gas) dari suhu 126˚C menjadi 470˚C sebelum masuk reaktor

b. Jenis : *Shell and Tube heat exchanger*

c. Kondisi Operasi

Temperature : 470ºC

Tekanan : 1 atm

d. Material Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

e. Jumlah Alat : 1 unit

f. Dimensi

Shell Side : Fluida dingin (Etanol - Air)

- Laju alir massa : 93687.7869 lb/jam
- T in : 258 ºF
- T out : 878 ºF
- ID : 41.04 inch
- Jarak *baffle* : 10.26 inch
- Jumlah *pass* : 8 *passes*

Tube Side : Fluida panas (Steam)

- Laju alir massa : 19545.2490 lb/jam
- T in : 896 ºF
- T out : 878 ºF
- OD : 1.00 inch
- ID : 0.67 inch
- BWG : 8
- Panjang : 6 ft
- Jumlah : 790 *tube*
- Pt : 1.25 inch
- Bentuk *pitch* : *square*
- Jumlah *pass* : 8 *passes*

g. Harga Satuan : Rp 2,394,996,951

4.1.13 Condenser Parsial CD-01

a. Fungsi : Mendinginkan keluaran dari reaktor fix bed sebelum masuknya ke flash drum dimana ada dua fasa yaitu fasa cair (etanol,air) dan gas (etilen) dengan suhu 80˚C

b. Jenis : *Shell and Tube heat exchanger*

c. Kondisi Operasi

Temperature : 80°C

Tekanan : 6 atm

d. Material Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

e. Jumlah Alat : 1 unit

f. Dimensi

Shell Side : Fluida dingin (Etanol - Air)

- Laju alir massa : 93687.7869 lb/jam
- T in : 158 °F
- T out : 176 °F
- ID : 22.66 inch
- Jarak *baffle* : 5.66 inch
- Jumlah *pass* : 2 *passes*

Tube Side : Fluida panas (Etilen)

- Laju alir massa : 469521.9344 lb/jam
- T in : 878 °F
- T out : 176 °F
- OD : 0.75 inch
- ID : 0.53 inch
- BWG : 12
- Panjang : 24 ft
- Jumlah : 400 *tube*
- Pt : 1.25 inch
- Bentuk *pitch* : *tringular*
- Jumlah *pass* : 2 *passes*

g. Harga Satuan : Rp 4,885,049,864

4.1.14 Cooler C-01

a. Fungsi : Mendinginkan produk atas (etilen) dari flash drum sebelum) masuknya ke tangki penyimpanan etilen (bola)

b. Jenis : *Double Pipe Heat Exchanger*

c. Kondisi Operasi

Temperature : 30°C

Tekanan : 27 atm

d. Material Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

e. Jumlah Alat : 1 unit

f. Dimensi

- Luas permukaan aktual : 195.6233 ft^2
- Panjang total HE : 182.2310 m
- Jumlah *hairpin* : 15 set
- OD *inner* : 0.1025 ft
- ID *inner* : 0.0846 ft
- OD *annulus* : 0.1640 ft
- ID *annulus* : 0.1462 ft

g. Evaluasi Desian

- Koefisien transfer panas bersih : 80.0328 Btu/jam.ft2.˚F
- ΔP *inner* : 0.0000019571 psi
- ΔP *annulus* : 0.2182 psi

h. Harga Satuan : Rp 526,940,658

4.1.15 Cooler C-02

a. Fungsi : Mendinginkan produk bawah (etanol-air) dari flash drum sebelum masuknya ke tangki limbah

b. Jenis : *Shell and Tube heat exchanger*

c. Kondisi Operasi

Temperature : 30°C

Tekanan : 6 atm

d. Material Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

e. Jumlah Alat : 1 unit

f. Dimensi

Shell Side : Fluida dingin (Air Pendingin)

- Laju alir massa : 50454.5178 lb/jam
- T in : 50 °F
- T out : 77 °F

- ID : 46.1100 inch
- Jarak *baffle* : 11.5275 inch
- Jumlah *pass* : 8 *passes*

Tube Side : Fluida panas (Etanol-Air)

- Laju alir massa : 70027.0924 lb/jam
- T in : 176 °F
- T out : 86 °F
- OD : 1.50 inch
- ID : 1.40 inch
- BWG : 18
- Panjang : 6 ft
- Jumlah : 950 *tube*
- Pt : 1.875 inch
- Bentuk *pitch* : *tringular*
- Jumlah *pass* : 8 *passes*

g. Harga Satuan : Rp 609,598,016

4.1.16 Pompa P-01

a. Fungsi : Mengalirkan Etanol 99% menuju MP-01

b. Jenis : *Centrifugal pump horizontal*

c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

d. Kapasitas : 110.1568 gpm

e. Efisiensi Pompa : 68%

f. Dimensi

Suction

NPS : 4 inch

Schedule : 40

Discharge

NPS : 3.5 inch

Schedule : 40

g. Daya : 10.0575 HP

h. NPSHA : 6.004 m

i. NPSHR : 0.2141 m
j. Jumlah : 2 unit, 1 unit *standby*
k. Harga Satuan : Rp 212,842,697

4.1.17 Pompa P-02

a. Fungsi : Mengalirkan air menuju MP-01
b. Jenis : *Centrifuge pump horizontal*
c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*
d. Kapasitas : 116.9141 gpm
e. Efisiensi Pompa : 70%
f. Dimensi
   Suction
   NPS : 4 inch
   Schedule : 40
   Discharge
   NPS : 3.5 inch
   Schedule : 40
g. Daya : 10.0575 HP
h. NPSHA : 6.004 m
i. NPSHR : 2.461 m
j. Jumlah : 2 unit, 1 unit *standby*
k. Harga Satuan : Rp 212,842,697

4.1.18 Pompa P-03

a. Fungsi : Mengalirkan campuran etanol dan air menuju VP-01
b. Jenis : *Centrifugal pump horizontal*
c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*
d. Kapasitas : 227.07 gpm
e. Efisiensi Pompa : 78%
f. Dimensi
   Suction
   NPS : 6 inch
   Schedule : 40

Discharge

NPS : 5 inch

Schedule : 40

g. Daya : 20.115 HP

h. NPSHA : 6.004 m

i. NPSHR : 3.831 m

j. Jumlah : 2 unit, 1 unit *standby*

k. Harga Satuan : Rp 260,370,678

4.1.19 Pompa P-04

a. Fungsi : Mengalirkan buangan dari FD-01 ke tangka pembuangan

b. Jenis : *Centrifugal pump horizontal*

c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

d. Kapasitas : 150.6451 gpm

e. Efisiensi Pompa : 78%

f. Dimensi

Suction

NPS : 5 inch

Schedule : 40

Discharge

NPS : 5 inch

Schedule : 40

g. Daya : 14.751 HP

h. NPSH A : 6.004 m

i. NPSH R : 2.914 m

j. Jumlah : 2 unit, 1 unit *standby*

k. Harga Satuan : Rp 260,370,678

4.2 Peralatan Utilitas

4.2.1 Screen

a. Fungsi : Menyaring benda-benda pengotor yang berukuran besar yang terbawa aliran sungai

b. Ukuran screen : 2 m x 3 m

c. Bahan Konstruksi : Besi

d. Ukuran Lubang : 1 cm x 1 cm

e. Harga Satuan : Rp 4,027,479,773

4.2.2 Reservoir

a. Fungsi : Menampung air sungai yang keluar dari screen

b. Bentuk : Persegi panjang

c. Bahan Konstruksi : Beton

d. Jumlah : 1 unit

e. Waktu Tinggal : 12 jam

f. Dimensi bak

Volume bak : 4380.6378 m^3

Panjang : 27.0137 m

Lebar : 18.0091 m

Tinggi : 9.0046 m

g. Harga Satuan : Rp 621,795,400

4.2.3 Bak Pengadukan Cepat (BPC)

a. Fungsi : Menggumpalkan partikel-partikel pengotor yang ada di dalam air sungai dengan penambahan koagulan ($Al_2(SO_4)_3$)

b. Jenis : Tangki silinder vertikal

c. Bahan Konstruksi : *Stainless Steel SA-240 Grade S Type 304*

d. Jumlah : 1 unit

e. Waktu Tinggal : 40 menit

f. Dimensi

OD : 5.4549 m

ID : 5.4295 m

Ketebalan : 0.0127 m

Tinggi : 10.8590 m

g. Harga Satuan : Rp 3,802,238,472

4.2.4 Bak Pengendap I

a. Fungsi : Mengendapkangumpalan - gumpalan yang lebih besar dari bak pengadukan cepat

b. Jenis : Tangki silinder vertical
c. Bahan Konstruksi : *Stainless Steel SA-240 Grade S Type 304*
d. Jumlah : 1 unit
e. Waktu Tinggal : 2 jam
f. Dimensi
   OD : 7.8320 m
   ID : 7.8307 m
   Ketebalan : 0.0254 m
   Tinggi : 15.6614 m
g. Harga Satuan : Rp 221,108,433

4.2.5 Bak Pengendap II

a. Fungsi : Mengendapkan flok-flok yang lebih halus partikel partikelnya, yang tidak dapat terendapkan pada bak sebelumnya
b. Bentuk : Bak persegi empat
c. Bahan Konstruksi : Beton
d.
e. Jumlah : 1 unit
f. Waktu Tinggal : 4 jam
g. Dimensi bak
   Volume bak : 1432.3720 m^3
   Panjang : 30.6574 m
   Lebar : 15.3278 m
   Tinggi : 3.0480 m
h. Harga Satuan : Rp 1,935,689,700

4.2.6 Tangki Filtrasi *(sand filter)*

a. Fungsi : Untuk menyaring partikel partikel halus yang tersisa dalam air yang berasal dari bak pengendap II
b. Jenis : Tangki silinder vertical
c. Bahan Konstruksi : *Stainless Steel SA-240 Grade S Type 304*
d. Jumlah : 1 unit
e. Media Penyaring : Pasir dan kerikil

f. Dimensi

OD : 0.4977 m

ID : 0.4974 m

Ketebalan : 0.0048 m

Tinggi : 0.9949 m

g. Harga Satuan : Rp 398,821,753

4.2.7 Bak Penampung Air Bersih

a. Fungsi : Menampung air yang keluar dari bak flitrasi

b. Bentuk : Empat persegi panjang

c. Bahan Konstruksi : Beton

d. Jumlah : 1 unit

e. Waktu Tinggal : 6 jam

f. Dimensi bak

Volume bak : 1939.0736 m^3

Panjang : 20.5874 m

Lebar : 13.7250 m

Tinggi : 6.8625 m

g. Harga Satuan : Rp 101,549,500

4.2.8 Tangki Demineralisasi *(ion exchanger)*

a. Fungsi : Untuk menghilangkan kesadahan air dan kandungan mineral dalam air dengan menggunakan resin penukar ion

b. Jenis : Tangki silinder vertical

c. Bahan Konstruksi : *Stainless Steel SA-240 Grade S Type 304*

d. Jumlah : 1 unit

e. Media : Resin sintesis

f. Tipe Resin : *Mixed cation and strong base anion (Chemical equivalent mixture)*

g. Kapasitas : 6.2807 m^3/jam

h. Dimensi

OD : 0.2810 m

ID : 0.2715 m

Ketebalan : 0.0048 m

Tinggi : 2,4000 m

i. Harga Satuan : Rp 398,821,753

4.2.9 Bak Umpan Boiler

a. Fungsi : Untuk menampung kondensat steam untuk *feed* boiler

b. Bentuk : Empat persegi panjang

c. Bahan Konstruksi : Beton

d. Jumlah : 1 unit

e. Waktu Tinggal : 1 jam

f. Dimensi bak

Volume bak : 66.9840 m^3

Panjang : 6.7049 m

Lebar : 4.4700 m

Tinggi : 2.2350 m

g. Harga Satuan : Rp 101,549,500

4.2.10 Bak Umpan Cooling Tower

a. Fungsi : Untuk menampung air sebagai *feed cooling tower*

b. Bentuk : Empat persegi panjang

c. Bahan Konstruksi : Beton

d. Jumlah : 1 unit

e. Waktu Tinggal : 1 jam

f. Dimensi bak

Volume bak : 11.1703 m^3

Panjang : 3.6906 m

Lebar : 2.4604 m

Tinggi : 1.2302 m

g. Harga Satuan : Rp 3,787,015,200

4.2.11 Bak Penampung Air Domestik

a. Fungsi : Untuk menampung air domestik

b. Bentuk : Empat persegi panjang

c. Bahan Konstruksi : Beton

d. Jumlah : 1 unit

e. Waktu Tinggal : 1 jam

f. Dimensi bak

Volume bak : 1.0945 m^3

Panjang : 1.7014 m

Lebar : 1.1343 m

Tinggi : 0.5671 m

g. Harga Satuan : Rp 3,749,520,000

4.2.12 Bak Penampung Limbah

a. Fungsi : Untuk menampung air limbah

b. Bentuk : Empat persegi panjang

c. Bahan Konstruksi : Beton

d. Jumlah : 1 unit

e. Waktu Tinggal : 12 jam

f. Dimensi bak

Volume bak : 83.8541 m^3

Panjang : 7.2262 m

Lebar : 4.8175 m

Tinggi : 2.4087 m

g. Harga Satuan : Rp 2,343,450,000

4.2.13 Spesifikasi Pompa PU-01

a. Fungsi : Memompa air dari badan sungai ke screen

b. Jenis : *Centrifugal pump horizontal*

c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

d. Kapasitas : 330,6491 m^3/jam ~ 1455,8045 gpm

e. Efisiensi Pompa : 82%

f. Dimensi

NPS : 8 inch

Schedule : 40

g. Daya : 2.95 HP

h. Jumlah : 1 unit, 1 unit *standby*

i. Harga Satuan : Rp 479,412,677

4.2.14 Spesifikasi Pompa PU-02

a. Fungsi : Memompa air dari screen ke reservoir

b. Jenis : *Centrifugal pump horizontal*

c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

d. Kapasitas : 330,6491 m^3/jam ~ 1455,8045 gpm

e. Efisiensi Pompa : 82%

f. Dimensi

NPS : 8 inch

Schedule : 40

g. Daya : 2.95 HP

h. Jumlah : 1 unit, 1 unit *standby*

i. Harga Satuan : Rp 479,412,677

4.2.15 Spesifikasi Pompa PU-03

a. Fungsi : Memompa air dari reservoir ke bak pengadukan cepat (BPC)

b. Jenis : *Centrifugal pump horizontal*

c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

d. Kapasitas : 330,6491 m^3/jam ~ 1455,8045 gpm

e. Efisiensi Pompa : 82%

f. Dimensi

NPS : 8 inch

Schedule : 40

g. Daya : 2.95 HP

h. Jumlah : 1 unit, 1 unit *standby*

i. Harga Satuan : Rp 479,412,677

4.2.16 Spesifikasi Pompa PU-04

a. Fungsi : Memompa air dari bak pengadukan cepat (BPC) ke bak pengendap I

b. Jenis : *Centrifugal pump horizontal*

c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

d. Kapasitas : 330,6491 m^3/jam ~ 1455,8045 gpm

e. Efisiensi Pompa : 82%

f. Dimensi

NPS : 8 inch

Schedule : 40

g. Daya : 2.95 HP

h. Jumlah : 1 unit, 1 unit *standby*

i. Harga Satuan : Rp 479,412,677

4.2.17 Spesifikasi Pompa PU-05

a. Fungsi : Memompa air dari pengendap I ke pengendap II

b. Jenis : *Centrifugal pump horizontal*

c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

d. Kapasitas : 330,6491 m^3/jam ~ 1455,8045 gpm

e. Efisiensi Pompa : 82%

f. Dimensi

NPS : 8 inch

Schedule : 40

g. Daya : 2.95 HP

h. Jumlah : 1 unit, 1 unit *standby*

i. Harga Satuan : Rp 479,412,677

4.2.18 Spesifikasi Pompa PU-06

a. Fungsi : Memompa air dari bak pengendap II ke tangki filtrasi

b. Jenis : *Centrifugal pump horizontal*

c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

d. Kapasitas : 330,6491 m^3/jam ~ 1455,8045 gpm

e. Efisiensi Pompa : 82%

f. Dimensi

NPS : 8 inch

Schedule : 40

g. Daya : 2.95 HP

h. Jumlah : 1 unit, 1 unit *standby*

i. Harga Satuan : Rp 479,412,677

4.2.19 Spesifikasi Pompa PU-07

a.	Fungsi	: Memompa air dari tangki filtrasi bak air bersih
b.	Jenis	: *Centrifugal pump horizontal*
c.	Bahan Konstruksi	: *Stainless steel SA-167 Grade 11 Type 316*
d.	Kapasitas	: 330,6491 m^3/jam ~ 1455,8045 gpm
e.	Efisiensi Pompa	: 82%
f.	Dimensi	
	NPS	: 8 inch
	Schedule	: 40
g.	Daya	: 2.95 HP
h.	Jumlah	: 1 unit, 1 unit *standby*
i.	Harga Satuan	: Rp 479,412,677

4.2.20 Spesifikasi Pompa PU-08

a.	Fungsi	: Memompa air bak air bersih ke tangki demineralisasi
b.	Jenis	: *Centrifugal pump horizontal*
c.	Bahan Konstruksi	: *Stainless steel SA-167 Grade 11 Type 316*
d.	Kapasitas	: 330,6491 m^3/jam ~ 1455,8045 gpm
e.	Efisiensi Pompa	: 54%
f.	Dimensi	
	NPS	: 8 inch
	Schedule	: 40
g.	Daya	: 2.95 HP
h.	Jumlah	: 1 unit, 1 unit *standby*
i.	Harga Satuan	: Rp 479,412,677

4.2.21 Spesifikasi Pompa PU-09

a.	Fungsi	: Memompa air dari bak air bersih ke bak penampung cooling tower
b.	Jenis	: *Centrifugal pump horizontal*
c.	Bahan Konstruksi	: *Stainless steel SA-167 Grade 11 Type 316*
d.	Kapasitas	: 9,3086 m^3/jam ~ 40.9844 gpm
e.	Efisiensi Pompa	: 81%

f. Dimensi

NPS : 8 inch

Schedule : 40

g. Daya : 2.01 HP

h. Jumlah : 1 unit, 1 unit *standby*

i. Harga Satuan : Rp 479,412,677

4.2.22 Spesifikasi Pompa PU-10

a. Fungsi : Memompa air dari tangki demineralisasi ke bak umpan boiler

b. Jenis : *Centrifugal pump horizontal*

c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

d. Kapasitas : 55.8200 m^3/jam ~ 245.7680 gpm

e. Efisiensi Pompa : 82%

f. Dimensi

NPS : 6 inch

Schedule : 40

g. Daya : 2.01 HP

h. Jumlah : 1 unit, 1 unit *standby*

i. Harga Satuan : Rp 402,954,621

4.2.23 Spesifikasi Pompa PU-11

a. Fungsi : Memompa air dari bak air bersih ke bak domestik

b. Jenis : *Centrifugal pump horizontal*

c. Bahan Konstruksi : *Stainless steel SA-167 Grade 11 Type 316*

d. Kapasitas : 0.9121 m^3/jam ~ 4,0158 gpm

e. Efisiensi Pompa :

f. Dimensi

NPS : 1.00 inch

Schedule : 40

g. Daya : 0.54 HP

h. Jumlah : 1 unit, 1 unit *standby*

i. Harga Satuan : Rp 134,318,207

4.2.24 Spesifikasi Pompa PU-12

a. Fungsi	: Memompa air dari bak air bersih ke bak limbah
b. Jenis	: *Centrifugal pump horizontal*
c. Bahan Konstruksi	: *Stainless steel SA-167 Grade 11 Type 316*
d. Kapasitas	: 244,8325 m^3/jam ~ 1077,9653 gpm
e. Efisiensi Pompa	: 82%
f. Dimensi	
NPS	: 8 inch
Schedule	: 40
g. Daya	: 2.01 HP
h. Jumlah	: 1 unit, 1 unit *standby*
i. Harga satuan	: Rp 479,412,677

BAB V

ASPEK KESELAMATAN, KESEHATAN KERJA DAN LINGKUNGAN

5.1 Deskripsi Singkat

Menurut Material Safety Data Sheet (MSDS), bahan baku pembuatan etilen adalah etanol, air, dan katalis *zeolit ZSM-5* serta adanya membran anorganik. Semua zat ini diberi label dengan kata sinyal "danger" yang menunjukkan bahwa zat tersebut harus digunakan, ditangani, dan diperlakukan dengan tepat dan hati-hati. Etanol mudah terbakar dan juga mengiritasi tubuh apabila terkena. Air berfungsi sebagai medium sendiri aman dalam penggunaannya selagi tidak ada ditambahkan bahan kandungan lainnya. Katalis *zeolit ZSM-5* bersifat mudah terbakar, iritan dan berbahaya jika terhirup. Membran yang sudah pernah di research orang lain mengandung bahan yang tidak berbahaya.

Produk etilen sendiri juga membawa sinyal kata "danger" dan bersifat iritasi pada kulit, mata, dan napas jika terhirup. Mengetahui simbol peringatan pada bahan mentah, aksesori, dan produk memungkinkan Anda menyesuaikan infrastruktur dan penanganan alat pelindung diri (APD) dengan cara yang tepat sasaran dan berdasarkan kebutuhan. Untuk melakukan aktivitas kerja yang aman dan nyaman, penting juga untuk memilih karyawan yang sesuai, kompeten, dan berminat serta memahami aspek keselamatan dan kesehatan kerja.

5.2 Pertimbangan Aspek Keselamatan Pabrik

Tabel 5.1 Identifikasi Hazard Berdasarkan MSDS

A. Identifikasi Hazard Bahan Kimia yang Digunakan pada Proses Berdasarkan MSDS								
	Hazard							**Pengelolaan dan APD yang digunakan**
	Explosive	***Flammable***	***Toxic***	***Corrosive***	***Irritant***	***Hazardous to Environment***	***Carcinogenic***	
Bahan Baku								

Etanol		√			√			• Untuk mata/wajah = Kacamata keselamatan dengan sisi-perisai sesuai dengan peralatan EN166 • Untuk kulit/tangan = Sarung tangan pelindung contoh KCL 898 Butoject@ (Kontak Penuh) • Untuk tubuh = Jas lengkap berupa flame retardant • Untuk pernapasan = Masker filter tipe A rekomendasikan
Air								*No Danger*
Zeolit ZSM-5		√	√		√			• Untuk mata/wajah = Kacamata pengaman sesuai dengan EN166 • Untuk kulit = Sarung tangan diperiksa sebelum digunakan. • Untuk tubuh = Jas Lab Kedap Air • Untuk pernapasan = Respirator partikep tipe P95 (AS)
Membran								No Danger
Produk								
Etilen	√	√	√		√			• *Untuk pernapasan = Respirator bersuplai penutup wajah sesuai standar US OSHA* • *Untuk mata = Kacamata perlindung percikan sesuai standar US OSHA* • *Untuk tangan = Sarung tangan tahan mekanis contoh Kevlar sesuai standar US OSHA*

								• *Untuk tubuh = Pelindung tubuh sesuai tugas dengan berpedoman standar US OSHA*

Tabel 5.2 Identifikasih Hazard Peralatan Proses

B. Identifikasi Hazard Peralatan Proses							
Peralatan	Tekanan	Temperature	Putaran Pengaduk	Elevasi	Kuantitas Bahan	Keterangan	Pengelolaan
Reaktor Fixed Bed Multitube (R-01)	√	√			√	P : 6 atm T : 470ºC	Reaktor dilengkapi dengan *Pressure relieve, Temperature Indicator controller* dan *Level Indicator Controller*
Tangki (T-01,02,03)		√			√	P : 1 atm T : 30ºC	Tangki dilengkapi dengan *Temperature Indicator*
Mixer (P-01)			√		√	P : 1 atm T : 30ºC	Mixer dilengkapi dengan *Level Indicator*
Vaporizer (VP-01)		√				P : 1 atm T : 126ºC	Vaporizer dilengkapi dengan *Temperature Indicator Controller*
Flash Drum (FD-01)		√				P : 6 atm T : 90ºC	Flash drum dilengkapi dengan *Pressure Indicator Controller*

Tabel 5. 3 Identifikasi Hazard Tata Letak dan Lokasi

C. Identifikasi Hazard Tata Letak Pabrik dan Lokasi Proses						
Peralatan	Hazard				Keterangan	Pengelolaan
	Ledakan	Kebakaran	Pelepasan Bahan	Operability & Maintenability		
Tata Letak Pabrik						
Letak Tangki Penyimpanan terhadap area proses	√	√	√			i. Tangki penyimpanan bahan baku ditempatkan di dekat dermaga untuk memudahkan proses loading/unloading. i. Gudang penyimpan produk diletakkan 500 m dari area proses supaya Ketika loading/unloading tidak membahayakan area proses
Vaporizer				√		i. Vaporizer berjenis DPHE diletakkan di pinggir area proses dengan ujung menghadap jalan /. Di samping HE (arah memanjang) diberi ruang selebar min 1,5 kali Panjang vaporizer untuk memberikan kecukupan *space* ketika pemisahan

						budle tube terhadap shell pada waktu *maintenance.*
Tangki Etilen			√	√	T = 30˚C; P = 27 atm	Tangki penyimpanan diberi tanggul dengan kapasitas tanggul minum 100% dari volume tangki. Tangki dilengkapi dengan water sprinkle. Tangki dilengkapi PRV yang dihubungkan dengan unit flare
Lokasi Proses						
Jarak antara area proses dengan jalan gedung kantor	√	√	√			Jarak antara proses produksi dengan kantor utama adalah 1 km, sesuai dengan Kemenperin No 40 Tahun 2016 dimana jarak permukiman penduduk dengan suatu industri minimal 2 km
Jarak antara area proses dengan jalan raya	√	√	√			
Jarak antara area proses dengan permukiman penduduk	√	√	√			

5.3 Pertimbangan Aspek Keselamatan dan Kesehatan Kerja

Tabel 5.4 Identifikasi Potensi Paparan Kimia

A. Identifikasi Potensi Paparan Kimia								
Jenis Paparan	*Hazard*						Keterangan	Pengelolaan
	Kanker	Kerusakan paru-paru	Kerusakan ginjal	Kerusakan organ tubuh	Mutasi gen	Iritasi		

Etanol						√	Berfungsi sebagai bahan baku utama	Operator menggunakan sarung tangan pelindung, masker dan baju pelindung
Zeolit ZSM-5				√	√	√	Sebagai katalis reaksi dehidrasi etanol	Operator menggunakan sarung tangan pelindung, masker dan baju pelindung
Produk								
Etilen				√	√	√	Produk	Operator menggunakan sarung tangan pelindung, masker dan baju pelindung serta bila terjadi kebocoran tangki mengikuti prosedur tanggap darurat

Tabel 5.5 Identifikasi Potensi Paparan Fisis

B. Identifikasi Potensi Paparan Fisis							
Jenis Paparan	*Hazard*					Keterangan	Pengelolaan
	Tuli	Kanker	ISPA	Sakit Kepala	Nyeri Otot		
Kebisingan	√					Mesin proses seperti motor,	Operator menggunakan *earplug*.

						pengaduk dan kompresor	
Panas				√	√	Area *boiler* penghasil *steam, heater sebagai alat pemanas dan reaktor sebagai tempat berlangsungnya reaksi*	Operator menggunakan *waerpack* yang sesuai seperti sarung tangan anti panas
Radiasi UV, radioaktif, gelombang elektromagnetik		√		√		Area dalam pabrik dan luar pabrik	Operator menggunakan *wearpack* yang sesuai dan kacamata antiradiasi
Debu		√	√	√		Area dalam pabrik dan luar pabrik	Operator menggunakan kacamata antidebu

5.4 Pertimbangan Aspek Lingkungan Pabrik

Tabel 5.6 Identifikasi Hazard Emisi Gas

A. Identifikasi *Hazard* Emisi Gas Yang diHasilkan dari Proses								
Emisi	Sumber	*Hazard*					Keterangan	Pengelolaan
		Toksik	Pemanasan Global	Pengikisan Ozon	Hujan Asam	Kerusakan Ekologi		
CO_2	*Boiler*		√					CO2 capture dilakukan dengan kolom scrubber, dengan larutan

								MDEA sebagai solvent penyerap
SO_2	*Boiler*				√	√		SO2 capture dilakukan dengan kolom scrubber, dengan larutan $Ca(OH)^2$ sebagai solvent penyerap

Tabel 5.7 Identifikasi Hazard Limbah Cair

B. Identifikasi *Hazard* Limbah Cair Yang diHasilkan dari Proses						
Limbah Cair	Sumber	*Hazard*			Keterangan	Pengelolaan
		Racun	Merusak ekosistem	Mencemari		
Air Limbah Proses	*Unit Vaporizer dan demineralisasi*					Adanya IPAL sebelum limbah dibuang ke badan air.
Limbah Domestik	*Unit demineralisasi*		√			

Tabel 5.8 Identifikasi Hazard Limbah Padat

C. Identifikasi *Hazard* Limbah Padat Yang diHasilkan dari Proses						
Limbah Padat	Sumber	*Hazard*			Keterangan	Pengelolaan
		Racun	Merusak ekosistem	Mencemari		
Katalis padat	Reaktor					Metode yang bisa digunakan pada pengurangan limbah katalis padat diantaranya regenerasi, pencucian, pengolahan dan daur ualng.

BAB 6
ANALISIS KELAYAKAN PABRIK

6.1 Manajemen Perusahaan

Perusahaan merupakan tempat berlangsungnya kegiatan produksi dan tempat berkumpulnya seluruh faktor produksi. Semua kegiatan ekonomi dalam bisnis memerlukan organisasi untuk mencapai tujuan mencapai keuntungan yang sebesar-besarnya. Keberhasilan suatu perusahaan tergantung pada kepemimpinan organisasi di dalam perusahaan tersebut. Pengelolaan organisasi dalam suatu perusahaan meliputi perencanaan, pelaksanaan, dan pengelolaan organisasi dalam suatu perusahaan. Di Indonesia, perusahaan mempunyai bentuk hukum yang berbeda-beda. Bentuk hukum yang dipilih pada saat perancangan pabrik etilen adalah perseroan terbatas (PT) karena berdasarkan jumlah modal yang tertanam cukup besar. Dalam perencanaan suatu perusahaan harus diawali dengan pemberian nama perusahaan dan lokasi berdirinya perusahaan dikarenakan nama perusahann akan menjadi sebuah objek citra tersendiri perusahan tersebut, berikut rincian tentang bentuk perusahaan:

Nama Perusahaan : PT Binsej Chemical Indonesia
Bentuk : Perseroan Terbatas (PT)
Lapangan Usaha : Industri Manufaktur Bahan Kimia
Lokasi Perusahaan : Kota Cilegon, Provinsi Banten

Perseroan Terbatas (PT) merupakan suatu bahan hukum usaha yang didirikan oleh beberapa orang. Badan hukum ini memiliki kekayaan, hak dan kewajiban tersendiri yang terpisah dari pendiri (pemegang saham) maupun pengurusnya (Dewan komisaris dan direktur).

Kelebihan perusahaan dengan bentuk Perseroan Terbatas adalah:

a. Kelangsungan perusahaan lebih terjamin karena tidak bergantung pada satu pihak dan kepemilikan hak saham bisa berganti-ganti.
b. Kekayaan perusahaan terpisah dari kekayaan pribadi pemilik saham.
c. Pengelolaan perusahaan terpisah dari pemilik saham (permilik perusahaan) sehingga perseroan diurus dan dipimpin oleh Dewan direksi.
d. Penambahan modal untuk perusahaan lebih mudah.
e. Pengelolaan perusahaan dapat dilakukan lebih efisien serta professional.

Dalam melakukan kegiatan usahanya Perseroan Terbatas diatur oleh:

1. Rapat Umum Pemegang Saham (RUPS)

   Rapat ini dilaksanakan sesuai dengan jangka waktu yang ditetapkan dalam akte pendirian perusahaan yang umumnya dilaksanakan setahun sekali. Badan ini mengangkat Dewan komisaris dan direktur serta memutuskan kebijaksanaan umum yang harus dijalankan oleh perusahaan.

2. Dewan Komisaris

   Dewan komisaris bertugas mewakili para pemegang saham dan berfungsi mengawasi direktur dalam menjalankan tugas agar tidak terjadi penyimpangan yang akan merugikan perusahaan maupun dalam menjalankan kebijaksanaan umum yang telah ditetapkan serta memberikan saran atau masukan kepada direktur.

3. Direktur

   Direktur diangkat dan diberhentikan oleh Rapat Umum Pemegang Saham (RUPS). Direktur merupakan orang yang bertanggung jawab dalam melaksanakan kebijaksanaan umum perusahaan yang telah ditetapkan oleh Rapat Umum Pemegang Saham.

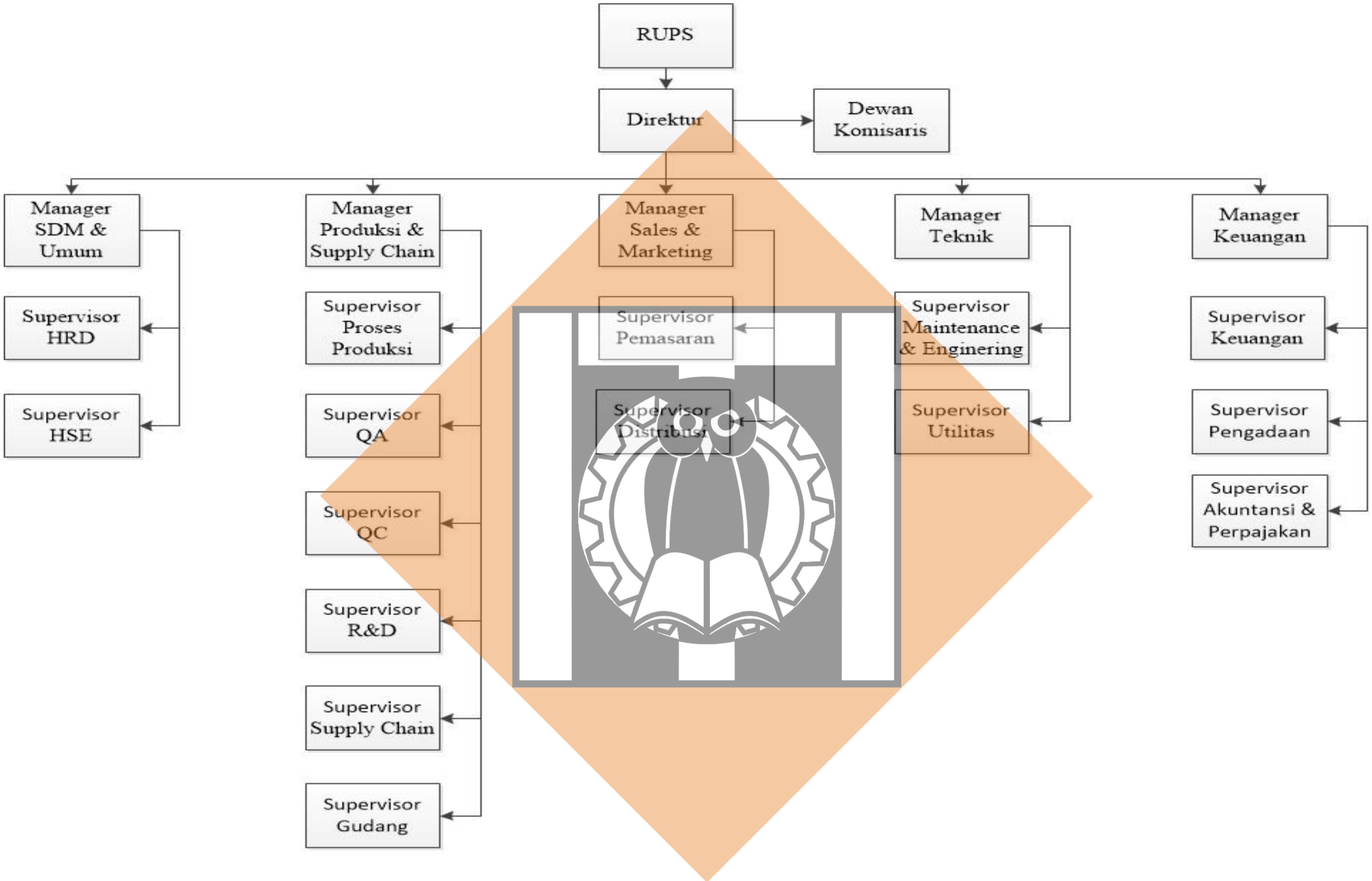
6.1.1 Diagram Organisasi

Untuk mencapai efisiensi perusahaan yang tinggi, diperlukan struktur organisasi yang baik. Struktur organisasi yang baik dapat menentukan kelancaran kegiatan sebuah perusahaan yang bertujuan mencapai laba yang maksimal, produksi yang berkelanjutan dan berkembang. Penyusunan struktur organisasi merupakan langkah awal perusahaan dalam memulai pelaksanaan kegiatan perusahaan untuk melaksanakan fungsi perencanaan, pengorganisasian, pengarahan, dan pengawasan.

Struktur organisasi disusun sebagaimana struktur organisasi dalam badan usaha yang bergerak dalam industri dan perdagangan, dimana unit dalam organisasi dibagi secara fungsional. Sistem organisasi dalam perusahaan ini adalah sistem organisasi garis dan staff yang mempunyai beberapa kelebihan:

a. Struktur yang jelas dan sederhana.
b. Pembagian tugas menjadi jelas antara pelaksana tugas pokok dan pelaksana tugas penunjang.
c. Disiplin kerja dapat terlaksana dengan baik.

d. Pengarahan dan informasi dapat diperoleh dengan mudah dengan melihat garis dalam sistem yang bersangkutan.
e. Pengambilan keputusan lebih efisien karena staf dapat memberi saran, pandangan dan lain-lainnya kepada pemimpinnya.
f. Mata rantai instruksi yang menghubungkan seluruh unit dalam organisasai berada dibawah organisasi yang jelas.



GAMBAR 6.1 DIAGRAM ORGANISASI

1. Rapat Umum Pemegang Saham

Merupakan kekuasaan tertinggi dalam perusahaan. Tugas dan wewenang RUPS adalah:

a. Mengangkat dan memberhentikan Dewan komisaris dan direktur serta mengesahkan anggota pemegang saham bila ada yang bergabung maupun mengundurkan diri.
b. Menetapkan tanggung jawab Dewan komisaris dan direktur atas mandat yang dapat dipercayakan kepada masyarakat.
c. Mengesahkan anggaran besar laba tahunan yang dibuat oleh Dewan komisaris dimana telah diperoleh untuk dibagikan dan dipakai kembali untuk penambahan modal demi kemajuan perusahaan.
d. Mengasahkan laporan pengajuan direktur mengenai keadaan dan jalannya perseroan, hasil yang sudah dicapai, perkiraan mengenai perkembangan perseroan dimasa yang akan datang, kegiatan utama perseroan dan perubahannya.

2. Dewan Komisaris

Dewan komisaris merupakan wakil dari pemegang saham yang berfungsi sebagai pengawas dalam Perseroan. Tugas dan wewenang Dewan komisaris sebagai berikut:

a. Memberikan pertanggung jawaban kepada RUPS.
b. Mewakili RUPS dalam melakukan pengawasan atas kebijakan direksi dalam menjalankan perseroan dan juga memberikan arahan/nasihat kepada direksi.
c. Melakukan pengawasan direktur secara berkelanjutan dan teratur.
d. Komisaris dapat menghentikan direktur untuk sementara jika didapat melakukan tindakan yang bertentangan dengan anggaran dasar.

3. Direktur

Direktur dipilih oleh RUPS untuk menjalankan kegiatan operasional perusahaan secara keseluruhan. Tugas dan wewenang direktur adalah sebagai berikut:

a. Sebagai penanggung jawab penuh atas kegiatan operasional perusahaan.
b. Bertanggung jawab atas kinerja perusahaan kepada RUPS.
c. Menetapkan kebijakan operasional Perusahaan.
d. Mengangkat dan memberhentikan karyawan.

4. Manager Produksi dan Supply Chain

Memiliki wewenang untuk merumuskan kebijaksanaan operasi pabrik, mengawasi kesinambungan operasional pabrik, dan bertanggung jawab terhadap supply chain bahan baku sampai menjadi produk serta melakukan pengembangan, pengujian, pemeriksaan semua bahan material, proses, dan mutu produk. Manager produksi dan supply chain membawahi:

a. Supervisior Proses Produksi

Tugas :

- Mengawasi dan bertanggung jawab atas kelancaran produksi.
- Bertanggung jawab atas kelancaran fungsional unit-unit sarana penunjang (utilitas).
- Mengawasi persediaan bahan baku dan penyimpanan hasil produksi serta transportasi hasil produksi.

b. Supervisior *Quality Assurance* (QA)

Tugas :

- Membuat perencanaan terkait kualitas produk yang hendak dibutuhkan.
- Melakukan pemantauan, analisis, dan uji coba terhadap produk sampel hingga saat diproduksi.
- Memberikan verifikasi terhadap bagaimana kualitas produk, kemudian diserahkan kepada tim produksi.

c. Supervisor *Quality Control* (QC)

Tugas :

- Memantau dan menguji perkembangan semua produk yang diproduksi oleh perusahaan.
- Memverifikasi kualitas produk.
- Memonitor setiap proses yang terlibat dalam produksi produk.
- Memastikan kualitas barang produksi sesuai standar agar lulus pemeriksaan.

d. Supervisor *Research and Development* (Rnd)

Tugas :

- Bertanggung jawab atas penelitian dan pengembangan proses produksi.
- Pemantauan dalam trial baik skala pilot maupun laboratorium.
- Melakukan registrasi produk ke lembaga badan berwenang baik berupa Nomor Izin Edar (NIE) maupun lainnya.

e. Supervisor Supply Chain

Tugas :

- Merancang produk baru (product development).
- Mendapatkan bahan baku (procurement).
- Merencanakan produksi dan persediaan (planning and control).
- Melakukan pengiriman (distribution).

f. Supervisor Gudang

Tugas :

- Menilai dan mengawasi segala kegiatan di dalam gudang.
- Mengatur alur distribusi barang.
- Menjaga kehandalan mesin produksi.

5. Manager Teknik

Memiliki wewenang dan bertanggung jawab atas pengkoordiniran segala kegiatan yang berhubungan dengan masalah teknik maupun proses penunjang pengoperasionalan pabrik baik di lapangan maupun di kantor. Manager teknik membawahi:

a. Supervisor Maintanance & Engineering

Tugas :

- Melakukan pengecekan dan permantauan segala yang berkaitan permasalahan teknik.
- Membuat kegiatan perusahaan berjalan dengan baik dengan meminimalisir gangguan.

b. Supervisor Utilitas

Tugas :

- Melakukan pengecekan dan pemantauan segala yang berkaitan permasalahan utilitas penunjang operasional pabrik.
- Membuat kegiatan perusahaan berjalan dengan baik dengan meminimalisir gangguan.

6. Manager Sales dan Marketing

Memiliki wewenang dan bertanggung jawab atas dalam perencanakan semua tahapan proses penjualan mulai dari menetapkan harga, mengoptimalkan produk, menentukan target pasar dalam apa target pembeli berdasarkan analisis data penjualan dan mendistribusikannya ke pelanggan. Manager sales dan marketing membawahi:

a. Supervisor pemasaran

Tugas :

- Menyusun dan melaksanakan strategi penjualan.
- Memastikan strategi penjualan berjalan.
- Melakukan fungsi koordinasi, support dan pengawasan.
- Mencapai target penjualan produk perusahaan.

b. Supervisor distribusi

Tugas :

- Melalukan perencanaan distribusi.
- Melakukan koordinasi persediaan dan pengelolaan pengiriman.
- Melalukan penjadwalan dan routing dalam pelaporan analisis distribusi.

7. Manager SDM dan Umum

Memiliki wewenang mengarahkan sistem dan mengatur sumber daya manusia diperusahaan. Manager SDM dan Umum membawahi:

a. Supervisor HRD

Tugas :

- Melakukan pengelolaan dan pengembangan sumber daya manusia (SDM), yaitu dalam hal perencancan, pelaksanaan dan pengawasan kegiatan sumber daya manusia, termasuk pengembangan kualitasnya dengan berpedoman pada kebijaksanaan dan prosedur yang berlaku di perusahaan.

b. Supervisor HSE

Tugas :

- Memberikan pelatihan dan pengarahan terhadap pekerja terutama tentang Keselamatan dan Kesehatan Kerja, pemakaian APAR, P3K, dan Tanggap Darurat.

8. Manager Keuangan

Berwewenang dalam pengaturan keuangan dan perpajakan dalam perseroan. Kepala Divisi Keuangan membawahi:

a. Supervisor keuangan

Tugas :

- Membuat laporan keuangan yang berkaitan dengan anggaran, hutang, piutang dan biaya.
- Melakukan analisis strategis dan membantu perencanaan strategis.

b. Supervisor akuntansi dan perpajakan

Tugas :

- Membantu kelancaran pelaksanaan kewajiban-kewajiban administrasi rutin perusahaan dalam rangka memenuhi kewajiban peraturan perpajakan.
- Membantu meneliti dan memeriksa laporan keuangan perusahaan untuk pelaporan SPT Tahunan Badan.
- Melakukan arsip perpajakan untuk seluruh unit bisnis.

c. Supervisor Pengadaan

Tugas :

- Melakukan negoisasi kontrak dengan supplier.
- Melakukan pengolaan persediaan barang dan jasa.
- Melakukan perencanaan pengadaan.

6.1.2 Perincian Jabatan dan Penggolongan Gaji

1. Jumlah Tenaga Kerja dan Gaji

Upah tenaga kerja disesuaikan dengan golongan tenaga kerja, bergantung pada kedudukan dalam struktur organisasi dan lamanya masa pengabdian.

Upah yang diterima karyawan terdiri dari

- Gaji pokok
- Tunjangan jabatan
- Tunjangan kinerja (persemester)
- Tunjangan makan dan transport (Natura)
- Tunjangan kesehatan
- Tunjangan seragam

- Bonus keuntungan perusahaan yang diberikan pertahun setelah menyelesaikan 1 tahun bekerja

Perbedaan golongan karyawan, membuat sistem penggajian dibagi menjadi 3, yaitu:

1. Sistem Bulanan
   Diberikan pada karyawan tetap maupun kontrak. Besarnya didasarkan pada kedudukan dalam organisasi dan keahliannya.
2. Sistem Harian (Outsourching PT/Yayasan)
   Diberikan kepada pekerja harian seperti buruh langsung atau pekerja yang dibutuhkan sewaktu-waktu saja.
3. Sistem Borongan (Outsourching Yayasan)
   Sistem borongan diberikan pada pekerja borongan yang besarnya tidak tetap, tergantung jenis pekerjaan yang dilakukan.

Selain gaji, karyawan juga diberikan gaji tambahan dengan penghitungan berdasarkan :

- Lembur di hari biasa (umumnya besarnya 1 ½ kali gaji perjam untuk tiap jam lemburan)
- Lembur di hari libur (umumnya besarnya 2 kali gaji perjam)
- Jika karyawan dipanggil bukan di jam kerjanya
- Tunjangan hari raya (umumnya sebesar sekali gaji pokok)

Berikut merupakan perincian gaji karyawan Pabrik Ethylene berdasarkan UMK Cilegon (databoks.katadata, 2024) dan Peranturan Menteri Keuangan no 250/PMK.02 tahun 2008 (Keungan, 2008).

UMK Cilegon	=	Rp4.815.108	
Pendidikan			
SMA	=	8%	Rp385,209
SMK	=	10%	Rp481,511
D3	=	15%	Rp722,266
S1	=	20%	Rp963,022
S2	=	35%	Rp1,685,288
Leader	=	45%	Rp2,166,799

Spv	=	50%	Rp2,407,554
Shift	=	5%	Rp240,755

Tabel 6.1 Rincian Gaji Karyawan

No.	Jabatan	Jumlah	Jenjang Pendidikan (minimum)	Gaji Bulanan	Total
1	Dewan Komisaris	2	S1	Rp20,223,454	Rp40,446,907
2	Direktur	1	S1	Rp24,075,540	Rp24,075,540
3	Manager	5	S2	Rp13,723,058	Rp68,615,289
4	Supervisor	13	S1	Rp10,593,238	Rp137,712,089
5	Sekretaris Direktur	1	S2	Rp6,500,396	Rp6,500,396
6	Sekretaris Manager	4	S1	Rp5,778,130	Rp23,112,518
Karyawan *Shift*					
7	**Proses**				
	- Leader	2	S1	Rp6,741,151	Rp13,482,302
	- Pelaksana	10	SMK	Rp5,537,374	Rp55,373,742
8	***Quality Aassurance***				
	- Leader	2	S1	Rp6,741,151	Rp13,482,302
	- Pelaksana	6	D3	Rp5,778,130	Rp34,668,778
9	***Quality Control***				
	- Leader	2	S1	Rp6,741,151	Rp13,482,302
	- Pelaksana	8	D3	Rp5,778,130	Rp46,225,037
10	**Utilitas**				
	- Leader	2	S1	Rp6,741,151	Rp13,482,302
	- Pelaksana	6	D3	Rp5,778,130	Rp34,668,778
11	**Maintenance & Engineering**				
	- Leader	2	S1	Rp6,741,151	Rp13,482,302
	- Pelaksana	10	D3	Rp5,778,130	Rp57,781,296
12	**Keamanan**				
	- Leader	2	D3	Rp6,018,885	Rp18,056,655
	- Pelaksana	8	SMA	Rp5,200,317	Rp41,602,533
Karyawan Non *Shift*					
13	**Rnd**				

No.	Jabatan	Jumlah	Jenjang Pendidikan (minimum)	Gaji Bulanan	Total
	- Leader	1	S1	Rp6,259,640	Rp6,259,640
	- Pelaksana	5	S1	Rp5,296,619	Rp26,483,094
14	**Supply Chain**				
	- Leader	1	S1	Rp5,778,130	Rp5,778,130
	- Pelaksana	2	S1	Rp5,296,619	Rp10,593,238
15	**Gudang**				
	- Leader	1	S1	Rp5,778,130	Rp5,778,130
	- Pelaksana	4	D3	Rp5,055,863	Rp20,223,454
16	**Distribusi**				
	- Leader	1	S1	Rp5,778,130	Rp5,778,130
	- Pelaksana	4	SMK	Rp4,815,108	Rp19,260,432
17	**Pemasaran**				
	- Leader	1	S1	Rp5,778,130	Rp5,778,130
	- Pelaksana	4	D3	Rp5,055,863	Rp20,223,454
18	**Kesehatan**				
	- Dokter	1	S1	Rp5,778,130	Rp5,778,130
	- Perawat	3	D3	Rp5,296,619	Rp15,889,856
19	**HSE**				
	- Leader	2	S1	Rp5,778,130	Rp11,556,259
	- Pelaksana	4	D3	Rp5,055,863	Rp20,223,454
20	**HRD**				
	- Leader	2	S1	Rp5,778,130	Rp11,556,259
	- Pelaksana	5	S1	Rp5,296,619	Rp26,483,094
21	**Rumah Tangga**				
	Cleaning Service	6	SMA	Rp4,237,295	Rp25,423,770
	Kebersihan Taman	3	SMA	Rp4,237,295	Rp12,711,885
	Sopir Perusahaan	3	SMA	Rp4,959,561	Rp14,878,684
22	**Keuangan**				
	Akutansi	2	S1	Rp5,296,619	Rp10,593,238
	Perpajakan	2	S1	Rp5,296,619	Rp10,593,238
	Pengadaan	2	S2	Rp6,018,885	Rp12,037,770
TOTAL		146			Rp960,132,535

No.	Jabatan	Jumlah	Jenjang Pendidikan (minimum)	Gaji Bulanan	Total

Gaji / tahun (a)	=	Rp11,521,590,422
Tunjangan Hari Raya (1 bulan gaji pokok)	=	Rp703,005,768
Tunjangan Makan dan Transport (5% a)	=	Rp576,079,521
Tunjangan Kesehatan (2,5% a)	=	Rp288,039,761
Total Gaji Per Tahun	=	Rp13,088,715,472

2. Fasilitas Bagi Karyawan

Selain sistem pengupahan dan pengaturan jadwal yang sudah disusun sedemikian rupa, perseroan/perusahaan yang baik dan taat hukum akan menyediakan fasilitas-fasilitas lain demi kesejahteraan karyawan, yaitu sebagai berikut :

a. Jaminan Keselamatan Kerja

Perusahaan wajib menyediakan fasilitas keselamatan kerja berupa alat pelindung diri (APD) yang disesuaikan dengan area dan risiko kerja, diantaranya:

- Helm atau topi pengamanan
- Jas laboratorium
- Sarung tangan
- Kaca mata pelindung
- Alat penyumbat telinga *(earplug)*
- Masker

b. Jaminan Asuransi

Perusahaan wajib mengikutsertakan seluruh karyawannya dalam program jaminan sosial tenaga kerja sesuai dengan ketetapan yang dibuat pemerintah.

c. Hak Cuti

Karyawan yang telah memenuhi masa kerja tertentu mendapatkan 12 hari kerja setiap tahunnya. Untuk karyawan *non-shift* mendapatkan libur pada hari libur nasional, sedangkan untuk karyawan *shift* jika masuk di hari libur nasional maka dianggap sebagai lembur. Dimana lembur dilakukan apabila ada keperluan mendesak dan atas persetujuan kepala bagian.

d. Fasilitas ibadah

Menyediakan fasilitas tempat dan memaklumi waktu tertentu untuk beribadah bagi karyawan.

e. Fasilitas Kesehatan

Memnyediakan seorang dokter dan perawat untuk bersiaga.

f. Fasilitas olahraga

6.1.3 Penggiliran Tugas

Pabrik beroperasi 330 hari setahun dan jam kerja 24 jam sehari. Selama hari kerja, unit produksi beroperasi dari Senin hingga Minggu. Perbaikan, perawatan dan penghentian proses (*shutdown)* dilakukan pada saat libur nasional. Untuk menunjang kelancaran proses produksi, maka waktu kerja karyawan produksi diatur dalam sistem *shift* dan *non shift.*

1. Pengaturan jadwal kerja

a. Karyawan *Shift*

Waktu kerja diatur dalam 3 *shift* dalam 24 jam. Tiap *shift* berdurasi selama 8 jam, namun untuk bagian pengamanan jam kerja dimulai satu jam sebelum jadi kerja *shift.* Sistem kerja yang dilakukan terbagi dalam 4 kelompok. Hal ini dilakukan untuk lebih mengefektifkan kinerja para karyawan.

Unit-unit yang termasuk dalam kerja *shift* antara lain:

- Unit proses
- Unit utilitas
- Unit *quality control* (QC)
- Unit *quality assurance* (QA)
- Unit keamanan
- Unit maintenance dan *engineering*

Jam kerja *Shift* dalam 24 jam adalah sebagai berikut:

- *Shift* 1 : 06.00 – 14.00 WIB
- *Shift* 2 : 14.00 – 22.00 WIB
- *Shift 3* : 22.00 – 06.00 WIB

Jam kerja untuk unit pengamanan adalah sebagai berikut:

- *Shift* 1 : 07.00 – 15.00 WIB
- *Shift* 2 : 15.00 – 23.00 WIB

- *Shift 3* : 23.00 – 07.00 WIB

Berikut merupakan jadwal kerja shift

Tabel 6.2 Jadwal Kerja Karyawan Shift

Hari Ke	1	2	3	4	5	6	7	8
Shift 1	A	A	B	B	C	C	D	D
Shift 2	B	B	C	C	D	D	A	A
Shift 3	C	C	D	D	A	A	B	B
Libur	D	D	A	A	B	B	C	C

Keterangan :

A : Grup kerja I

B : Grup Kerja II

C : Grup kerja III

D : Grup kerja IV

b. Karyawan *Non Shift*

Untuk karyawan *Non Shift* unit produksi adalah supervisor dan karyawan yang bukan termasuk unit produksi kecuali departemen Rnd, supply chain dan gudang. Waktu kerja karyawan adalah 8 jam perhari atau 40 jam seminggu dengan 1 jam istirahat setiap hari kecuali hari Jum'at ditambah 30 menit istirahatnya. Biasanya waktu kerjanya disesuaikan dengan waktu kerja kantor yaitu hari Senin – Jum'at, domulai dari jam 07.30 – 16.30 WIB. Rincian waktu kerja karyawan *Non Shift* sebagai berikut:

Tabel 6.3 Jadwal Kerja Karyawan Non Shift

Hari	Jam Masuk	Jam Istirahat	Jam Pulang
Senin	07.30	12.00 – 13.00	16.30
Selasa	07.30	12.00 – 13.00	16.30
Rabu	07.30	12.00 – 13.00	16.30
Kamis	07.30	12.00 – 13.00	16.30
Jumat	07.30	11.30 – 13.00	16.30

6.2 Kelayakan Ekonomi

Analisa ekonomi pabrik *Ethylene Dari Dehidrasi Etanol Kapasitas 85.000 Ton/Tahun* dibuat untuk mendapatkan gambaran kelayakan suatu penanaman modal dalam suatu kegiatan produksi. Dengan analisa ekonomi peninjauan kebutuhan modal investasi, besarnya laba yang akan diperoleh, lamanya investasi modal kembali dan titik impas terhadap volume produksi dapat diprediksi atau diperkirakan.

Analisa ekonomi dimulai dengan menganalisa kapasitas produksi, jenis bahan baku, bahan alat dan harga alat-alat proses serta penunjang produksi. Perkiraan harga peralatan diambil berdasarkan indeks harga tahun-tahun sebelumnya di dapat dari *Chemical Engineering Plant Cost Index.*

Berdasarkan indeks harga tahun 2017 sebesar 567.5 (Cheresources, 2022), prediksi tahun-tahun mendatang menggunakan regresi linear didapatkan bahwa indeks harga pada tahun 2027 adalah sekitar 724.76. Sedangkan untuk perkiraan harga peralatan didapatkan dari situs alibaba.com dan matche.com.

6.2.1 Asumsi dan Parameter

Asumsi dan parameter yang digunakan dalam analisa ekonomi pendirian pabrik *Ethylene* adalah sebagai berikut:

Tabel 6.4 Asumsi & Parameter Analisa Ekonomi

Pembangunan fisik pabrik	=	2025	
Pabrik mulai beroperasi	=	2027	masa konstruksi 2 tahun
Jumlah hari kerja pertahun	=	330	hari
Umur alat utama	=	10	tahun
Shut down alat pertahun	=	35	hari
Nilai tukar rupiah	=	15623	rupiah
Suku bunga bank	=	7.90%	per tahun
Kenaikan harga bahan baku dan produksi	=	10%	per tahun
Salvage Value (Nilai Rongsok)	=	10%	DFCI tanpa harga tanah

6.2.2 Fixed Capital

Berikut merupakan komponen-komponen biaya yang termasuk dalam ketegori modal teteap pada pendirian pabrik *Ethylene.*

Tabel 6.5 Modal tetap langsung

A.	**Modal Investasi Tetap Langsung (Direct Fixed Capital Investment / DFCI)**		
1. Peralatan Utama dan Penunjang		A	Rp 373,183,969,926
2. Pemasangan mesin dan peralatan (termasuk isolasi & pengecatan)		25% A	Rp 93,295,992,482
3. Instrumentasi dan kontrol terpasang		10% A	Rp 37,318,396,993
4. Sistem perpipaan		5% A	Rp 18,659,198,496
5. Instalasi listrik terpasang		6% A	Rp 18,659,198,496
6. Bangunan		13% A	Rp 22,391,038,196
7. Perbaikan		3%	Rp 48,513,916,090
8. Tanah		1%	Rp 37,500,000,000
9. Fasilitas Pelayanan		15% A	Rp 55,977,595,489
Sub Total		A'	Rp 705,499,306,168
	DFCI tidak terduga	20% A'	Rp 141,099,861,234
Total Modal Tetap Langsung (DFCI)		**B**	Rp 846,599,167,401

Tabel 6.6 Modal tetap tidak langsung

B..	**Modal Investasi Tetap Tidak Langsung (Indirect Fixed Capital Investment / IFCI)**	
1. Prainvestasi	12% B	Rp 101,591,900,088
2. Keteknikan dan kepengawasan	12% B	Rp 101,591,900,088
3. Biaya Konstraktor dan konstruksi	5% B	Rp 42,329,958,370
4. Bunga pinjaman selama masa konstruksi	8% B	Rp 67,727,933,392
5. Trial Run		Rp-
Sub Total	B'	Rp 313,241,691,938
IFCI tidak terduga	20% B'	Rp 62,648,338,388
Total Modal Tetap Tidak Langsung (IFCI)	**C**	**Rp 375,890,030,326**

Total Modal Investasi Tetap (FCI)

FCI = DFCI + IFCI = Rp 1,222,489,197,727

6.2.3 Modal Kerja *(Working Capital Investment)*

Modal kerja merupakan modal yang digunakan untuk membiayai seluruh kegiatan perusahaan dari awal produksi hingga terkumpulnya hasil penjualan untuk memenuhi kebutuhan perputaran beban biaya operasional.

Modal ini dihitung dalam kurun waktu 3 bulan dengan jumlah hari kerja sebanyak 90 hari.

Tabel 6.7 Biaya kebutuhan bahan baku

Tabel Persediaan Kebutuhan Bahan Baku WCI

Bahan Baku	Kebutuhan (Kg/jam)	Harga (Rp/Kg)	Biaya
Etanol 99% (kg)	17802.1535	Rp 18,299	Rp 703,641,512,920
Katalis (kg)	161880.7022	Rp 22,874	Rp 7,998,047,963,274
Total Biaya Bahan Baku		a*	Rp 8,701,689,476,194

Tabel 6.8 Biaya kebutuhan sarana penunjang

Tabel Persediaan Kebutuhan Sarana Penunjang WCI

Bahan Baku	Kebutuhan	Harga (/satuan)	Biaya
Solar (lt/jam)	249563.89	Rp 9,956	Rp 223,616,530,815
Listrik PLN (kwh)	9751.56	Rp 1,115	Rp 978,569,133
Air Keseluruhan	275476.45	Rp 4,832	Rp 119,787,544,024
Downtherm	234269.00	Rp 91,495	Rp 1,929,090,055,101
Resin Ion Exchanger	5.71	Rp 22,874	Rp 11,754,116
Total biaya sarana penunjang		b*	Rp 224,595,099,948

Biaya pengemasan & distribusi produk	0,5% a*	Rp 10,877,111,845
Biaya pengawasan mutu	0.5% a*	Rp 10,877,111,845
Biaya pemeliharaan & perbaikan	2% DFCI	Rp 4,232,995,837
Gaji karyawan (3 bulan gaji)		Rp 960,132,535

Sub Total WCI Rp 8,953,231,928,204

WCI tidak terduga 20% sub total WCI Rp 1,790,646,385,641

Total Modal kerja (WCI) = Sub total WCI + WCI tidak terduga

= Rp 10,743,878,313,845

Total Modal Investasi (TCI) = FCI + WCI

= Rp 11,966,367,511,573

6.2.4 Biaya Produksi

Biaya produksi terdiri dari *Direct Manufacturing Cost* (DMC) dan *Fixed Manufacturing Cost* (FMC). Kedua biaya tersebut dipengaruhi oleh *fixed cost* dan *variable cost*. *Fixed cost* merupakan biaya yang besarnya tidak pengaruhi oleh kapasitas produksi, sedangkan *variable cost* adalah biaya yang dipengaruhi oleh besarnya kapasitas produksi.

Tabel 6.9 Biaya produksi tahun ke-1 dan ke-2

TAHUN				I		II	
KAPASITAS PRODUKSI				80%		90%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
A.	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
A.	Biaya Bahan Baku				Rp 2,254,615,081,330		Rp 2,790,086,163,146
b.	Gaji Karyawan			Rp 13,088,715,472		Rp 14,397,587,019	
c.	Biaya Pemeliharaan dan Perbaikan (kenaikan 5% per tahun)	2%	DFCI	Rp 16,931,983,348		Rp 17,778,582,515	
d.	Biaya Royalti dan Paten	0,5	TS		Rp 93,324,427,944		Rp 115,488,979,581
e.	Biaya Laboratorium	0,	B		Rp 11,273,075,407		Rp 13,950,430,816
f.	Biaya pengemasan produk	0,5%	BB		Rp 11,273,075,407		Rp 13,950,430,816
g.	Biaya sarana penunjang			Rp 3,744,599,373	Rp 724,833,090,680	Rp 4,119,059,310	Rp 896,980,949,717

TAHUN				I		II	
KAPASITAS PRODUKSI				80%		90%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
h.	Biaya Start Up			Rp 9,841,271,432, 647		Rp 10,825,398,575 ,912	
	Total Biaya Manufacturing Langsung (DMC)			**Rp 9,875,036,730, 839**	**Rp 3,095,318,750,767**	**Rp 10,861,693,804 ,756**	**Rp 3,830,456,954,075**
	Biaya Plant Overhead	20%	(b+c)	**Rp 6,004,139,764**		**Rp 6,435,233,907**	
Biaya Manufacturing Tetap (FMC)							
2.	**Depresiasi**			Rp 78,656,030,587		Rp 78,656,030,587	
3.	Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th	0,1%		Rp 59,891,038		Rp 65,880,142	
a.	Biaya asuransi (kenaikan 10 %) pertahun	0,5%	DFCI	Rp 4,232,995,837		Rp 4,656,295,421	
	Total Biaya Manufacturing Tetap (FMC)			**Rp 82,948,917,462**		**Rp 83,378,206,149**	

6.2.5 Pengeluaran Umum

Merupakan pengeluaran biaya untuk menunjang beroperasinya suatu pabrik seperti biaya administrasi, biaya distribusi dan penjualan serta bunga bank beserta cicilan pokoknya.

Tabel 6.10 Biaya pengeluaran umum

B.	**Pengeluaran Umum (General Expenses)**						
a.	Biaya administrasi	5%	b	Rp 654,435,774		Rp 719,879,351	
b.	Biaya distribusi dan penjualan	10%	f		Rp 1,127,307,541		Rp 1,395,043,082

c.	Bunga Bank + Cicilan Pokok			Rp 221,932,500,000		Rp 206,136,875,000	
	Total Pengeluaran Umum			**Rp 222,586,935,774**	**Rp 1,127,307,541**	**Rp 206,856,754,351**	**Rp 1,395,043,082**
Total Biaya				**Rp 10,186,576,723,839**	**Rp 3,096,446,058,308**	**Rp 11,158,363,999,163**	**Rp 3,831,851,997,156**
Total Biaya Produksi (TPC)				**Rp 13,283,022,782,147**		**Rp 14,990,215,996,319**	

6.2.6 Penjualan dan Keuntungan

Penjualan merupakan aktivitas penting bagi perusahaan terutama untuk mencapai keuntungan. Penjualan merupakan suatu kegiatan perusahaan untuk dapat mengembangkan berbagai strategi bisnis yang diarahkan untuk memuaskan kebutuhan dan juga keinginan konsumen untuk mendapat laba sebesar-besarnya. Penjualan juga merupakan sumber utama pendapatan suatu perusahaan yang melakukan transaksi jual beli.

Tabel berikut merupakan proyeksi penjualan dan keuntungan pabrik etilen selama 10 tahun dengan kapasitas 85.000 ton/tahun.

Tabel 6.11 Penjualan dan Laba 10 tahun

Tahun	**Penjualan**	**Pengeluaran**	**Laba sebelum pajak**	**PPH 25%**	**Laba setelah pajak**
1	Rp 18,664,885,588,800	Rp 13,283,022,782,147	Rp 5,381,862,806,653	Rp 1,345,465,701,663	Rp 4,036,104,990
2	Rp 23,097,795,916,140	Rp 14,990,215,996,319	Rp 8,107,579,919,821	Rp 2,026,894,979,955	Rp 6,080,684,939,865
3	Rp 28,230,639,453,060	Rp 16,912,233,431,761	Rp 11,318,406,021,299	Rp 2,829,601,505,325	Rp 8,488,804,515,974
4	Rp 31,053,703,398,366	Rp 18,559,641,371,180	Rp 12,494,062,027,186	Rp 3,123,515,506,797	Rp 9,370,546,520,390
5	Rp 34,159,073,738,203	Rp 20,214,563,664,506	Rp 13,944,510,073,697	Rp 3,486,127,518,424	Rp 10,458,382,555,273
6	Rp 37,574,981,112,023	Rp 22,226,919,572,003	Rp 15,348,061,540,020	Rp 3,837,015,385,005	Rp 11,511,046,155,015

Tahun	Penjualan	Pengeluaran	Laba sebelum pajak	PPH 25%	Laba setelah pajak
7	Rp 41,332,479,223,225	Rp 24,440,449,327,455	Rp 16,892,029,895,771	Rp 4,223,007,473,943	Rp 12,669,022,421,828
8	Rp 45,465,727,145,548	Rp 26,875,267,228,517	Rp 18,590,459,917,031	Rp 4,647,614,979,258	Rp 13,942,844,937,773
9	Rp 50,012,299,860,102	Rp 29,505,721,680,089	Rp 20,506,578,180,014	Rp 5,126,644,545,003	Rp 15379,933,635,010
10	Rp 55,013,529,846,113	Rp 32,314,868,991,339	Rp 22,698,660,854,774	Rp 5,674,665,213,693	Rp 17,023,995,641,080

(Berdasarkan UU No.36 Tahun 2008 Wajib pajak badan usaha dalam negeri adalah 25%)

Jumlah nominal aliran masuk = laba setelah pajak + depresiasi + salvage value

Tabel 6.12 Nominal aliran masuk

Tahun	Laba setelah pajak	Depresiasi	Salvage value+ tanah	Cash in Nominal
1	Rp 4,036,104,990	Rp 78,656,030,587	Rp-	Rp 4,115,053,135,576
2	Rp 6,080,684,939,865	Rp 78,656,030,587	Rp-	Rp 6,159,340,970,452
3	Rp 8,488,804,515,974	Rp 78,656,030,587	Rp-	Rp 8,567,460,546,561
4	Rp 9,370,546,520,390	Rp 78,656,030,587	Rp-	Rp 9,449,202,550,976
5	Rp 10,458,382,555,273	Rp 78,656,030,587	Rp 703,035,000	Rp 10,537,741,620,859
6	Rp 11,511,046,155,015	Rp 78,656,030,587	Rp-	Rp 11,589,702,185,602
7	Rp 12,669,022,421,828	Rp 78,656,030,587	Rp-	Rp 12,747,678,452,415
8	Rp 13,942,844,937,773	Rp 78,656,030,587	Rp-	Rp 14,021,500,968,360
9	Rp 15379,933,635,010	Rp 30,878,862,421	Rp-	Rp 15,410,812,497,431
10	Rp 17,023,995,641,080	Rp 30,878,862,421	Rp 93,546,611,179	Rp 17,148,421,114,679

6.2.7 *Break Event Point* (BEP)

Break Event Point (BEP) atau titik impas adalah besarnya persen kapasitas produksi dimana nilai total penjualan bersih sama dengan nilai total biaya yang dikeluarkan perusahaan dalam kurun waktu satu tahun. BEP berfungsi sebagai pengendali total produksi penjualan dan mengendalikan keuangan pada tahun buku berjalan.

$$BEP = \frac{Total\ fixed\ cost}{Total\ sales - Total\ variable\ cost} \times 100\%$$

Tabel 6.13 Break Event Point (BEP)

Tahun	Total			BEP (%)
	Fixed Cost (Rp)	Variabel Cost (Rp)	Penjualan (Rp)	
1	Rp 10,185,264,546,720	Rp 3,096,446,058,308	Rp 18,664,885,588,800	65.43
2	Rp 11,156,920,604,332	Rp 3,831,851,997,156	Rp 23,097,795,916,140	57.92
3	Rp 12,227,271,034,256	Rp 4,683,374,663,191	Rp 28,230,639,453,060	51.93
4	Rp 13,406,182,733,925	Rp 5,151,712,129,510	Rp 31,053,703,398,366	51.76
5	Rp 14,545,759,163,525	Rp 5,666,883,342,461	Rp 34,159,073,738,203	51.06
6	Rp 15,991,234,620,924	Rp 6,233,571,676,707	Rp 37,574,981,112,023	51.03
7	Rp 17,581,195,881,268	Rp 6,856,928,844,378	Rp 41,332,479,223,225	51.00
8	Rp 19,330,088,437,712	Rp 7,542,621,728,816	Rp 45,465,727,145,548	50.98
9	Rp 21,206,025,010,203	Rp 8,296,883,901,697	Rp 50,012,299,860,102	50.84
10	Rp 23,322,038,649,923	Rp 8,989,736,296,409	Rp 55,013,529,846,113	50.68

6.2.8 Analisis Ekonomi

Dalam analisis ekonomi akan didapat perkiraan-perkiraan mengenai jumlah investasi yang meliputi :

1. Net Present Value (NPV)
2. Net Cash Flow Present Value (NCFPV)
3. Internal Rate Of Return (IRR) Suatu indikator untuk menilai kelayakan pendirian pabrik.
4. Minimum Payback Period (MPP) jangka waktu minimum pengembalian modal investasi.

Tabel 6.14 MPP & NCFPV

Tahun	NCF nominal (Rp)	Faktor Discount 1/(1+0.0790)^n	NCF PV (Rp)	Akumulasi
0	-Rp 11,966,367,511,573	1.00	-Rp 11,966,367,511,573	-Rp 11,966,367,511,573
1	Rp 4,115,053,135,576	0.93	Rp 3,813,765,649,283	-Rp 8,152,601,862,290
2	Rp 6,159,340,970,452	0.86	Rp 5,290,434,687,021	-Rp 2,862,167,175,269
3	Rp 8,567,460,546,561	0.80	Rp 6,820,053,462,604	Rp 3,957,886,287,335
4	Rp 9,449,202,550,976	0.74	Rp 6,971,229,490,468	Rp 10,929,115,777,803
5	Rp 10,537,741,620,859	0.68	Rp 7,205,105,135,064	Rp 18,134,220,912,868
6	Rp 11,589,702,185,602	0.63	Rp 7,344,184,999,458	Rp 25,478,405,912,326
7	Rp 12,747,678,452,415	0.59	Rp 7,486,537,210,500	Rp 32,964,943,122,826
8	Rp 14,021,500,968,360	0.54	Rp 7,631,729,149,753	Rp 40,596,672,272,579
9	Rp 15,410,812,497,431	0.50	Rp 7,773,785,163,537	Rp 48,370,457,436,115
10	Rp 17,148,421,114,679	0.47	Rp 8,016,959,564,511	Rp 56,387,417,000,627
Total			**Rp 56,387,417,000,627**	

Berdasarkan tabel diatas, jangka waktu pengembalian investasi modal dengan suku bunga 7.90% adalah 2 tahun atau 6 bulan, perhitungan dilakukan dengan metode interpolasi.

Dilakukan perhitungan IRR seperti pada tabel berikut:

Tabel 6. 15 Internal Rate of Return

Tahun	Net Cash Flow	Bunga 7,90% 1/(1+I)^n	Present Value	Bunga 57.82% 1/(1+I)^n	Present Value	Bunga 25% 1/(1+I)^n	Present Value
0	-Rp11,966,367,511,573	1.00	-Rp11,966,367,511,572.700	1.00	-Rp11,966,367,511,572.700	1.00	-Rp11,925,272,914,658.200
1	Rp4,115,053,135,576	0.93	Rp3,813,765,649,282.980	0.63	Rp2,607,494,003,541.700	0.80	Rp3,429,024,498,176.100
2	Rp6,159,340,970,452	0.86	Rp5,290,434,687,020.940	0.40	Rp2,473,033,437,541.070	0.64	Rp4,069,152,300,340.680
3	Rp8,567,460,546,561	0.80	Rp6,820,053,462,604.170	0.25	Rp2,179,694,990,118.780	0.51	Rp4,504,287,326,486.360
4	Rp9,449,202,550,976	0.74	Rp6,971,229,490,467.880	0.16	Rp1,523,304,209,821.730	0.41	Rp3,974,011,188,329.360
5	Rp10,537,741,620,859	0.68	Rp7,205,105,135,064.440	0.10	Rp1,076,432,744,884.950	0.33	Rp3,544,190,858,958.710
6	Rp11,589,702,185,602	0.63	Rp7,344,184,999,458.190	0.06	Rp750,169,671,157.216	0.26	Rp3,118,412,532,221.690
7	Rp12,747,678,452,415	0.59	Rp7,486,537,210,500.060	0.04	Rp522,836,858,983.676	0.21	Rp2,743,994,581,565.610
8	Rp14,021,500,968,360	0.54	Rp7,631,729,149,752.680	0.03	Rp364,399,249,518.749	0.17	Rp2,414,556,631,839.760
9	Rp15,410,812,497,431	0.50	Rp7,773,785,163,536.630	0.02	Rp253,779,405,071.203	0.13	Rp2,123,086,672,622.870
10	Rp17,148,421,114,679	0.47	Rp8,016,959,564,511.160	0.01	Rp178,938,111,455.942	0.11	Rp1,889,418,237,313.230
Total			**Rp 56,387,417,000,626.500**		**-Rp36,284,829,477.640**		**Rp18,951,265,823,528.500**

Berdasarkan tabel diatas, didapat persen IRR sebesar = 57.88%. dengan bunga pinjaman sebesar 7.90% serta nilai NCPV bertanda positif maka dapat disimpulkan bahwa Proyek investasi Pabrik Ethylene adalah ***FEASIBLE*** atau **LAYAK UNTUK DIDIRIKAN**.

DAFTAR PUSTAKA

Aries, R.S., and Newton, R.D., 1955, “Chemical Engineering Cost”, Mc. Graw Hill Book co., New York

BPS. (2023, Oktober 13). Retrieved from Badan Pusat Statistik: http://www.bps.go.id/

Brown, G.G., 1950, “Unit Operation”, John Wiley and Sons Inc, New York.

Brownell, L.E and Young, E.H., 1959., “Equipment Design”, John Willey & Sons, inc., New York.

Coupard, Vincent., 2017. United States, Patent No US 009663414B2

Kern, D.Q., 1950, “Process Heat Transfer”, McGraw-Hill International Book Company Inc., New York.

Keungan, P. M. (2008). PMK 03 no 250.

Perry, R.H., and Green, D.W., 1984, “Perry’s Chemical Engineer’s Handbook”, 6th ed., McGraw-Hillo Book Company, New York.

Peters, M. S., & Timmerhaus, K. D. (1991). Plant Design and Economics for Chemical Engineers 4th Ed. New York: McGraw-Hill Book Company.

Severn. (2004). Diktat Utilitas. In *Boiler Design* (pp. 140-170).

Smith, J.M., Van Ness, H.G., and Abbott, M., 1997, “Introduction to Chemical Engineering Thermodynamics”, Sixth Edition., New York : Mc Graw Hill Book Companies, Inc.

TLV. (2023). *Calculator: Superheated Steam Table by Pressure*. Retrieved October 2023, from TLV A Steam Specialist Company: https://www.tlv.com/global/TI/calculator/steam-table-pressure.html

Yaws, C.L., 1999, “Chemical Properties Handbook Physical, Thermodynamic, Enviromental, Transport, Safety, and Health Related Properties For Organic and Inorganic Chemicals”, New York : Mc Graw Hill Book Companies, Inc.

LAMPIRAN 1 DATA

L1.1 Spesifikasi Bahan Baku dan Produk

Tabel L1.1 Spesifikasi Bahan Baku Etanol

Spesifikasi	Katerangan
Wujud	Cairan, Tidak Berwarna
Rumus Molekul	C_2H_5OH
Bobot Mulekul	46.07 g/mol
Spesific Gravity	0.790-0.793 g/cm^3 at 20°C
pH	7.0 at 20°C
Viskositas	1.2 mPa.s cp at 20°C
Titik Lebur	-114.5°C
Titik Didih	78.3 °C at 1 atm
Kapasitas Panas	N/A
Tekanan Uap	59 hPa at 20°C
Kemurnian	≤ 100%
Penanganan dan Penyimpanan	**Tindakan pencegahan untuk penanganan yang aman :** Kenakan pakaian pelindung. Jangan menghirup zat/campuran. Hindari terbentuknya uap/aerosol. **Kondisi penyimpanan yang aman :** Simpan wadah tertutup rapat di tempat yang kering dan berventilasi baik. Jauhkan dari panas dan sumber api.
Efek pada air	Tidak diketahui
Toksisitas	LD50 Tikus: 10.470 mg/kg, LC50 Tikus: 124,7 mg/l; 4 h ; uap

Karsinogenisitas	Tidak ada efek karsinogenik

Tabel L1.2 Spesifikasi Produk Ethylene

Spesifikasi	Katerangan
Wujud	Gas, tidak berwarna, berbau manis
Rumus Molekul	C_2H_4
Bobot Mulekul	28.05 g/mol
Density	1.261 kg/m^3
Spesific Gravity	0.98
pH	N/A
Viskositas	N/A
Titik Beku	-169°C
Titik Didih	-104°C
Kapasitas Panas	N/A
Tekanan Uap	N/A
Kemurnian	>99%
Penanganan dan Penyimpanan	**Tindakan pencegahan untuk penanganan yang aman :** Hindari kontak langsung dengan Ethylene. Jangan makan atau minum saat menangani bahan kimia. Waspadai tanda-tanda pusing atau kelelahan; karena kekurangan oksigen, paparan terhadap konsentrasi Ethylene yang fatal dapat terjadi tanpa gejala peringatan yang berarti. **Kondisi penyimpanan yang aman :** Simpan di tempat yang kering,

	berventilasi baik, jauh dari sumber panas, jauh dari area lalu lintas padat dan pintu keluar darurat.
Efek pada air	Tidak diketahui
Toksisitas	LC 50 (inhalation, mouse) = 96 pph, LCLo (inhalation, mammal) = 950.000 ppm/5 menit
Karsinogenisitas	Tidak ada efek karsinogenik

L1.2 Literatur

L1.2.1 Patent US 20130090510A1 (2013)

US 20130090510A1

(19) **United States**
(12) **Patent Application Publication**
COUPARD et al.

(10) **Pub. No.: US 2013/0090510 A1**
(43) **Pub. Date: Apr. 11, 2013**

(54) **PROCESS FOR DEHYDRATION OF DILUTE ETHANOL INTO ETHYLENE WITH LOW ENERGY CONSUMPTION WITHOUT RECYCLING OF WATER**

(71) Applicants: **TOTAL RESEARCH & TECHNOLOGY**, SENEFFE (BE); **IFP ENERGIES NOUVELLES**, RUEIL-MALMAISON CEDEX (FR)

(72) Inventors: **Vincent COUPARD**, Villeurbanne (FR); **Natacha TOUCHAIS**, Vienne (FR); **Stephanie FLEURIER**, Lyon (FR); **Helena GONZALEZ PENAS**, Lyon (FR); **Philip DE SMEDT**, Sint-Niklaas (BE); **Walter VERMEIREN**, Sint Niklaas (BE); **Cindy ADAM**, Wierde (BE); **Delphine MINOUX**, Nivelles (BE)

(73) Assignees: **IFP ENERGIES NOUVELLES**, RUEIL-MALMAISON CEDEX (FR); **TOTAL RESEARCH & TECHNOLOGY FELUY**, SENEFFE (BE)

(21) Appl. No.: **13/646,002**

(22) Filed: **Oct. 5, 2012**

(30) **Foreign Application Priority Data**

Oct. 7, 2011 (FR) 11/03.075

Publication Classification

(51) **Int. Cl.**
C07C 1/24 (2006.01)

(52) **U.S. Cl.**
USPC **585/640**; 585/639

(57) **ABSTRACT**

A process for dehydration of an ethanol feedstock into ethylene, comprising the vaporization of said dilute hydrated ethanol feedstock in an exchanger, with heat exchange with the effluent that is obtained from a last reactor, with said mixture being introduced into said vaporization stage at a pressure that is lower than the pressure of the effluent that is obtained from the last reactor, the compression of the mixture that is vaporized in a compressor, the introduction of the vaporized and compressed mixture, into at least one adiabatic reactor that contains at least one dehydration catalyst.

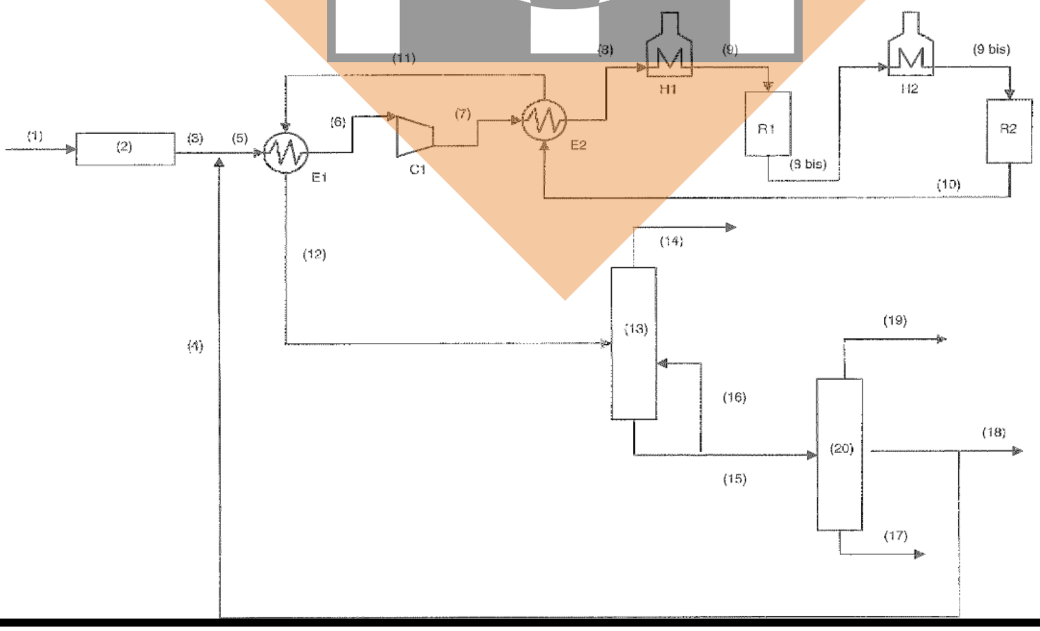
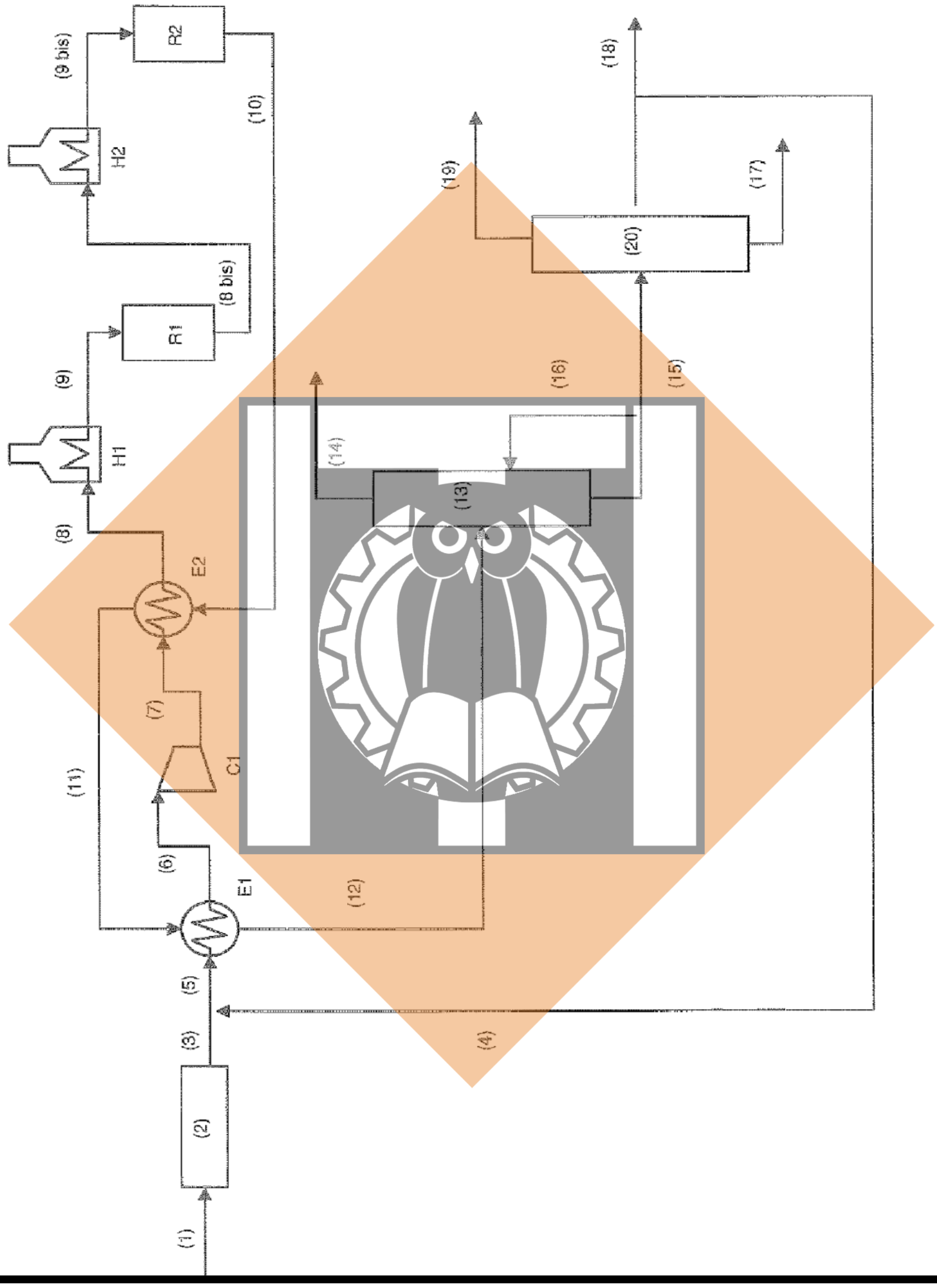


Figure 1


(1)
(2)
(3)
(4)
(5)
E1
(6)
C1
(7)
E2
(8)
H1
(9)
R1
(8 bis)
H2
(9 bis)
R2
(10)
(11)
(12)
(13)
(14)
(15)
(16)
(17)
(18)
(19)
(20)

PROCESS FOR DEHYDRATION OF DILUTE ETHANOL INTO ETHYLENE WITH LOW ENERGY CONSUMPTION WITHOUT RECYCLING OF WATER

FIELD OF THE INVENTION

[0001] This invention relates to a process for transformation of ethanol into ethylene and in particular to a process for dehydration of ethanol.

PRIOR ART

[0002] The reaction of dehydration of ethanol into ethylene is known and has been presented in detail since the end of the 19th century. It is known that this reaction is very endothermic, balanced, and shifted toward ethylene at high temperature. The temperature drop that corresponds to the total conversion of pure ethanol is 380° C. The reference catalyst that is often used is a monofunctional acid catalyst. The gamma-alumina is the most cited catalyst. "The Dehydration of Alcohols over Alumina. I: The Reaction Scheme," H. Knözinger, R. Köhne, Journal of Catalysis (1966), 5, 264-270 is considered to be the basic publication on the works on dehydration of alcohols including ethanol. The zeolites are also used for this application, and in particular ZSM5 from the 1980s, such as, for example, in "Reactions of Ethanol over ZSM-5," S. N. Chaudhuri & al., Journal of Molecular Catalysis 62: 289-295 (1990).

[0003] The U.S. Pat. No. 4,232,179 describes a process for dehydration of ethanol into ethylene in which the heat that is necessary to the reaction is supplied by the introduction into the reactor of a coolant mixed with the feedstock. The coolant is either water vapor that is obtained from an outside source, or an outside stream that comes from the process, or the recycling of a portion of the effluent of the dehydration reactor, i.e., the ethylene that is produced. The introduction of the mixture of the feedstock with said coolant makes it possible to provide the heat that is necessary for keeping the temperature of the catalytic bed at a compatible level with the desired conversion levels. In the case where the coolant is the effluent from the dehydration reactor, a compressor for recycling said effluent is necessary. However, the recycling of the ethylene that is produced by the reaction is a drawback because the introduction of ethylene modifies the balance of the dehydration reaction. In addition, the ethylene participates in secondary oligomerization reactions, transfer of hydrogen and disproportionation of olefins that are reactions of an order that is higher than 0 relative to their reagent. The increase in the ethylene concentration from the beginning of the reaction multiplies the formation of secondary products. The loss of ethylene is therefore more significant, which reflects a lowering of selectivity.

[0004] The patent application WO 2007/134415 A2 describes a process for dehydration of ethanol into ethylene that is improved relative to that of the U.S. Pat. No. 4,232,179 that makes possible a reduced investment cost, owing to a reduced number of pieces of equipment, and a reduced operational cost, owing to the non-use of water vapor external to the process. In this process, at least a portion of the effluent of the dehydration reactor (mixture of ethylene that is produced and water vapor) and the superheated water vapor obtained from the water that is produced by the dehydration of ethanol and condensed in the reactor are used as a coolant and enter within the dehydration reactor by mixing with ethanol. Furthermore, said patent application is silent on the pressure condition that is to be complied with between the ethanol feedstock and the effluent for the purpose of maximizing the heat exchange.

[0005] The U.S. Pat. No. 4,396,789 also describes a process for dehydration of ethanol into ethylene in which the ethanol and the water vapor acting as coolant are introduced into the first reactor at a temperature that is between 400 and 520° C. and at a high pressure of between 20 and 40 atm, in such a way that the effluent produced by the dehydration reaction is drawn off from the last reactor at a pressure that is at least higher than 18 atm, said reaction product, i.e., ethylene, after cooling, being able to undergo the final cryogenic distillation stage without an intermediate compression stage. Said process is also characterized by a heat exchange between said product of the dehydration reaction and the feedstock that is introduced into the first reactor, with said reaction product being used for vaporizing the feedstock that comes into the first reactor. The unreacted ethanol, at least a portion of the water that is formed during the reactions of the process, and the water that is added for the final washing of gases are recycled to ensure the complete conversion of the ethanol.

[0006] One objective of the invention is to provide a process for dehydration of ethanol into ethylene in which the feedstock is introduced into stage a) for vaporization of the feedstock at low pressure, less than the reaction pressure, such that said process does not require any coolant that is external to the process. In particular, the feedstock is introduced into stage a) for vaporization of the feedstock at a pressure that is lower than the pressure of the effluent at the outlet of the last reactor so as to maximize the heat exchange between the feedstock and the effluent that is obtained from the last reactor, i.e., to exchange the entire vaporization enthalpy of the feedstock and the condensation enthalpy of said effluent.

[0007] Another objective of the invention is to provide a process for dehydration of ethanol into ethylene of high purity, whereby said process makes it possible to increase the selectivity of ethylene with a specific consumption per ton of ethylene that is produced that is significantly lowered relative to the processes of the prior art.

SUMMARY AND ADVANTAGE OF THE INVENTION

[0008] The invention describes a process for dehydration of an ethanol feedstock, which comprises a percent by mass of ethanol of between 2 and 55% by weight, into ethylene comprising:

[0009] a) The vaporization of said dilute ethanol feedstock in an exchanger, owing to an exchange of heat with the effluent that is obtained from the last reactor, with said ethanol feedstock being introduced into said vaporization stage at a pressure that is lower than the pressure of the effluent that is obtained from the last reactor,

[0010] b) The compression of said feedstock that is vaporized in a compressor,

[0011] c) The introduction of said vaporized and compressed feedstock, at an entrance temperature of between 350 and 550° C. and at an entrance pressure of between 0.3 and 1.8 MPa, in at least one adiabatic reactor that contains at least one dehydration catalyst and in which the dehydration reaction takes place,

[0012] d) The separation of the effluent that is obtained from the last adiabatic reactor of stage c) into an effluent that comprises ethylene at a pressure that is lower than 1.6 MPa and an effluent that comprises water,

[0013] e) The purification of at least a portion of the effluent that comprises water that is obtained from stage d) and the separation of at least one stream of unconverted ethanol, with no recycling of said stream of purified water that is obtained from said stage e) being done upstream from stage a).

[0014] The process uses an ethanol feedstock that is already diluted and in no case requires recycling of the purified water that is obtained from stage e) upstream from stage a), the water playing the role of diluent and coolant for the dehydration reaction.

[0015] This invention offers the advantage relative to the processes of the prior art for maximizing the heat exchange between the feedstock and the effluent that is obtained from the last reactor, i.e., to exchange the entire vaporization enthalpy of the feedstock and the major portion of the condensation enthalpy of said effluent owing to the introduction of the feedstock into the vaporization stage a) at a pressure that is lower than that of the effluent at the outlet of the last reactor.

DESCRIPTION OF THE INVENTION

[0016] The ethanol feedstock that is treated in the process according to the invention is optionally obtained by a process for the synthesis of alcohol from fossil resources, such as, for example, from carbon, natural gas, and carbon waste (plastic waste, municipal waste, etc.).

[0017] Said feedstock can also advantageously come from non-fossil resources. Preferably, the ethanol feedstock that is treated in the process according to the invention is an ethanol feedstock that is produced from a renewable source that is obtained from biomass and is often called "bioethanol." Said ethanol feedstock is a feedstock that is produced by biological means, preferably by fermentation of sugar obtained from, for example, sugar-producing crops such as sugarcane (saccharose, glucose, fructose and sucrose), beet scraps, or else amylase plants (starch), or lignocellulosic biomass or hydrolyzed cellulose (majority glucose and xylose, galactose), containing variable amounts of water.

[0018] Said feedstock is advantageously obtained by fermentation from three sources: 1) The sucrose from cane sugar or beet scraps, 2) The starch that is present in the grains and the tubers, and 3) The cellulose and hemicellulose that are present in wood, the herbs and other lignocellulosic biomasses, starch, cellulose and hemicellulose having to be hydrolyzed into sugars before undergoing a fermentation stage.

[0019] At the end of fermentation, the fermented ethanol is concentrated in a first column called a "beer column." The ethanol feedstock that is treated in the process according to the invention advantageously comes from this beer column whose concentration in ethanol at the top of the column is compatible with the concentration of said dilute ethanol feedstock that is used in the process, with said feedstock comprising between 2 and 55% by weight of ethanol. The use of a dilute hydrated ethanol feedstock thus makes it possible not to concentrate the ethanol in a conventionally more intense manner in a second so-called "rectification" column that concentrates the ethanol toward its azeotrope.

[0020] For a more complete description of the standard fermenting processes, it is possible to refer to the work 'Les Biocarburants, Etat des lieux, perspectives et enjeux du développement [The Biofuels: Assessment, Perspectives and Development Issues], Daniel Ballerini, Editions Technip.'

[0021] Said feedstock can advantageously also be obtained by fermentation of synthesis gas. Said feedstock can also advantageously be obtained by hydrogenation of the corresponding acids or esters. In this case, the acetic acid or the acetic esters are advantageously hydrogenated using hydrogen in ethanol. The acetic acid can advantageously be obtained by carbonylation of methanol or by fermentation of the carbohydrates.

[0022] Preferably, the ethanol feedstock that is treated in the process according to the invention is an ethanol feedstock that is produced from a renewable source that is obtained from the biomass.

[0023] According to the invention, the ethanol feedstock that is used is an ethanol feedstock that comprises a percent by mass of ethanol that is between 2 and 55% by weight. Said ethanol feedstock is said to be diluted. Preferably, said ethanol feedstock comprises a percent by mass of ethanol of between 2 and 35% by weight. Preferably, said ethanol feedstock is hydrated. Said ethanol feedstock also advantageously comprises, in addition to water, a content of alcohols other than ethanol, such as, for example, methanol, butanol and/or isopentanol that is less than 10% by weight, and preferably less than 5% by weight, a content of oxidized compounds other than the alcohols, such as, for example, ethers, acids, ketones, aldehydes, and/or esters that are advantageously less than 1% by weight, and a nitrogen and sulfur content, organic and mineral, advantageously less than 0.5% by weight, with the percentages by weight being expressed relative to the total mass of the ethanol that is present in said feedstock.

[0024] The ethanol feedstock that is used according to the invention advantageously undergoes a pretreatment stage prior to the vaporization stage a) of said feedstock. Said pretreatment stage makes it possible to eliminate the impurities that are contained in said feedstock in such a way as to limit the deactivation of the dehydration catalyst that is placed downstream, and in particular the metal cations, the metal anions, the compounds that contain nitrogen and the compounds that contain sulfur. The oxidized compounds that are present in said feedstock are not substantially eliminated.

[0025] Said pretreatment stage is advantageously implemented by means that are known to one skilled in the art, such as, for example, the use of at least one resin, by the adsorption of impurities on solids preferably at a temperature of between 20 and 200° C., by a concatenation that comprises a first hydrogenolysis stage that operates at a temperature of between 20 and 200° C., followed by a stage for recovery on acid solid at a temperature of between 20 and 200° C. and/or by distillation. In the case of the use of at least one resin, said resin is preferably acidic and is used at a high temperature of between 20 and 200° C. Said resin can optionally be preceded by a basic resin.

[0026] In the case where the pretreatment stage is implemented by the adsorption of impurities on solids, said solids are advantageously selected from among the molecular sieves, activated carbon, alumina and zeolites.

[0027] Said pretreatment stage of the ethanol feedstock makes it possible to produce a purified ethanol fraction in which the metal and organic impurities have been eliminated, so as to obtain a purified feedstock that responds to the level of impurities that are compatible with the dehydration catalyst.

Stage a)

[0028] According to the invention, the dehydration process comprises a stage a) for vaporization of said ethanol feedstock, optionally pretreated, in an exchanger owing to a heat exchange with the effluent that is obtained from the last adiabatic reactor, with said ethanol feedstock being introduced into said vaporization stage at a pressure that is lower than the pressure of the effluent that is obtained from the last reactor.

[0029] Preferably, said ethanol feedstock being introduced into said vaporization stage at a pressure of between 0.1 and 1.4 MPa [sic].

[0030] Preferably, at least a portion and preferably all of an unreacted ethanol stream that is obtained from stage e) for purification of the effluent that comprises water is also introduced, mixed with said dilute hydrated ethanol feedstock, optionally pretreated, in the exchanger of the vaporization stage a).

[0031] Preferably, said ethanol feedstock is mixed with at least a portion of an unreacted ethanol stream that is obtained from stage e) for purification of the effluent comprising water, after the pretreatment stage of said ethanol feedstock.

[0032] Preferably, said ethanol feedstock, optionally mixed with at least a portion of an unreacted ethanol stream that is obtained from stage e), is introduced into said vaporization stage a) at a pressure that is lower than the pressure of the effluent that is obtained from the last reactor.

[0033] In a preferred manner, said ethanol feedstock that is optionally mixed with at least a portion of an unreacted ethanol stream that is obtained from stage e) is introduced into said vaporization stage a) at a pressure that is less than the pressure of the effluent that is obtained from the last reactor, with said pressure also being between 0.1 and 1.4 MPa.

[0034] Thus, the pressure at which said ethanol feedstock, optionally mixed in said vaporization stage a), is introduced is subjected to two conditions: said pressure is to be lower than the pressure of the effluent that is obtained from the last reactor, and, in this interval, said pressure is also advantageously to be between 0.1 and 1.4 MPa.

[0035] Actually, an essential criterion of this invention is the adjustment of the pressure upstream from the vaporization stage a) of said ethanol feedstock, optionally mixed with at least a portion of an unreacted ethanol stream that is obtained from stage e), in such a way as to maximize the heat exchange between said feedstock, optionally mixed, with said unreacted ethanol stream and the effluent that is obtained from the last adiabatic reactor. The introduction of said ethanol feedstock, optionally mixed with at least a portion of said unreacted ethanol stream, in the vaporization stage a) at this specific pressure level that is lower than the pressure of the effluent that is obtained from the last reactor and preferably between 0.1 and 1.4 MPa, makes it possible to benefit from a vaporization temperature of the possible feedstock mixture that is lower than the condensation temperature of the effluent that is obtained from the last adiabatic reactor. Thus, the major portion of the latent heat and the major portion of the condensation enthalpy of the aqueous phase of the effluent that is obtained from the last adiabatic reactor is recovered for vaporizing said ethanol feedstock, optionally mixed with at least a portion of said unreacted ethanol stream that is obtained from stage e), without an external heat supply. The entire vaporization enthalpy of said ethanol feedstock, optionally mixed with at least a portion of an unreacted ethanol stream that is obtained from stage e), is therefore exchanged with the condensation enthalpy of said effluent.

[0036] The pressure of said ethanol feedstock, optionally mixed with at least a portion of said unreacted ethanol stream that is obtained from stage e), at its vaporization, is advantageously selected in such a way that the temperature difference between the effluent that is obtained from the last adiabatic reactor that is condensed and said feedstock mixture that evaporates is always at least higher than 2° C., and preferably at least higher than 3° C.

Stage b)

[0037] According to the invention, said ethanol feedstock, optionally mixed with at least a portion of said unreacted ethanol stream that is obtained from stage e), vaporized, undergoes compression in a compressor. The compression stage b) is advantageously implemented in any type of compressor that is known to one skilled in the art. In particular, the compression stage b) is advantageously implemented in a compressor of the radial compressor type with an integrated multiplier or in a compressor that comprises one or more fans with a radial wheel that are arranged in series without intermediate cooling.

[0038] The compression stage b) of said ethanol feedstock that is optionally mixed with at least a portion of said unreacted ethanol stream, vaporized, makes it possible to prevent the supply of coolant that is external to the process for ensuring the vaporization of said mixture of said feedstock. Thus, only the streams that are obtained from the process are used. The compression stage b) therefore makes it possible to produce a heat pump that is integrated in said process, using the streams that are obtained from the process, and not involving external coolant.

[0039] The combination of the specific operating conditions of stage a) and stage b) makes it possible to recover the major portion of the latent heat of said effluent and the major portion of the condensation enthalpy of the aqueous phase of the effluent that is obtained from the last adiabatic reactor for vaporizing the entire ethanol feedstock that is optionally mixed with at least a portion of said unreacted ethanol stream that is obtained from stage e), minimalizing the loss of latent heat and the condensation enthalpy by an external cooling and thus minimizing the supply of external heat.

[0040] The pressure of said ethanol feedstock that is optionally mixed with at least a portion of said unreacted ethanol stream that is obtained from stage e), vaporized at the end of the compression stage b), is advantageously between 0.3 and 1.8 MPa. The exit pressure of said optional mixture of said feedstock is adequate for producing the temperature condition that is necessary to the exchange of stage a): in stage a), the vaporization temperature of said optional mixture of said feedstock is to be lower than the condensation temperature of the effluent that is obtained from the last reactor.

[0041] Said ethanol feedstock that is optionally mixed with at least a portion of said unreacted ethanol stream that is obtained from stage e), vaporized and compressed, obtained from compression stage b), is optionally heated in a gas single-phase-type exchanger, owing to a heat exchange with the effluent that is obtained from the last adiabatic reactor of stage c). In said gas single-phase-type exchanger, said optional mixture of said feedstock, vaporized and compressed, is superheated, and the effluent that is obtained, in the gaseous state, from the last adiabatic reactor of stage c) is "de-superheated" without being condensed.

[0042] Said optional mixture of said feedstock is advantageously superheated to a temperature of between 250 and 375° C. and preferably between 280 and 360° C. At the end of said gas single-phase-type exchanger, the effluent that is obtained, in the gaseous state, from the last adiabatic reactor of stage c) advantageously has a temperature of between 150 and 320° C., and preferably between 200 and 300° C.

[0043] Thus, the use of different exchangers, of the gas single-phase type and the gas/liquid vaporizer type, and vaporization, at a pressure that is lower than the pressure of the output effluent of the last reactor, of said ethanol feedstock that is optionally mixed with at least a portion of said unreacted ethanol stream that is obtained from stage e), makes possible the condensation of at least 80% of the water vapors that are present in the effluent that is obtained from the last reactor.

[0044] Said optional mixture of feedstock—vaporized, compressed and optionally heated in said gas single-phase-type exchanger—is next advantageously introduced into a furnace in such a way as to bring it to an entrance temperature in at least one adiabatic reactor that is compatible with the temperature of the dehydration reaction.

Stage c)

[0045] According to the invention, said ethanol feedstock that is optionally mixed with at least a portion of said unreacted ethanol stream that is obtained from stage e), vaporized and compressed, and optionally heated, is introduced at an entrance temperature of between 350 and 550° C. and at an entrance pressure of between 0.3 and 1.8 MPa in at least one adiabatic reactor that contains at least one fixed catalyst bed for dehydration and in which the dehydration reaction takes place.

[0046] The effluent that is obtained from the last adiabatic reactor of stage c) advantageously has, at the outlet of the last adiabatic reactor of stage c), a temperature of between 270 and 450° C., and preferably between 300 and 410° C.

[0047] The effluent that is obtained from the last adiabatic reactor of stage c) advantageously has, at the outlet of the last adiabatic reactor of stage c), a pressure of between 0.2 and 1.6 MPa.

[0048] Said pressure at the outlet of the last adiabatic reactor of stage c) is also advantageously higher than the pressure at which said ethanol feedstock is introduced into said vaporization stage, in such a way as to recover the major portion of the latent heat and the major portion of the condensation enthalpy of the aqueous phase of the effluent that is obtained from the last adiabatic reactor.

[0049] Stage c), in which the dehydration reaction takes place, is advantageously carried out in one or two reactors.

[0050] In the case where stage c) is implemented in an adiabatic reactor, said ethanol feedstock that is optionally mixed with at least a portion of said unreacted ethanol stream that is obtained from stage e), vaporized and compressed, and optionally heated, is advantageously introduced into said reactor at an entrance temperature of between 350 and 550° C., and preferably between 400 and 500° C., and at an entrance pressure of between 0.3 and 1.8 MPa. The effluent that is obtained from said adiabatic reactor advantageously has a temperature of between 270 and 450° C., and preferably between 340 and 420° C., and an exit pressure that is advantageously between 0.2 and 1.6 MPa.

[0051] In the case where stage c) is implemented in two adiabatic reactors, said ethanol feedstock that is optionally mixed with at least an unreacted ethanol stream that is obtained from stage e), vaporized and compressed, and optionally heated, is advantageously introduced into the first reactor at an entrance temperature of between 350 and 550° C. and preferably at a temperature of between 370 and 500° C., and at an entrance pressure of between 0.4 and 1.8 MPa.

[0052] The effluent that is obtained from the first adiabatic reactor advantageously exits from said first reactor at a temperature of between 270 and 450° C., and preferably between 290 and 390, and at a pressure of between 0.3 and 1.7 MPa.

[0053] Said effluent is next advantageously introduced into a furnace in such a way that the entrance temperature of said effluent in the second adiabatic reactor is between 350 and 550° C., and preferably between 400 and 500° C. Said effluent has an entrance pressure in said second reactor that is advantageously between 0.3 and 1.7 MPa.

[0054] The effluent that is obtained from the second adiabatic reactor exits from said second adiabatic reactor at a temperature that is advantageously between 270 and 450° C., and preferably between 340 and 420° C. The exit pressure of said effluent that is obtained from the second adiabatic reactor is advantageously between 0.2 and 1.6 MPa.

[0055] The entrance temperature of the reactor(s) can advantageously be gradually increased to prevent the deactivation of the dehydration catalyst.

[0056] The dehydration reaction that takes place in at least one adiabatic reactor of stage c) of the process according to the invention is advantageously performed at an hourly speed by weight that is between 0.1 and 20 h-1 and preferably between 0.5 and 15 h-1. The hourly speed by weight is defined as being the ratio of the mass flow rate of the pure ethanol feedstock to the mass of the catalyst.

[0057] The dehydration catalyst that is used in stage c) is a catalyst that is known to one skilled in the art. Said catalyst is preferably an amorphous acid catalyst or a zeolitic acid catalyst.

[0058] In the case where the dehydration catalyst that is used in stage c) is a zeolitic catalyst, said catalyst comprises at least one zeolite that is selected from among the zeolites that have at least pore openings containing 8, 10 or 12 oxygen atoms (8 MR, 10 MR or 12 MR). It is actually known to define the size of the pores of the zeolites by the number of oxygen atoms that form the annular cross-section of the channels of the zeolites, called "member ring" or MR in English. In a preferred manner, said zeolitic dehydration catalyst comprises at least one zeolite that has a structural type that is selected from among the structural types MFI, MEL, FAU, MOR, FER, SAPO, TON, CHA, EUO and BEA. Preferably, said zeolitic dehydration catalyst comprises an MFI-structural-type zeolite and in a preferred manner a ZSM-5 zeolite.

[0059] The zeolite that is employed in the dehydration catalyst that is used in stage c) of the process according to the invention can advantageously be modified by dealuminification or desilication according to any method of dealuminification or desilication known to one skilled in the art.

[0060] The zeolite that is employed in the dehydration catalyst that is used in stage c) of the process according to the invention or the final catalyst can advantageously be modified by an agent of the type to attenuate its total acidity and to improve its hydrothermal resistance properties. Preferably, said zeolite or said catalyst advantageously comprises phosphorus, preferably added in phosphate ($^{3-}PO_4$) form followed by a vapor treatment after neutralization of the excess acid by a basic precursor, such as, for example, sodium Na or calcium

Ca. In a preferred manner, said zeolite comprises a phosphorus content of between 0.5 and 4.5% by weight relative to the total mass of the catalyst.

[0061] Preferably, the dehydration catalyst that is used in stage c) of the process according to the invention is the catalyst that is described in the patent applications WO/2009/098262, WO/2009/098267, WO/2009/098268 or WO/2009/098269.

[0062] In the case where the dehydration catalyst that is used in stage c) is an amorphous acid catalyst, said catalyst comprises at least one porous refractory oxide that is selected from among alumina, alumina that is activated by a deposit of mineral acid, and silica-alumina.

[0063] Said amorphous or zeolitic dehydration catalyst that is used in stage c) of the process according to the invention can advantageously also comprise at least one oxide-type matrix that is also called a binder. According to the invention, matrix is defined as an amorphous or poorly crystallized matrix. Said matrix is advantageously selected from among the elements of the group that is formed by clays (such as, for example, among the natural clays such as kaolin or bentonite), magnesia, aluminas, silicas, silica-aluminas, aluminates, titanium oxide, boron oxide, zirconia, aluminum phosphates, titanium phosphates, zirconium phosphates, and carbon. Preferably, said matrix is selected from among the elements of the group that is formed by the aluminas, the silicas, and the clays.

[0064] Said dehydration catalyst that is used in stage c) of the process according to the invention is advantageously shaped in the form of grains of different shapes and sizes. It is advantageously used in the form of cylindrical or multilobar extrudates such as bilobar, trilobar and multilobar extrudates of straight or twisted shape, but it can optionally be manufactured and used in the form of crushed powder, tablets, rings, balls, wheels, or spheres. Preferably, said catalyst is in the form of extrudates.

[0065] Said dehydration catalyst that is used in stage c) of the process according to the invention is advantageously implemented in at least one reactor, in a fixed bed, or in a moving bed.

[0066] In stage c) of the process according to the invention, the catalysts that are used and the operating conditions are selected in such a way as to maximize the production of ethylene. The overall dehydration reaction that is implemented in stage c) of the process according to the invention is as follows:

$$C_2H_5OH \rightarrow CH_2{=}CH_2 + H_2O$$

[0067] The conversion of the ethanol feedstock in stage c) is advantageously greater than 90%, preferably 95%, and in a preferred manner greater than 99%.

[0068] The conversion of the ethanol feedstock is defined, in percentage, by the following formula: [1−(hourly output mass of ethanol/hourly input mass of ethanol)]× x 100.

[0069] The hourly input and output mass of ethanol is measured conventionally by gas phase chromatography of the aqueous phase.

[0070] Stage c), in which the dehydration reaction takes place, is advantageously carried out in one or two reactors. A preferred reactor is a radial reactor that operates in upward or downward mode. During stage c) of the process according to the invention, the transformation of the feedstock is accompanied by the deactivation of the dehydration catalyst by coking and/or by adsorption of inhibiting compounds. The dehydration catalyst is therefore to periodically undergo a regeneration stage. Preferably, the reactor is used in an alternate regeneration mode, also called a swing reactor, so as to alternate the reaction and regeneration phases of said dehydration catalyst. The objective of this regeneration treatment is to burn the organic deposits as well as the radicals that contain nitrogen and sulfur, contained at the surface and within said dehydration catalyst.

[0071] The regeneration of the dehydration catalyst that is used in said stage c) is advantageously carried out by oxidation of coke and inhibiting compounds under a stream of air or in an air/nitrogen mixture, for example by using a recirculation of the combustion air with or without water so as to dilute oxygen and to control regeneration exothermy. In this case, it is possible to advantageously adjust the content of oxygen at the inlet of the reactor by a supply of air. Regeneration takes place at a pressure between atmospheric pressure (0 bar relative) and the reaction pressure. The regeneration temperature is advantageously selected from between 400 and 600° C.; it can advantageously vary during regeneration. The end of the regeneration is detected when there is no longer oxygen consumption, a sign of the total combustion of the coke.

[0072] Preferably, the effluent that is obtained from the last adiabatic reactor of stage c) is not recycled upstream from stage c), in at least one adiabatic reactor.

[0073] The effluent that is obtained from the last adiabatic reactor of stage c) is optionally sent into a gas single-phase-type exchanger in which it is "de-superheated" without being condensed by heat exchange with the vaporized and compressed feedstock that is obtained from stage b), in which it is heated. Said "de-superheated" effluent is next advantageously sent into a second gas/liquid-type exchanger in which it is partially condensed by a heat exchange that is used to evaporate the feedstock.

Stage d)

[0074] According to the invention, the effluent that is obtained from the last adiabatic reactor of stage c) undergoes a separation stage d) into an effluent that comprises ethylene at a pressure that is lower than 1.6 MPa and an effluent that comprises water.

[0075] Stage d) for separation of said effluent that is obtained from the last adiabatic reactor of stage c) can advantageously be implemented by any method that is known to one skilled in the art, such as, for example, by a gas/liquid separation zone, and preferably a gas/liquid separation column. The effluent of the gas/liquid separation zone that comprises ethylene at a pressure that is lower than 1.6 MPa next advantageously undergoes compression. Said compression makes it possible to raise the pressure of said effluent to a pressure that is advantageously between 2 and 4 MPa that is necessary for its final purification.

[0076] Preferably, the effluent that comprises ethylene that is separated at the end of stage d) is not recycled in at least one adiabatic reactor of stage c). The non-recycling of the ethylene that is separated at the end of stage d) in at least one adiabatic reactor of stage c) does not alter the selectivity of ethylene of the process according to the invention.

[0077] At least a portion of the effluent that comprises water that is obtained from stage d) is optionally recycled in separation stage d). In the case where at least a portion of the effluent that comprises water is recycled, said portion of the effluent that comprises water is advantageously cooled using cold fluid or a fluid that is obtained from the process and is preferably purified according to the known purification methods described below.

Stage e)

[0078] According to the invention, at least a portion of the effluent that comprises water that is obtained from separation stage d) undergoes a purification stage e), and with no recycling of said stream of purified water that is obtained being done upstream from stage a).

[0079] The purification stage e) can advantageously be implemented by any purification method that is known to one skilled in the art. By way of example, the purification stage e) can advantageously be implemented by use of ion-exchange resins, molecular sieves, membranes, by adding chemical agents for adjusting the pH, such as, for example, soda or amines, and by adding chemical agents for stabilizing the products, such as, for example, polymerization inhibitors that are selected from among bisulfites and surfactants.

[0080] At least one purified water stream and at least one unconverted ethanol stream are next separated. The separation can advantageously be implemented by any separation method that is known to one skilled in the art. By way of example, the separation can advantageously be implemented by distillation, the use of molecular sieves, membranes, vapor stripping or heat stripping or by absorption with solvent, such as, for example, glycol-containing solvents.

[0081] A stream that contains light gases, preferably acetaldehyde and methanol, can advantageously also be separated.

[0082] At least a portion of said unreacted ethanol stream that is obtained from the purification stage e) of the effluent that comprises water is advantageously recycled and mixed, upstream from vaporization stage a), with the ethanol feedstock that is optionally pretreated.

BRIEF DESCRIPTION OF THE FIGURES

[0083] FIG. **1** diagrammatically shows the process for dehydration of ethanol in the case of the dehydration of a dilute ethanol feedstock that contains between 2 and 35% by weight of ethanol, with the remainder being primarily water.

[0084] The dilute ethanol feedstock is introduced into a pretreatment zone (**2**) via the pipe (**1**). The pretreated ethanol feedstock (**3**) is next mixed in the pipe (**5**) with a portion of the unreacted ethanol stream that is obtained from the purification zone (**20**), via the pipe (**4**). The pretreated ethanol feedstock that is mixed with a portion of the unreacted ethanol stream is introduced via the pipe (**5**) at a pressure of between 0.1 and 1.4 MPa, in a gas/liquid exchanger E**1** in which said mixture undergoes heat exchange with the effluent that is obtained from the last adiabatic reactor R**2** that penetrates the exchanger via the pipe (**11**). The latent heat and/or condensational enthalpy of the effluent that is obtained from the last adiabatic reactor R**2** is used to vaporize the ethanol feedstock that is mixed with the unreacted ethanol stream, without an external heat supply.

[0085] The ethanol feedstock that is mixed with the unreacted ethanol stream, vaporized, is next sent via the pipe (**6**) into a compressor C**1**.

[0086] Said mixture of the feedstock, vaporized and compressed, is next sent via the pipe (**7**) into a gas single-phase-type exchanger E**2**, in which said mixture is heated owing to a heat exchange with the effluent that is obtained from the last adiabatic reactor R**2** that is introduced into E**2** via the pipe (**10**). In said gas single-phase-type exchanger, said vaporized and compressed feedstock is superheated, and the effluent that is obtained, in the gaseous state, from the last adiabatic reactor R**2** is "de-superheated" without being condensed.

[0087] Said mixture of the feedstock—vaporized, compressed and heated in the gas single-phase-type exchanger E**2**—is next introduced into a furnace H**1** via the pipe (**8**) in such a way as to bring it to an entrance temperature in the first adiabatic reactor R**1** that is compatible with the temperature of the dehydration reaction. The effluent that is obtained from the first reactor R**1** is sent into a second furnace H**2** via the pipe (**8*b***) before being introduced into the second reactor R**2** via the pipe (**9*b***).

[0088] The effluent that is obtained from the second reactor R**2** next undergoes the two successive exchanges that are described above in the exchangers E**2** and E**1** via the pipes (**10**) and (**11**).

[0089] The effluent that is obtained from the exchanger E**1** is sent via the pipe (**12**) into a gas/liquid separation column (**13**) where it is separated into an effluent that comprises ethylene (**14**) and an effluent that comprises water (**15**). A portion of the effluent that comprises water is recycled after cooling in the column (**13**) via the pipe (**16**).

[0090] The portion of the effluent that comprises non-recycled water in the column (**13**) is sent via the pipe (**15**) into a purification and separation zone (**20**). At least one stream of purified water (**17**) and at least one stream of unconverted ethanol (**4**) and (**18**) are next separated. A stream that contains the light gases (**19**) is also separated.

[0091] A portion of said unreacted ethanol stream that is obtained from the stage (**20**) for purification of the effluent that comprises water is recycled via the pipe (**4**) and is mixed upstream from the exchanger E**1**, with the pretreated ethanol feedstock (**3**).

[0092] The following examples illustrate the invention without limiting its scope.

EXAMPLES

Example 1

In Accordance with the Invention

[0093] Example 1 illustrates a process according to the invention in which stage c) is implemented in an adiabatic reactor.

[0094] The ethanol feedstock under consideration is produced by fermentation of wheat, without extracting gluten, by a dry-milling-type process according to the English term and only distilled in the first beer column to remain dilute.

[0095] The dilute ethanol feedstock whose composition is provided in Column 1 of Table 1 is pretreated on a resin TA 801 at a temperature of 140° C. The characteristics of the pretreated ethanol feedstock are also provided in Column 2 of Table 1.

Stage a)

[0096] Said dilute and pretreated ethanol feedstock is introduced, at a flow rate of 160,735 kg/h, into a mixture with 132 kg/h of unconverted ethanol that is obtained from stage e), in an exchanger El at a pressure that is equal to 0.31 MPa.

[0097] No stream of purified water that is obtained from stage e) is recycled and mixed with said ethanol feedstock.

TABLE 1

Characteristics of the Ethanol Feedstock Before and After Pretreatment			
	(1)	(2)	Unit
Ethanol Content	26.3	26.3	% by Weight
Acetaldehyde	0.0169	0.0169	% by Weight
Aldehydes	0.0175	0.0175	% by Weight
Esters	0.003	0.003	% by Weight
Higher Alcohols	0.2144	0.2144	% by Weight
Methanol	0.0038	0.0038	% by Weight
1-Propanol	0.0604	0.0604	% by Weight
2-Methyl-1 Propanol	0.0551	0.0551	% by Weight
1 Butanol	0.0018	0.0018	% by Weight
2-Methyl-1 Butanol	0.0256	0.0256	% by Weight
3-Methyl-1 Butanol	0.0715	0.0715	% by Weight
Nitrogen Compounds	0.005	0	% by Weight
Water Content	73.2	73.2	% by Weight

(1): Feedstock Ethanol
(2): After Pretreatment

[0098] For the sake of simplicity, the description of the impurities in the pretreated feedstock was removed from the text below.

[0099] In stage a), the majority of the latent heat of the aqueous phase of the effluent that is obtained from the adiabatic reactor of stage c) is recovered for vaporizing the mixture of the feedstock and the unconverted ethanol, without an external heat supply. Thus, 90.1% of the water that is contained in said effluent that is obtained from the adiabatic reactor of stage c) is in liquid aqueous form. Thus, 88.5 MW is exchanged between the mixture of the feedstock and two other streams and the effluent of the reactor.

[0100] The temperature at the beginning of the vaporization of said feedstock is equal to 126° C. (at 0.27 MPa) and the final condensation temperature of said effluent that is obtained from the adiabatic reactor is—the effluent is—117° C. (at 0.41 MPa) [sic].

Stage b)

[0101] The mixture of the feedstock and the unconverted ethanol, vaporized, obtained from the exchanger, is next compressed in a radial compressor with an integrated multiplier such that the pressure of said mixture of the feedstock and the unconverted ethanol, vaporized at the end of the compression, is equal to 0.63 MPa.

[0102] The mixture of the feedstock and the unconverted ethanol, vaporized and compressed, is next heated in a gas single-phase-type exchanger E**2**, owing to a heat exchange with the effluent that is obtained from the adiabatic reactor of stage c). In said gas single-phase-type exchanger, said mixture of the feedstock and the unconverted ethanol, vaporized and compressed, is superheated to a temperature of 345° C., and the effluent that is obtained, in the gaseous state, of the adiabatic reactor of stage c) is "de-superheated" without being condensed and has a temperature of 269° C.

Stage c)

[0103] Said mixture of the feedstock and the unconverted ethanol—vaporized, compressed and heated in said gas single-phase-type exchanger—is next introduced into a furnace in such a way as to bring it to an entrance temperature in said adiabatic reactor that is compatible with the temperature of the dehydration reaction, i.e., at a temperature of 500° C.

[0104] Said mixture of the feedstock and the unconverted ethanol—vaporized, compressed and heated—is introduced into the adiabatic reactor at an entrance pressure of 0.53 MPa.

[0105] The adiabatic reactor contains a fixed bed of dehydration catalyst, with said catalyst comprising 80% by weight of ZSM-5 zeolite that is treated with H_3PO_4 in such a way that the P_2O_5 content is 3.5% by weight.

[0106] The temperature and pressure conditions of the streams entering and exiting from the adiabatic reactor of stage c) are provided in Table 2:

TABLE 2

Operating Conditions of Dehydration Stage c).			
	Unit	Entrance	Exit
Pressure	MPa	0.53	0.50
Hourly Speed by Weight	h^{-1}	7	7
Reaction Temperature	° C.	500	384

[0107] The conversion of the ethanol feedstock in stage c) is 99.4%.

Stage d)

[0108] The effluent that is obtained from the adiabatic reactor of stage c) next undergoes the two heat exchanges described above and is sent into a gas/liquid separation column. An effluent that comprises ethylene at a pressure that is equal to 0.39 MPa is separated as well as an effluent that comprises water. This separation is carried out by the use of a gas/liquid separation column, with recycling of the water that is produced at the bottom of the column toward the top of the column and after cooling and injection of neutralizing agent.

[0109] The effluent that comprises ethylene next undergoes compression for raising its pressure to 2.78 MPa before its final purification. The separated ethylene is not recycled in said adiabatic reactor.

Stage e)

[0110] A purified water stream and an unconverted ethanol stream as well as a stream that contains light gases are next separated by conventional low-pressure distillation from the raw water.

Stage f)

[0111] A portion of the unconverted ethanol stream is recycled upstream from the vaporization stage a).

[0112] Information regarding the different streams, in kg/h, is given in Tables 3 and 4:

TABLES 3 and 4

Composition of the Primary Streams.					
		Description of the Stream			
		Pretreated Ethanol Feedstock	Stream Entering Into R1	Stream Exiting From R1	Effluent Comprising Ethylene
		Stream No. Corresponding to the Figure			
		3	9	10	14
Total Mass Flow Rate	kg/h	160,735	160,866	160,866	26,076

TABLES 3 and 4-continued

Composition of the Primary Streams.

Mass Flow Rate, by Components	kg/h				
Ethylene		0	0	25,154	25,124
Ethane		0	0	21	21
C3		0	0	88	88
C4		0	0	504	503
Oxidized Compounds (Other than Ethanol)		0	0	110	27
Ethanol		42,559	42,685	267	6
H_2O		118,176	118,182	134,722	307

		Description of the Stream			
		Effluent That Comprises Water	Unconverted Ethanol Recycling	Purged Water	Light Gases
		Stream No. Corresponding to the Figure			
		15	4	17	19
Total Mass Flow Rate	kg/h	134,790	132	134,544	114
Mass Flow Rate, By Components	kg/h				
Ethylene		31	0	0	30.9
Ethane		0	0	0	0.0
C3		0	0	0	0.1
C4		1	0	0	0.5
Oxidized Compounds (Other than Ethanol)		83	0	0	83
Ethanol		261	126	135	0.0
H_2O		134,415	6	134,409	0.0

[0113] The compounds C3 and C4 are C3 and C4 hydrocarbon compounds.

[0114] The selectivity of the process in terms of ethylene is 97%.

[0115] It is calculated in the following way: (Ethylene that is contained in the effluent that comprises ethylene)/(0.61 *amount of converted ethanol) where the amount of converted ethanol is the ethanol that is contained in the pretreated ethanol feedstock that is subtracted from the ethanol that is contained in the streams of purged water and in the effluent that comprises ethylene. 0.61 g is the maximum amount of ethylene that is obtained by dehydrating 1 g of pure ethanol.

[0116] Information on the energy balance of the diagram according to Example 1 in accordance with the invention is given in Table 5:

TABLE 5

Energy Balance

Energy Exchanged Inside the System		Energy Provided to the System by an External Supply		
Amount of Heat Exchanged in the First Exchanger (El)	Amount of Heat Exchanged in the Second Exchanger (E2)	Amount of Heat Exchanged in the Furnace	Electricity Required for Compression	Amount of Heat Extracted on the Gas/Liquid Separation Column
MW	MW	MW	MW	MW
88.5	10.8	15.4	8.4	13.8

[0117] The estimation of the primary energy consumption was carried out by using the following bases:

[0118] Effectiveness of 0.8 on the furnaces

[0119] Effectiveness of 0.375 on the production of electricity.

[0120] The diagram according to Example 1 in accordance with the invention has an equivalent primary energy consumption or a specific consumption of 6 GJ equivalent per ton of ethylene that is produced.

Example 2

In Accordance with the Invention

[0121] Example 2 illustrates a process according to the invention in which stage c) is implemented in two adiabatic reactors.

Stage a)

[0122] The same pretreated ethanol feedstock as the one that is used in Example 1 is introduced at a flow rate of 160,735 kg/h into an exchanger E**1** at a pressure that is equal to 0.31 MPa, mixed with 132 kg/h of unconverted ethanol, obtained from stage e). No flow of purified water obtained from stage e) is recycled and mixed with said feedstock.

Stage b)

[0123] The heat exchange that is described in Example 1 takes place, and the mixture of the feedstock and the unconverted ethanol, vaporized, is next compressed in a compressor of the same type as that of Example 1 in such a way that the pressure of said mixture of the feedstock and the unconverted ethanol, vaporized at the end of the compression, is equal to 0.69 MPa. 90.2% of the water that is contained in the effluent that is obtained from the last reactor is in liquid aqueous form. Thus, 88.9 MW is exchanged between mixing the feedstock and the two streams and the effluent that is obtained from the last reactor.

[0124] The temperature at the beginning of the vaporization of said feedstock is equal to 126° C. (at 0.27 MPa) and the final condensation temperature of said effluent that is obtained from the adiabatic reactor is—the effluent is—117° C. (at 0.41 MPa) [sic].

Stage c)

[0125] The mixture of the feedstock and the unconverted ethanol, vaporized and compressed, is next heated in a gas single-phase-type exchanger E**2** owing to a heat exchange with the effluent that is obtained from the second adiabatic reactor of stage c). In said gas single-phase-type exchanger, the mixture of the feedstock and the unconverted ethanol, vaporized and compressed, is superheated to a temperature of 353° C., and the effluent that is obtained, in the gasesous state, of the adiabatic reactor of stage c) is "de-superheated" without being condensed and has a temperature of 275° C.

[0126] The mixture of the feedstock and the unconverted ethanol—vaporized, compressed and heated in said gas single-phase-type exchanger—is next introduced into a furnace in such a way as to bring it to an entrance temperature in the first adiabatic reactor that is compatible with the temperature of the dehydration reaction, i.e., to a temperature of 400° C.

[0127] The mixture of the feedstock and the unconverted ethanol—vaporized, compressed and heated—is introduced into the first adiabatic reactor at an entrance pressure of 0.62 MPa.

[0128] The effluent that is obtained from the first adiabatic reactor exits from said first reactor at a temperature of 318° C. and is next introduced into a furnace in such a way that the entrance temperature of said effluent in the second adiabatic reactor is 405° C. Said effluent has an entrance pressure in said second reactor of 0.53 MPa.

[0129] The effluent that is obtained from the second adiabatic reactor exits from said second adiabatic reactor at a temperature of 380° C. and at a pressure of 0.50 MPa.

[0130] The two adiabatic reactors each contain a fixed catalyst bed for dehydration, whereby said catalyst is identical in the two reactors and identical to the one that is used in Example 1.

[0131] The temperature and pressure conditions of the streams that enter and exit from the adiabatic reactors of stage c) are provided in Table 6:

TABLE 6

Operating Conditions of Dehydration Stage c).

	Unit	Reactor 1		Reactor 2	
		Entrance	Exit	Entrance	Exit
Pressure	MPa	0.59	0.56	0.53	0.50
Hourly Speed by Weight	h^{-1}	14		14	
Reaction Temperature	° C.	400	318	405	380

[0132] The conversion of the ethanol feedstock at the end of stage c) is 99.4%.

Stage d)

[0133] The effluent that is obtained from the second adiabatic reactor of stage c) next undergoes the two heat exchanges described above and is sent into a gas/liquid separation column. An effluent that comprises ethylene at a pressure that is equal to 0.39 MPa is separated as well as an effluent that comprises water. This separation is carried out by use of a gas/liquid separation column, with recycling of the water that is produced at the bottom of the column toward the top of the column after cooling and injection of a neutralizing agent.

[0134] The effluent that comprises ethylene next advantageously undergoes compression for raising its pressure to 2.78 MPa before its final purification. The separated ethylene is not recycled in the first or the second adiabatic reactor.

Stage e)

[0135] A purified water stream and an unconverted ethanol stream as well as a stream containing light gases are next separated by conventional low-pressure distillation of the raw water.

Stage f)

[0136] A portion of the unconverted ethanol stream is recycled upstream from the vaporization stage a).

[0137] Information regarding the different streams, in kg/h, is given in Tables 7 and 8:

TABLES 7 AND 8

Composition of the Primary Streams.

		Description of the Stream				
		Pretreated Ethanol Feedstock	Stream Entering into R1	Stream Entering into R2	Stream Exiting from R2	Effluent that Comprises Ethylene
Stream No.		3	9	9b	10	14
Total Mass Flow Rate	kg/h	160,735	160,867	160,867	160,867	26,076
Mass Flow Rate, by Components	kg/h					
Ethylene		0	0	17,929	25,155	25,123
Ethane		0	0	10	21	21
C3		0	0	11	88	88
C4		0	0	137	504	503
Oxidized Compounds (Other than Ethanol)		0	0	599	110	27
Ethanol		42,559	42,685	12,183	267	6
H_2O		118,176	118,182	129,998	134,727	307

		Description of the Stream			
		Effluent that Comprises Water	Unconverted Ethanol Recycling	Purged Water	Light Gases
Stream No.		15	4	17	19
Total Mass Flow Rate	kg/h	134,790	132	134,544	114
Mass Flow Rate, by Components	kg/h				
Ethylene		31	0	0	31
Ethane		0	0	0	0
C3		0	0	0	0

TABLES 7 AND 8-continued

Composition of the Primary Streams.				
C4	1	0	0	1
Oxidized Compounds (Other than Ethanol)	83	0	0	83
Ethanol	261	126	135	0
H_2O	134,415	6	134,409	0

[0138] The selectivity of the process in terms of ethylene is 97%. It is calculated in the same way as for Example 1.

[0139] Information regarding the energy balance of the diagram according to Example 1 in accordance with the invention is given in Table 9:

TABLE 9

Energy Balance					
Energy Exchanged Inside the System		Energy Provided to the System by an External Supply			
Amount of Heat Exchanged in the First Exchanger El MW	Amount of Heat Exchanged in the Second Exchanger E2 MW	Amount of Heat Exchanged in the First Furnace MW	Amount of Heat Exchanged in the Second Furnace MW	Electricity Required for Compression MW	Amount of Heat Extracted on the Gas/Liquid Separation Column MW
88.9	9.9	4.6	8.4	9.5	14.0

[0140] The estimation of the primary energy consumption was carried out by using the same bases as for Diagram 1.

[0141] The diagram according to Example 2 in accordance with the invention has an equivalent primary energy consumption or specific consumption of 6 GJ equivalent per ton of ethylene produced.

Example 3

For Comparison

[0142] Example 3 illustrates a process in which the dehydration reaction is implemented in an adiabatic reactor and in which the feedstock, mixed with an unconverted ethanol stream, is introduced at low pressure into the vaporization stage a), and said mixture, vaporized, at the outlet of the exchanger does not undergo compression stage b). In this example, the separated ethylene is not recycled in said adiabatic reactor that contains the dehydration catalyst.

[0143] The same pretreated ethanol feedstock as that used in Example 1 is introduced, with a flow rate of 160,618 kg/h, into an exchanger at a pressure that is equal to 0.65 MPa, mixed with 132 kg/h of unconverted ethanol that is obtained from stage e). The mixture of the ethanol feedstock with the unconverted ethanol is partially vaporized by heat exchange between said mixture and the effluent that is obtained from the adiabatic reactor. Only a portion of the condensation enthalpy of the aqueous phase of the effluent can be used to partially vaporize said mixture of the ethanol feedstock with the unconverted ethanol. Thus, only 33.3% by weight of said mixture is vaporized, and only 12% of the aqueous effluent is condensed, which corresponds to an exchanged heat amount of 31.8 MW. So as to totally vaporize said mixture, an additional amount of heat of 58.2 MW is to be provided by an outside heat source: said partially vaporized mixture is next totally vaporized in an evaporator-type exchanger, using the vapor as a coolant.

[0144] Said partially vaporized mixture, which is then evaporated in said evaporator-type exchanger, is next introduced into a furnace in such a way as to bring it to an entrance temperature in said adiabatic reactor that is compatible with the temperature of the dehydration reaction, i.e., at a temperature of 500° C.

[0145] Said vaporized and heated feedstock is introduced into the adiabatic reactor at an entrance pressure of 0.53 MPa.

[0146] The adiabatic reactor contains a fixed catalyst bed for dehydration, whereby said catalyst is identical to that which is used in Example 1.

[0147] The temperature and pressure conditions in said adiabatic reactor are as follows:

TABLE 10

Operating Conditions of Dehydration Stage c).	Unit	Entrance	Exit
Pressure	MPa	0.53	0.50
Hourly Speed by Weight	h^{-1}	7	7
Reaction Temperature	° C.	500	383

[0148] The conversion of the ethanol feedstock is 99.4%.

[0149] The effluent that is obtained from the adiabatic reactor of stage c) next undergoes the heat exchange described above: it is cooled up to 144° C. and should be cooled in an exchanger that uses an outside refrigerant fluid for reaching 117° C. before being sent into a gas/liquid separation column. This exchanger is a cooler that operates with water. An amount of heat of 68 MW should thus be exchanged between the effluent of the reactor and the refrigerant fluid. An effluent that comprises ethylene at a pressure that is equal to 0.38 MPa is separated as well as an effluent that comprises water. This separation is carried out by the use of a gas/liquid separation column, with recycling of the water that is produced at the bottom of the column toward the top of the column after cooling and injection of neutralizing agent.

[0150] The effluent that comprises ethylene next advantageously undergoes compression for raising its pressure to 2.78 MPa before its final purification. The separated ethylene is not recycled in said adiabatic reactor.

[0151] A purified water stream and an unconverted ethanol stream as well as a stream that contains light gases are next separated by conventional low-pressure distillation from the raw water.

[0152] A portion of the unconverted ethanol stream is recycled upstream from vaporization stage a).

[0153] Information regarding the different streams, in kg/h, is provided in Tables 11 and 12:

TABLES 11 AND 12

Composition of the Primary Streams

		Description of the Stream				
		Pretreated Ethanol Feedstock	Stream Entering into R1	Stream Exiting from R1	Effluent that Comprises Ethylene	Effluent that Comprises Water
Total Mass Flow Rate	kg/h	160,618	160,750	160,750	26,015	134,734
Mass Flow Rate, by Components	kg/h					
Ethylene		0	0	25,088	25,057	31
Ethane		0	0	21	21	0
C3		0	0	88	88	0
C4		0	0	502	502	1
Oxidized Compounds (Other than Ethanol)		0	0	110	28	82
Total Mass Flow Rate	kg/h	160,618	160,750	160,750	26,015	134,734
Ethanol		42,446	42,572	266	6	260
H_2O		118,172	118,178	134,675	314	134,361

		Description of the Stream		
		Unconverted Ethanol Recycling	Purged Water	Light Gases
Total Mass Flow Rate	kg/h	132	134,489	113
Mass Flow Rate, by Components	kg/h			
Ethylene		0	0	31
Ethane		0	0	0
C3		0	0	0
C4		0	0	1
Oxidized Compounds (Other than Ethanol)		0	0	82
Ethanol		126	134	0
H_2O		6	134,355	0

[0154] The selectivity of the process in terms of ethylene is 97%. It is calculated in the same way as for Example 1.

[0155] Information regarding the energy balance of the diagram according to Example 3 that is not in accordance with the invention is given in Table 13:

TABLE 13

Energy Balance

Energy	Energy Provided to the System by an External Supply			
Exchanged Inside the System Amount of Heat Exchanged on the 1st Exchanger MW	Amount of Heat Exchanged on the Evaporator MW	Amount of Heat Exchanged on the Furnace MW	Amount of Heat Extracted on the Cooler MW	Amount of Heat Extracted on the Gas/Liquid Separation Column MW
31.9	58.2	33.0	68.0	13.6

[0156] The estimation of the primary energy consumption was carried out by using the same bases as for Diagram 1, by considering in addition an effectiveness of 0.9 on the vapor production.

[0157] This Diagram 3 has an equivalent primary energy consumption, or a specific consumption, of 15.2 GJ equivalent per ton of ethylene that is produced. The vaporization of the feedstock that is mixed with an unconverted ethanol stream and a purified water stream, carried out in Diagram 1 of Example 1 according to this invention, at low pressure, makes it possible to reduce in a significant way the equivalent primary energy consumption: Diagram 1 had a primary energy consumption of 6 GJ equivalent per ton of ethylene.

Example 4

For Comparison

[0158] Example 4 illustrates a process in which the dehydration reaction is implemented in an adiabatic reactor and in which the feedstock, mixed with an unconverted ethanol stream, is introduced into vaporization stage a), and said mixture, vaporized, at the outlet of the exchanger, does not undergo compression stage b).

[0159] Example 4 is based on the fact that a portion of the effluent that is obtained from the adiabatic reactor, comprising ethylene and water, is compressed and recycled at the inlet of the first reactor, this for the purpose of recycling a portion of the coolant that is the water that is directly in vapor form without condensation and revaporization. This recycling contains ethylene, however, and consequently, secondary reactions of oligomerization, hydrogen transfer, and disproportionation of the olefins will take place in a larger amount on the reactor, leading to an overall loss of ethylene production on the reactor and therefore a reduction in ethylene selectivity.

[0160] A feedstock with 35% by weight of pretreated ethanol is introduced at a rate of 121,508 kg/h in an exchanger at a pressure that is equal to 0.65 MPa, mixed with 128 kg/h of unconverted ethanol. In this example, the recycling of a portion of the effluent of the reactor that contains water to the input of the reactor makes it possible to ensure a dilution rate of the ethanol at the inlet of the reactor that is comparable to the preceding examples. Said mixing of the ethanol feedstock and the unconverted ethanol stream is partially vaporized by heat exchange with the effluent that is obtained from the adiabatic reactor. Only a portion of the condensation enthalpy of the aqueous phase of the effluent can be used for partially vaporizing said mixture. Thus, only 38.4% by weight of said mixture is vaporized and only 22.9% by weight of the aqueous effluent is condensed, which corresponds to an amount of exchanged heat of 34.5 MW.

[0161] Said partially vaporized mixture is next mixed with a portion of the effluent that is obtained from the adiabatic reactor that comprises ethylene and water, previously compressed, whose flow rate is 49,765 kg/h. The supply of heat that is linked to said recycled and compressed effluent is not adequate to vaporize the entire mixture of the ethanol feedstock mixed with the unconverted ethanol: 68% by weight of said mixture is vaporized. So as to totally vaporize said mixture, it is necessary to provide an additional 31.5 MW by an external heat source: said partially vaporized mixture is next vaporized totally in an evaporator-type exchanger that uses vapor as a coolant.

[0162] Said mixture that is vaporized and heated in said evaporator-type exchanger is next introduced into a furnace in such a way as to bring it to an entrance temperature in said adiabatic reactor that is compatible with the temperature of the dehydration reaction, i.e., to a temperature of 490° C.

[0163] Said vaporized and heated feedstock is introduced into the adiabatic reactor at an entrance pressure of 0.53 MPa.

[0164] The adiabatic reactor contains a fixed catalyst bed for dehydration, whereby said catalyst is identical to the one that is used in Example 1.

[0165] The temperature and pressure conditions in said adiabatic reactor are as follows:

TABLE 14

Operating Conditions.

	Unit	Entrance	Exit
Pressure	MPa	0.53	0.50
Hourly Speed by Weight	h^{-1}	7	7
Reaction Temperature	° C.	490	393

[0166] The conversion of the ethanol feedstock is 99.5%.

[0167] The effluent that is obtained from the adiabatic reactor next undergoes the heat exchange that is described above, and is cooled to 117° C. by an outside source before being sent into a gas/liquid separation column. This exchanger can be a cooler that operates with water. An amount of heat of 37.3 MW should thus be exchanged between the effluent of the reactor and the refrigerant fluid. An effluent that comprises ethylene at a pressure that is equal to 0.43 MPa is separated as well as an effluent that comprises water. This separation is carried out by the use of a gas/liquid separation column, with recycling of the water that is produced at the bottom of the column to the top of the column after cooling and injection of neutralizing agent.

[0168] The effluent that comprises ethylene next advantageously undergoes compression for raising its pressure to 2.78 MPa before its final purification. The separated ethylene is not recycled in said adiabatic reactor.

[0169] A purified water stream and an unconverted ethanol stream as well as a stream that contains light gases are next separated by conventional low-pressure distillation of the raw water.

[0170] An unconverted ethanol stream is recycled upstream from the vaporization stage a).

[0171] Information regarding the different streams, in kg/h, is given in Table 15:

TABLE 15

Composition of the Primary Streams.

		Description of the Streams				
		Pretreated Ethanol Feedstock	Exchanger Entrance Recombined Load	Recycling of the Effluent Obtained from the Reactor	Stream Entering the Reactor	Stream Exiting from the Reactor
Total Mass Flow Rate	kg/h	121,508	121,636	49,765	171,401	171,400
Mass Flow Rate, by Components	kg/h					

TABLE 15-continued

Composition of the Primary Streams.					
Ethylene	0	0	9,433	9,433	32,488
Ethane	0	0	8	8	28
C3	0	0	264	264	909
C4	0	0	753	753	2,595
Oxidized Compounds (Other than Ethanol)	0	0	28	28	98
Ethanol	42,451	42,573	91	42,664	313
H_2O	79,057	79,063	39,111	118,174	134,707
C4+	0	0	77	77	264

		Description of the Streams			
		Effluent Going to the Exchanger	Effluent that Comprises Ethylene	Effluent that Comprises Water	Recycling of Unconverted Ethanol
otal Mass Flow Rate	kg/h	121,636	26,012	95,624	128
Mass Flow Rate, by Components	kg/h				
Ethylene		23,056	23,025	30	0
Ethane		20	20	0	0
C3		645	644	1	0
C4		1,842	1,840	2	0
Oxidized Compounds (Other than Ethanol)		69	23	47	0
Ethanol		222	6	216	122
H_2O		95,596	269	95,327	6
C4+		187	185	2	0

Description of the Streams		Purged Water	Light Gases
Total Mass Flow Rate	kg/h	95,414	82
Mass Flow Rate, by Components	kg/h		
Ethylene		0	30
Ethane		0	0
C3		0	1
C4		0	2
Oxidized Compounds (Other than Ethanol)		0	47
Ethanol		95	0
H_2O		95,321	0
C4+		0	2

[0172] The selectivity of the process in terms of ethylene is 89%. It is calculated in the same way as for Example 1. The loss of selectivity that is linked to the recycling of the effluent that is obtained from the adiabatic reactor comprising ethylene and water is noted, with the preceding diagrams not implementing the recycling of said effluent that comprises ethylene, making it possible to obtain selectivity in terms of ethylene of 97%.

[0173] Information regarding the energy balance of the diagram according to Example 4 that is not in accordance with the invention is given in Table 16:

TABLE 16

Energy Balance					
	Energy Provided to the System by an External Supply				
Energy Exchanged Inside the System Amount of Heat Exchanged on the First Exchanger MW	Amount of Heat Exchanged on the Evaporator MW	Amount of Heat Exchanged on the Furnace MW	Electricity Required for the Compressor MW	Amount of Heat Extracted on the Cooler MW	Amount of Heat Extracted on the Gas/Liquid Separation Column MW
34.5 Internal Source	31.5 External Source	34.8 External Source	0.37 External Source	37.3 External Source	13.6 External Source

[0174] The estimation of the primary energy consumption was carried out by using the same bases as for Diagram 1, by considering in addition an effectiveness of 0.9 on the vapor production.

[0175] This Diagram 4 has an equivalent primary energy consumption or a specific consumption of 13.5 GJ equivalent per ton of ethylene that is produced. The vaporization of the feedstock mixed with an unconverted ethanol stream, carried out in Diagram 1 of Example 1 according to this invention, at low pressure, makes it possible to reduce in a significant way the equivalent primary energy consumption: Diagram 1 had a primary energy consumption of 6 GJ equivalent per ton of ethylene.

[0176] Without further elaboration, it is believed that one skilled in the art can, using the preceding description, utilize the present invention to its fullest extent. The preceding preferred specific embodiments are, therefore, to be construed as merely illustrative, and not limitative of the remainder of the disclosure in any way whatsoever.

[0177] The entire disclosures of all applications, patents and publications, cited herein and of corresponding French application No. 11/03075, filed Oct. 7, 2011, are incorporated by reference herein.

[0178] The preceding examples can be repeated with similar success by substituting the generically or specifically described reactants and/or operating conditions of this invention for those used in the preceding examples.

[0179] From the foregoing description, one skilled in the art can easily ascertain the essential characteristics of this invention and, without departing from the spirit and scope thereof, can make various changes and modifications of the invention to adapt it to various usages and conditions.

1. Process for dehydration of an ethanol feedstock, which comprises a percent by mass of ethanol of between 2 and 55% by weight, into ethylene comprising:

a) The vaporization of said dilute ethanol feedstock in an exchanger, owing to an exchange of heat with the effluent that is obtained from the last reactor, said ethanol feedstock being introduced into said vaporization stage at a pressure that is lower than the pressure of the effluent that is obtained from the last reactor,

b) The compression of said feedstock that is vaporized in a compressor,

c) The introduction of said vaporized and compressed feedstock, at an entrance temperature of between 350 and 550° C. and at an entrance pressure of between 0.3 and 1.8 MPa, in at least one adiabatic reactor that contains at least one dehydration catalyst and in which the dehydration reaction takes place,

d) The separation of the effluent that is obtained from the last adiabatic reactor of stage c) into an effluent that comprises ethylene at a pressure that is lower than 1.6 MPa and an effluent that comprises water,

e) The purification of at least a portion of the effluent that comprises water that is obtained from stage d) and the separation of at least one stream of unconverted ethanol, with no recycling of said stream of purified water that is obtained from said stage e) being done upstream from stage a).

2. Process according to claim **1**, in which said ethanol feedstock is an ethanol feedstock that is produced from a renewable source that is obtained from biomass or waste containing hydrocarbon or carbohydrate components.

3. Process according to claim **1**, in which said ethanol feedstock comprises a percent by mass of ethanol of between 2 and 35% by weight.

4. Process according to claim **1**, in which at least one unreacted ethanol stream that is obtained from stage e) for purification of the effluent that comprises water is also introduced into the exchanger of vaporization stage a).

5. Process according to claim **1**, in which said ethanol feedstock is also introduced into said stage a) for vaporization at a pressure of between 0.1 and 1.4.

6. Process according to claim **1**, in which the pressure of the ethanol feedstock, vaporized and compressed at the end of compression stage b), is between 0.3 and 1.8 MPa.

7. Process according to claim **1**, in which said ethanol feedstock, vaporized and compressed and obtained from compression stage b), is heated in a gas single-phase-type exchanger, owing to a heat exchange with the effluent that is obtained from the last adiabatic reactor of stage c).

8. Process according to claim **1**, in which the effluent that is obtained from the last adiabatic reactor of stage c) has a temperature of between 270 and 450° C. at the outlet of the last adiabatic reactor of stage c).

9. Process according to claim **1**, in which the effluent that is obtained from the last adiabatic reactor of stage c) has a pressure of between 0.2 and 1.6 MPa at the outlet of the last adiabatic reactor of stage c).

10. Process according to claim **1**, in which in stage c), the dehydration reaction is carried out in one or two reactors.

11. Process according to claim **1**, in which in stage c), the dehydration reaction is carried out in fixed-bed reactors characterized by a downward or upward flow.

12. Process according to claim **1**, in which in stage c), the dehydration reaction takes place, is carried out in at least one fixed-bed or moving-bed reactor that is characterized by a radial flow.

13. Process according to claim **1**, in which said dehydration catalyst that is used in stage c) is an amorphous acid catalyst or a zeolitic acid catalyst.

14. Process according to claim **1**, in which at least a portion of said unreacted ethanol stream that is obtained from stage e) is recycled and mixed with the ethanol feedstock that is upstream from stage a) for vaporization of said feedstock.

* * * * *

Tabel L2.4 Entalpi Steam yang digunakan

Steam / Uap	P = 600 kPA	T = 480 ^{o}C = 753.15 K
H vap (Kj/Kg)	3457.79	

L1.2.2 Patent US 009663414B2 (2017)

US009663414B2

(12) **United States Patent**
Coupard et al.

(10) **Patent No.: US 9,663,414 B2**
(45) **Date of Patent: May 30, 2017**

(54) **PROCESS FOR DEHYDRATION OF ETHANOL TO ETHYLENE AT LOW ENERGY CONSUMPTION**

(71) Applicants: **IFP ENERGIES NOUVELLES**, Rueil-Malmaison (FR); **TOTAL RESEARCH & TECHNOLOGY FELUY**, Seneffe (BE)

(72) Inventors: **Vincent Coupard**, Villeurbanne (FR); **Natacha Touchais**, Vienne (FR); **Thomas Plennevaux**, Lyons (FR); **Emilie Kobel**, Chasse sur Rhone (FR); **Stephanie Fleurier**, Lyons (FR); **Walter Vermeiren**, Houthalen-Helchteren (BE); **Delphine Minoux**, Nivelles (BE); **Philip De Smedt**, Sint-Niklaas (BE); **Cindy Adam**, Wierde (BE); **Nikolai Nesterenko**, Nivelles (BE)

(73) Assignees: **TOTAL RESEARCH & TECHNOLOGY FELUY**, Seneffe (BE); **IFP ENERGIES NOUVELLES**, Rueil-Malmaison (FR)

(*) Notice: Subject to any disclaimer, the term of this patent is extended or adjusted under 35 U.S.C. 154(b) by 30 days.

(21) Appl. No.: **14/646,877**

(22) PCT Filed: **Nov. 18, 2013**

(86) PCT No.: **PCT/FR2013/052767**
§ 371 (c)(1),
(2) Date: **May 22, 2015**

(87) PCT Pub. No.: **WO2014/083260**
PCT Pub. Date: **Jun. 5, 2014**

(65) **Prior Publication Data**

US 2015/0299068 A1 Oct. 22, 2015

(30) **Foreign Application Priority Data**

Nov. 27, 2012 (FR) 12 03201

(51) **Int. Cl.**
C07C 1/24 (2006.01)
C07C 29/76 (2006.01)
C07C 41/09 (2006.01)

(52) **U.S. Cl.**
CPC ***C07C 1/24*** (2013.01); ***C07C 29/76*** (2013.01); ***C07C 41/09*** (2013.01);
(Continued)

(58) **Field of Classification Search**
CPC C07C 1/24; C07C 29/76; C07C 41/09; C07C 11/04; C07C 31/08; C07C 43/06;
(Continued)

(56) **References Cited**

U.S. PATENT DOCUMENTS

4,396,789 A * 8/1983 Barrocas C07C 1/24 585/639
9,000,236 B2 4/2015 Minoux et al.
(Continued)

FOREIGN PATENT DOCUMENTS

WO 2010060981 A1 6/2010

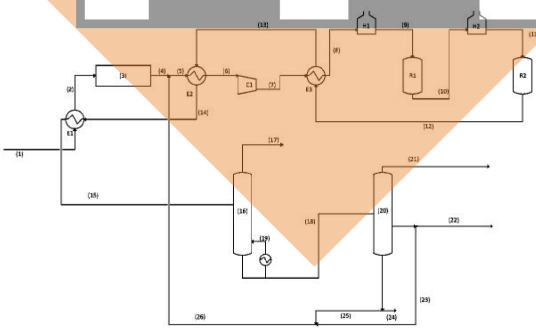
OTHER PUBLICATIONS

International Search Report and Search Opinion from PCT/FR2013/052767 dated Apr. 28, 2014.

Primary Examiner — Sharon Pregler
(74) *Attorney, Agent, or Firm* — Millen White Zelano and Branigan, PC; Csaba Henter; Anthony Zelano

(57) **ABSTRACT**

A process for dehydration of an ethanol feedstock to ethylene by:
a) preheating ethanol feedstock by heat exchange with effluent from e),
(Continued)



b) pretreating the ethanol feedstock to produce pretreated ethanol feedstock,
c) vaporizing a vaporization feedstock containing pretreated ethanol feedstock and at least a portion of the flow of treated water recycled in an exchanger to produce a vaporized feedstock,
d) compressing said vaporized feedstock to produce a compressed feedstock,
e) dehydrating said compressed feedstock in at least one adiabatic reactor,
f) separating the effluent from the last adiabatic reactor of e) into an effluent containing ethylene and an effluent containing water,
g) purifying at least a portion of the effluent containing water from 0 and separating at least one flow of treated water and at least one flow of unconverted ethanol,
h) recycling at least a portion of the flow of treated water from g) upstream of c).

19 Claims, 1 Drawing Sheet

(52) **U.S. Cl.**
CPC *C07C 2527/173* (2013.01); *C07C 2529/40* (2013.01); *C07C 2529/85* (2013.01)

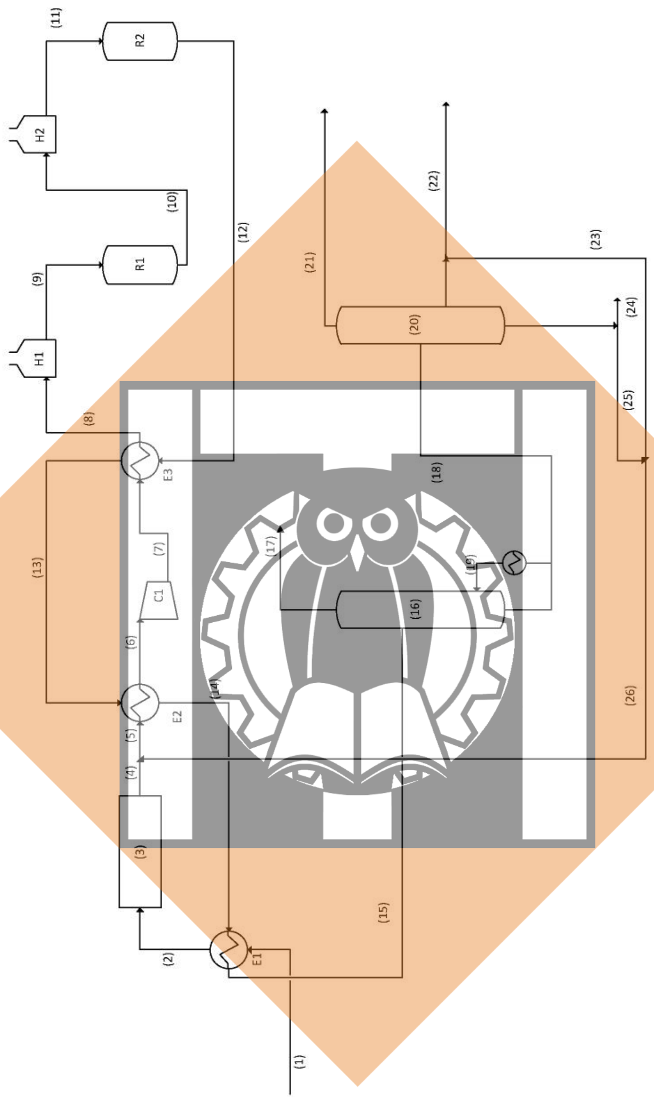
(58) **Field of Classification Search**
CPC C07C 2527/173; C07C 2529/40; C07C 2529/85
USPC 585/638, 639, 640, 641
See application file for complete search history.

(56) **References Cited**

U.S. PATENT DOCUMENTS

9,085,502 B2 * 7/2015 Coupard C07C 1/24 585/639
2011/0313213 A1 12/2011 Minoux et al.

* cited by examiner


R1
R2
H1
H2
E1
E2
E3
C1
(1)
(2)
(3)
(4)
(5)
(6)
(7)
(8)
(9)
(10)
(11)
(12)
(13)
(14)
(15)
(16)
(17)
(18)
(19)
(20)
(21)
(22)
(23)
(24)
(25)
(26)

PROCESS FOR DEHYDRATION OF ETHANOL TO ETHYLENE AT LOW ENERGY CONSUMPTION

FIELD OF THE INVENTION

The present invention relates to a process for converting ethanol to ethylene and in particular to a process for dehydration of ethanol.

PRIOR ART

The dehydration reaction of ethanol to ethylene has been known and described in detail since the end of the 19th century. "The Dehydration of Alcohols over Alumina. I: The reaction scheme", H. Knözinger, R. Köhne, Journal of Catalysis (1966), 5, 264-270 is regarded as the basic publication for operations of dehydration of alcohols, including ethanol. It is known that this reaction is very endothermic, equilibrated and displaced towards ethylene at high temperature. The temperature drop corresponding to the complete conversion of pure ethanol in an adiabatic reactor is 380° C. At lower temperature, ethanol is converted to diethyl ether (DEE). This reaction "intermediate" may be present in processes for dehydration of ethylene in which the conversion is partial, or between two reactors in multi-reactor processes. DEE can then be converted to ethylene at higher temperature. The reference catalyst often used is a monofunctional acid catalyst, gamma alumina being the catalyst mentioned most. Zeolites are also used for this application, in particular ZSM5 since the 1980s, for example in "Reactions of ethanol over ZSM-5", S. N. Chaudhuri et al., Journal of Molecular Catalysis 62: 289-295 (1990).

U.S. Pat. No. 4,232,179 describes a process for dehydration of ethanol to ethylene in which the heat required for the reaction is supplied by introducing into the reactor a heat-transfer fluid mixed with the feedstock. The heat-transfer fluid is either steam originating from an external source, or an external flow originating from the process, or from recycling a portion of the effluent from the dehydration reactor, i.e. the ethylene produced. Introduction of the mixture of the feedstock with said heat-transfer fluid makes it possible to supply the heat required for maintaining the temperature of the catalyst bed at a level compatible with the desired conversions. In the case where the heat-transfer fluid is the effluent from the dehydration reactor, a compressor for recycling said effluent is necessary. However, the recycling of the ethylene produced by the reaction is a drawback, as the introduction of the ethylene alters the equilibrium of the dehydration reaction. Moreover, ethylene participates in the secondary reactions of oligomerization, of hydrogen transfer and of disproportionation of the olefins, which are reactions of an order greater than 0 with respect to their reagent. The increase in the ethylene concentration from the start of the reaction increases the formation of by-products. The loss of ethylene is therefore greater, which is reflected in a reduction in selectivity.

Patent application WO 2007/134415 describes a process for dehydration of ethanol to ethylene that is improved compared with that of U.S. Pat. No. 4,232,179, making reduced capital expenditure possible, owing to a reduced number of items of equipment and reduced operating costs, because it does not use steam external to the process. In this process, at least a portion of the effluent from the dehydration reactor (mixture of ethylene produced and steam) and superheated steam obtained from the water produced by the dehydration of the ethanol and condensed in the reactor are used as heat-transfer fluid and enter the dehydration reactor mixed with the ethanol. Said patent application says nothing regarding the pressure condition to be respected between the ethanol feedstock and the effluent with the aim of maximizing heat exchange.

U.S. Pat. No. 4,396,789 also describes a process for dehydration of ethanol to ethylene in which ethanol and steam acting as heat-transfer fluid are introduced into the first reactor at a temperature comprised between 400 and 520° C. and at a high pressure comprised between 20 and 40 atm, so that the effluent produced by the dehydration reaction is withdrawn from the last reactor at a pressure at least above 18 atm, said reaction product, i.e. ethylene, being able to undergo, after cooling, the final cryogenic distillation step without an intermediate compression step. Said process is also characterized by heat exchange between said product of the dehydration reaction and the feedstock introduced into the first reactor, said reaction product being used to vaporize the feedstock entering the first reactor. The unconverted ethanol, at least a portion of the water formed during the reactions of the process and the water added for the final scrubbing of the gases are recycled to ensure complete conversion of the ethanol.

Patent application WO 2011/002699 discloses a process for dehydration of an ethanol feedstock to ethylene comprising vaporization of a mixture of ethanol and water and reaction of this mixture in an adiabatic reactor. This application does not address the problem of maximizing heat recovery in order to reduce the energy consumption of the process.

An objective of the invention is to provide a process for dehydration of ethanol to ethylene in which the ethanol feedstock is pretreated using an acidic solid in order to limit the quantity of organic nitrogen, which shortens the catalyst's life, and to convert the ethanol partially to DEE.

An objective of the invention is to provide a process for dehydration of ethanol to high-purity ethylene, said process making it possible to increase the selectivity for ethylene with a specific consumption per tonne of ethylene produced that is lowered significantly compared with the processes of the prior art, as it does not require a heat-transfer fluid external to said process.

SUMMARY AND BENEFIT OF THE INVENTION

The invention describes a process for dehydration of an ethanol feedstock to ethylene comprising in particular a step of pretreatment which reduces the level of organic or basic nitrogen contained in said feedstock and converts a fraction of the ethanol to DEE, and a step of vaporizing the pretreated ethanol feedstock, mixed with at least a portion of a flow of recycled treated water, in an exchanger by means of heat exchange with the effluent from the last dehydration reactor.

Said invention offers the advantage, over the processes of the prior art, of increasing the cycle time of the ethanol dehydration catalyst by trapping the cationic or anionic impurities, the basic, complexing, and chelating impurities, the inorganic or organic impurities, such as for example the nitrogen present in the feedstock in basic form, for example in the form of ammonia and/or organic and basic species, for example in the form of amine, amide, imine or nitrile during the pretreatment step. Trapping the nitrogen-containing compounds has in particular the effect of improving the activity of the acid catalysts used in dehydration.

The present invention also offers the advantage, over the processes of the prior art, of maximizing the heat exchange between the feedstock and the effluent from the last dehydration reactor, i.e. of exchanging all of the enthalpy of vaporization of the feedstock and most of the enthalpy of condensation of said effluent owing to the introduction of the feedstock in the vaporization step c) at a pressure below the pressure of the effluent leaving the last reactor.

The applicant discovered, surprisingly, that said step of pretreatment carried out under the operating conditions according to the invention led to partial conversion of ethanol to DEE and made it possible to reduce the energy consumption of ethylene production significantly.

DESCRIPTION OF THE INVENTION

The invention relates to a process for dehydration of an ethanol feedstock to ethylene comprising:

a) a step of preheating said ethanol feedstock to a temperature comprised between 100 and 130° C. by heat exchange with the effluent from step e),
b) a step of pretreating the ethanol feedstock on an acidic solid operating at a temperature comprised between 100 and 130° C. so as to produce a pretreated ethanol feedstock,
c) a step of vaporizing a vaporization feedstock comprising said pretreated ethanol feedstock and at least a portion of the flow of treated water recycled according to step h) in an exchanger by means of heat exchange with the effluent from the last reactor of step e), said vaporization feedstock being introduced into said vaporization step at a pressure comprised between 0.1 and 1.4 MPa so as to produce a vaporized feedstock,
d) a step of compressing said vaporized feedstock in a compressor so as to produce a compressed feedstock,
e) a step of dehydrating said compressed feedstock in at least one adiabatic reactor containing at least one dehydration catalyst and in which the dehydration reaction takes place, operating at an inlet temperature comprised between 350 and 550° C. and at an inlet pressure comprised between 0.3 and 1.8 MPa,
f) a step of separating the effluent from the last adiabatic reactor of step e) into an effluent comprising ethylene at a pressure below 1.6 MPa and an effluent comprising water,
g) a step of purifying at least a portion of the effluent comprising water from step f) and separating at least one flow of treated water and at least one flow of unconverted ethanol,
h) a step of recycling at least a portion of the flow of treated water from step g) upstream of step c).

Feedstock

According to the invention, the feedstock treated in the dehydration process is an ethanol feedstock.

Said ethanol feedstock is advantageously a concentrated ethanol feedstock. By concentrated ethanol feedstock is meant an ethanol feedstock comprising a percentage by weight of ethanol greater than or equal to 35% by weight. Preferably, said concentrated ethanol feedstock comprises a percentage by weight of ethanol comprised between 35 and 99.9% by weight.

The ethanol feedstock comprising less than 35% by weight of ethanol can be concentrated by any means known to a person skilled in the art, for example by distillation, absorption, or pervaporation.

Said ethanol feedstock also advantageously comprises, in addition to water, a content of alcohols other than ethanol, such as for example methanol, butanol and/or isopentanol, below 10% by weight, and preferably below 5% by weight, a content of oxygenated compounds other than alcohols, such as for example ethers, acids, ketones, aldehydes and/or esters, below 1% by weight and a content of nitrogen and of sulphur, organic and mineral, below 0.5% by weight, the percentages by weight being expressed relative to the total weight of said feedstock.

The ethanol feedstock treated in the process according to the invention is optionally obtained by a process of synthesis of alcohol starting from fossil resources such as for example from coal, natural gas or carbon-containing waste.

Said feedstock can also advantageously originate from non-fossil resources. Preferably, the ethanol feedstock treated in the process according to the invention is an ethanol feedstock produced from a renewable source obtained from biomass, often called "bioethanol". Bioethanol is a feedstock produced by a biological route, preferably by fermentation of sugars obtained for example from crops of sugar-containing plants such as sugar cane (saccharose, glucose, fructose and sucrose), beets, or else from amylaceous plants (starch) or from lignocellulosic biomass or from hydrolysed cellulose (predominantly glucose, and xylose, galactose), containing variable quantities of water.

For a more complete description of the classical fermentation processes, reference may be made to the work "Les Biocarburants, État des lieux, perspectives et enjeux du développement" [Biofuels, appraisal, prospects and development challenges], Daniel Ballerini, Editions Technip.

Said feedstock can also advantageously be obtained from synthesis gas.

Said feedstock can also advantageously be obtained by hydrogenation of the corresponding acids or esters. In this case, acetic acid or acetic esters are advantageously hydrogenated with hydrogen to ethanol. Acetic acid can advantageously be obtained by carbonylation of methanol or by fermentation of carbohydrates.

Preferably, the ethanol feedstock treated in the process according to the invention is an ethanol feedstock produced from a renewable source obtained from biomass.

Preheating Step a)

According to the invention, the ethanol feedstock undergoes a preheating step a) in a heat exchanger so as to produce a preheated ethanol feedstock, by means of heat exchange with the effluent from the dehydration step e) to bring it under the required temperature conditions, between 100 and 130° C., preferably between 110° C. and 130° C., for the pretreatment step b). The pressure of the ethanol feedstock is adjusted, in such a way that the latter is still liquid at the end of the preheating step a), to a value comprised between 0.1 and 3 MPa.

Pretreatment Step b)

According to the invention, the preheated ethanol feedstock undergoes a pretreatment step b) so as to produce a pretreated ethanol feedstock. Said pretreatment step makes it possible to remove the nitrogen-containing compounds present in said preheated feedstock so as to limit the deactivation of the dehydration catalyst located downstream.

Said pretreatment step b) is carried out on an acidic solid, preferably an acid resin, and at a temperature comprised between 100 and 130° C., preferably between 110° C. and 130° C.

Said pretreatment step b) makes it possible to remove the basic and/or organic impurities, and the cationic species, in order to obtain a pretreated ethanol feedstock corresponding to the level of impurities compatible with the dehydration catalyst.

The pretreatment on the acidic solid under the operating conditions according to the invention makes it possible to convert between 3% by weight and 20% by weight, preferably between 8 and 12% by weight of the ethanol present in said feedstock to DEE, the percentage by weight being determined relative to the total weight of ethanol present in said feedstock at the inlet of the pretreatment step b).

The acidic solid includes all the acidic solids known to a person skilled in the art: silica-aluminas, acid clays, zeolites, sulphated zirconias, acid resins, etc. The main thing is that the acidic solid has a high exchange capacity for capturing, as far as possible, the basic and cationic species and an acidity strength high enough to carry out the partial conversion of ethanol to DEE.

Acidic solids that are commonly available commercially are clays treated with acids to make them acidic (such as montmorillonite) and zeolites, having a ratio of silica to alumina in the crystal lattice from 2.5 to 100 molar. The acid resins comprise sulphonic groups, grafted on an organic support composed of aromatic and/or haloaliphatic chains. Preferably the acidic solids have an exchange capacity of at least 0.1 mmol H^+ equivalent per gram.

The acid resin includes acidic sulphonic groups and is prepared by polymerization or co-polymerization of aromatic vinyl groups followed by sulphonation, said aromatic vinyl groups being selected from styrene, vinyl toluene, vinyl naphthalene, vinyl ethyl benzene, methyl styrene, vinyl chlorobenzene and vinyl xylene, said resin having a level of cross-linking comprised between 20 and 35%, preferably between 25 and 35% and preferably equal to 30% and an acid strength, determined by potentiometry on neutralization with a KOH solution, from 0.2 to 6 mmol H^+ equivalent per gram and preferably between 0.2 and 2.5 mmol H^+ equivalent per gram.

Said acidic ion-exchange resin contains between 1 and 2 sulphonic end groups per aromatic group. Its size is comprised between 0.15 and 1.5 mm. By size of the resin is meant the diameter of the smallest sphere circumscribing the particle of resin. The classes of resin size are measured by sieving through suitable sieves according to a technique known to a person skilled in the art.

A preferred resin is a resin consisting of aromatic monovinyl and aromatic polyvinyl copolymers, and very preferably a copolymer of divinyl benzene and polystyrene having a level of cross-linking comprised between 20 and 45%, preferably between 30 and 40%, and preferably equal to 35% and an acid strength, representing the number of active sites of said resin, determined by potentiometry on neutralization with a KOH solution, comprised between 1 and 10 mmol H^+ equivalent per gram and preferably comprised between 3.5 and 6 mmol H^+ equivalent per gram. For example, the resin is a TA801 resin sold by the company Axens.

The acidic solids can be regenerated from time to time once the exchange capacity is almost saturated by adsorption of basic and cationic species in situ or ex situ. In the case of inorganic acidic solids such as clays and zeolites, regeneration can consist of simple heating at high temperature in order to desorb the basic species in the presence of an inert or oxygen-containing flow. The cations can be removed by ion exchange. The acid resins can be regenerated by ion exchange, typically by liquid-phase treatment with an acid. The acidic solids can also be used once until saturation and replaced with virgin solid.

The acidic solid can be used alone or mixed with other types of acidic solids. Mixtures of different acidic solids or sequences of acidic solids can be used in order to optimize the capacity for adsorbing the basic and cationic species and the capacity for partial conversion of ethanol to DEE.

The pretreatment described above can advantageously be supplemented with a pretreatment using an anion exchange resin. This resin can for example be a resin loaded with sodium, or trimethylammonium, characterized by an exchange capacity measured in mg(OH^-)/litre. This resin can for example be the resin Amberlite IRN78. This additional resin makes it possible to retain the sulphate ions SO_4^{2-} in order to prolong the catalyst's life.

Vaporization Step c)

The mixture comprising said pretreated ethanol feedstock and at least a portion of the flow of treated water recycled according to the recycling step h) is called the vaporization feedstock.

Preferably, said vaporization feedstock also comprises at least one flow of unconverted ethanol from the step g) of purifying the effluent comprising water.

According to the invention, the dehydration process comprises a step c) of vaporizing said vaporization feedstock so as to produce a vaporized feedstock. Said vaporization is performed by means of heat exchange with the effluent from the dehydration step e) in a heat exchanger.

Preferably, said vaporization feedstock is introduced into said vaporization step c) at a pressure below the pressure of the effluent from the dehydration step e).

The pressure of said vaporization feedstock upstream of the vaporization step c), an essential criterion of the present invention, is advantageously selected to be as high as possible, so that the temperature difference in the heat exchanger between the effluent from the dehydration step e), which is condensing, and said vaporization feedstock, which is evaporating, is at least greater than or equal to 2° C., and preferably at least greater than or equal to 3° C., so as to maximize the heat exchange between said vaporization feedstock and said effluent from the dehydration step e).

This temperature difference in the heat exchanger is called the temperature approach.

Surprisingly, at a given pressure, the vaporization temperature of the vaporization feedstock is lowered compared with that of a feedstock obtained by a sequence of operations that would not include the pretreatment step b). For a given condensation temperature of the effluent from the dehydration step e) and a fixed temperature approach, the pressure upstream of the vaporization step c) can therefore be adjusted to a value higher than it would have been in a sequence of operations not including the pretreatment step b).

The adjustment of said pressure upstream of the vaporization step c) to the highest possible value, within the limits defined in the preceding paragraph, makes it possible to minimize the energy required for compression during the compression step d) of the process according to the invention.

Said vaporization feedstock is introduced into said vaporization step c) at a pressure comprised between 0.1 and 1.4 MPa, preferably between 0.2 and 0.6 MPa.

The introduction of said vaporization feedstock into the vaporization step c) at this level of specific pressure comprised between 0.1 and 1.4 MPa, preferably between 0.2 and 0.6 MPa, below the pressure of the effluent leaving the last reactor of the dehydration step e), makes it possible to take advantage of a vaporization temperature of said vaporization feedstock that is lower than the temperature of condensation of the effluent from the last adiabatic reactor. Thus, most of the latent heat of the aqueous phase of the effluent from the last adiabatic reactor is recovered for vaporizing said vaporization feedstock, without external heat supply. All of the enthalpy of vaporization of said vaporization feedstock is therefore exchanged with the enthalpy of condensation of said effluent.

Compression Step d)

According to the invention, said vaporized feedstock undergoes compression in a compression step d) so as to produce a compressed feedstock. Said compression step d) is advantageously carried out in any type of compressor known to a person skilled in the art. In particular, the compression step d) is advantageously performed in a compressor of the radial compressor type with an integrated multiplier or in a compressor comprising one or more blowers with a radial impeller placed in series without intermediate cooling or in a compressor of the positive-displacement type with or without lubrication.

As step b) makes it possible, surprisingly, to operate at higher pressure upstream of step d), the level of compression necessary in step d) is reduced in order to reach a given pressure at the end of said step d), thus reducing the energy consumption of said step d).

The compression step d) makes it possible to include a heat pump integrated in said process, using the streams from the process, and without using an external heat-transfer fluid.

The combination of the specific operating conditions of step c) and step d) makes it possible to avoid the supply of heat-transfer fluid external to the process to ensure the vaporization of said vaporization feedstock, recovering most of the latent heat of the aqueous phase of the effluent from the last adiabatic reactor for vaporizing the vaporization feedstock. Thus, only the streams from the process are used.

The pressure of said compressed feedstock at the end of the compression step d) is advantageously comprised between 0.3 and 1.8 MPa, preferably between 0.5 and 1.3 MPa. The outlet pressure of said feedstock is high enough for the condensation temperature of the effluent from the last reactor to be above the vaporization temperature of the feedstock entering step c), which is a necessary condition for the feasibility of step c).

Said compressed feedstock from the compression step d) is optionally heated to an outlet temperature comprised between 250 and 420° C. and preferably comprised between 280 and 410° C. in an exchanger of the single-phase gas type, by means of heat exchange with the effluent from the last adiabatic reactor of step e). In said exchanger of the single-phase gas type, said compressed feedstock is superheated and the effluent leaving, in the gaseous state, the last adiabatic reactor of step e) is "desuperheated" without being condensed. After said exchanger of the single-phase gas type, the effluent leaving, in the gaseous state, the last adiabatic reactor of step e) advantageously has a temperature comprised between 180 and 260° C.

Thus, the use of the different exchangers, of the single-phase gas type and gas/liquid evaporator type, and the vaporization, at a pressure below the pressure of the effluent leaving the last reactor, of said vaporization feedstock, makes it possible to condense at least 80% of the steam present in the effluent from the last reactor of the dehydration step e).

Said compressed feedstock, optionally heated in said exchanger of the single-phase gas type, is then advantageously introduced into a furnace so as to bring it to an inlet temperature in at least one adiabatic reactor compatible with the temperature of the dehydration reaction. This exchanger of the single-phase gas type is an exchanger of a type of technology known to a person skilled in the art which makes it possible to minimize the feedstock losses while having a large exchange surface area. This gas/gas exchange at low pressure produces a low heat flux density through the wall of the exchanger (low transfer coefficient), which necessitates having a large exchange surface area. Moreover, the pressure loss must be minimized in order to limit the load on the compressor of step d). For example, this exchanger can be an exchanger with pressurized plates in a shell, of the Packinox type supplied by Alfa Laval.

Dehydration Step e)

According to the invention, said compressed feedstock, optionally heated, undergoes a dehydration step e) in at least one adiabatic reactor which contains at least one fixed bed of dehydration catalyst and in which the dehydration reaction takes place.

The dehydration step e) is advantageously carried out in one or two reactors.

In the case where step e) is carried out in one adiabatic reactor, said compressed feedstock, optionally heated, is advantageously introduced into said reactor at an inlet temperature comprised between 350 and 550° C. and preferably between 400 and 500° C., and at an inlet pressure comprised between 0.3 and 1.8 MPa, and preferably between 0.4 and 0.8 MPa.

The effluent from said adiabatic reactor of step e) advantageously has a temperature comprised between 270 and 450° C. and preferably between 340 and 430° C., and an outlet pressure comprised between 0.2 and 1.6 MPa and preferably between 0.3 and 0.8 MPa.

In the case where step e) is carried out in two adiabatic reactors, said compressed feedstock, optionally heated, is advantageously introduced into the first reactor at an inlet temperature comprised between 350 and 550° C. and preferably at a temperature comprised between 370 and 500° C., and at an inlet pressure comprised between 0.3 and 1.8 MPa, and preferably between 0.4 and 1.1 MPa.

The effluent from the first adiabatic reactor advantageously leaves said first reactor at a temperature comprised between 270 and 450° C. and preferably between 290 and 390° C., and at a pressure comprised between 0.3 and 1.7 MPa and preferably between 0.3 and 1.0 MPa.

Said effluent is then advantageously introduced into a furnace so that the inlet temperature of said effluent in the second adiabatic reactor is comprised between 350 and 550° C. and preferably between 400 and 500° C. Said effluent has an inlet pressure in said second reactor advantageously comprised between 0.3 and 1.7 MPa and preferably between 0.3 and 0.9 MPa.

The effluent from the second adiabatic reactor leaves said second adiabatic reactor at a temperature advantageously comprised between 270 and 450° C. and preferably between 340 and 430° C. The outlet pressure of said effluent from the second adiabatic reactor is advantageously comprised between 0.2 and 1.6 MPa and preferably between 0.3 and 0.8 MPa.

The inlet temperature of the reactor or reactors can advantageously be increased gradually to avoid deactivating the dehydration catalyst.

The dehydration reaction that takes place in at least one adiabatic reactor of step e) of the process according to the invention advantageously operates at a weight hourly space velocity comprised between 0.1 and 20 h^{-1} and preferably between 0.5 and 15 h^{-1}. The weight hourly space velocity is defined as the ratio of the mass flow rate of the pure ethanol feedstock to the mass of the catalyst.

The dehydration catalyst used in step e) is a catalyst known to a person skilled in the art. Said catalyst is preferably an amorphous acid catalyst or a zeolitic acid catalyst.

In the case where the dehydration catalyst used in step e) is a zeolitic catalyst, said catalyst comprises at least one zeolite selected from the zeolites having at least pore openings containing 8, 10 or 12 oxygen atoms (8 MR, 10 MR or 12 MR). It is known in fact to define the pore size in zeolites by the number of oxygen atoms forming the annular section of the channels in the zeolites, called "member ring" or MR. Preferably, said zeolitic dehydration catalyst comprises at least one zeolite having a structural type selected from the structural types MFI, FAU, MOR, FER, SAPO, TON, CHA, EUO, MEL and BEA. Preferably, said zeolitic dehydration catalyst comprises a zeolite of the MFI structural type and preferably a zeolite ZSM-5.

The zeolite used in the dehydration catalyst used in step e) of the process according to the invention can advantageously be modified by dealumination or desilication according to any method of dealumination or desilication known to a person skilled in the art.

The zeolite used in the dehydration catalyst used in step e) of the process according to the invention or the final catalyst can advantageously be modified with an agent of a nature such as to attenuate its total acidity and improve its properties of hydrothermal resistance. Preferably, said zeolite or said catalyst advantageously comprises phosphorus, preferably added in the form of H_3PO_4 followed by a steam treatment after neutralization of the excess acid with a basic precursor such as for example calcium Ca. Preferably, said zeolite has a phosphorus content comprised between 1 and 4.5% by weight, preferably between 1.5 and 3.1% by weight relative to the total weight of the catalyst.

Preferably, the dehydration catalyst used in step e) of the process according to the invention is the catalyst described in patent applications WO/2009/098262, WO/2009/098267, WO/2009/098268, or WO/2009/098269.

In the case where the dehydration catalyst used in step e) is an amorphous acid catalyst, said catalyst comprises at least one porous refractory oxide selected from alumina, alumina activated with a deposit of mineral acid, and silica-alumina.

Said amorphous or zeolitic dehydration catalyst used in step e) of the process according to the invention can advantageously also comprise at least one matrix of the oxide type, also called binder. By matrix is meant, according to the invention, an amorphous matrix, a crystalline matrix, or a matrix comprising amorphous and crystalline components. Said matrix is advantageously selected from the elements of the group formed by clays (such as for example from the natural clays such as kaolin or bentonite), magnesia, aluminas, silicas, silica-aluminas, aluminates, titanium oxide, boron oxide, zirconia, aluminium phosphates, titanium phosphates, zirconium phosphates, and carbon, used alone or mixed. Preferably, said matrix is selected from the elements of the group formed by aluminas, silicas and clays.

Said dehydration catalyst used in step e) of the process according to the invention is advantageously formed in the form of grains of different shapes and sizes. It is advantageously used in the form of cylindrical or multilobed extrudates, such as bilobed, trilobed, multilobed of straight or twisted shape, but can optionally be manufactured and used in the form of crushed powder, pellets, rings, beads, wheels, or spheres. Preferably, said catalyst is in the form of extrudates.

Said dehydration catalyst used in step e) of the process according to the invention is advantageously used in at least one reactor, in a fixed bed or in a moving bed.

In step e) of the process according to the invention, the catalysts used and the operating conditions are selected so as to maximize the production of ethylene. The overall reactions of dehydration used in step e) of the process according to the invention are as follows:

$$2C_2H_5OH \rightarrow 2CH_2{=}CH_2 + 2H_2O$$

$$CH_3CH_2OCH_2CH_3 \rightarrow 2CH_2{=}CH_2 + H_2O$$

The conversion of the ethanol feedstock in step e) is above 90%, preferably 95% and more preferably above 99%.

A conversion below 90% has the effect of lowering the overall yield of the process, as a larger quantity of DEE not converted to ethylene is lost in the downstream separation steps.

The conversion of the ethanol feedstock is defined, as a percentage, by the following formula:

$$[1-(\text{hourly weight of ethanol at outlet/hourly weight of ethanol at inlet})]\times 100.$$

The hourly weight of ethanol at inlet and at outlet is measured conventionally, for example by chromatography.

Step e), in which the dehydration reaction takes place, is advantageously carried out in one or two reactors. A preferred reactor is a radial reactor operating in ascending or descending mode. During step e) of the process according to the invention, conversion of the feedstock is accompanied by deactivation of the dehydration catalyst by coking and/or by adsorption of inhibiting compounds. The dehydration catalyst must therefore undergo a regeneration step periodically. Preferably, the reactor is used in an alternating regeneration mode, also called swing reactor, in order to alternate the phases of reaction and of regeneration of said dehydration catalyst. The objective of this regeneration treatment is to burn the organic deposits as well as the species containing nitrogen and sulphur, contained at the surface and within said dehydration catalyst. The pretreatment step b) used in this invention makes it possible to reduce the quantity of basic and organic impurities, as well as the cationic species that will alter the catalyst's cycle life. The removal of these species thus makes it possible to limit the number of catalyst regenerations.

The regeneration of the dehydration catalyst used in said step e) is advantageously carried out by oxidation of the coke and of the inhibiting compounds under an air flow or under an air/nitrogen mixture, for example using recirculation of the combustion air with or without water in order to dilute the oxygen and control the exothermic nature of regeneration. In this case, the oxygen content at the reactor inlet can advantageously be adjusted with an additional supply of air. The regeneration takes place at a pressure comprised between atmospheric pressure and the reaction pressure.

The regeneration temperature is advantageously selected to be between 400 and 600° C.; it can advantageously vary over the course of regeneration. The end of the regeneration is detected when there is no longer consumption of oxygen, a sign of total combustion of the coke.

The effluent from the last adiabatic reactor of step e) is optionally sent to an exchanger of the single-phase gas type in which it is "desuperheated" without being condensed by heat exchange with the compressed feedstock from step d), which for its part is superheated.

Said "desuperheated" effluent is then advantageously sent to a second exchanger of the gas/liquid type in which it is partially condensed by a heat exchange that serves to vaporize the vaporization feedstock.

Said effluent is then cooled again by heat exchange with the ethanol feedstock during the step a) of preheating the ethanol feedstock.

Separation Step f)

According to the invention, the effluent from the last adiabatic reactor of step e) undergoes a step f) of separation into an effluent comprising ethylene at a pressure below 1.6 MPa, preferably below 0.8 MPa and an effluent comprising water.

The step f) of separating said effluent from the last adiabatic reactor of step e) can advantageously be performed using any method known to a person skilled in the art, such as for example by a zone for gas/liquid separation, and preferably a gas/liquid separating column.

The effluent comprising ethylene at a pressure below 1.6 MPa then advantageously undergoes a compression. Said compression makes it possible to bring the pressure of said effluent back up to a pressure advantageously comprised between 2 and 4 MPa, necessary for its final purification.

At least a portion of the effluent comprising water from step f) is optionally recycled to the separation step f). This recycling makes it possible to increase the effectiveness of step f) by absorbing a portion of the unconverted feedstock. In the case where at least a portion of the effluent comprising water is recycled, said portion of the effluent comprising water is advantageously cooled with a cold fluid or with a fluid from the process and is preferably treated according to the known methods of purification described below.

Purification Step g)

According to the invention, at least a portion of the effluent comprising water from the separation step f) undergoes a purification step g). The purification step g) can advantageously be performed using any method of purification known to a person skilled in the art. As an example, the purification step g) can advantageously be carried out by the use of ion exchange resins, by adding chemicals to adjust the pH, such as for example soda or amines, and by adding chemicals to stabilize the products, such as for example polymerization inhibitors selected from the bisulphites and surfactants.

At least one flow of treated water and at least one flow of unconverted ethanol are then separated. The separation can advantageously be performed using any method of separation known to a person skilled in the art. As an example, the separation can advantageously be carried out by distillation, by the use of molecular sieves, by steam or heat stripping or by solvent absorption, such as for example with glycol-containing solvents.

A flow containing the light gases, preferably acetaldehyde and methanol, can advantageously also be separated.

Recycling Step h)

According to the invention, at least a portion of the flow of treated water from the purification step g) is recycled upstream of the vaporization step c) according to the recycling step h).

The flow of treated water from step g) plays the role of thermal reaction diluent.

The dilution of the pretreated ethanol feedstock by adding at least a portion of the flow of treated water from step g) is performed in a weight ratio of diluent to feedstock advantageously comprised between 1 and 4 with the aim of lowering the ethanol partial pressures in the reactor or reactors and of making the process more selective for ethylene.

At least a portion of said flow of unconverted ethanol from the step g) of purifying the effluent comprising water is advantageously recycled and mixed, upstream of the vaporization step c), with the pretreated ethanol feedstock, and mixed with at least a portion of the flow of treated water recycled according to the recycling step h).

BRIEF DESCRIPTION OF THE FIGURES

FIG. **1** is a diagrammatic representation of the process for dehydration of ethanol in the case of dehydration of a concentrated ethanol feedstock with recycling of at least a portion of the water treated during step h) of the process.

The ethanol feedstock (**1**) is preheated in an exchanger E**1** with the effluent of the last adiabatic reactor R**2**, which enters via pipeline (**14**). The preheated ethanol feedstock is then introduced into a pretreatment zone (**3**) via pipeline (**2**). The pretreated ethanol feedstock (**4**) is then mixed in pipeline (**5**) with a portion of the flow of treated water from the purification zone (**20**), which is recycled so as to serve as reaction diluent via pipelines (**25**) and (**26**). The ethanol feedstock is also mixed with a portion of the flow of unconverted ethanol from the purification zone (**20**), via pipeline (**23**), then (**26**). This mixture, constituting the vaporization feedstock, is introduced via pipeline (**5**) into a gas/liquid exchanger E**2**, in which said mixture undergoes heat exchange with the effluent from the last adiabatic reactor R**2**, which enters the exchanger via pipeline (**13**) so as to produce a vaporized feedstock. The latent heat, also called enthalpy of condensation, of the effluent from the last adiabatic reactor R**2** is used to vaporize the vaporization feedstock, without external heat supply.

The vaporized feedstock is then sent via pipeline (**6**) to a compressor C**1**.

Said vaporized and compressed feedstock is then sent via pipeline (**7**) to an exchanger E**3** of the single-phase gas type, in which said feedstock is heated by means of heat exchange with the effluent from the last adiabatic reactor R**2**, which is introduced into E**3** via pipeline (**12**). In said exchanger of the single-phase gas type, said vaporized and compressed feedstock is superheated and the effluent leaving, in the gaseous state, the last adiabatic reactor R**2** is "desuperheated", without being condensed.

Said vaporized and compressed feedstock, heated in the exchanger of the single-phase gas type E**3**, is then introduced into a furnace H**1** via pipeline (**8**) so as to bring it to an inlet temperature in the first adiabatic reactor R**1** compatible with the temperature of the dehydration reaction. The effluent from the first reactor R**1** is sent to a second furnace H**2** via pipeline (**10**) before being introduced into the second reactor R**2** via pipeline (**11**).

The effluent from the second reactor R**2** then undergoes the three successive exchanges described above in exchangers E**3**, E**2** and E**1** via pipelines (**12**), (**13**) and (**14**).

The effluent from exchanger E**1** is sent via pipeline (**15**) to a gas/liquid separating column (**16**), where it is separated into an effluent comprising ethylene (**17**) and an effluent comprising water (**18**). A portion of the effluent comprising water is recycled after cooling to column (**16**) via pipeline (**19**).

The portion of the effluent comprising water not recycled to column (**16**) is sent via pipeline (**18**) to a step (**20**) of purification and separation. At least one flow of treated water

(**24**) and (**25**) and at least one flow of unconverted ethanol (**22**) and (**23**) are then separated. A flow containing the light gases (**21**) is also separated.

All (optionally a portion) of said flow of unconverted ethanol from the purification step (**20**) is recycled via pipeline (**23**) and is mixed with the flow of treated water recycled via pipeline (**25**) in pipeline (**26**). The mixture of these two streams is incorporated upstream of exchanger E**2** with the pretreated ethanol feedstock (**4**).

The following examples illustrate the invention without limiting its scope.

EXAMPLES

Example 1

According to the Invention

Example 1 illustrates a process according to the invention.

The ethanol feedstock under consideration is produced by fermentation of wheat, without extraction of glutens, by a process of the dry milling type.

Step a)

Said ethanol feedstock is introduced, at a flow rate of 45,664 kg/h, into an exchanger E**1** at a pressure equal to 1.15 MPa and is heated, remaining in the liquid phase, to a temperature of 120° C. against the effluent from the last adiabatic reactor of step e).

Step b)

The heated ethanol feedstock is pretreated on TA801 resin to remove the traces of nitrogen-containing compounds. During this pretreatment, a portion of the ethanol is converted to DEE. The characteristics of the raw ethanol feedstock and of the pretreated feedstock are given in Table 1.

TABLE 1

Characteristics of the ethanol feedstock before and after pretreatment (percentages by weight)

	ETHANOL FEEDSTOCK	ETHANOL AFTER PRETREATMENT
ETHANOL	91.2%	82.1%
H_2O	8.7%	10.5%
DEE	0%	7.3%
NITROGEN-CONTAINING COMPOUNDS	0.005%	0.000%

Step c)

The vaporization feedstock, constituted by the pretreated ethanol feedstock mixed with 141,252 kg/h of treated water and of unconverted ethanol recycled according to step h), is depressurized and introduced into an exchanger E**2** at a pressure equal to 0.27 MPa. The bubble point of this feedstock at this pressure is 127° C. taking into account the presence of DEE. The vaporization feedstock enters exchanger E**2** at 113° C. and is therefore already vaporized at 8.6% by weight. The pressure at the inlet of exchanger E**2** was adjusted in such a way that the temperature approach with the flow from the last adiabatic reactor of step e) is at a minimum of 15° C.

In step c), most of the latent heat of the aqueous phase of the effluent from the last adiabatic reactor of step e) is recovered for vaporizing the vaporization feedstock, without external heat supply. Thus, 93.6 MW is exchanged between said vaporization feedstock and said effluent.

Step d)

The vaporized feedstock is then compressed in a radial compressor with an integrated multiplier so that the pressure of said vaporized feedstock is equal to 0.695 MPa at the end of the compression.

The compressed feedstock is then heated in an exchanger E**3** of the single-phase gas type, by means of heat exchange with the effluent from the adiabatic reactor of step e). In said exchanger of the single-phase gas type, said compressed feedstock is superheated to a temperature of 405° C. and the effluent leaving, in the gaseous state, the last adiabatic reactor of step e) is "desuperheated" without being condensed, and has a temperature of 253° C.

Step e)

Said compressed feedstock, heated in said exchanger of the single-phase gas type, is then introduced into a furnace so as to bring it to an inlet temperature in the first adiabatic reactor of step e) compatible with the temperature of the highly endothermic reaction of dehydration and of conversion of DEE to ethylene, i.e. to a temperature of 440° C. The outlet temperature of the last adiabatic reactor of step e) is 420° C.

The trapping of the nitrogen-containing compounds in the pretreatment step b) makes it possible to reduce the inlet temperature of the first adiabatic reactor of step e) significantly.

Said compressed and heated feedstock is introduced into the first adiabatic reactor at an inlet pressure of 0.595 MPa. The pressure of the effluent at the outlet of the last adiabatic reactor of step e) is 0.500 MPa. The dehydration step e) is carried out at a weight hourly space velocity of 7 h^{-1}.

The adiabatic reactor contains a fixed bed of dehydration catalyst, said catalyst comprising 80% by weight of zeolite ZSM-5 treated with H_3PO_4 so that the content of phosphorus P is 3% by weight.

The conversion of the ethanol feedstock in step e) is 95%.

Step f)

The effluent from the last adiabatic reactor of step e) then undergoes the three heat exchanges described above and is sent to a gas/liquid separating column. An effluent comprising ethylene at a pressure equal to 0.36 MPa is separated, as well as an effluent comprising water. This separation is carried out using a gas/liquid separating column, with recycling of the water produced at bottom of the column to the top of the column and after cooling and injection of neutralizing agent.

The effluent comprising ethylene then undergoes a compression to bring its pressure back up to 2.78 MPa prior to its final purification.

Step g)

A flow of treated water and a flow of unconverted ethanol as well as a flow containing the light gases are then separated by conventional low-pressure distillation of the raw water.

Step h)

A portion of the flow of treated water and a portion of the flow of unconverted ethanol are recycled upstream of the vaporization step c) in the proportions described in step c). The different streams, in kg/h, are presented in Table 2 and in Table 3.

TABLE 2

Composition of the main streams (1/2)					
Description of the flow		Pretreated ethanol feedstock	Flow entering R1	Flow leaving R2	Effluent comprising ethylene
Corresponding flow No. in the figure		4	9	12	17
Total mass flow rate	kg/h	45664	186916	186916	25692
Mass flow rate by components	kg/h				
ethylene		0	0	25087	25087
ethane		0	0	8	8
C3		0	0	93	93
C4		0	0	87	87
DEE		3352	3352	14	14
ethanol		37504	39310	2187	151
H_2O		4808	143730	158602	198
oxygenated compounds (other than ethanol)		0	325	586	42
Other minor components		0	199	252	12

TABLE 3

Composition of the main streams (2/2)					
Description of the flow		Effluent comprising water	Ethanol and water recycle	Purged water	Light gases
Corresponding flow No. in the figure		18	26	24	21
Total mass flow rate	kg/h	161224	141252	19007	965
Mass flow rate by components	kg/h				
ethylene		0	0	0	0
ethane		0	0	0	0
C3		0	0	0	0
C4		0	0	0	0
DEE		0	0	0	0
ethanol		2036	1806	3	227
H_2O		158404	138922	18987	495
oxygenated compounds (other than ethanol)		544	325	6	213
Other minor components		240	199	11	30

Compounds **C3** and **C4** are **C3** and **C4** hydrocarbon-containing compounds.

The selectivity of the process for ethylene is 99%.

It is calculated as follows: (Ethylene contained in the effluent comprising ethylene)/(0.61*quantity of ethanol converted) where the quantity of ethanol converted is the ethanol contained in the ethanol feedstock before pretreatment subtracted from the ethanol contained in the streams of purged water and in the effluent comprising ethylene. 0.61 g is the maximum quantity of ethylene obtained on dehydrating 1 g of pure ethanol.

The energy balance of the scheme according to Example 1 according to the invention is presented in Table 4:

TABLE 4

Energy balance					
Energy exchanged within the system			Energy supplied to the system by external supply		
Quantity of heat exchanged in the first exchanger (E1) MW	Quantity of heat exchanged in the second exchanger (E2) MW	Quantity of heat exchanged in the third exchanger (E3) MW	Quantity of heat exchanged in the furnace MW	Power required for compression MW	Quantity of heat extracted on the gas/liquid separating column MW
4.21	93.6	18.32	10.4	10.9	22.53

The primary energy consumption was estimated on the following basis:

efficiency of 0.8 for the furnaces

efficiency of 0.375 for electricity generation

The scheme according to Example 1 according to the invention has an equivalent primary energy consumption or specific consumption of 6.0 GJ equivalent per tonne of ethylene produced.

Example 2

Comparison

Example 2 illustrates a process in which the steps a) and b) of preheating and pretreatment do not take place. The ethanol is not converted to DEE and the process starts at step c): exchanger **E1** is no longer present.

Step c)

The vaporization feedstock, constituted by the unpretreated ethanol feedstock mixed with 141,258 kg/h of treated water and of unconverted ethanol recycled according to step h), is introduced at a flow rate of 186,922 kg/h into exchanger **E2** at a pressure equal to 0.24 MPa.

The pressure was lowered by 0.03 MPa compared with Example 1. Without the presence of DEE, the bubble point of the vaporization feedstock at 0.27 MPa is 115° C. (127° C. in Example 1). The inlet pressure is altered by 0.03 MPa so as to maintain a minimum temperature approach of 15° C. with the effluent from the last adiabatic reactor of step e).

In step c), most of the latent heat of the aqueous phase of the effluent from the adiabatic reactor of step e) is recovered for vaporizing the vaporization feedstock, without external heat supply. Thus, 98 MW is exchanged between the vaporization feedstock and the effluent from the reactor.

Step d)

The vaporized feedstock is then compressed in a radial compressor with an integrated multiplier so that the pressure of said vaporized feedstock at the end of the compression is equal to 0.695 MPa.

The compressed feedstock is then heated in an exchanger **E3** of the single-phase gas type, by means of heat exchange with the effluent from the last adiabatic reactor of step e). In said exchanger of the single-phase gas type, said compressed feedstock is superheated to a temperature of 405° C. and the effluent leaving, in the gaseous state, the last adiabatic reactor of step e) is "desuperheated" without being condensed and has a temperature of 269° C.

Step e)

Said compressed feedstock, heated in said exchanger of the single-phase gas type, is then introduced into a furnace in order to bring it to an inlet temperature in the first adiabatic reactor of step e) compatible with the temperature of the dehydration reaction, i.e. to a temperature of 470° C. The outlet temperature of the last adiabatic reactor of step e) is 420° C.

Said compressed and heated feedstock is introduced into the adiabatic reactor at an inlet pressure of 0.595 MPa. The pressure of the effluent at the outlet of the last adiabatic reactor of step e) is 0.500 MPa. The dehydration step e) is carried out at a weight hourly space velocity of 7 h^{-1}.

The conversion of the ethanol feedstock in step e) is 95%.

Step f)

The effluent from the last adiabatic reactor of step e) then undergoes the two heat exchanges described above and is sent to a gas/liquid separating column. An effluent comprising ethylene at a pressure equal to 0.39 MPa is separated, as well as an effluent comprising water. This separation is performed using a gas/liquid separating column, with recycling of the water produced at the bottom of the column to the top of the column and after cooling and injection of neutralizing agent.

The effluent comprising ethylene then undergoes a compression to bring its pressure back up to 2.78 MPa prior to its final purification.

Step g)

The raw water from step f) is then neutralized with soda, then undergoes conventional low-pressure distillation to be separated into three streams: a flow of treated water, a flow of unconverted ethanol and a flow containing the light gases.

Step h)

A portion of the flow of treated water and a portion of the flow of unconverted ethanol are recycled upstream of the vaporization step c).

The different streams, in kg/h, are presented in Table 5 and Table 6.

TABLE 5

Composition of the main streams (1/2)					
Description of the flow		Ethanol feedstock	Flow entering R1	Flow leaving R2	Effluent comprising ethylene
Corresponding flow No. in the figure		4	9	12	17
Total mass flow rate	kg/h	45664	186922	186922	25964
Mass flow rate by components	kg/h				
ethylene		0	0	25087	25087
ethane		0	0	8	8
C3		0	0	93	93
C4		0	0	87	87
DEE		0	0	14	14
ethanol		41671	43496	2187	151
H_2O		3993	142947	158602	311
oxygenated compounds (other than ethanol)		0	413	586	62
Other minor components		0	66	258	151

TABLE 6

TABLE 6-continued

Composition of the main streams (1/2)					
Description of the flow		Effluent comprising water	Ethanol and water recycle	Purged water	Light gases
Corresponding flow No. in the figure		18	26	24	21
Total mass flow rate	kg/h	160958	141258	19007	693
Mass flow rate by components	kg/h				
ethylene		0	0	0	0
ethane		0	0	0	0
C3		0	0	0	0
C4		0	0	0	0
DEE		0	0	0	0
ethanol		2036	1825	3	208
H_2O		158291	138954	18987	350
oxygenated compounds (other than ethanol)		524	413	6	105
Other minor components		107	66	11	30

Compounds **C3** and **C4** are C3 and C**4** hydrocarbon-containing compounds.

The selectivity of the process for ethylene is 99%.

The energy balance of the scheme according to Example 2 is presented in Table 7.

TABLE 7

Energy balance				
Energy exchanged within the system		Energy supplied to the system by an external supply		
Quantity of heat exchanged in the first exchanger (E2) MW	Quantity of heat exchanged in the second exchanger (E3) MW	Quantity of heat exchanged in the furnace MW	Electricity required for compression MW	Quantity of heat extracted on the gas/liquid separating column MW
98.0	17.1	13.9	12.4	22.53

The scheme according to Example 2, as a comparison with the invention, has an equivalent primary energy consumption or specific consumption of 7.23 GJ equivalent per tonne of ethylene produced.

Without pretreatment, the primary energy consumption therefore increases by 1.2 GJ equivalent per tonne of ethylene produced.

The invention claimed is:

1. A process for dehydrating an ethanol feedstock to ethylene comprising:

a) a step of preheating said ethanol feedstock to a temperature between 100 and 130° C. by heat exchange with the effluent from step e),

b) a step of pretreating the ethanol feedstock on an acidic solid operating at a temperature between 100 and 130° C. so as to produce a pretreated ethanol feedstock,

c) a step of vaporizing a vaporization feedstock comprising said pretreated ethanol feedstock and at least a portion of the flow of treated water recycled according to step h) in an exchanger by heat exchange with the effluent from the last reactor of step e), said vaporization feedstock being introduced into said vaporization step at a pressure between 0.1 and 1.4 MPa so as to produce a vaporized feedstock,

d) a step of compressing said vaporized feedstock in a compressor so as to produce a compressed feedstock,

e) a step of dehydrating said compressed feedstock in at least one adiabatic reactor containing at least one dehydration catalyst and in which the dehydration reaction takes place, operating at an inlet temperature between 350 and 550° C. and at an inlet pressure between 0.3 and 1.8 MPa,

f) a step of separating the effluent from the last adiabatic reactor of step e) into an effluent comprising ethylene at a pressure below 1.6 MPa and an effluent comprising water,

g) a step of purifying at least a portion of the effluent comprising water from step f) and separating at least one flow of treated water and at least one flow of unconverted ethanol, and

h) a step of recycling at least a portion of the flow of treated water from step g) upstream of step c).

2. The process according to claim **1**, wherein said ethanol feedstock is an ethanol feedstock produced starting from a renewable source obtained from biomass.

3. The process according to claim **1**, wherein the vaporization feedstock also comprises at least one flow of unconverted ethanol from step g) of purifying the effluent comprising water.

4. The process according to claim **1**, wherein the pressure of the compressed feedstock is between 0.3 and 1.8 MPa.

5. The process according to claim **1**, wherein said compressed feedstock is heated in an exchanger of the single-phase gas type, by heat exchange with the effluent from the last adiabatic reactor of step e).

6. The process according to claim **1**, wherein the effluent from the last adiabatic reactor of step e) has a temperature between 270 and 450° C. at the outlet of the last adiabatic reactor of step e).

7. The process according to claim **1**, wherein the effluent from the last adiabatic reactor of step e) has a pressure between 0.2 and 1.6 MPa at the outlet of the last adiabatic reactor of step e).

8. The process according to claim **1**, wherein the dehydration step e) is carried out in one or two reactors.

9. The process according to claim **1**, wherein said dehydration catalyst in step e) is an amorphous acid catalyst or a zeolitic acid catalyst.

10. The process according to claim **1**, wherein said ethanol feedstock comprises a percentage by weight of ethanol greater than or equal to 35% by weight.

11. The process according to claim **10**, wherein said ethanol feedstock comprises a percentage by weight of ethanol between 35 and 99.9% by weight.

12. The process according to claim **1**, wherein the pretreatment step b) is supplemented with a pretreatment by an anion exchange resin.

13. The process according to claim **1**, wherein the acidic solid is a silica-alumina, acid clay, zeolite, sulphated zirconia or acid resin.

14. The process according to claim **1**, wherein the acidic solid has an exchange capacity for capturing basic and cationic species of at least 0.1 mmol H^+ equivalent per gram.

15. The process according to claim **1**, wherein the acidic solid has acidic sulphonic groups and has been prepared by polymerization or co-polymerization of aromatic vinyl groups followed by sulphonation, said aromatic vinyl groups being selected from the group consisting of styrene, vinyl toluene, vinyl naphthalene, vinyl ethyl benzene, methyl styrene, vinyl chlorobenzene and vinyl xylene, said resin having a level of cross-linking between 20 and 35%.

16. The process according to claim **1**, wherein the acidic solid is a copolymer of divinyl benzene and polystyrene having a level of cross-linking between 20 and 45%.

17. The process according to claim **1**, wherein the acidic solid is a TA801 resin.

18. The process according to claim **1**, wherein step b) is performed at a temperature between 110 and 130° C.

19. The process according to claim **1**, wherein step b) is performed at a temperature between 110 and 120° C.

* * * * *

LAMPIRAN 2 NERACA MASSA DAN ENERGI

L2.1 Informasi Umum

Tabel L2.1 Berat Molekul Masing-Masing Komponen

Komponen	Rumus Molekul	MR (g/mol)
Etilen	C2H4	28.054
Etanol	C2H5OH	46.069
Water	H2O	18.015

Tabel L2.2 Panas Spesifik Fasa Gas Bahan Baku dan Produk

Komponen	Rumus Molekul	Cp of Gas (J/mol K)				
		A	BT	CT^2	DT^3	ET^4
Etilen	C_2H_4	32.083	-1.48E-02	2.48E-04	-2.38E-07	6.88E-11
Etanol	C_2H_5OH	27.091	1.11E-01	1.10E-04	-1.50E-07	4.66E-11
Water	H_2O	33.933	-8.42E-03	2.99E-05	-1.78E-08	3.69E-12

Tabel L2.3 Panas Spesifik Fasa Cair Bahan Baku dan Produk

Komponen	Rumus Molekul	Cp of Liquid (J/mol K)			
		A	BT	CT^2	DT^3
Etilen	C_2H_4	-6.94E+00	2.05E+00	-4.86E-03	4.50E-06
Etanol	C_2H_5OH	8.76E+01	-4.84E-04	-4.54E-06	1.19E-09
Water	H_2O	9.21E+01	-4.00E-02	-2.11E-04	5.35E-07

Tabel L2.5 Entalpi Dowtherm A yang digunakan

Dowtherm A	T in = 30ºC = 303.15 K	T = 70 ºC = 343.15 K
Cp 20ºC (Kj/kg K)	1.601	
Cp 70ºC (Kj/kg K)	1.715	
ΔH (Kj/kg)	524.8136	
ΔH (Kj/kg) @20ºC	461.1250	

L2.1.1 Perhitungan Kebutuhan Bahan Baku

Kapasitas Produksi = 85000 ton/tahun

Jumlah hari = 330 hari/tahun

Jumlah jam = 24 jam/hari

$$Kapasitas\ Produksi = \frac{85000\ ton}{1\ tahun} \times \frac{1000\ kg}{1\ ton} \times \frac{1\ tahun}{330\ jam} \times \frac{1\ hari}{24\ jam} = 10732.323\ kg/jam$$

Kemurnian produk = 99 %

Kebutuhan Etanol

$$Mol\ Etanol\ Umpan = \frac{Massa\ Etanol\ \ Umpan}{BM\ Etanol}$$

$$Mol\ Etanol\ Umpan = \frac{17802.1535\ \frac{kg}{jam}}{46.069\ \frac{kg}{kmol}} = 386.4237\ \frac{kmol}{jam}$$

$$Mol\ Etanol\ Bereaksi = Konversi\ x\ Mol\ Etanol\ Umpan$$

$$Mol\ Etanol\ Bereaksi = 99\%\ x\ 386.4237\ \frac{kmol}{jam} = 382.5595\ \frac{kmol}{jam}$$

$$Mol\ Etanol\ Sisa = Mol\ Etanol\ Umpan - \ Mol\ Etanol\ Bereaksi$$

$$Mol\ Etanol\ Sisa = 386.4237\ \frac{kmol}{jam} - \ 382.5595\ \frac{kmol}{jam} = 3.8642\ \frac{kmol}{jam}$$

Jumlah Etilen

$$Mol\ Etilen\ Hasil\ Reaksi = \frac{1}{1}\ x\ Mol\ Etanol\ Bereaksi$$

$$Mol\ Etilen\ Hasil\ Reaksi = \frac{1}{1}\ x\ 382.5595\ \frac{kmol}{jam} = 382.5595\ \frac{kmol}{jam}$$

Jumlah Air yang Terbentuk

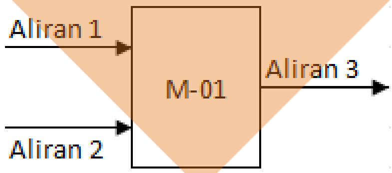
$$Mol\ Air\ Hasil\ Reaksi = \frac{1}{1}\ x\ Mol\ Etanol\ Bereaksi$$

$$Mol\ Air\ Hasil\ Reaksi = \frac{1}{1}\ x\ 382.5595\frac{kmol}{jam} = 382.5595\frac{kmol}{jam}$$

	Etanol	→	Etilen	+	Water
Reaksi	C_2H_5OH	→	C_2H_4	+	H_2O

	Etanol	→	Etilen	+	Water
Reaksi	C_2H_5OH	→	C_2H_4	+	H_2O
Awal	386.4237		0.0000		0.0000
Bereaksi	382.5595		382.5595		382.5595
Sisa (kgmol)	3.8642		382.5595		382.5595

L2.2 Neraca Mixer (M-01)



GAMBAR L2.1 DIAGRAM ALIR MIXER (M-01)

Temperatur operasi = 30ºC

Tekanan Operasi = 1 atm

Fungsi = Tempat pencampuran aliran dari TP-01 dan aliran dari TP-02

.

L2.2.1 Neraca Massa Mixer (M-01)

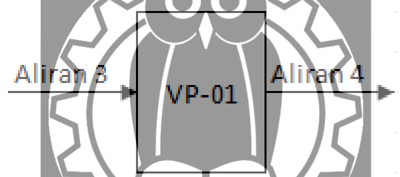
Tabel L2.6 Neraca Massa Mixer (M-01)

Komponen	**Laju alir (kg/jam)**		
Aliran	**1**	**2**	**3**
Etanol	17802.1535	-	17802.1535
H_2O	-	24693.9623	24693.9623

L2.2.2 Neraca Energi Mixer (M-01)

Pada Mixer (M-01) ini tidak melibatkan perubahan entalpi sehingga tidak dilakukan kalkulasi neraca energi.

L2.3 Neraca Vaporizer (VP-01)



GAMBAR L2.2 DIAGRAM ALIR VAPORIZER (VP-01)

Temperatur operasi = $126^{o}C$

Tekanan Operasi = 1 atm

Fungsi = Menguapkan Etanol 42 % dari fase cair ke fase gas.

L2.3.1 Neraca Massa Vaporizer (VP-01)

Tabel L2.7 Neraca Massa Vaporizer (VP-01)

Komponen	**Laju alir (kg/jam)**	
Aliran	**3**	**4**
Etanol	17802.1535	17802.1535

H_2O	24693.9623	24693.9623

L2.3.2 Neraca Energi Vaporizer (VP-01)

Kondisi Operasi Masuk Aliran 3

Tin = 30ºC = 303.15 K

Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tin} CpdT$$

Tabel L2.8 Neraca Energi Aliran 3 Masuk Vaporizer (VP-01)

Komponen	Massa (kJ/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H in (KJ/jam)
Etanol	17802.1535	386.4237	537.0986	207547.6103
H_2O	24693.9623	1370.7445	377.4714	517416.8662
Total				724964.4766

Kondisi Fase Liquid

Tin = 95.25ºC = 368.40 K
Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tin} CpdT$$

Tabel L2.9 Neraca Energi Fase Liquid Vaporizer (VP-01)

Komponen	Massa (kJ/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H in (KJ/jam)
Etanol	17802.1535	386.4237	7332.0158	2833264.6772
H_2O	24693.9623	1370.7445	4906.9625	6726191.8789
Total				9559456.5561

Kondisi Perubahan Fase Komponen

T = 95.25ºC = 368.40 K

$$H = n\ x\ Hfg$$

Tabel L2.10 Neraca Energi Perubahan Fase Komponen Vaporizer (VP-01)

Komponen	Massa (Kg/jam)	mol (Kmol/jam)	Hfg (kJ/Kmol)	H Laten (kJ/jam)
Etanol	17802.1535	386.4237	39065.9925	15096025.4315
H_2O	24693.9623	1370.7445	39721.1225	54447510.4163
Total				69543535.8478

Kondisi Fase Gas

Tin = 95.25ºC = 368.40 K
Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tin} CpdT$$

Tabel L2.11 Neraca Energi Fase Gas Vaporizer (VP-01)

Komponen	Massa (kJ/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H in (KJ/jam)
Etanol	17802.1535	386.4237	5010.2973	1936097.6220
H_2O	24693.9623	1370.7445	2377.2901	3258657.3416
Total				5194754.9635

Kondisi Operasi Keluar Aliran 4

Tout = 126ºC = 399.15 K

Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tout} CpdT$$

Tabel L2.12 Neraca Energi Aliran 4 Keluar Vaporizer (VP-01)

Komponen	Massa (kg/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H out (KJ/jam)
Etanol	17802.1535	386.4237	7421.7766	2867950.4081
H_2O	24693.9623	1370.7445	3428.3384	4699376.0892
Total				7567326.4973

Kebutuhan Pemanas

$$H\ in\ +\ Q\ Pemanas\ =\ H\ out\ +\ \Delta H\ Laten\ +\ \Delta H\ Liquid\ +\ \Delta H\ Gas$$

$$Q\ pemanas = 724964.4766\frac{kj}{jam} + 69543535.8478\frac{kj}{jam} + 9559456.5561\frac{kj}{jam} + 5194754.9635\frac{kj}{jam} - 7567326.4973\frac{kj}{jam} = 91140109.3881\frac{kj}{jam}$$

$$Massa\ Pemanas = \frac{Q\ Pemanas}{H\ vap}$$

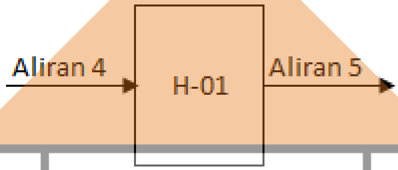
$$Massa\ Pemanas = \frac{91140109.3881\frac{kj}{jam}}{3457.79\frac{kj}{kg}} = 26357.9076\frac{kg}{jam}$$

Tabel L2. 13 Neraca Energi Total Vaporizer (VP-01)

Neraca Energi Total Vaporizer		
Komponen	Masuk (Kj/jam)	Keluar (Kj/jam)
H in	724964.4766	-
ΔH Liquid	-	9559456.5561
ΔH Laten	-	69543535.8478

ΔH Gas	-	5194754.9635
H Out	-	7567326.4973
Q Pemanas	91140109.3881	-
Total	91865073.8647	91865073.8647

L2.4 Neraca Heater (H-01)



GAMBAR L2.3 DIAGRAM ALIR HEATER (H-01)

Temperatur operasi = 470°C

Tekanan Operasi = 1 atm

Fungsi = Memanaskan bahan baku etanol 42% (fase gas) setelah vaporizer suhu 126˚C menjadi 470˚C.

L2.4.1 Neraca Massa Heater (H-01)

Tabel L2.14 Neraca Massa Vaporizer (VP-01)

Komponen	**Laju alir (kg/jam)**	
Aliran	**4**	**5**
Etanol	17802.1535	17802.1535
H_2O	24693.9623	24693.9623

L2.4.2 Neraca Energi Heater (H-01)

Fungsi = Memanaskan bahan etanol menuju reaktor.

Kondisi Operasi Masuk

Tin = 126ºC = 399.15 K

Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tin} CpdT$$

Tabel L2.15 Neraca Energi Masuk Heater (H-01)

Komponen	Massa (kg/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H in (KJ/jam)
Etanol	17802.1535	386.4237	7421.7766	2867950.4081
H2O	24693.9623	1370.7445	3428.3384	4699376.0892
Total				7567326.4973

Kondisi Operasi Keluar

Tout = 470ºC = 743.15 K

Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tout} CpdT$$

Tabel L2. 16 Neraca Energi Keluar Heater (H-01)

Komponen	Massa (Kg/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H out (Kj/jam)
Etanol	17802.1535	386.4237	42799.1857	16538619.7904
H2O	24693.9623	1370.7445	15819.1716	21684042.6177
Total				38222662.4081

Kebutuhan Pemanas

$$Q\ Pemanas = \Delta Hout - \Delta Hin$$

$$Q\ pemanas = 38222662.4081 \frac{kj}{jam} - 7567326.4973 \frac{kj}{jam} = 30655335.9108 \frac{kj}{jam}$$

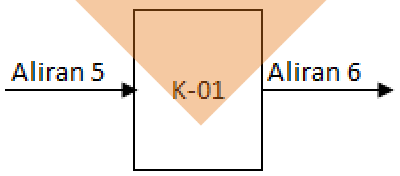
$$Massa\ Pemanas = \frac{Q\ Pemanas}{H\ vap}$$

$$Massa\ Pemanas = \frac{30655335.9108 \frac{kj}{jam}}{3457.79 \frac{kj}{kg}} = 8865.5864 \frac{kg}{jam}$$

Tabel L2.17 Neraca Energi Total (H-01)

Neraca Energi Total Heater E-01		
Komponen	Masuk (KJ/jam)	Keluar (Kj/jam)
H in	7567326.4973	-
H out	-	38222662.4081
Q pemanas	30655335.9108	-
Total	38222662.4081	38222662.4081

L2.5 Neraca Kompresor (K-01)



GAMBAR L2.4 DIAGRAM ALIR KOMPRESOR (K-01)

Temperatur operasi = 470ºC

Tekanan Operasi = 6 atm

Fungsi = Menaikkan tekanan umpan dari 1 atm menjadi 6 atm menuju ke reaktor dalam bentuk fase gas.

L2.5.1 Neraca Massa Kompresor (K-01)

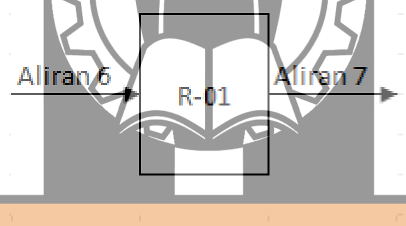
Tabel L2.18 Neraca Massa Kompresor (K-01)

Komponen	Laju alir (kg/jam)	
Aliran	**5**	**6**
Etanol	17802.1535	17802.1535
H_2O	24693.9623	24693.9623

L2.5.2 Neraca Energi Kompresor (K-01)

Pada Kompresor (K-01) ini tidak melibatkan perubahan entalpi sehingga tidak dilakukan kalkulasi neraca energi.

L2.6 Neraca Reaktor (R-01)



GAMBAR L2.5 DIAGRAM ALIR REAKTOR (R-01)

Temperatur operasi = 470ºC

Tekanan Operasi = 6 atm

Fungsi = Mereaksikan etanol 42% membentuk etilen.

L2.6.1 Neraca Massa Reaktor (R-01)

Tabel L2.19 Neraca Massa Reaktor (R-01)

Komponen	Laju alir (kg/jam)

Aliran	6	7
Etilen	-	10732.3232
Etanol	17802.1535	178.0215
H_2O	24693.9623	31585.7711

L2.6.2 Neraca Energi Reaktor (R-01)

Kondisi Operasi Masuk

Tin = 470ºC = 743.15 K

Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tin} CpdT$$

Tabel L2.20 Neraca Energi Aliran Masuk Reaktor (R-01)

	Komponen	Massa (kg/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H in (KJ/jam)
Aliran 6	Etanol	17802.1535	386.4237	42799.1857	16538619.7904
	H_2O	24693.9623	1370.7445	15819.1716	21684042.6177
Total					38222662.4081

Perhitungan Panas Reaksi

Komponen	Hf @298 K
Etilen	52.3
Etanol	-234.81
H_2O	-241.8

Menghitung Qr Reaktan

Qr Reaktan = n reaktan x Hf @298 K

Tabel L2.21 Neraca Energi Reaktan Reaktor (R-01)

Komponen	Massa (kg/jam)	mol (Kmol/jam)	Hf @ 298 K	Qr Reaktan (Kj/jam)
Etanol	17802.1535	386.4237	-234.8100	-90736.1495
H_2O	24693.9623	1370.7445	-241.8000	-331446.0218
Total				-422182.1713

Menghitung Qr Produk

Qr Produk = n produk x Hf @298 K

Tabel L2.22 Neraca Energi Produk Reaktor (R-01)

Komponen	Massa (kg/jam)	mol (Kmol/jam)	Hf @ 298 K	Qr Reaktan (Kj/jam)
Etilen	10732.3232	382.5595	52.3000	20007.8600
Etanol	178.0215	3.8642	-234.8100	-907.3615
H_2O	31585.7711	1753.3040	-241.8000	-423948.9004
Total				-404848.4019

$$Q\ Reaksi = Qr\ Produk - Qr\ Reaktan$$

$$Q\ Reaksi = -404848.4019\ \frac{kj}{jam} - (-422182.1713)\frac{kj}{jam} = 17333.7694\frac{kj}{jam}$$

Kondisi Operasi Keluar

Tin = 470ºC = 743.15 K

Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tout} CpdT$$

Tabel L2.23 Neraca Energi Aliran Keluar Reaktor (R-01)

	Komponen	Massa (kg/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H out (KJ/jam)
Aliran 9	Etilen	10732.3232	382.5595	27993.9922	10709366.6624
	Etanol	178.0215	3.8642	42799.1857	165386.1979
	H_2O	31585.7711	1753.3040	15819.1716	27735816.4541
Total					38610569.3144

Kebutuhan Pemanas

$$\Delta Hin + Q\ pemanas = \Delta Hout + Q\ Reaksi$$

$$Q\ pemanas = 38610569.3144 \frac{kj}{jam} + 17333.7694 \frac{kj}{jam} - 38222662.4081 \frac{kj}{jam} = 405240.6757 \frac{kj}{jam}$$

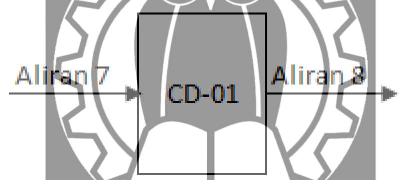
$$Massa\ Pemanas = \frac{Q\ Pemanas}{H\ Vap}$$

$$Massa\ Pemanas = \frac{405240.6757 \frac{kj}{jam}}{3457.79 \frac{kj}{kg}} = 117.1964 \frac{kg}{jam}$$

Tabel L2.24 Neraca Energi Total Reaktor (R-01)

Neraca Energi Total Reaktor		
Komponen	Masuk (KJ/jam)	Keluar (Kj/jam)
H in	38222662.4081	-
H out	-	38610569.3144
Q reaksi	-	17333.7694
Q pemanas	405240.6757	-
Total	38627903.0838	38627903.0838

L2.7 Neraca Condenser Parsial (CD-01)



GAMBAR L2.6 DIAGRAM ALIR CONDENSER PARSIAL (CD-01)

Temperatur operasi = 80°C

Tekanan Operasi = 6 atm

Fungsi = Mendinginkan keluaran dari reaktor fix bed multitube sebelum masuknya ke flash drum dimana ada dua fasa yaitu fasa cair (Etanol,air) dan gas (etilen).

L2.7.1 Neraca Massa Condenser Parsial (CD-01)

Tabel L2.25 Neraca Massa Condenser Parsial (CD-01)

Komponen	**Laju alir (kg/jam)**

Aliran	7	8
Etilen	10732.3232	10732.3232
Etanol	178.0215	178.0215
H_2O	31585.7711	31585.7711

L2.7.2 Neraca Energi Condenser Parsial (CD-01)

Kondisi Operasi Masuk Aliran 7

Tin = 470ºC = 743.15 K
Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tin} CpdT$$

Tabel L2.26 Neraca Energi Aliran 7 Masuk Condenser Parsial (CD-01)

Komponen	Massa (kJ/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H in (KJ/jam)
Etilen	10732.3232	382.5595	27993.9922	10709366.6624
Etanol	178.0215	3.8642	42799.1857	165386.1979
H_2O	31585.7711	1753.3040	15819.1716	27735816.4541
Total				38610569.3144

Kondisi Fase Gas

Tin = 97.75ºC = 370.62 K
Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tin} CpdT$$

Tabel L2.27 Neraca Energi Fase Gas Condenser Parsial (CD-01)

Komponen	Massa (kJ/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H in (KJ/jam)

Etanol	178.0215	3.8642	5179.7140	20015.6427
H_2O	31585.7711	1753.3040	2452.8250	4300547.8650
Total				4320563.5077

Kondisi Perubahan Fase Komponen

T = 97.47°C = 370.62 K

$$H = n\ x\ Hfg$$

Tabel L2.28 Neraca Energi Perubahan Fase Komponen Condenser Parsial (CD-01)

Komponen	Massa (Kg/jam)	mol (Kmol/jam)	Hfg (kJ/Kmol)	H Laten (kJ/jam)
Etanol	178.0215	3.8642	39019.3980	150780.2023
H_2O	31585.7711	1753.3040	39619.4434	69464927.4158
Total				69615707.6181

Kondisi Fase Liquid

Tin = 97.47°C = 370.62 K
Tref = 25°C = 298.15 K

$$H = n \int_{Treff}^{Tin} CpdT$$

Tabel L2.29 Neraca Energi Fase Liquid Condenser Parsial (CD-01)

Komponen	Massa (kJ/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H in (KJ/jam)
Etanol	178.0215	3.8642	8130.8687	31419.6039
H_2O	31585.7711	1753.3040	5451.6975	9558482.8515
Total				9589902.4554

Kondisi Operasi Keluar Aliran 8

Tout = 80ºC = 353.15 K

Tref = 25ºC = 298.15 K

$$H = n\int_{Treff}^{Tout} CpdT$$

Tabel L2.30 Neraca Energi Aliran 8 Keluar Condenser Parsial (CD-01)

Komponen	Massa (kg/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H out (KJ/jam)
Etilen	10732.3232	382.5595	8078.4814	3090499.5205
Etanol	178.0215	3.8642	6093.8897	23548.2342
H_2O	31585.7711	1753.3040	4136.2978	7252187.4459
		Total		10366235.2006

Kebutuhan Pendingin

$$H\ in\ + \Delta H\ Laten\ +\ \Delta H\ Liquid\ +\ \Delta H\ Gas\ =\ H\ out\ +\ Q\ Pendingin$$

$$Q\ pendingin = 38610569.3144\frac{kj}{jam} + 69615707.6181\frac{kj}{jam} + 9589902.4554\frac{kj}{jam} + 4320563.5077\frac{kj}{jam} - 10366235.2006\frac{kj}{jam} = 111770507.6950\frac{kj}{jam}$$

$$Massa\ Pendingin = \frac{Q\ Pendingin}{\Delta \text{H}\ pendingin}$$

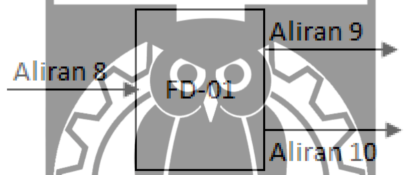
$$Massa\ Pendingin = \frac{111770507.6950\frac{kj}{jam}}{524.8136\frac{kj}{kg}} = 212971.8203\frac{kg}{jam}$$

Tabel L2.31 Neraca Energi Total Condenser Parsial (CD-01)

Neraca Energi Total Sub Cooler Condenser		
Komponen	Masuk (Kj/jam)	Keluar (Kj/jam)

H in	38610569.3144	-
ΔH Gas	4320563.5077	-
ΔH Laten	69615707.6181	-
ΔH Liquid	9589902.4554	-
H Out	-	10366235.2006
Q Pendingin	-	111770507.6950
Total	122136742.8956	122136742.8956

L2.8 Neraca Flash Drum (FD-01)



GAMBAR L2.7 DIAGRAM ALIR FLASH DRUM (FD-01)

Temperatur operasi = 80ºC

Tekanan Operasi = 6 atm

Fungsi = Tempat memisahkan campuran etilen dengan etanol dan air

.

L2.8.1 Neraca Massa Flash Drum (FD-01)

Tabel L2.32 Neraca Massa Flash Drum (FD-01)

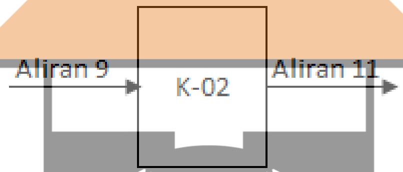
Komponen	Laju alir (kg/jam)		
Aliran	8	9	10

Etilen	10732.3232	10732.3232	0.0000
Etanol	178.0215	0.0000	178.0215
H_2O	31585.7711	0.0000	31585.7711

L2.8.2 Neraca Energi Flash Drum (FD-01)

Pada Flash Drum (FD-01) ini tidak melibatkan perubahan entalpi sehingga tidak dilakukan kalkulasi neraca energi.

L2.9 Neraca Kompresor (K-02)



GAMBAR L2.8 DIAGRAM ALIR KOMPRESOR (K-02)

Temperatur operasi = 80ºC

Tekanan Operasi = 27 atm

Fungsi = Menaikkan tekanan umpan dari 6 atm menjadi 27 atm menuju ke tangki produk (etilen) dalam bentuk fase gas.

L2.9.1 Neraca Massa Kompresor (K-02)

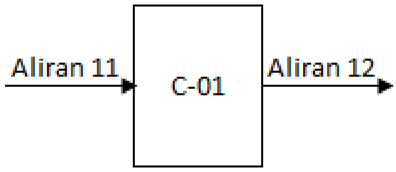
Tabel L2.33 Neraca Massa Kompresor (K-02)

Komponen	**Laju alir (kg/jam)**	
Aliran	**9**	**11**
Etilen	17802.1535	17802.1535

L2.9.2 Neraca Energi Kompresor (K-02)

Pada Kompresor (K-02) ini tidak melibatkan perubahan entalpi sehingga tidak dilakukan kalkulasi neraca energi.

L2.10 Neraca Cooler (C-01)



GAMBAR L2.9 DIAGRAM ALIR COOLER (C-01)

Temperatur operasi = 30ºC

Tekanan Operasi = 27 atm

Fungsi = Mendinginkan produk atas (etilen) dari flash drum sebelum masuknya ke tangki penyimpanan etilen (bola).

.

L2.10.1 Neraca Massa Cooler (C-01)

Tabel L2.34 Neraca Massa Cooler (C-01)

Komponen	**Laju alir (kg/jam)**	
Aliran	**11**	**12**
Etilen	17802.1535	17802.1535

L2.10.2 Neraca Energi Cooler (C-01)

Fungsi = Mendinginkan produk atas keluaran flash drum-01 sebelum masuk ke TP-03

Kondisi Operasi Masuk

Tin = 80ºC = 353.15 K

Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tin} CpdT$$

Tabel L2.35 Neraca Energi Masuk Cooler (C-01)

Komponen	Massa (kg/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H in (KJ/jam)
Etilen	10732.3232	382.5595	2537.3004	970668.2985
Total				970668.2985

Kondisi Operasi Keluar

Tout = 30ºC = 303.15 K

Tref = 25ºC = 298.15 K

$$H = n \int_{Treff}^{Tout} CpdT$$

Tabel L2.36 Neraca Energi Keluar Cooler (C-01)

Komponen	Massa (Kg/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H out (Kj/jam)
Etilen	10732.3232	382.5595	220.7231	84439.7070
Total				84439.7070

Kebutuhan Pendingin

$$Q\ Pendingin = \Delta Hin - \Delta Hout$$

$$Q\ pendingin = 970668.2985 \frac{kj}{jam} - 84439.7070 \frac{kj}{jam} = 886228.5915 \frac{kj}{jam}$$

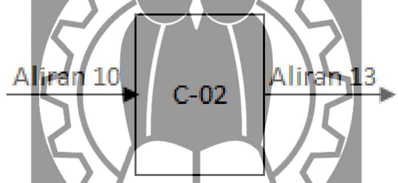
$$Massa\ Pendingin = \frac{Q\ Pendingin}{\Delta H\ Pendingin}$$

$$Massa\ Pendingin = \frac{886228.5915 \frac{kj}{jam}}{461.1250 \frac{kj}{kg}} = 1921.8838 \frac{kg}{jam}$$

Tabel L2.37 Neraca Energi Total (C-01)

Neraca Energi Total Cooler C-01		
Komponen	Masuk (KJ/jam)	Keluar (Kj/jam)
H in	970668.2985	-
H out	-	84439.7070
Q pendingin	-	886228.5915
Total	970668.2985	970668.2985

L2.11 Neraca Cooler (C-02)



GAMBAR L2.10 DIAGRAM ALIR COOLER (C-02)

Temperatur operasi = 30ºC

Tekanan Operasi = 6 atm

Fungsi = Mendinginkan produk bawah (etanol-air) dari flash drum sebelum masuknya ke tangki limbah.

.

L2.11.1 Neraca Massa Cooler (C-02)

Tabel L2.38 Neraca Massa Cooler (C-02)

Komponen	**Laju alir (kg/jam)**	
Aliran	**10**	**13**

Etanol	178.0215	178.0215
H2O	31585.7711	31585.7711

L2.11.2 Neraca Energi Cooler (C-02)

Fungsi = Mendinginkan produk bawah keluaran flash drum-01 sebelum masuk tempat pembuangan

Kondisi Operasi Masuk

Tin = 80ºC = 353.15 K

Tref = 25ºC = 298.15 K

$$H = n\int_{Treff}^{Tin} CpdT$$

Tabel L2.39 Neraca Energi Masuk Cooler (C-02)

Komponen	Massa (kg/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H in (KJ/jam)
Etanol	178.0215	3.8642	6093.8897	23548.2342
H_2O	31585.7711	1753.3040	6556.1705	11494959.8332
Total				11518508.0674

Kondisi Operasi Keluar

Tout = 30ºC = 303.15 K

Tref = 25ºC = 298.15 K

$$H = n\int_{Treff}^{Tout} CpdT$$

Tabel L2.40 Neraca Energi Keluar Cooler (C-02)

Komponen	Massa (Kg/jam)	mol (Kmol/jam)	Cp dT (Kj/Kmol)	H out (Kj/jam)

Etanol	178.0215	3.8642	537.0986	2075.4761
H_2O	31585.7711	1753.3040	549.3707	963213.9052
Total				965289.3813

Kebutuhan Pendingin

$$Q\ Pendingin = \Delta Hin - \Delta Hout$$

$$Q\ pendingin = 11518508.0674 \frac{kj}{jam} - 965289.3813 \frac{kj}{jam} = 10553218.6862 \frac{kj}{jam}$$

$$Massa\ Pendingin = \frac{Q\ Pendingin}{\Delta H\ Pendingin}$$

$$Massa\ Pendingin = \frac{10553218.6862 \frac{kj}{jam}}{461.1250 \frac{kj}{kg}} = 22885.8115 \frac{kg}{jam}$$

Tabel L2.41 Neraca Energi Total (C-02)

Neraca Energi Total Cooler C-01		
Komponen	Masuk (KJ/jam)	Keluar (Kj/jam)
H in	11518508.0674	-
H out	-	965289.3813
Q pendingin	-	10553218.6862
Total	11518508.0674	11518508.0674

LAMPIRAN 3 PERHITUNGAN SPESIFIKASI ALAT

L3.1 Perhitungan Spesifikasi Alat Proses

L3.1.1 Perhitungan Tangki Penyimpanan TP-01

Fungsi : Tempat penyimpanan etanol

Bentuk : Silinder tegak torispherical head dan flat bottom

Bahan : Stainless Steel SA-167 Tipe 304

Kondisi Operasi

Tekanan : 1 atm

Temperature : 30ºC = 303 k

Kapasitas : 17802.1535 kg/jam selama 3 hari.

Jumlah : 6 unit

Densitas : 766.9836 kg/m^3

Menghitung Volume Tangki

Kebutuhan bahan baku/hari = 17802.1535 kg/jam x 24 jam = 427251.6847 kg/hari

Kebutuhan bahan baku/3 hari = 427251.6847 kg/hari x 3 hari = 1281755.0542 kg/3 hari

$$\text{Volume Cairan NaOH} = \frac{\text{Laju Alir Massa}}{\text{Densitas}}$$

= 278.5272 m3

Faktor Keamanan = 20%

1.2

Volume Tangki = 334.2327 m3

Penentuan Ukuran Tangki

Menentukan Diameter Tangki

Diasumsikan H/D = 2, maka H = 2D dan asumsi tangki berbentuk silinder tegak.

H	=	2	D
	=	$1/4 \times \pi \times ID^2 \times H$	
	=	$1/4 \times \pi \times ID^2 \times 2ID$	
	=	$1/4 \times \pi \times 2ID^3$	
ID^3	=	212.8871	m3
ID	=	5.9710	m
	=	235.0803	inch
H	=	11.9421	m
	=	470.1606	inch

Menentukan Ketebalan Tangki

•Tinggi Cairan dalam Tangki

H tangki	=	11.9421	m
H cairan (hL)	=	$\frac{VL}{\pi/4 \times ID^2}$	$VL = 1/4 \times \pi \times D^2 \times H$ cairan
	=	9.9517	m
	=	391.8005	inch
P Hidrostatik	=	$hL \times \rho$ cairan x g	
	=	74801.5659	N/m^2
	=	0.7382	atm
	=	10.8491	psi
Faktor keamanan	=	20%	
	=	1.2	
P total	=	Faktor Keamanan x (P operasi + P hidrostatik)	
P Desain	=	1.7382	atm
	=	25.5450	psi

$$\text{Tebal tangki} = \frac{P \times ID}{2 \times f \times E - 0{,}6 \times P} + C$$

Keterangan

E =	efisiensi maksimum sambungan	=	80%	*Double welded butt joint tanpa diradiografi*
f =	Max Allowable stress	=	18750	psi
C =	faktor korosi	=	0.0125	inch/tahun
P =	tekanan desain	=	25.5450	psi
D =	Diameter tangki	=	235.0803	inch

Umur tangki	=	10	tahun
C, 10 tahun	=	0.125	Inch
Tebal tangki	=	0.3254	Inch
Di pilih tebal standar	=	1.1250	Inch
	=	0.0286	M

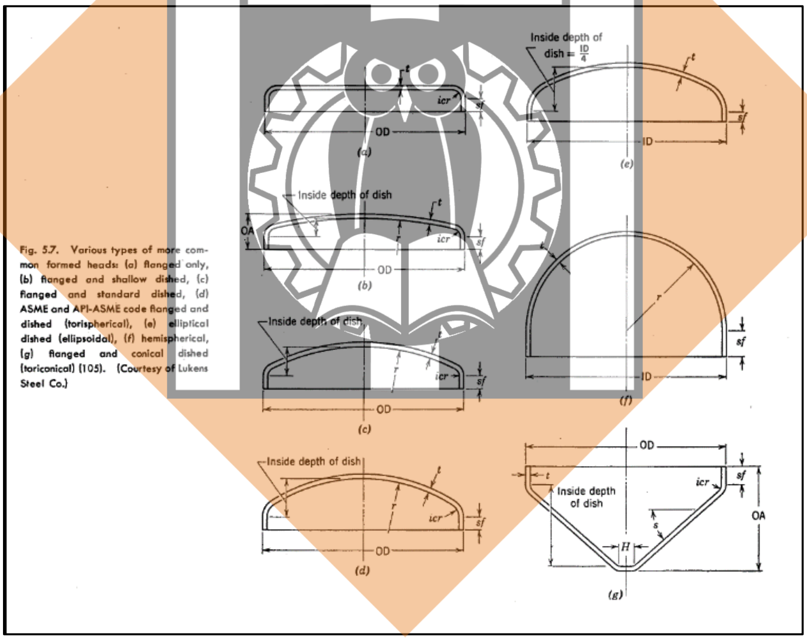


Fig. 5.7. Various types of more common formed heads: (a) flanged only, (b) flanged and shallow dished, (c) flanged and standard dished, (d) ASME and API-ASME code flanged and dished (torispherical), (e) elliptical dished (ellipsoidal), (f) hemispherical, (g) flanged and conical dished (toriconical) [105]. (Courtesy of Lukens Steel Co.)

GAMBAR L3.1 UKURAN BERBAGAI MACAM HEAD (BROWNELL&YOUNG FIG 5.7)

Menentukan Diameter dan Tinggi Tangki Sesungguhnya

Outside Diameter (OD)	=	ID + (2 x ts)	
	=	237.3303	inch
OD tabel standar	=	240	inch

	=	6.0960	M
Inside Diameter (ID)	=	OD - (2 x ts)	
	=	237.75	inch
	=	6.0388	M

Menentukan Ukuran Head

Table 5.4. Dimensions of Standard Flanged-only Heads for All Diameters
(Courtesy of Lukens Steel Company)

Gage (Thickness) t	Standard Straight Flange (in.) sf	Inside-corner Radius (in.) icr
3/16	1½–2	9/16
1/4	1½–2½	3/4
5/16	1½–3	15/16
3/8	1½–3	1 1/8
7/16	1½–3½	1 5/16
1/2	1½–3½	1 1/2
5/8	1½–3½	1 7/8
3/4	1½–3½	2 1/4
7/8	1½–4	2 5/8
1	1½–4	3
1 1/8	1½–4½	3 3/8
1 1/4	1½–4½	3 3/4
1 3/8	1½–4½	4 1/8
1 1/2	1½–4½	4 1/2
1 3/4	1½–4½	5 1/4
2	1½–4½	6

GAMBAR L3.2 STANDAR DIMENSI UNTUK HEAD FLANGED

Berdasarkan (Brownell & Young, 1959) Persamaan 7.76 , Persamaan 7.77 , dan tabel 5.8.

•Tebal Head (th) dan Tebal Bottom tangki (tb)

Inside Corner Radius (icr)	=	14.4375	inch
Crown radius (r)	=	180	inch
Icr/r	=	0.0802	
	=	8.021 %	
Jika icr/r ≥ 6% maka			
W	=	Faktor Stress Intensification untuk Torispherical Head	
W	=	1/4 x [3 + (r/icr)^1/2]	

= 1.6327 Inch

Tebal head (th) = $\dfrac{\text{P desain x r x W}}{\text{(2 x f x E) - (0,2 x P desain)}}$ + C

= 0.3746 inch

Tebal head standar = 0.4375 inch

= Design-good

Std Sraight Flange = 1.5 s.d 3.5 inch

(sf) = 2.5 inch

Diambil nilai tengahnya

Menentukan Tinggi Head/Tinggi Tangki Penutup (OA)

Berdasarkan pada (Brownell & Young, 1959) Figure 5.8.

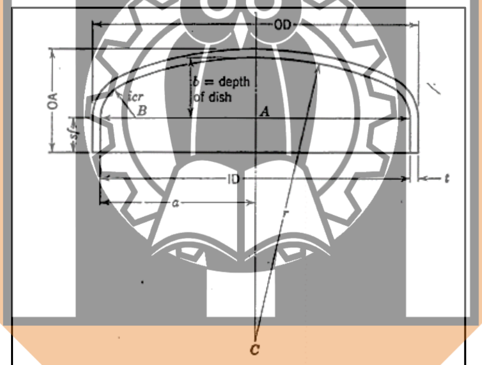


Fig. 5.8. Dimensional relationships for flanged and dished heads.

$$a = \frac{ID}{2}$$

$$b = r - \sqrt{(BC)^2 - (AB)^2}$$

$$AB = \frac{ID}{2} - (icr)$$

$$BC = r - (icr)$$

$$AC = \sqrt{(BC)^2 - (AB)^2}$$

$$OA = t + b + sf$$

GAMBAR L3.3 HUBUNGAN DIMENSI FLANGE DAN DISHED HEAD

ID = 237.75 inch

6.0388 m

a = r = jari-jari = 118.875 inch

= 3.0194 m

AB = a - icr	=	104.4375	inch
	=	2.6527	m
BC = rc - icr	=	165.5625	inch
	=	4.2053	m
AC = $((BC)^2-(AB)^2)^{\wedge}½$	=	128.4669	inch
	=	3.2631	m
b = rc - AC	=	51.5331	inch
	=	1.3089	m
OA (Tinggi head)	=	54.4706	inch
th + b + sf	=	1.3836	m

•Tinggi Tangki

H total = H + OA	=	524.6312	inch
	=	13.3256	m

L3.1.2 Perhitungan Tangki Penyimpanan TP-02

Fungsi : Tempat penyimpanan air

Bentuk : Silinder tegak torispherical head dan flat bottom.

Bahan : Stainless Steel SA-167 Tipe 304

Kondisi Operasi

Tekanan : 1 atm

Temperature : 30ºC = 303 k

Kapasitas : 24693.9623 kg/jam selama 1 hari.

Jumlah : 3 unit

Densitas : 1022.8753 kg/m^3

Menghitung Volume Tangki

Kebutuhan bahan baku/hari = 24693.9623 kg/jam x 24 jam = 592655.0950 kg/hari

Volume Cairan	=	$\frac{\text{Laju Alir Massa}}{\text{Densitas}}$	
	=	193.1337	m3
Faktor Keamanan	=	20%	
		1.2	
Volume Tangki	=	231.7605	m3

Menentukan Ukuran Tangki

Menentukan Diameter Tangki

Diasumsikan H/D = 2, maka H = 2D dan Asumsi tangki berbentuk silinder tegak

H	=	2	
	=	$1/4 \times \pi \times ID^2 \times H$	
	=	$1/4 \times \pi \times ID^2 \times 2ID$	
	=	$1/4 \times \pi \times 2ID^3$	
ID^3	=	147.6181	m3
ID	=	5.2850	m
	=	208.0717	inch
H	=	10.5700	m
	=	416.1435	inch

Menentukan Ketebalan Tangki

•Tinggi Cairan dalam Tangki

H tangki	=	10.5700	m	
H cairan (hL)	=	$\frac{VL}{\pi/4 \times ID^2}$		$VL = 1/4 \times \pi \times D^2 \times H$ cairan
	=	10.5700	m	
	=	416.1435	inch	

P Hidrostatik	=	hL x ρ cairan x g	
	=	105955.9404	N/m²
	=	1.0457	atm
	=	15.3676	psi
Faktor keamanan	=	20%	
	=	1.2	
P total	=	2.0457 x (P operasi + P hidrostatik)	
P Desain	=	2.2503	atm
	=	33.0700	psi

•Tebal Tangki

$$\text{Tebal tangki} = \frac{P \times ID}{2 \times f \times E - 0,6 \times P} + C$$

Keterangan

E =	efisiensi maksimum sambungan	=	80%	*Double welded butt joint tanpa diradiografi*
f =	Max Allowable stress	=	18750	psi
C =	faktor korosi	=	0.0125	inch/tahun
P =	tekanan desain	=	33.0700	psi
D =	Diameter tangki	=	104.0359	inch

Umur tangki	=	10	tahun
C, 10 tahun	=	0.125	inch

Tebal tangki	=	0.3547	inch
Di pilih tebal standar	=	0.875	inch
	=	0.0222	m

Menentukan Diameter dan Tinggi Tangki Sesungguhnya

Outside Diameter (OD)	=	ID + (2 x ts)	
	=	209.8217	inch
OD tabel standar	=	216	inch
	=	5.4864	m

Inside Diameter (ID)	=	OD - (2 x ts)	
	=	214.2500	inch
	=	5.4419	m

Menentukan Ukuran Head

•Tebal Head (th) dan Tebal Bottom tangki (tb)

Inside Corner Radius (icr)	=	13.00	inch		
Crown radius (r)	=	170	inch		
Icr/r	=	0.0765			
	=	7.6471%			
Jika icr/r ≥ 6% maka					
W	=	Faktor Stress Intensification untuk Torispherical Head			
W	=	1/4 x [3 + (r/icr)^1/2]			
	=	1.6541	inch		
Tebal head (th)	=	$\frac{P\ desain \times r \times W}{(2 \times f \times E) - (0{,}2 \times P\ desain)} + C$			
	=	0.4350	inch		
Tebal head standar	=	0.4375	inch		
	=	Design-good			
Std Sraight Flange	=	1.5	s.d	4	inch
(sf)	=	2.75	inch		
		Diambil nilai tengahnya			

•Tinggi Head / Tinggi tangki Penutup (OA)

ID	=	214.2500	inch
		5.4419	m
a = r = jari-jari	=	107.1250	inch
	=	2.7210	m

AB = a – icr	=	94.1250	inch
	=	2.3908	m
BC = rc – icr	=	157.0000	inch
	=	3.9878	m
AC = $((BC)^2-(AB)^2)$^½	=	125.6562	inch
	=	3.1917	m
b = rc – AC	=	44.3438	inch
	=	1.1263	m
OA (Tinggi head)	=	47.5313	inch
th + b + sf	=	1.2073	m

•Tinggi Tangki

H total = H + OA	=	463.6747	inch
	=	11.7773	m

L3.1.3 Perhitungan Tangki Penyimpanan TP-03

Fungsi : Tempat penyimpanan Etilen

Bentuk : Tangki bulat *(spherical).*

Bahan : Stainless Steel SA 204 Grade A

Kondisi Operasi

Tekanan : 27 atm

Temperature : 30ºC = 303 k

Kapasitas : 257575.7576 kg/jam selama 1 hari.

Jumlah : 1 unit

Densitas : 209.0144 kg/m^3

Menghitung Volume Gas

Kebutuhan bahan baku/hari = 10732.3232 kg/jam x 24 jam = 257575.7576 kg/hari

Volume Gas Etilen	=	$\frac{\text{Laju Alir Massa}}{\text{Densitas}}$	
	=	51.3473	m3

Menentukan Ukuran Tangki

Menentukan Diameter Tangki

Tangki berbentuk bola			
V tangki spheris	=	$4/3 \times \pi \times r^3$	
Jari-jari (r)	=	$((V \times 3)/(4 \times \pi))^{\wedge}⅓$	
	=	2.3061	m
	=	90.7921	inch
Diameter bola	=	4.6123	m
	=	181.5848	inch

Menentukan Tebal Tangki

P Opeasi	=	27	atm
	=	396.9000	psi
Faktor keamanan	=	20%	
	=	1.2	
P total	=	Faktor Keamanan x P operasi	
	=	32.4	atm
	=	476.2800	psi

Tebal tangki = $\frac{P \times r}{\quad}$

2 x f x E - 0.2P

Keterangan

r = jari jari dalam tangka	=	90.7921	inch
P = Tekanan desain	=	476.2800	psi
f = Max Allowable stress	=	18750	psi/inch^2
E = efisiensi maksimum sambungan	=	80%	

Tebal tangki	=	1.4414	inch
Tebal tangki standar	=	1.5	inch

Menentukan Diameter Sesungguhnya

Outside Diameter	=	ID + (2 x ts)	
	=	184.5848	inch
	=	4.6885	m

Menentukan Nozzle

Nozzle In

ρ	=	209.0144	kg/m3
	=	460.7931	lbm/ft3
Debit umpan (qf)	=	51.3473	m3/jam
	=	0.5036	ft3/s
Diopt	=	3.9 x qf^0.45 x ρ^0.13	
	=	6.3573	inch

Dari brownell, appendix K untuk pipa yang mendekati ukuran dipilih :

Nominal Pipe Size	=	8	inch
OD Pipe	=	8.625	inch
Schedule Number	=	40 ST	40 S
ID Pipe	=	7.981	inch

Nozzle Out

ρ	=	209.0144	kg/m3
	=	460.7931	lbm/ft3

Debit umpan (qf)	=	51.3473	m3/jam
	=	0.5036	ft3/s
Diopt	=	3.9 x qf^0.45 x ρ^0.13	
	=	6.3573	inch

Dari brownell, appendix K untuk pipa yang mendekati ukuran dipilih :

Nominal Pipe Size	=	10	inch
OD Pipe	=	10.075	inch
Schedule Number	=	40 ST	40 S
ID Pipe	=	10.02	inch

L3.1.4 Perhitungan Mixer M-01

Fungsi : Tempat pencampuran etanol dan air

Bentuk : Tangki silinder vertikal

Bahan : Stainless steel SA-167 grade 11 Tipe 316

Kondisi Operasi

Tekanan : 1 atm

Temperature : 30ºC = 303 k

Kapasitas : 42496.1158 kg/jam selama 1 jam.

Jumlah : 1 unit

Densitas : 922.3752 kg/m^3

Penentuan Volume Tangki (Vt)

Bahan masuk = 42496.1158 Kg

$$\text{Volume Cairan} = \frac{\text{Laju Alir Massa}}{\text{Densitas}}$$

= 46.0725 m³

Faktor Keamanan = 15%

1.15

Volume Tangki = 52.9834 m3

Penentuan Ukuran Tangki

Menentukan Diameter Tangki

Diasumsikan H/D = 2, maka H = 2D dan Asumsi tangki berbentuk silinder tegak

H = 2

= $1/4 \times \pi \times ID^2 \times H$

= $1/4 \times \pi \times ID^2 \times 2ID$

= $1/4 \times \pi \times 2ID^3$

ID^3 = 33.7303 m3

ID = 3.2310 m

= 129.2409 inch

H = 6.4620 m

= 258.4817 inch

Menentukan Ketebalan Tangki

•Tinggi Cairan dalam Tangki

H tangki = 6.4620 m

H cairan (hL) = $\dfrac{VL}{\pi/4 \times ID^2}$

= 5.6192 m

= 224.7667 inch

P Hidrostatik = hL x ρ cairan x g

= 50793.2167 N/m²

= 0.5013 atm

= 0.5079 bar

= 7.3672 psi

Faktor keamanan = 15%

= 1.15

P total = Faktor Keamanan x (P operasi + P hidrostatik)

P Desain = 1.7265 atm

= 25.3724 psi

•Tebal Tangki

Tebal tangki = $\dfrac{P \times ID}{2 \times f \times E - 0{,}6 \times P} + C$ *Brownell and Young, hal 254, Persamaan 13.1*

Keterangan

E = efisiensi maksimum sambungan	=	80%	*Double welded butt joint tanpa diradiografi*
f = Max Allowable stress	=	18750	psi
C = faktor korosi	=	0.0125	inch/tahun
P = tekanan desain	=	25.3724	psi
D = Diameter tangki	=	129.2409	inch

Umur tangki	=	10	tahun	
C, 10 tahun	=	0.125	inch	
Tebal tangki	=	0.2344	inch	
Di pilih tebal standar	=	0.4375	inch	*Brownell and Young, hal 93, Tabel 5.8*
	=	0.0109	m	

Menentukan Diameter dan Tinggi Tangki Sesungguhnya

Outside Diameter (OD)	=	ID + (2 x ts)	
	=	130.1159	inch
OD tabel standar	=	132	inch
	=	3.3	m
Inside Diameter (ID)	=	OD - (2 x ts)	
	=	131.1250	inch
	=	3.2781	m

Menentukan Ukuran Head

•Tebal Head (th) dan Tebal Bottom tangki (tb)

Inside Corner Radius (icr)	=	8	inch
Crown radius (r)	=	130	inch
Icr/r	=	0.0615	
	=	6.15%	

Jika icr/r ≥ 6% maka

W	=	Faktor Stress Intensification untuk Torispherical Head
W	=	1/4 x [3 + (r/icr)^1/2]
	=	1.7578 inch

Tebal head (th) = $\dfrac{\text{P desain x r x W}}{(2 \text{ x f x E}) - (0{,}2 \text{ x P desain})} + C$

= 0.3183 inch

Tebal head standar	=	0.375 inch
	=	Design-good
Std Sraight Flange	=	1.5 s.d 3 inch
(sf)	=	2.25 inch
		Diambil nilai tengahnya

•Tinggi Head / Tinggi tangki Penutup (OA)

Perhitungan Berdasarkan Brownell and Young, hal 87, Figure 5.8

ID	=	131.1250	inch
		3.2781	m
a = r = jari-jari	=	65.5625	inch
	=	1.6391	m
AB = a - icr	=	57.5625	inch
	=	3.0781	m
BC = rc - icr	=	122.0000	inch
	=	3.0500	m
AC = ((BC)²-(AB)²)^½	=	107.5665	inch
	=	2.6892	m
b = rc - AC	=	22.4335	inch
	=	0.5608	m
OA (Tinggi head)	=	25.0585	inch
th + b + sf	=	0.6265	m

•Volume Head (Vh)

Bentuk head dipilih *Flange and Dish Head (Torispherical)*

Bagian Lengkung *Torispherical head* (Vh')

Dianggap icr/r = 6% (tanpa bagian *sraight flange* (sf))

Vh'	=	$0.000049 \times ID^3$	
	=	110.4721	$Inch^3$
	=	0.0640	ft^3
	=	0.0018	m^3

Bagian *Straight flange head* (Vsf)

Volume *torispherical head* bagian (Vsf) dihitung sebagai bentuk silinder dengan ketinggian (H) = sf

Vsf	=	$\pi/4 \times ID^2 \times sf$	
	=	30383.8919	$Inch^3$
	=	17.5923	ft^3
	=	0.4980	m^3
V head total (Vh)	=	30494.3640	$Inch^3$
	=	17.6562	ft^3
	=	0.4998	m^3

•Tinggi Tangki

H total = H + (2OA)	=	308.5987	inch
	=	7.7150	m

Menentukan Luas permukaan Mixer

Luas permukaan (L)	=	L shell + L head	
	=	$(\pi \times OD \times H\ shell) + (2 \times (\pi \times De^2 / 4))$	*Brownell and Young, Hal 88 Persamaan 5.12*
Diameter ekivalen head (De)	=	$OD + (OD/42) + (2 \times sf) + (2/3 \times icr)$	
	=	144.9762 inch	
	=	3.6244 m	
L	=	87.6281 m2	

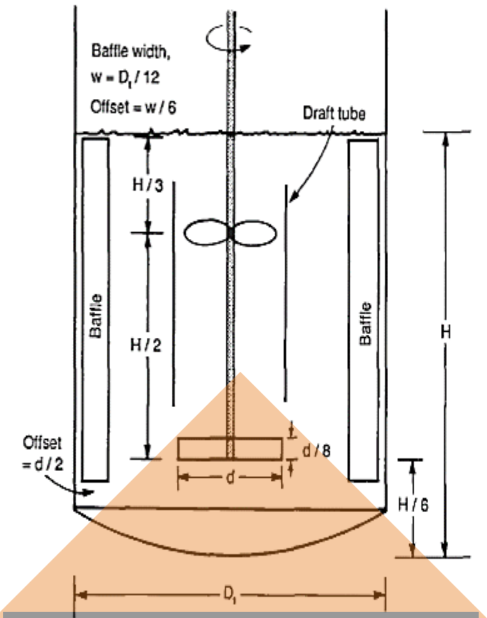


Figure 10.1. A basic stirred tank design, not to scale, showing a lower radial impeller and an upper axial impeller housed in a draft tube. Four equally spaced baffles are standard. H = height of liquid level, D_t = tank diameter, d = impeller diameter. For radial impellers, $0.3 \leq d/D_t \leq 0.6$.

GAMBAR L3.3 DESAIN BASIS TANGKI BERPENGADUK

Perancangan Pengaduk Mixer

Jenis pengaduk	=	Propeller	
Alasan	=	Dipergunakan untuk Cp 0.1-1000 Cp	

- **Menentukan Diameter Tangki**

Untuk turbin dengan 6 blades, D vessel / D¡ = 3 Mc Cabe, hal 235

Diketahui :

D vessel	=	ID	
	=	131.125	inch
	=	3.2781	m
D¡	=	Dvessel/3	
	=	43.7083	inch
	=	1.0927	m

- **Menentukan Tinggi blade (tb) dan lebar blade (wb)**

Tinggi blade (tb)	=	0.2 x D¡	Brown, hal 507
	=	8.7417	inch
	=	0.2185	m

	=	21.8542	cm
Lebar blade (wb)	=	0.1 x D_i	
	=	4.3708	inch
	=	0.1093	m
	=	10.9271	cm

- **Menentukan Lebar Baffle**

Jumlah baffle	=	4	
Lebar baffle	=	D vessel/12	Wallas, hal 287-288
	=	10.9271	inch
	=	0.2732	m

- **Menentukan Offset Top dan Offset Bottom**

Lebar blade (wb)	=	4.3708	inch
Diameter Impeller (D_i)	=	43.7083	inch
Offset top	=	Wb/6	
	=	0.7285	inch
	=	0.0182	m
Offset bottom	=	D_i/2	
	=	21.8542	inch
	=	0.5464	m

- **Menentukan Jarak Impeller dari Dasar Vessel**

Volume liquid	=	1/4 x π x ID² x H Cairan	
	=	46.0725	m³
Inside Diameter (ID)	=	131.1250	inch
	=	3.2781	m
H cairan (hL)	=	$\frac{VL}{\pi/4 \times ID^2}$	VL = 1/4 x π x D² x H cairan
	=	5.4588	m
H cairan maksimum	=	HL + OA	
	=	6.0853	m

Tinggi tepi bawah blade dari dasar vessel (Zi)/D untuk turbin degan 6 blades berdasarkan Brown.507

Zi/D_i = 0.75 s.d 1.3

	=	1.025	*diambil saja nila tengahnya*
Zi	=	44.8010	inch
	=	1.1200	m

- **Menentukan Banyak Pengaduk yang Digunakan**

Menetukan tinggi cairan maksimum yang dapat di jangkau pengaduk (zL)

ZL/D¡	=	2.7	s.d 3.9
	=	3.3	*diambil saja nila tengahnya*
ZL	=	144.2375	inch
	=	3.6059	m
Banyak Pengaduk (NT)	=	H cairan maksimum/ZL	
	=	1.6876	
	=	2	

- **Menentukan Jakarak Impeller ke-2** *Wallas, hal. 288, gambar 10.1*

Jarak Impeller 1 ke Impeller 2 (ΔH)	=	H cairan maksimum/2	
	=	3.0427	m
	=	121.7061	inch
Jarak Impeller 2 dari dasar tangki (Zi')			
Zi'	=	Zi + ΔH	
	=	4.1627	m
	=	166.5072	inch
Jarak impeller 2 ke permukaan cairan			
	=	H cairan maksimum - Zi'	
	=	1.9226	m
	=	76.9051	inch
Jarak Impeller 2 tanpa ada cairan dihitung			
	=	Total tinggi tangki - Zi'	
	=	5.7923	m

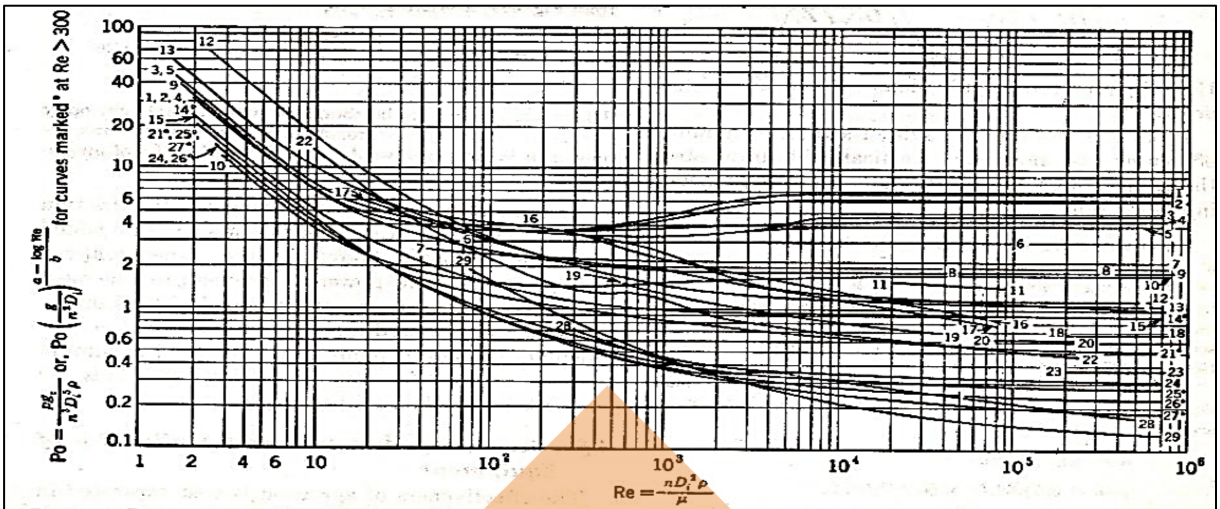


FIG. 477. Power consumption of various agitators expressed in terms of Po as a function of Reynolds number, Re. For curves marked with ° surface effects become important and the Froude number $Fr = g/n^2D_i$ is included as indicated for $Re > 300$.

GAMBAR L3.5 KONSUMSI DAYA UNTUK BERBAGAI MACAM PENGADUK

Menentukan Daya Pengadukan

- **Bilangan Reynold (Re)**

Nre	=	n x $(Di)^2$ x ρ / μ		*Brown, hal 507*	
ρ campuran	=	0.9224	g/cm^3	57.5562	lb/cu.ft
Di	=	43.7083	inch	3.6424	Ft
μ	=	0.8803	Cp	0.0006	lbmass/ft.s
gc	=	32.2	$lbm.ft/lbf.s^2$		
Nre	=	1290610.5568			

harga n diperoleh dari trial & error menggunakan fig 477 garis no 3

untuk turbine dengan 6 curved blades

Np dari grafik = 4.8

- **Menentukan Numper of Power (Np)**

Np	=	$(P \times gc)/(\rho \times n^3 \times Di^5)$		*Brown, hal 507*	
V cairan	=	46.0725	m3	12172.3862	US Gallon

P = 0.5 HP/1000 gallon (Wallas, hal 292, untuk blending dan mixture)

P	=	6.0862	HP
	=	3347.4062	ft.lb/s
	=	462.9767	kg.m/s

Np	=	$(462.9767 \times 9.8)/(922.3752 \times n^3 \times 1.0927^5)$
	=	$\frac{3.1576}{n^3}$

Untuk Np grafik	=	4.8	
n	=	0.8697	Rps
	=	52.1823	Rpm

- **Menentukan Tenaga Pengadukan**

N re Sesungguhnya, dengan harga Np= 4.8 dan n=52.1823rpm

Nre = 1122449.749

Karena Re yang diperoleh >10.000 maka P tidak bergantung terhadap Re

Sehingga rumus P adalah =

P	=	$(K_T \times \rho \times n^3 \times Di^5)/gc$	*Mc Cabe, Pers 9-24, hal 245*
K_T	=	4.8	*Mc Cabe, Pers 9-24, hal 245, untuk curved turbin 6 blade*

		462.976	
P	=	7	Kg.m/s
	=	6.0862	HP
Effisiensi motor 80%	=	7.6077	HP
1 impeller dibutuhkan P	=	3.8039	HP
2 impeller dibutuhkan P	=	7.6077	HP

TABLE 9.3
Values of constants K_L and K_T in Eqs. (9.21) and (9.23) for baffled tanks having four baffles at tank wall, with width equal to 10 percent of the tank diameter

Type of impeller	K_L	K_T
Propeller, three blades		
Pitch 1.0[40]	41	0.32
Pitch 1.5[35]	55	0.87
Turbine		
Six-blade disk[35] ($S_3 = 0.25$, $S_4 = 0.2$)	65	5.75
Six curved blades[40] ($S_4 = 0.2$)	70	4.80
Six pitched blades[39] (45°, $S_4 = 0.2$)	—	1.63
Four pitched blades[35] (45°, $S_4 = 0.2$)	44.5	1.27
Flat paddle, two blades[40] ($S_4 = 0.2$)	36.5	1.70
Anchor[35]	300	0.35

GAMBAR L3.6 NILAI KONSTANTA KL DAN KT

Menentukan Poros Pengaduk

- **Panjang Poros Pengaduk**

Tinggi mixer total (H total)	=	7.7150	m
Tinggi impeller dari dasar tangki (Zi)	=	1.1200	m
Panjang poros tangki (LP)	=	H Total - Zi	
	=	6.5949	m
	=	263.7976	Inch

Panjang poros antara bearing dan motor diambil = 0.3 m = 11.81 inch

Tebal head (Th)	=	0.375	inch
Panjang poror total (Lpt)	=	275.9826	inch
	=	6.8996	m
Jarak impeller 2 ke bearing	=	H total - Zi	
	=	3.5523	m

- **Diameter poros**

Pt	=	1.5 x P

K	=	1	
B	=	1.5	
Pt	=	11.4116	HP
Torsi (T)	=	826966.9672	inch.lb
Bending Moment (M)	=	Fm x Lp	
Fm (momen puntir)	=	Torsi maksimum/0.75 x ri	
Torsi maksimum	=	1240450.4509	inch.lb
Di	=	43.7083	inch
	=	1.0927	m
ri = jari-jari impeller = Di/2	=	0.5464	m
Fm (moment puntir)	=	3027219.4034	lb
M (Bending moment)	=	798573279.3531	inch.lb

$$D = \sqrt[3]{\frac{5{,}09}{S} \cdot \sqrt{(K \cdot T)^2 + (B \cdot M)^2}}$$

D poros	=	41.5497	inch
D standar	=	42	inch
jari-jari poros	=	21	inch

L3.1.5 Perhitungan Kompresor (K-01)

Fungsi : Menaikkan tekanan umpan dari 1 atm menjadi 6 atm menuju ke reaktor dalam bentuk fase gas

Bentuk : *Reciprocate*

Jumlah : 1 unit

Bahan : Carbon Steel (SA-285)

Kondisi Operasi

T_{in} = 126 °C = 399.15 K

Laju alir masuk, m = 42496.1158 kg/jam

= 708.2686 kg/menit

= 1561.4631 lb/menit

Tekanan masuk kompresor, P_1 = 1 atm

= 1 x 14.696 = 14,696 lb/in^2

Tekanan keluar kompresor, P_2 = 6 atm

= 6 x 14,696 = 88.176 lb/in^2

Menghitung kecepatan volumetrik umpan

Laju alir umpan, m = 42496.1158 kg/jam

= 42496.1158 / 60 = 708.2686 kg/min

= 708.2686 / 2.20462 = 1561.4631lb/min

Daya Kompresor, P

$$H_{AD} = \frac{ZRT_1}{(k-1)k}\left[\left(\frac{P_2}{P_1}\right)^{(k-1)/k} - 1\right]$$

(Rule Of Thumb 3rd Ed ,Hal 115)

$$HP = \frac{W \times H_{AD}}{E_A \times 33000}$$

(Rule Of Thumb 3rd Ed ,Hal 118)

Dimana :

Z = Faktor Kompresibilitas = 1.4145

k = Cp / Cv = Cp / (Cp - R) = 3.6519

(k-1)/k = (3.6519 – 1)/ 3.6519 = 0.7262

R = 1,544 x BM campuran = 1.544 x 29.7672 = 45.961/mol

T = temperatur bahan masuk (°R)

= ((126-32)/1,8)+459,7 = 511.9222°R

W = Laju alir Umpan (lb/min) = 1561.4631 lb/min

EA = 0.8

H_{AD} = 122521.0221

HP kompresor = 7246.6688

Digunakan kompresor yang di jual dipasaran dengan daya = 7300 HP

= 5500 kW

L3.1.6 Perhitungan Kompresor (K-02)

Fungsi : Menaikkan tekanan umpan dari 6 atm menjadi 27 atm menuju ke TP-03 dalam bentuk fase gas

Bentuk : *Reciprocate*

Jumlah : 1 unit

Bahan : Carbon Steel (SA-285)

Kondisi Operasi

T_{in} = 30 °C = 303.15 K

Laju alir masuk, m = 10732.3232 kg/jam

= 178.8721 kg/menit

= 394.3449 lb/menit

Tekanan masuk kompresor, P_1 = 6 atm

= 6 x 14.696 = 88.176 lb/in^2

Tekanan keluar kompresor, P_2 = 27 atm

= 27 x 14,696 = 396.792 lb/in^2

Menghitung kecepatan volumetrik umpan

Laju alir umpan, m = 10732.3232 kg/jam

= 10732.3232 / 60 = 178.8721 kg/min

= 178.8721 / 2.20462 = 394.3449 lb/min

Daya Kompresor, P

$$H_{AD} = \frac{ZRT_1}{(k-1)k}\left[\left(\frac{P_2}{P_1}\right)^{(k-1)/k} - 1\right]$$

(Rule Of Thumb 3rd Ed ,Hal 115)

$$HP = \frac{W \times H_{AD}}{E_A \times 33000}$$

(Rule Of Thumb 3rd Ed ,Hal 118)

Dimana :

Z = Faktor Kompresibilitas = 0.9015

k = Cp / Cv = Cp / (Cp - R) = 3.6519

(k-1)/k = (3.6519 – 1)/ 3.6519 = 0.7262

R = 1,544 x BM campuran = 1.544 x 27.7735 = 42.882/mol

T = temperatur bahan masuk (°R)

= ((30-32)/1,8)+459,7 = 458.5889°R

W = Laju alir Umpan (lb/min) = 394.3449 lb/min

EA = 0.8

H_{AD} = 58661.4749

HP kompresor = 876.2445

Digunakan kompresor yang di jual dipasaran dengan daya = 730 HP

= 550 kW

L3.1.7 Perhitungan Blower (BL-01)

Fungsi : Mengalirkan gas keluaran Vaporizer (V-01)

Bentuk : *Centrifugal Multiblade Backward Curved Blower*

Jumlah : 1 unit

Bahan : Carbon Steel (SA-285)

Kondisi Operasi

T_{in} = 126°C = 399.15 K

Laju alir masuk, m = 42496.1158 kg/jam

= 93687.8983 lb/jam

Tekanan masuk blower, P_1 = 1 atm

= 101.33 kPa

= 14.70 psi

Tekanan keluar blower, P2 = 1 atm

= 101.33 kPa

= 14.70 psi

ρ gas campuran = 2.8776 kg/m^3

= 0.1796 lb/ft^3

Menghitung kecepatan volumetrik umpan, Q

Laju alir umpan (Q) = W / ρ

= 42496.1158 / 2.8776 = 14768.1256 m^3/jam

= 521531.4353 ft^3/jam

= 8692.1906 ft^3/min

Daya Blower, P

Daya (kW) = $2.72x10^{-5}QP$ (*Perry, hal 10-46)*

Daya (kW) $= 2.72 \times 10^{-5}$ x 14768.1256 m^3/jam x 1 atm

= 0.4017 kW

= 0.5387 HP

Efisiensi blower pada Perry hal 10-46 sebesar 40 s.d 80%

Diambil efisiensi blower pada = 0.60

Daya blower aktual = Daya / Efisiensi

= 0.5387 HP / 0.60

= 0.8978 HP

Digunakan blower dengan daya = 1 HP

L3.1.8 Perhitungan Blower (BL-02)

Fungsi : Untuk mengalirkan produk atas (gas) keluaran Flash Drum (FD-01)

Bentuk : *Centrifugal Multiblade Backward Curved Blower*

Jumlah : 1 unit

Bahan : Carbon Steel (SA-285)

Kondisi Operasi

T_{in} = 80°C = 353.15 K

Laju alir masuk, m = 10732.3232 kg/jam

= 23660.7226 lb/jam

Tekanan masuk blower, P_1 = 6 atm

= 607.9500 kPa

= 88.1757 psi

Tekanan keluar blower, P2 = 6 atm

= 607.9500 kPa

= 88.1757 psi

ρ gas campuran = 25.3716 kg/m^3

= 1.5839 lb/ft^3

Menghitung kecepatan volumetrik umpan, Q

Laju alir umpan (Q) = W / ρ

= 10732.3232 / 25.3716 = 423.0051 m^3/jam

= 14938.2857 ft^3/jam

= 248.9714 ft^3/min

Daya Blower, P

Daya (kW) = 2.72x10^{-5}QP (*Perry, hal 10-46*)

Daya (kW) = 2.72x10^{-5} x 423.0051 m^3/jam x 6 atm

= 0.0690 kW

= 0.0926 HP

Efisiensi blower pada Perry hal 10-46 sebesar 40 s.d 80%

Diambil efisiensi blower pada = 0.60

Daya blower aktual = Daya / Efisiensi

= 0.0926 HP / 0.6090

= 0.1543 HP

Digunakan blower dengan daya = 1 HP

L3.1.9 Perhitungan Reaktor Fixed Bed Multitube (R-01)

Fungsi : Tempat terjadinya reaksi dehidrasi dan pembentukan produk etilen

Bentuk : Tangki Silinder

Bahan : Stainless Steel SA-167 Tipe 304

Kondisi operasi

Temperatur : 470 °C = 743.15 K

Tekanan : 6 atm = 88.1757 psia

Jumlah : 1 unit

Perancangan Dimensi Reaktor

Menentukan Volume Reaktor

Komposisi	Massa (Kg/Jam)	Massa (lb/Jam)	Mol (Kmol/Jam)	Mol (lbmol/Jam)
Etanol	17802.1535	39246.6277	386.4237	851.9097
Air	24693.9623	54440.3093	1370.7445	3021.9433
Total	42496.1158	93686.9369	1757.1682	3873.8530

Rumus :

V gas = n.R.T/ P *(J.M .Smitth and H.C. Van Nees.pers 13. hal. 71)*

Dimana :

R = 10.7310 ft3 . lbf/in2 . lbmol . °R

M = 42496.1158 kg/jam

= 42496.1158 kg/jam x 2.2046 = 93686.9369 lb/jam

n = 1757.1682kmol/jam

= 1757.1682 x 2.2046 = 3873.8530 lbmol/jam

T = 470 °C

= 470 °C x 1.8 + 491.67 = 1337.67°Ra

P = 6 atm

= 6 x 14.696 = 88.176 lb/in^2

GHSV = 4000/jam

Waktu tinggal = 1/GHSV = 0.9 detik

Maka : V gas = n x R x T / P

= 630640.6030 ft^3/jam

= 630640.6030 / 35.314 = 17858.0904m^3/jam

= 630640.6030 / 3600 = 175.1779 ft^3/detik

= 175.1779 x 0.9 detik = 157.6602 ft^3

Menentukan Perhitungan Tube

- Menentukan volume tube
  V gas = 157.6602 ft^3
  ε = 0.44

  V tube = V gas / ε = 358.3185 ft^3

- Menentukan kebutuhan katalis zsm-5 zeolit
  Densitas = 2300 kg/m^3

  = 2.300 x 2.68 / 35.31 = 174.5681 lb/ft^3

  V katalis = V tube – V gas

  = 358.3185 – 157.6602 = 200.6584 ft^3

  Berat katalis = V katalis x ρ

  = 200.6584 x 174.5681 = 35028.5533 lb

- Menentukan panjang tube yang berisi katalis
  V tube = $\pi \times di^2 \times L / 4$

  Pipa yang digunakan : NPS 3 in, Sch 40, l = 20 ft

  Berdasarkan *tabel. 11. Appendix Donald Q. Kern. hal. 844*, didapatkan data :

  a' = 7..38 in²

  = 7.38 x 0.0833^2 = 0.0512 ft²

  L = V tube / flow area

  = 358.3185 / 0.0512

  = 6997.1776 ft

- Menentukan jumlah tube

Nt = L / Panjang tube standard

= 6997.1776 ft / 20 = 349.8589 = 350 buah

Maka : Kecepatan gas (vo) = V / τ

vo = 157.6602 ft^3 / 0.9 detik = 175.1779 ft^3/detik

Rate setiap 1 tube = Kecepatan gas vo / N

= 175.1779 ft^3/detik / 350 buah = 0.5005 ft^3/detik

Volume setiap panjang tube yang berisi katalis = a' × l × ε

Dimana :

a' = Flow area (ft^2) = 0.0512 ft^2

l = Panjang tube standard yang berisi katalis (ft) = 20 ft

ε = Porosity = 0.44

V = a' × l × ε

= 0.0512 ft^2 x 20 ft x 0.44

= 0.4506 ft^3

- Menentukan luas tube

Direncanakan susunan tube berbentuk segitiga (trianguler pitch) dengan sudut 60°

Berdasarkan *tabel. 11. Appendix Donald Q. Kern. hal. 844*, didapat :

OD = 3.5 in

PT = OD + ¼ . OD = 4.375 in = 0.3644 ft

Tinggi segitiga :

t = PT × sin 60°

= 4.375 in x 0.866 = 3.7888 in

= 3.7888 in x 0.0833 = 0.3156 ft

Luas segitiga (triangular pitch) :

A = 1/2 × PT × t

= ½ x 4.375 in x 3.788 in = 8.2879 in^2

= 8.2879 in^2 x 0.0833^2 = 0.0575 ft^2

Dengan Nt = 350 buah, maka :

Luas tube = Nt × Luas segitiga

= 350 x 8.2879 in^2 = 2900.7617 in^2

= 2900.7617 in^2 x 0.0833^2 = 20.1281 ft^2

Asumsi : Luas tube = 80% × Luas total

Luas total = Luas tube / 0.8

= 2900.7617in^2 / 0.8 = 3625.9521 in^2

= 3625.9521 in^2 x 0.0833^2 = 25.1601 ft^2

Menentukan Dimensi Reaktor

- Menentukan diameter reaktor
  Luas total = π/4 × di²

  ID² = Luas total / (π/4)

  = 3625.9521 in^2 / 3.14 /4 = 4616.7057 in^2

  ID = 4616.7057$^{1/2}$ in^2 = 67.9463 in

  = 67.9463 in x 0.0833 = 5.6599 ft

  H = 4.5 . ID

  = 4.5 x 67.9463 in = 305.7585 in

  = 305.7585 in x 0.0833 = 25.4697 ft

  hg = H = 305.7585 in

- Menentukan tekanan design
  Diketahui : ρ campuran = 18.1060 lb/ft³

  Berdasarakan rumus :

  P desain = P operasi + P hidrostatik

  P hidrostatik = ρ campuran (hg – 1) /144

  = 18.1060 lb/ft³ x (25.4697 ft – 1) / 144

  = 3.0767 lb/ft^2 (psia)

  P design = 88.1760 psia + 3.0767 psia = 91.2527 psia

  = 91.2527 psia – 14.7 = 76.5527 psig

  Untuk faktor keamanan maka Pdesain ditambah 10%

  Pdesain = (100% + 15%) × (Poperasi + Phidrostatik)

  = 1.5 x 91.2527 psia = 104.9406 psia

  = 104.9406 psia – 14.7 = 90.2406 psig

- Menentukan tebal silinder reaktor
  ts = P × ID / [2 (f × E – 0.6 × P)] + C *(Brownell & Young. pers. 13.1. hal. 254)*

  = 90.2406 psig x 67.9463 in / [2 (18750 × 0.8 – 0.6 × 90.2406)] + 0.125

= 0.3301 in , maka diambil standar ts = 0.3750 in

OD = ID + 2 . ts

= 67.9463 in + 2 x 0.3750 = 68.6963 in

Berdasarkan tabel 5.7. Brownell & Young hal. 90, didapatkan :

OD'standard = 72 in = 6 ft

ID baru = OD' - 2 .ts

= 72 – 2 x 0.3750 = 71.2500 in = 5.9351 ft

Tebal silinder reaktor setelah di standardisari :

ts = P × ID / [2 (f × E – 0.6 × P)] + C *(Brownell & Young. pers. 13.1. hal. 254)*

ts = 0.3424 in

ts standar = 0.3750 in

- Menentukan tinggi silinder reaktor
  H standard = 4.5 . di baru

  = 4.5 x 71.2500 in = 320.6250 in

  = 320.6250 in x 0.0833 = 26.7081 ft

Menentukan Dimensi tutup

Menentukan tebal tutup atas dan tebal tutup bawah

Berdasarkan tabel 5.7. Brownell & Young hal. 90, didapatkan data :

r = 72 in

icr = 4.375 in

direncanakan tutup atas dan tutup bawah berbentuk standard dished heaad, sehingga tebal tutup atas = tebal tutup bawah

tha = thb = 0.885 × P × r / (f × E - 0.1 × P) + C

= 0.885 × 90.2406× 72 / (18750 × 0.8 - 0.1 × 90.2406) + 0.125

= 0.5086 in, maka diambil standar tha = thb = 0.6250 in

Menentukan tinggi tutup atas dan tinggi tutup bawah

Berdasarkan tabel 5.8. Brownell & Young hal. 93, didapatkan data :

sf = 2.5 in

icr = 4.375 in

direncanakan tutup atas dan tutup bawah berbentuk standard dished head, sehingga tinggi tutup atas = tinggi tutup bawah

a = ID/2

b = r - ((BC)² - (AB)²)^(1/2)

AB = ID/2 - icr = a - icr

BC = r - icr

AC =((BC)² - (AB)²)^(1/2)

ha = ts + b + sf (Brownell & Young. Fig. 5.8. hal. 87)

Dimana :

ID = diameter dalam = 71.2500 in

ts = tebal silinder = 0.3750 in

th = tebal tutup = 0.6250 in

r = 72 in

icr = 4.375 in

Sehingga :

a = ID/2 = 71.2500 in / 2 = 35.6250 in

AB = ID/2 - icr = a – icr

= 35.6250 – 4.375 = 31.2500 in

BC = r - icr

= 72 – 4.375 = 67.6250 in

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

= $(67.6250^2 – 31.2500^2)^{1/2}$ = 59.9715 in

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

= 72 – 59.9715 = 12.0285 in

OA = th+ b + sf *(Brownell & Young. Fig. 5.8. hal. 87)*

= 0.6250 + 12.0285 + 2.5

= 15.1535 in

Menentukan Tinggi reaktor Total

Rumus : H total = H + (2OA)

Dimana :

H total = Tinggi reaktor

OA = Tinggi head

H = Tinggi silinder

Maka :

H total = H + (2OA) = 350.9320 in

Perancangan Nozzle

Perancangan Nozzle pada tutup atas

Diketahui :

Rate bahan masuk = 42496.1158 kg/jam

= 42496.1158 kg/jam x 2.2046

= 93686.9369 lb/jam

ρ campuran = 18.1060 lb/ft^3

Maka :

Rate volumetrik (Q) = m / ρ

= 93686.9369 lb/jam / 18.1060 lb/ft^3

= 5174.3610 ft^3/jam

= 5174.3610 ft^3/jam / 3600

= 1.4373 ft^3/detik

Pemilihan diameter nozzle berdasarkan diameter pipa :

ID optimum = $3{,}9\ (Q)^{0{,}45}\ .\ (\rho)^{0{,}13}$ *(Klaus D. Timmerhaus, pers. 15. hal. 496)*

ID = $3{,}9 \times 1.4373^{0{,}45}$ ft^3/detik x $18.1060^{0{,}13}$ lb/ft^3

= 6.6908 in ≈ 8 in

Dipilih pipa standard berdasarkan *App. K. Brownell & Young. hal. 389*, yaitu :

D Nominal = 8 in

OD = 8.625 in

ID = 7.981 in

= 7.981 in x 0.0833 = 0.6648 ft

Laju alir fluida (v)

= Q/A = Q/((π/4) x ID^2)

= 1.4373 / ((3.14/4) x 0.6648^2)

= 4.1406 ft/dtik

Perancangan Nozzle pada silinder

- Nozzle untuk pemasukan dan pengeluaran pendingin
  Diketahui :

  Rate pemanas masuk = 117.8056 kg/jam = 259.7142 lb/jam

  ρ pemanas = 0.9922 g/cm^3 = 61.9409 lb/ft^3

  Rate volumetrik (Q) = m / ρ

  Q = 259.7142 lb/jam / 61.9409 lb/ft^3

  = 4.1929 ft^3/jam

  = 4.1929 ft^3/jam / 3600

  = 0.0012ft^3/detik

  Pemilihan diameter nozzel berdasarkan diameter pipa :

  Asumsi aliran fluida adalah turbulen NRE > 2100, maka :

  ID optimum = 3,9 (Q)^{0,45} . (ρ)^{0,13} *(max. peters, pers. 15. hal. 496)*

  ID = 3.9 x 0.0012^{0,45} ft^3/detik x 61.9409^{0,13} lb/ft^3

  = 0.3190 in ≈ 0.375 in

  Dipilih pipa standard berdasarkan *App. K. Brownell & Young. hal. 389*, yaitu :

  D Nominal = 0.375 in

  OD = 0.675 in

  ID = 0.493 in = 0.493 in x 0.0833 = 0.0411 ft

  Laju alir fluida (v) = Q/A = Q/((π/4) x ID^2)

  = 0.0012 ft^3/detik x 4 /3,14 x 0.0411^2 ft^2

  = 0.8793 ft/dtik

Perancangan Nozzle pada pengeluaran produk

Komposisi	Massa (Kg/Jam)	Massa (lb/Jam)	Mol (Kmol/Jam)	Mol (lbmol/Jam)
Etilen	10732.3232	23660.4798	382.5595	843.3906
Etanol	178.0215	392.4663	3.8642	8.5191
Air	31585.7711	69633.9909	1753.3040	3865.3339
Total	**42496.1158**	**93686.9369**	**2139.7277**	**4717.2436**

Rumus :

V gas = n.R.T/ P (*J.M .Smitth and H.C. Van Nees.pers 13. hal. 71)*

Dimana :

R = 10.7310 ft3 . lbf/in2 . lbmol . ºR

m = 42496.1158Kg/jam = 93686.9369 lb/jam

n = 2139.7277 Kmol/jam = 4717.2436 lbmol/jam

T = 470 °C = (270 x 1,8) + 491,67 = 1337.67°Ra

P = 6 atm = 6 x 14.696 = 88.1760 lb/in^2

Maka :

V gas = 767939.6564 ft^3/jam

= 767939.6564 ft^3/jam / 3600 = 213.3166 ft^3/detik

= 213.3166 ft^3/detik x 1.8 detik = 383.9698 ft^3

Diketahui :

Rate bahan keluar = 42496.1158 kg/jam = 93686.9369 lb/jam

ρ campuran = 17.5941 lb/ft^3

rate volumetrik (Q) = m / ρ

= 93686.9369 lb/jam / 17.5941 lb/ft^3

= 5324.9151 ft^3/jam = 5324.9151 ft^3/jam / 3600

= 1.4791 ft^3/detik

Pemilihan diameter nozzle berdasarkan diameter pipa :

Asumsi aliran fluida adalah turbulen NRE > 2100, maka :

ID optimum = 3.9 .(Q)0.45 . (ρ)0.13 *(Klaus D. Timmerhaus, pers. 15. hal. 496)*

ID = 3.9 x 1.4791$^{0.45}$ ft^3/detik x 17.5941$^{0.13}$ lb/ft^3

= 6.7525 in ≈ 8 in

Dipilih pipa standard berdasarkan *App. K. Brownell & Young. hal. 388*, yaitu :

D Nominal = 8 in

OD = 8.625 in

ID = 7.891 in = 7.891 in x 0.0833 = 0.6648 ft

Laju alir fluida (v) = Q/A = Q/((π/4) x ID^2) *(Christie J. Geankoplis, hal. 49)*

= 1.4791 ft^3/detik x 4 / 3.14 x 0.6648^2 ft^2

= 4.2610 ft/detik

L3.1.10 Perhitungan Flash Drum (FD-01)

Fungsi : Memisahkan etilen dengan etanol dan air

Bentuk : *Vertical Vessel*

Bahan : Stainless Steel SA-167 Tipe 304

Kondisi operasi

Temperatur : 80 °C = 353 K

Tekanan : 6 atm = 88.1757 psia

Jumlah : 1 unit

Densitas uap

Komponen gas

Komponen	m (kg)	BM	n (kmol)	yi	BM camp.
Etilen	10732.3232	28.054	382.6140	1	28.054
	10732.3232		**382.6140**		**28,054**

Maka :

ρ_{gas} = (P x BM camp) / (R x T)

ρ_{gas} = 6 x 28,054 / 0.08206 x 353 = 5.8109 kg/m^3

Densitas Cair

Komponen	m (kg)	BM	n (kmol)	xi	ρ (kg/m3)	ρ camp.
Etanol	178.0215	46,069	3.8642	0.0022	1.6111	0.0035
Air	31585.7711	18,015	1753.3040	0.9978	973.4952	973.4952
	31763.7926		**1757.1682**	1.0000		**971.3579**

$\rho_L = 971.3579\ kg/m^3$

$\rho_{cairgas}$ $= \rho L \times (QL / (QL + M\ gas)) + \rho_{gas} \times (M\ gas / (QL + M\ gas))$

$= 727.5106 kg/m^3$

Penentuan Laju Alir Volumetrik Cairan dan uap

• Laju Alir Volumetrik Cairan (QL)

$$Q_L = \frac{W_L}{\rho_L}$$

dimana :

Q_L = Laju Volumetrik cairan (m^3/jam)

W_L = Laju alir massa cairan

= 31763.7926 kg/jam

ρ_L = densitas cairan

$= 971.3579\ kg/m^3$

Maka :

Q_L = 31763.7926kg/jam / 971.3579 kg/m^3

Q_L = 31763.7926 m^3/jam

= 0.0091 m^3/detik

• Laju Alir Volumetrik Uap (Qv)

$$Q_v = \frac{Wv}{\rho v}$$

dimana :

Q_v = Laju Volumetrik uap (m^3/jam)

W_v = Laju alir massa uap

= 10732.3232 kg/jam

ρ_v = densitas uap

= 5.8109 kg/m^3

Maka :

Q_v = 10732.3232 kg/jam / 5.8109 kg/m^3

Q_v = 1846.9448 m^3/jam

= 0.5130 m^3/detik

Penentuan Volume Vessel

• Volume Cairan,VL

V_L = QL x t

dimana :

V_L = volume ruang cair (m^3)

t = waktu tinggal (menit)

ditetapkan : 5 menit = 300 detik

maka :

V_L = QL x t

V_L = 0.0091 m^3/detik x 300

V_L = 2.7250 m^3

● **Volume Uap, Vv**

V_v = Qv x t

dimana :

V_v = volume ruang uap (m^3)

t = waktu tinggal uap

= 5 menit = 300 detik

maka :

V_v = Qv x t

V_v = 0.5130 m^3/detik x 300 detik

V_v = 153.9121 m^3

Jadi volume total = $V_L + V_v$

= 2.7250 m^3 + 153.9121 m^3

= 156.6371 m^3

Faktor keamanan = 20%, sehingga :

= 1,2 x 156.6371 m^3

$= 187.9645\ m^3$

$= 187.9645 \times 264{,}172 = 49654.9636$ gal

Penentuan dimensi Vessel

Tinggi cairan

Ditetapkan H/D = 3, maka H = 3 x D

dimana :

H = tinggi cairan (m)

D = diameter vessel (m)

maka :

Volume Vessel = $\frac{\pi.ID^2.H}{4}$

$$D^3 = \frac{V}{(1/4).\pi.3}$$

$$D^3 = \frac{187.9645\,m^3}{2{,}3550}$$

$$D^3 = 79.8151\,m$$

Maka

D = 4.3055 m = 169.5098 in

H = 3 x 4.3055 m

H = 12.9166 m = 508.5294 in

Penentuan ketinggian ruang cair dan uap

$$H_L = \frac{VL}{(1/4).\pi.D^2}$$

$$H_v = H - HL$$

dimana :

H_L = Ketinggian ruang cair (m)

H_v = Ketinggian ruang uap (m)

maka :

H_L = 2.7250 m^3 / 14.5521

H_L = 0.1873 m

H_v = 12.9166 m – 0.1873 m

H_v = 12.7294 m

Penentuan tebal dinding Vessel (ts)

$$t_s = \frac{P_{design}\ (OD_t\ /\ 2)}{(f.E) - (0{,}6P_{design})} + C$$

dimana :

P hidrostatik (ρgh) = 0.0176 atm

t_s = tebal vessel (inch)

P tekanan design = 7.2211 atm

= 106.1503 psi

E = efisiensi sambungan (doeble welded) = 0.8

Korosi yang diizinkan = 0,0125/tahun, diperkirakan umur alat 10 tahun maka:

C = 0.1250

Berdasarkan data, untuk bahan Stainless Steel SA-167 Grade 11 tipe 316

(tabel 13.1, Brownell and Young)

F = Maximum allowable working stress = 18750 psi

maka :

r = 1/2 D

ID = 169.5098 in

r = 84.7549 in

maka nilai ts :

$$ts = \frac{P_{design}\ (ID_T)}{2(f.E) - (0{,}6\,P_{design})} + C$$

t_s = 0.7273in

Digunakan tebal dinding standar

t_s = 0.75 (3/4) in *(Tabel 5.6 Brownell and Young)*

Menentukan Dimensi Head Vessel

Bentuk : Torispherical head

Bahan : Stainless Steal SA 167 Grade 3 Tipe 304

(tabel 13.1, Brownell and Young)

• **Menentukan Tebal Head**

P design = 106.1503 psi

F = Allowable working Stress = 18750 psi

E (double welded) = Effisiensi sambungan = 0.8

C = Faktor korosi = 0.125

OD = ID + 2 t

ID = 169.5098 + 2 x 0.75

OD = 171.0098 in

= 4.3437 m

Digunakan OD standar, OD = 180 in *(Dari tabel 5.7, Brownell and Young)* ID standar = 180 – 1,5 = 178.5

icr = 11 in

r = 170 in

Maka :

icr/r = 11/170

= 0.0647

Maka perhitungan tebal head :

W = $^{1}/_{4}$ [3 + (r/icr)^{0,5}] = 1.7328 *(Pers. 7.76, Brownell and Young)*

$$\frac{P_{design}\ rc.W}{2f.E - 0,2P_{design}} + C$$

th =

= 1.1681 in

Digunakan tebal penutup vessel standar

th = 1.25 (1 1/4) in

• Menentukan Tinggi Head

Dari Brownell & Young, tabel 5.6, untuk th = 1.25 inch diperoleh

s_f (standart straight flange) = 1 1/2 - 4 1/2 inch,

diambil s_f = (1.5 + 4.5) /2 =3 *(Tabel 5.6, Brownell and Young)*

a	= ID / 2	= 89.25 in
AB	= (ID / 2) - (icr)	= 78.25 in
BC	= r - (icr)	= 159 in
AC	= $(BC^2 - AB^2)^{0,5}$	= 138.4122 in
b	= r - $(BC2 - AB2)^{0,5}$	= 31.5878 in

Dari Fig 5.8 (Brownell and Young)

Maka :

Tinggi Head : OA = $t_h + b + s_f$

OA = 1.25 + 31.5878 + 3

$OA = 35.8378$ in

$OA = 0.9103$ m

Menentukan ukuran Total Vessel

- **Menentukan Tinggi Total Vessel**

$H_t = H_s + 2H_h$

Dimana :

H_t = Tinggi total vessel

H_s = Tinggi vessel

= 12.9166 m

H_h = Tinggi Head = OA = 35.8378 in

= 0.9103 m

$2H_h$ = 71.6756 in

= 1.8206 m

Maka :

H_t = 12.9166 + 1.8206 = 14.7372 m

= 580.2050 in

Perancangan Noozle

- **Noozle untuk aliran gas**

Di opt $= 3{,}9 x q f^{0.45} x \rho^{0.13}$ *(pers. 13,15 Peters)*

Dimana :

D_{iopt} = optimum inside diameter ,in

qf = W / ρ

qf = fluid flow rate ,ft^3/s

ρ = fluid density , lb/ft^3

W gas = 894.3603 kg/jam

= 0.5481 lb/s

ρ = 5.8109 kg/m^3

= 0.3631 lb/ft^3

qf = 0.5481 / 0.3631= 1.5098 ft^3/s

Maka Di opt = 3,9 x $1.5098^{0,45}$ x $0,3631^{0,13}$ = 4.1151 in

Diameter yang digunakan = 4.25 in

- **Noozle untuk aliran cair**

$D_{iopt} = 3{,}9 x qf^{0.45} x \rho^{0.13}$ *(pers. 13,15 Peters)*

Dimana :

D_{iopt} = optimum inside diameter ,in

qf = W / ρ

qf = fluid flow rate ,ft^3/s

ρ = fluid density , lb/ft^3

W cair = 2646.9827 kg/jam

$= 1.6223$ lb/s

ρ $= 971.3579$ kg/m^3

$= 60.6893$ lb/ft^3

qf $= 1.6223 / 60.6893 = 0.0267$ft^3/s

D_{iopt} $= 1.3033$ in

Diameter yang digunakan = 1.5 in

L3.1.11 Perhitungan Vaporizer VP-01

Fungsi	: Memanaskan serta menguapkan bahan baku menjadi bahan baku etanol 42% (fase gas) dari suhu 30˚C menjadi 126˚C
Bentuk	: Horizontal heater
Bahan	: *Stainless steel* SA-167 *grade* 11 Tipe 316
Tekanan	: 2.6 atm

- Fluida dingin (Campuran) ;

  t1 : 30 °C = 86 °F

  t2 : 126 °C = 258 °F

- Fluida Panas (*Superheated Steam*) ;

  T1 : 480 °C = 896 °F

  T2 : 470 °C = 878 °F

Laju alir massa (Ws)	: 42496.1158 Kg/Jam
	: 93687.7869 lb/jam
Kebutuhan panas yang diserap (Q)	: 91140109.3881 KJ/Jam
	: 86437279.7437 Btu/jam
Kebutuhan steam (Wt)	: 26357.9076 Kg
	: 58109.1703 lb

Menetukan True Temperatur Difference (Δtrue):

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

LMTD = 711.7967 °F

Luas permukaan transfer panas (A)

Dari Tabel 8 Kern, untuk :

steam-light organics adalah 100-200 BTU/jam.ft².F

Maka:

Diambil Ud = 150 Btu/jam.ft².ºF)

Untuk : Fluida panas : steam (tube side)

Fluida dingin : etanol-air (shell side)

$$A = \frac{Q_C}{U_D.LTMDkoreksi}$$

$= 809.5689\ ft^2$ *$<200\ ft^2$ Shell and tube heat exchanger*

Menentukan Tc average dan tc average

Tc average $= (T_1+T_2)/2$ = 887 ºF

tc average $= (t_1+t_2)/2$ = 172.4 ºF

Dari Hal 843, Kern , diperoleh Informasi :

Panjang pipa, L = 6 ft

Pemilihan route fluida :

Shell side = etanol-air (cold fluid)

Tube side = steam (hot fluid)

aliran = counter current

Dipilih penukar panas shell and tube, L = 6 ft dengan spesifikasi :

Berdasarkan Pers. 14.34 Brownell & Young:

ODt =	1.0000	in
f = allowable stress =	18750.0000	psi
E = efisiensi penyambungan =	0.8000	
c = faktor korosi =	0.1250	in/10 tahun
P = tekanan desain (faktor keamanan 20%) =	45.2519	psi

$$tm = \left[\frac{Pi.ODt}{2.(FE-0.6P)}\right] + C$$

tm = 0.1265 In
(tebal minimum tube)

Menentukan Jumlah Tube (Nt) :

Dari Kern tabel 10 hal 843, dipilih:

ODt =	1.0000	in
Panjang tube =	6.0000	ft
t =	0.1265	in
BWG =	8.0000	
IDt	0.6700	in
a't	0.3550	in^2
Surface per lin (ao) =	0.2618	ft^2/ln ft

Jumlah tube, Nt $= \dfrac{A}{a_o.L}$

$= \dfrac{809.5689\ ft^2}{0.2618\ X\ 6\ ft}$

Nt = 515.3864 buah

Dari Kern tabel 9, hal 841, dipilih (square pitch):

Nt = 520.0000 buah

N = 2.0000 passes

Pt = 1.2500 in, square pitch

IDs = 32.6400

Menentukan Ud terkoreksi

A terkoreksi = Nt x ao x L

A terkoreksi = 816.8160 ft^2

Ud terkoreksi $= \dfrac{Q}{\text{A terkoreksi x LMTD}}$

$= \dfrac{86437279.7437\ Btu/jam}{816.8160\ ft^2 \times 711.7967\ ^oF}$

Ud terkoreksi = 148.6691 Btu/jam.ft².°F)

Menentukan faktor kekotoran (dirty factor)					
Shell Side (etanol dan air)			**Tube Side (steam)**		
Data:			**Data:**		
tc avg =	172.4000	F	Tc avg =	887.0000	F
viscositas, μ =	0.81087	Cp	viscositas, μ =	0.0286	cp
	1.9623	lb/ft.hr		0.0693	lb/ft.hr

,Cp = 1.0788 Btu/lb.°F
Konduktivitas panas, k = 0.2987 Btu/ft².jam(°F/ft)

Menentukan flow area, As:

IDs (in diameter) = 32.6400 In
B (bafflespace) = 1/4*IDs= 8.1600 In
Pt (harga pitch) = 1.2500 In
c (harga pitch - OD) = 0.2500 In

$$as = \frac{IDs \times C' \times B}{Pt \times 144}$$

as = $\frac{32.6400 \times 0.2500 \times 8.1600}{1.2500 \times 144}$

as = 0.3699 ft²

Menentukan kecepatan massa, Gs:

Gs = Ws/As = $\frac{93687.7869 \text{ lb/jam}}{0.3699 \text{ ft}^2}$

= 253264.9948 lb/jam.ft²

Menentukan bilangan Reynolds, Re_s :

Untuk Odt 1, De = 0.9900 in
fig 28 Kern 0.1650 ft
Re_s = De x Gs/μ
$Re_s = \frac{0.1650 \times 253264.9948}{1.9623}$
Re_s = 21295.7340

dari Kern Fig 28, diperoleh harga J_H = 100.0000

Menentukan koefisien perpindahan pipa lapisan luar, ho:

$$ho = \frac{jH \times K}{De}\left(\frac{C \times \mu}{K}\right)^{\frac{1}{3}}(\phi_s)$$

ho/Φs = 347.7077 Btu/ft².jam(°F/ft)

$$Tw = tc + \frac{(hio/\phi t)}{(hio/\phi t) + (ho/\phi s)}(Tc - tc)$$

Tw = 176.9667 °F

Dari *Chemical properties*, μw = 0.38006 Cp
0.9197 lb/ft.hr

Φs = (μ/μw) $^{0.14}$= 1.1119

ho = 386.6237 Btu/ft².jam(°F/ft)

Cp = 185.5856 Btu/lb.°F
Konduktivitas panas, k = 12.0601 Btu/ft².jam(°F/ft)

Menentukan flow area, at:

n (jumlah passes) = 2.0000
Nt (jumlah tube) = 520.0000
A't (flow area per tube) = 0.3550 in²

$$at = \frac{Nt \times a't}{n \times 144}$$

at = $\frac{520.0000 \times 0.3550}{2.0000 \times 144}$

at = 0.6410 ft²

Menentukan kecepatan massa, G_t :

Gt = Wt/At
= $\frac{58109.1703 \text{ lb/jam}}{0.6410 \text{ ft}^2}$
= 90657.8605 lb/jam.ft²

Menentukan bilangan Reynolds, Re_s :

IDt = 0.6700 in
Table 10, Kern 0.1117 ft
Re_t = Dt x Gt/μ =
= $\frac{0.1117 \times 90657.8605}{0.0693}$
Re_t = 146126.6647
L/D = 53.7313
dari Kern Fig 24, diperoleh harga J_H = 490.0000

Menentukan koefisien perpindahan pipa lapisan dalam, hi:

$$hi = \frac{jH \times k}{D}\left(\frac{C \times \mu}{k}\right)^{\frac{1}{3}}(\phi_t)$$

hi/Φt = 54061.3998 Btu/ft².jam(°F/ft)
Tw = 176.9667 F
μw = 0.0693 lb/ft.hr

$\Phi t = (\mu/\mu w)^{0.14}$
$\Phi t = (0.0693/0.0693)^{0.14}$
= 1.0000

hi = 54061.3998 Btu/ft².jam(°F/ft)

$$hio = hi\left(\frac{IDt}{ODt}\right)$$

hio = 54061.3998 x $\frac{(0.1117)}{(1.0000/6)}$
hio = 36221.1379 Btu/ft².jam(°F/ft)

Menentukan Clean Overall Coefficient, Uc:

$$Uc = \frac{hio \times ho}{hio + ho}$$

$$Uc = \frac{36221.1379 \times 386.6237}{36221.1379 + 386.6237}$$

= 382.5405 Btu/ft2.jam(°F/ft)

Menentukan faktor kekotoran, Rd:

$$Rd = \frac{Uc - Udact}{Uc \times Udact}$$

$$Rd = \frac{382.5405 - 148.6691}{382.5405 \times 148.6691} = \mathbf{0.004}$$

Rd >= 0.002, jadi syarat terpenuhi

Rd air =0.001 Rd etanol = 0.001

Menentukan Perubahan Tekanan

Shell Side (etanol dan air)

Re_s = 21295.7340

fs (fig 29, Kern) = 0.0500 ft^2/in^2

$$\Delta Ps = [8 j_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{L_b}\right)\left(\frac{u}{u_w}\right) - 0{,}14]\frac{pv_s^2}{2}$$

ΔPs = 0.7616 psi

$$v_s = \frac{W_s}{p_s \, x \, a_s}$$

Vs = 0.0280 m/s

Karena ΔPs < 10 psi, maka alat tersebut layak.

Tube Side (steam)

Re_t = 146126.6647

ft (fig 26, Kern) = 0.0000825 ft^2/in^2

$$\Delta P_t = Np[8 j_f \left(\frac{L}{d_i}\right)\left(\frac{u}{u_w}\right)^{-m} + 2{,}5]\frac{pv_t^2}{2}$$

ΔPt = 0.7277 psi

$$Vt = \frac{W_t}{p_t \, x \, a_t}$$

Vt = 0.0215 m/s

Karena ΔPT < 10 psi, maka alat tersebut layak.

L3.1.12 Perhitungan Heater H-01

Fungsi : Memanaskan bahan baku etanol 42% (fase gas) dari suhu 126˚C menjadi 470˚C sebelum masuk reactor

Bentuk : Horizontal heater

Bahan : *Stainless steel* SA-167 *grade* 11 Tipe 316

Tekanan : 2.6 atm

- Fluida dingin (Campuran) ;

  t1 : 126 °C = 258 °F

  t2 : 470 °C = 878 °F

- Fluida Panas (*Superheated Steam*) ;

  T1 : 480 °C = 896 °F

  T2 : 470 °C = 878 °F

Laju alir massa (Ws) : 42496.1158 Kg/Jam

: 93687.7869 lb/jam

Kebutuhan panas yang diserap (Q) : 30655335.9108 KJ/Jam

: 29073520.5778 Btu/jam

Kebutuhan steam (Wt) : 8865.5864 Kg

: 19545.2490 lb

Menetukan True Temperatur Difference (Δtrue):

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

LMTD = 169.9238 °F

Luas permukaan transfer panas (A)

Dari Tabel 8 Kern, untuk :

steam-light organics adalah 100-200 BTU/jam.ft².F

Maka:

Diambil Ud = 150 Btu/jam.ft².ºF)

Untuk : Fluida panas : steam (tube side)

Fluida dingin : etanol-air (shell side)

$$A = \frac{Q_C}{U_D.LTMDkoreksi}$$

$= 1226.5047\ ft^2$ *<200 ft² Shell and tube heat exchanger*

Menentukan Tc average dan tc average

Tc average = $(T_1+T_2)/2$ = 887 ºF

tc average = $(t_1+t_2)/2$ = 568.4ºF

Dari Hal 843, Kern , diperoleh Informasi :

Panjang pipa, L = 6 ft

Pemilihan route fluida :

Shell side = etanol-air (cold fluid)

Tube side = steam (hot fluid)

aliran = counter current

Dipilih penukar panas shell and tube, L = 6 ft dengan spesifikasi :

Berdasarkan Pers. 14.34 Brownell & Young:

ODt =	1.0000	in
f = allowable stress =	18750.0000	psi
E = efisiensi penyambungan =	0.8000	
c = faktor korosi =	0.1250	in/10 tahun
P = tekanan desain (faktor keamanan 20%) =	45.2519	psi

$$tm = \left[\frac{Pi.ODt}{2.(FE - 0.6P)}\right] + C$$

tm = 0.1265 In

(tebal minimum tube)

Menentukan Jumlah Tube (Nt) :

Dari Kern tabel 10 hal 843, dipilih:

ODt =	1.0000	in
Panjang tube =	6.0000	ft
t =	0.1265	in
BWG =	8.0000	
Idt	0.6700	in
a't	0.3550	in^2
Surface per lin (ao) =	0.2618	ft^2/ln ft

Jumlah tube, Nt $= \dfrac{A}{a_o.L}$

$= \dfrac{1226.5047\ ft^2}{0.2618 \text{ X } 6\ ft}$

Nt = 780.8153 buah

Dari Kern tabel 9, hal 841, dipilih (square pitch):

Nt =	790.0000	buah
N =	8.0000	passes
Pt =	1.2500	in, square pitch
IDs =	41.0400	

Menentukan Ud terkoreksi

A terkoreksi = Nt x ao x L

A terkoreksi = 1240.9320 ft^2

Ud terkoreksi $= \dfrac{Q}{\text{A terkoreksi x LMTD}}$

$= \dfrac{29073520.5778 \text{ Btu/jam}}{1240.9320\ ft^2 \text{ x } 169.9238\ ^oF}$

Ud terkoreksi = 137.8781 Btu/jam.ft².°F)

Menentukan faktor kekotoran (dirty factor)

Shell Side (etanol dan air)

Data:

tc avg =	568.4000	F
viscositas, μ =	0.10519	Cp
	0.2545	lb/ft.hr
,Cp =	126.8524	Btu/lb.°F
Konduktivitas panas, k =	0.3828	Btu/ft².jam(°F/ft)

Menentukan flow area, As:

IDs (in diameter) =	41.0400	In
B (bafflespace) = 1/4*IDs=	10.2600	In
Pt (harga pitch) =	1.2500	In
c (harga pitch - OD) =	0.2500	In

$$as = \frac{IDs \times C' \times B}{Pt \times 144}$$

as = $\frac{41.0400 \times 0.2500 \times 10.2600}{1.2500 \times 144}$

as = 0.5848 ft²

Tube Side (steam)

Data:

Tc avg =	887.0000	F
viscositas, μ =	0.0286	cp
	0.0693	lb/ft.hr
Cp =	185.5856	Btu/lb.°F
Konduktivitas panas, k =	6.3923	Btu/ft².jam(°F/ft)

Menentukan flow area, at:

n (jumlah passes) =	8.0000	
Nt (jumlah tube) =	790.0000	
A't (flow area per tube) =	0.3550	in²

$$at = \frac{Nt \times a't}{n \times 144}$$

at = $\frac{790.0000 \times 0.3550}{8.0000 \times 144}$

at = 0.2434 ft²

Menentukan kecepatan massa, Gs:

Gs = Ws/As = $\frac{93687.7869 \text{ lb/jam}}{0.5848 \text{ ft}^2}$

= 160199.3551 lb/jam.ft²

Menentukan kecepatan massa, G_t :

Gt = Wt/At

= $\frac{19545.2490 \text{ lb/jam}}{0.2434 \text{ ft}^2}$

= 80285.7083 lb/jam.ft²

Menentukan bilangan Reynolds, Re_s :

Untuk Odt 1, De =	0.9900	in
fig 28 Kern	0.1650	ft

Re_s = De x Gs/μ

$Re_s = \frac{0.1650 \times 160199.3551}{0.2545}$

Re_s = 103841.9489

dari Kern Fig 28, diperoleh harga J_H = 280.0000

Menentukan bilangan Reynolds, Re_s :

IDt =	0.6700	in
Table 10, Kern	0.1117	ft

Re_t = Dt x Gt/μ =

= $\frac{0.1117 \times 80285.7083}{0.0693}$

Re_t = 129408.3349

L/D = 53.7313

dari Kern Fig 24, diperoleh harga J_H = 480.0000

Menentukan koefisien perpindahan pipa lapisan luar, ho:

$$ho = \frac{jH \times K}{De}\left(\frac{C \times \mu}{K}\right)^{\frac{1}{3}}(\phi_s)$$

ho/Φs = 2848.8236 Btu/ft².jam(°F/ft)

$$Tw = tc + \frac{(hio/\phi t)}{(hio/\phi t) + (ho/\phi s)}(Tc - tc)$$

Tw = 592.5820 °F

Menentukan koefisien perpindahan pipa lapisan dalam, hi:

$$hi = \frac{jH \times k}{D}\left(\frac{C \times \mu}{k}\right)^{\frac{1}{3}}(\phi_t)$$

hi/Φt = 34684.6179 Btu/ft².jam(°F/ft)

Tw =	592.5820	F
μw =	0.0286	lb/ft.hr

$\Phi t = (\mu/\mu w)^{0.14}$

$\Phi t = (0.0693/0.0286)^{0.14}$

Dari *Chemical properties*, μw = 0.07067 Cp

0.1710 lb/ft.hr

$\Phi s = (\mu/\mu w)^{0.14} =$ 1.0573

ho = 3011.9508 Btu/ft².jam(°F/ft)

Menentukan Clean Overall Coefficient, Uc:

$$Uc = \frac{hio \times ho}{hio + ho}$$

$$Uc = \frac{26299.4017 \times 3011.9508}{26299.4017 + 3011.9508}$$

= 2702.4513 Btu/ft2.jam(°F/ft)

= 1.1317

hi = 39252.8384 Btu/ft².jam(°F/ft)

$$hio = hi\left(\frac{IDt}{ODt}\right)$$

$$hio = 39252.8384 \times \frac{(0.1117)}{(1.0000/6)}$$

hio = 26299.4017 Btu/ft².jam(°F/ft)

Menentukan faktor kekotoran, Rd:

$$Rd = \frac{Uc - Udact}{Uc \times Udact}$$

$$Rd = \frac{2702.4513 - 137.8781}{2702.4513 \times 137.8781} = \mathbf{0.007}$$

<u>Rd >= 0.002, jadi syarat terpenuhi</u>

Rd air =0.001 Rd etanol = 0.001

Menentukan Perubahan Tekanan

Shell Side (etanol dan air)

Re_s = 103841.9489

fs (fig 29, Kern) = 0.0280 ft²/in²

$$\Delta Ps = \left[8 j_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{L_b}\right)\left(\frac{u}{u_w}\right) - 0{,}14\right]\frac{pv\frac{2}{s}}{2}$$

ΔPs = 0.0022 psi

$$v_s = \frac{W_s}{p_{s\,x}\,a_s}$$

Vs = 0.0109 m/s

Karena ΔPs < 10 psi, maka alat tersebut layak.

Tube Side (steam)

Re_t = 129408.3349

ft (fig 26, Kern) = 0.00007 ft²/in²

$$\Delta P_t = Np\left[8 j_f \left(\frac{L}{d_i}\right)\left(\frac{u}{u_w}\right)^{-m} + 2{,}5\right]\frac{pv\frac{2}{t}}{2}$$

ΔPt = 0.4220 psi

$$Vt = \frac{W_t}{p_{t\,x}\,a_t}$$

Vt = 0.0332 m/s

Karena ΔPT < 10 psi, maka alat tersebut layak.

L3.1.13 Perhitungan Condenser Parsial CD-01

Fungsi : Mendinginkan keluaran dari reaktor fix bed multitube sebelum masuknya ke flash drum dimana ada dua fasa yaitu fasa cair (Etanol,air) dan gas (etilen) dengan suhu 80°C

Bentuk : Horizontal Condensor Parsial

Bahan : *Stainless steel* SA-167 *grade* 11 Tipe 316

Tekanan : 6 atm

- Fluida dingin (Campuran) ;

  t1 : 70 °C = 158 °F

  t2 : 80 °C = 176 °F

- Fluida Panas (*Superheated Steam*) ;

  T1 : 470 °C = 878 °F

  T2 : 80 °C = 176 °F

Laju alir massa (Ws) : 42496.1158 Kg/Jam

: 93687.7869 lb/jam

Kebutuhan panas yang diserap (Q) : 111770507.6950 KJ/Jam

: 106003149.4979 Btu/jam

Kebutuhan steam (Wt) : 212971.8203 Kg

: 469521.9344 lb

Menetukan True Temperatur Difference (Δtrue):

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

LMTD = 186.7036 °F

Luas permukaan transfer panas (A)

Dari Tabel 8 Kern, untuk :

steam-light organics – water adalah 75-150 BTU/jam.ft².F

Maka:

Diambil Ud = 100 Btu/jam.ft².°F)

Untuk : Fluida panas : etilen (tube side)

Fluida dingin : etanol-air (shell side)

$$A = \frac{Q_C}{U_D.LTMDkoreksi}$$

$= 5677.6180 \text{ ft}^2$ *<200 ft² Shell and tube heat exchanger*

Menentukan Tc average dan tc average

Tc average = $(T_1+T_2)/2$ = 527 °F

tc average = $(t_1+t_2)/2$ = 167 ºF

Dari Hal 843, Kern , diperoleh Informasi :

Panjang pipa, L = 24 ft

Pemilihan route fluida :

Shell side = etanol-air (cold fluid)

Tube side = etilen (hot fluid)

aliran = counter current

Dipilih penukar panas shell and tube, L = 12 ft dengan spesifikasi :

Berdasarkan Pers. 14.34 Brownell & Young:

ODt =	1.5000	in
f = allowable stress =	18750.0000	psi
E = efisiensi penyambungan =	0.8000	
c = faktor korosi =	0.1250	in/10 tahun
P = tekanan desain (faktor keamanan 20%) =	104.4274	psi

$$tm = \left[\frac{Pi.ODt}{2.(FE-0.6P)}\right] + C$$

tm = 0.1276 In

(tebal minimum tube)

Menentukan Jumlah Tube (Nt) :

Dari Kern tabel 10 hal 843, dipilih:

ODt =	0.7500	in
Panjang tube =	24.0000	ft
t =	0.1276	in
BWG =	12.0000	
Idt	0.5335	in
a't	0.5470	in^2
Surface per lin (ao) =	0.6455	ft^2/ln ft

Jumlah tube, Nt $= \dfrac{A}{a_o.L}$

$$= \frac{5677.6180\ ft^2}{0.6455\ X\ 24\ ft}$$

Nt = 366.4871 buah

Dari Kern tabel 9, hal 841, dipilih (square pitch):

Nt = 400.0000 buah
N = 2.0000 passes
Pt = 1.2500 in, triangular pitch
IDs = 22.6600

Menentukan Ud terkoreksi

A terkoreksi = Nt x ao x L

A terkoreksi = 6196.8000 ft^2

$$\text{Ud terkoreksi} = \frac{Q}{\text{A terkoreksi x LMTD}}$$

$$= \frac{106003149.4979 \text{ Btu/jam}}{6196.8000 \text{ ft}^2 \text{ x } 186.7036\ ^\circ\text{F}}$$

Ud terkoreksi = 91.6218 Btu/jam.ft².ºF)

Menentukan faktor kekotoran (dirty factor)	
Shell Side (etanol dan air)	**Tube Side (etilen)**
Data: tc avg = 167.0000 F viscositas, μ = 0.8453 Cp 2.0457 lb/ft.hr ,Cp = 0.4417 Btu/lb.ºF Konduktivitas panas, k = 0.0438 Btu/ft².jam(ºF/ft) **Menentukan flow area, As:** IDs (in diameter) = 22.6600 In B (bafflespace) = 1/4*IDs= 5.6650 In Pt (harga pitch) = 1.2500 In c (harga pitch - OD) = 0.5000 In $as = \frac{IDs \times C' \times B}{Pt \times 144}$ $as = \frac{22.6600 \times 0.5000 \times 5.6650}{1.2500 \times 144}$ as = 0.3566 ft²	**Data:** Tc avg = 527.0000 F viscositas, μ = 0.0131 cp 0.0317 lb/ft.hr Cp = 20.4118 Btu/lb.ºF Konduktivitas panas, k = 2.0338 Btu/ft².jam(ºF/ft) **Menentukan flow area, at:** n (jumlah passes) = 2.0000 Nt (jumlah tube) = 400.0000 A't (flow area per tube) = 0.5470 in² $at = \frac{Nt \times a't}{n \times 144}$ $at = \frac{400.0000 \times 0.5470}{2.0000 \times 144}$ at = 0.7597 ft²
Menentukan kecepatan massa, Gs:	**Menentukan kecepatan massa, G_t :**
$Gs = Ws/As = \frac{93687.7869 \text{ lb/jam}}{0.3566 \text{ ft}^2}$ = 262739.6766 lb/jam.ft² **Menentukan bilangan Reynolds, Re_s :** Untuk Odt 1, De = 0.7300 in fig 28 Kern 0.0304 ft Re_s = De x Gs/μ $Re_s = \underline{0.0304 \times 262739.6766}$	Gt = Wt/At $= \frac{469521.9344 \text{lb/jam}}{0.7597 \text{ ft}^2}$ = 618017.9027 lb/jam.ft² **Menentukan bilangan Reynolds, Re_s :** IDt = 0.5335 in Table 10, Kern 0.0222 ft Re_t = Dt x Gt/μ = $= \underline{0.0222 \times 618017.9027}$

2.0457
Re_s = 3906.5361

dari Kern Fig 28, diperoleh harga J_H = 35.0000

Menentukan koefisien perpindahan pipa lapisan luar, ho:

$$ho = \frac{jH \times K}{De}\left(\frac{C \times \mu}{K}\right)^{\frac{1}{3}}(\phi_s)$$

ho/Φs = 138.2938 Btu/ft².jam(°F/ft)

$$Tw = tc + \frac{(hio/\phi t)}{(hio/\phi t) + (ho/\phi s)}(Tc - tc)$$

Tw = 172.2383°F

Dari *Chemical properties*, μw = 0.37566 Cp
0.9091 lb/ft.hr

$\Phi s = (\mu/\mu w)^{0.14}$ = 1.1202

ho = 154.9227 Btu/ft².jam(°F/ft)

Menentukan Clean Overall Coefficient, Uc:

$$Uc = \frac{hio \times ho}{hio + ho}$$

$$Uc = \frac{6661.8597 \times 154.9227}{6661.8597 + 154.9227}$$

= 151.4018 Btu/ft2.jam(°F/ft)

0.0317
Re_t = 433846.4872
L/D = 1079.7337
dari Kern Fig 24, diperoleh harga J_H = 150.0000

Menentukan koefisien perpindahan pipa lapisan dalam, hi:

$$hi = \frac{jH \times k}{D}\left(\frac{C \times \mu}{k}\right)^{\frac{1}{3}}(\phi_t)$$

hi/Φt = 9365.9306 Btu/ft².jam(°F/ft)

Tw = 172.2383 F
μw = 0.0317 lb/ft.hr

$\Phi t = (\mu/\mu w)^{0.14}$
$\Phi t = (0.0317/0.0317)^{0.14}$

= 1.0000

hi = 9365.9306 Btu/ft².jam(°F/ft)

$$hio = hi\left(\frac{IDt}{ODt}\right)$$

$$hio = 9365.9306 \times \frac{(0.0222)}{(0.7500/24)}$$

hio = 6661.8597 Btu/ft².jam(°F/ft)

Menentukan faktor kekotoran, Rd:

$$Rd = \frac{Uc - Udact}{Uc \times Udact}$$

$$Rd = \frac{151.4018 - 91.6218}{151.4018 \times 91.6218} = \mathbf{0.004}$$

Rd >= 0.002, jadi syarat terpenuhi
Rd air =0,001 Rd etanol = 0,001

Menentukan Perubahan Tekanan

Shell Side (etanol dan air)

Re_s = 3906.5361
fs (fig 29, Kern) = 0.0410 ft²/in²

$$\Delta Ps = [8j_f\left(\frac{D_s}{d_e}\right)\left(\frac{L}{L_b}\right)\left(\frac{u}{u_w}\right) - 0,14]\frac{pv_s^2}{2}$$

ΔPs = 5.8391 psi

$$v_s = \frac{W_s}{p_{s\,x\,a_s}}$$

Vs = 0.2708 m/s

Karena ΔPs < 10 psi, maka alat tersebut layak.

Tube Side (etilen)

Re_t = 433846.4872
ft (fig 26, Kern) = 0.0000015 ft²/in²

$$\Delta P_t = Np[8j_f\left(\frac{L}{d_i}\right)\left(\frac{u}{u_w}\right)^{-m}) + 2{,}5]\frac{pv_t^2}{2}$$

ΔPt = 0.0188 psi

$$Vt = \frac{W_t}{p_{t\,x\,a_t}}$$

Vt = 0.0796 m/s

Karena ΔPT < 10 psi, maka alat tersebut layak.

L3.1.14 Perhitungan *Cooler* (C-01)

Fungsi : Mendinginkan *temperature* keluaran gas hidrogen dari reaktor

Bentuk : Horizontal heater

Bahan : *Stainless steel* SA-167 *grade* 11 Tipe 316

Tekanan : 6 atm

- Fluida dingin (Campuran) ;
  t1 : 10 °C = 50 °F
  t2 : 25 °C = 77 °F
- Fluida Panas (*Superheated Steam*) ;
  T1 : 80 °C = 176 °F
  T2 : 30 °C = 86 °F

Laju alir massa (Ws) : 10732.3232 Kg/Jam
: 23660.6944lb/jam

Kebutuhan panas yang diserap (Q) : 886228.5915 KJ/Jam
: 840499.1962 Btu/jam

Kebutuhan air pendingin (Wt) : 1921.8838 Kg
: 4237.0236 lb

1. Menetukan True Temperatur Difference (Δtrue):

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

LMTD = 62.2775 °F

2. Koefisien Transfer Panas kotor (UD)

Berdasarkan *Kern (1950) Hal 840. Tabel 8*, Harga koefisien transfer panas kotor (UD) untuk sistem :

light organics (LNO) – Water adalah 75-150 Btu/Jam.ft².°F

Maka diamnil nilai UD = 100 Btu/Jam.ft².°F

3. Menentukan Luas Permukaan Transfer Panas (A)

$$A = \frac{Q_C}{U_D.LMTD}$$

A = 134.9603 ft²

Karena nilai A<200 ft² maka dipilih double pipe exchanger

4. Menentukan Temperature Rata-Rata

$$Tav = \frac{T_1 + T_2}{2}$$

$$tav = \frac{t_1 + t_2}{2}$$

Didapat nilai Tavg dan tavg:

$T_{av} = \frac{80 + 30}{2} = 55°C$

$t_{av} = \frac{10 + 25}{2} = 17.5°C$

5. Menentukan Alokasi Fluida

Pemilihan aliran fluida

Annulus : Etilen

Inner Pipe : Air pendingin

dimana kita periksa tabel fouling factor fluida, disini didapatkan etanol dan air maka dipilih :

1. Natural gas-gasoline processing streams (Natural gas) : 0.001

2. Liquid (Refrigerant liquids) : 0.001

Maka, Ro yang digunakan :

Ro,etilen : (0.001 jam.ft^2.°F/Btu)

Ro,air : (0.001 jam.ft^2.°F/Btu)

Ro,total : (0.001 jam.ft^2.°F/Btu)

6. Menentukan Dimensi Double Pipe

Kern (1950), Appendix Tabel 12 Hal 846 :

Aliran counter current

Dipilih penukar panas double pipe, L = 20 ft

dengan spesifikasi :

External surface m2 per m length (annulus) : 0.5236 ft^2/ft

External surface m2 per m length (inner pipe) : 0.3272 ft^2/ft

Pipa (annulus)

ODa : 0.1640 ft

IDa : 0.1462 ft

Pipa (innerpipe)

ODp : 0.1025 ft

IDp : 0.0846 ft

7. Menentukan koefisien perpindahan panas

Inner Pipe : Fluida Panas (*Etilen*)

Data :

tc avg = 131 °F

Annulus : Fluida Dingin (Air Pendingin)

Data :

tc avg = 63.5 °F

Inner Pipe : Fluida Panas (*Etilen*)

μ = 0.0298 lb/ft.jam

Cp = 20.4118 Btu/lb. °F

k = 2.0338 Btu/jam.ft^2. °F

Menentukan *flow area,* a_{inner}:

π = 3.14

D = IDinner = 0.0846 ft

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

a_{inner} = 0.0056 ft²

$G_{inner} = W/a_{inner}$ = 4206765.6198 lb/jam.ft²

Re = D x G_p/μ

Re = 11953490.2615

L/D = 20 ft/ 0.0846 ft = 236.2791

dari *Kern* Fig. 24 Hal. 834, diperoleh harga:

$JH_{(grafik)}$ = 0.001

$JH_{inner} = JH_i \times Re_i$

= 11953.49026

Menentukan koefisien transfer panas (hi)

$$hi = j_H \frac{k}{D}\left(\frac{cp.\mu}{k}\right)^{1/3} \phi_p$$

hi = 126397.3247 Btu/jam.ft2.˚F

Annulus : Fluida Dingin (Air Pendingin)

μ = 1.2258 lb/ft.jam

Cp = 0.4392 Btu/lb. °F

k = 0.0404 Btu/jam.ft^2. °F

Menentukan *flow area,* a_a:

π = 3.14

D_2 = 0.1462 ft

D_1 = 0.1025 ft

$$a_{inner} = \frac{\pi}{4}(D_2{}^2 - D_1{}^2)$$

α_a = 0.0645 ft²

Menentukan *diameter equivalent*, De:

$$De = \frac{(D_2^2 - D_1^2)}{D_1}$$

De = 0.1058 ft

$G_a = W/a_a$ = 65738.7832 lb/jam.m²

Re_a = De x G_a/μ

Re_a = 5676.2092

L/D = 20 ft/ 0.1462 ft = 136.8350

dari *Kern* Fig. 24, Hal. 834 diperoleh harga:

$JH_{(grafik)}$ = 0.035

$JH_{annulus} = JH_a \times Re_a$

=198.6673

Menentukan koefisien transfer panas (ho)

Diasumsikan, Φa = 1

$$ho = j_H \frac{k}{De}\left(\frac{cp.\mu}{k}\right)^{1/3} \phi_a$$

Inner Pipe : Fluida Panas (*Etilen*)

Annulus : Fluida Dingin (Air Pendingin)

ho = 97.0134 Btu/jam.ft2.˚F

8. Menentukan koefisien perpindahan panas dalam-luar (hio)

$$hio = hi\,\frac{IDp}{ODp}$$

hio = 104353.6313 Btu/jam.ft2.˚F

9. Menentukan *Clean Overall Coefficient* (Uc)

$$Uc = \frac{hio \times ho}{hi + ho}$$

Uc = 80.0328 Btu/jam.ft2.˚F

10. Menentukan *Design Overall Coeffiecient* (U_D)

$$\frac{1}{U_D} = \frac{1}{U_C} + Rd$$

$$U_D = \frac{U_C}{(Rd \times U_C) + 1}$$

U_D = 68.9899 Btu/jam.ft2.˚F

11. Menentukan Luas Permukaan Bidang Tranfer Panas

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

Diketahui :

Q = 840499.1962 Btu/jam

U_D = 68.9899 Btu/jam.ft^2.˚F

LMTD = 62.2775 °F

$$A = \frac{840499.1962 \text{ Btu/jam}}{(68.9899 \frac{\text{Btu}}{\text{Jam. ft2}}.°F \times 62.2775\ °F)}$$

A = 195.6233 ft^2

12. Menentukan Panjang Total

P_{total} = A / eksternal *surface area inner pipe*

Diketahui ;

A = 195.6233 ft^2

Eks. A *Inner Pipe* = 0.3272 ft^2/ft

Ptotal = 195.6233 ft^2/ 0.3272 ft^2/ft

= 597.8706 ft

= 182.2310 m

Jumlah Hairpin = Ptotal / (2 x P_{pipa})

= 182.2310 m / (2 x 6.0960 m)

= 14.9468 = 15 set

13. Revisi U_D dan R_D

A_{aktual} = 2 x set hairpin x P_{pipa} x *Eksternal surface area*

A_{aktual} = 2 x 15 set x 20 ft x 0.3272 ft^2/ft

A_{aktual} = 196.3200 ft^2

$$U_D = \frac{Q}{A_{aktual} \times LMTD}$$

Diketahui :

Q = 840499.1962 Btu/jam

LMTD = 62.2775 °F

Uc = 80.0328 Btu/jam.ft2.˚F

Sehingga :

U_D = 68.7450 Btu/jam.ft2. °F

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

$$Rd = \frac{(80.0328 \text{ Btu/jam.ft2.°F} - 68.7450 \text{ Btu/jam.ft2.°F})}{(80.0328 \text{ Btu/jam.ft2.°F} \times 68.7450 \text{ Btu/jam.ft2.°F})}$$

Rd = 0.0021 Btu/jam.ft2.°F

Maka, spesifikasi memenuhi syarat, dengan Rd aktual > Rd_{min}/ teoritis.

14. Menentukan *Pressure Drop*

Inner Pipe : Fluida Panas (*Etilen*)	*Annulus :* Fluida Dingin (Air Pendingin)
Data: Re = 11953490.2615 $f = 0.0035 + \frac{0.264}{(\text{Re}_p)^{0.42}}$ f = 0.0038	D_2 = 0.1462 ft D_1 = 0.1025 ft $De' = (D_2 - D_1)$ De' = 0.0436 ft Ga = 65738.7832 lb/jam.ft²

Inner Pipe : Fluida Panas (*Etilen*)	*Annulus* : Fluida Dingin (Air Pendingin)
G_{inner} = 4206765.6198 lb/jam.ft²	μ = 1.2258 lb/ft.jam
L_{inner} = 20 ft	$\boxed{\mathrm{Re}'_a = \frac{De' \times G_a}{\mu}}$
g = 32.1740 ft/s²	
D = 0.0846 ft	$\mathrm{Re}'_a = 2340.1258$
$\Delta Finner = \frac{4.f.G_p^2.L_p}{2.g.\rho^2.D}$	$\boxed{f = 0.0035 + \frac{0.264}{(\mathrm{Re}'_a)^{0.42}}}$
$\Delta F_{iinner} = 0.00000000000698$ ft	f = 0.0137
$\boxed{\Delta P_{inner} = \frac{\Delta F_{inner} \times \rho \times g}{}}$	L_a = 20 ft
$\Delta P_{inner} = 0.00001957$ psi	g = 32.1740 ft/s²
Karena ΔP_{inner} < 10 psi, maka alat tersebut layak.	$\boxed{\Delta F_a = \frac{4.f.G_a^2.L_a}{2.g.s^2.De'}}$
Memenuhi Syarat	$\Delta F_a = 0.4996$ ft
	$F_t = 3\left(\frac{\frac{Ga^2}{\rho}}{2g}\right)$
	Δ Ft = 0.0040 ft
	$\Delta P_a = \frac{(\Delta F_a - \Delta F_l) \times \rho \times g}{}$
	$\Delta P_a = 0.2182$ psi

Inner Pipe : Fluida Panas (*Etilen*)	*Annulus :* Fluida Dingin (Air Pendingin)
	Karena ΔP_a < 10 psi, maka alat tersebut layak. Memenuhi Syarat.

L3.1.14 Perhitungan Cooler C-02

Fungsi : Mendinginkan produk bawah (etanol-air) dari flash drum sebelum masuknya ke tangki limbah

Bentuk : Horizontal Condensor Parsial

Bahan : *Stainless steel* SA-167 *grade* 11 Tipe 316

Tekanan : 6 atm

- Fluida dingin (Campuran) ;
  t1 : 10 °C = 50 °F
  t2 : 25 °C = 77 °F
- Fluida Panas (*Superheated Steam*) ;
  T1 : 80 °C = 176 °F
  T2 : 30 °C = 86 °F

Laju alir massa (Ws) : 31763.7926 Kg/Jam
: 70027.0924 lb/jam

Kebutuhan panas yang diserap (Q) : 10553218.6862 KJ/Jam
: 10008672.6019 Btu/jam

Kebutuhan air pendingin (Wt) : 22885.8115 Kg
: 50454.5178 lb

Menetukan True Temperatur Difference (Δtrue):

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

LMTD = 62.2775 °F

Luas permukaan transfer panas (A)

Dari Tabel 8 Kern, untuk :

steam-light organics – water adalah 75-150 BTU/jam.ft².F

Maka:

Diambil Ud = 75 Btu/jam.ft².°F)

Untuk : Fluida panas : etanol (tube side)

Fluida dingin : air pendingin (shell side)

$$A = \frac{Q_C}{U_D.LTMDkoreksi}$$

$= 2209.0836\ ft^2$ *<200 ft² Shell and tube heat exchanger*

Menentukan Tc average dan tc average

Tc average $= (T_1+T_2)/2$ = 131 °F

tc average $= (t_1+t_2)/2$ = 63.5 °F

Dari Hal 843, Kern , diperoleh Informasi :

Panjang pipa, L = 6 ft

Pemilihan route fluida :

Shell side = air pendingin (cold fluid)

Tube side = etanol (hot fluid)

aliran = counter current

Dipilih penukar panas shell and tube, L = 6 ft dengan spesifikasi :

Berdasarkan Pers. 14.34 Brownell & Young:

ODt =	1.5000	in
f = allowable stress =	18750.0000	psi
E = efisiensi penyambungan =	0.8000	
c = faktor korosi =	0.1250	in/10 tahun
P = tekanan desain (faktor keamanan 20%) =	87.0228	psi

$$tm = \left[\frac{Pi.ODt}{2.(FE - 0.6P)}\right] + C$$

tm = 0.1294 In

(tebal minimum tube)

Menentukan Jumlah Tube (Nt) :

Dari Kern tabel 10 hal 843, dipilih:

ODt =	1.5000	in
Panjang tube =	6.0000	ft
t =	0.1294	in
BWG =	18.0000	
Idt	1.4000	in
a't	1.5400	in^2
Surface per lin (ao) =	0.3925	ft^2/ln ft

$$\text{Jumlah tube, Nt} = \frac{A}{a_o.L}$$

$$= \frac{2209.0836\ ft^2}{0.3925\ X\ 6\ ft}$$

Nt = 938.0397 buah

Dari Kern tabel 9, hal 841, dipilih (square pitch):

Nt =	950.0000	buah
N =	8.0000	passes
Pt =	1.8750	in, triangular pitch
IDs =	46.1100	

Menentukan Ud terkoreksi

A terkoreksi = Nt x ao x L

A terkoreksi = 2237.2500 ft^2

$$\text{Ud terkoreksi} = \frac{Q}{\text{A terkoreksi x LMTD}}$$

$$= \frac{10008672.6019\ Btu/jam}{2237.2500\ ft^2 \times 62.2775\ ^oF}$$

Ud terkoreksi = 71.8341 Btu/jam.ft².ºF)

Menentukan faktor kekotoran (dirty factor)						
Shell Side (air pendingin)				**Tube Side (etanol)**		
Data:				**Data:**		
tc avg =	63.5000	F		Tc avg =	131.0000	F
viscositas, µ =	0.50653	cp		viscositas, µ =	0.6344	cp
	1.2258	lb/ft.hr			1.5354	lb/ft.hr
,Cp =	0.4017	Btu/lb.ºF		Cp =	0.0770	Btu/lb.ºF
Konduktivitas panas, k =	-0.0117	Btu/ft².jam(ºF/ft)		Konduktivitas panas, k =	0.0609	Btu/ft².jam(ºF/ft)
Menentukan flow area, As:				**Menentukan flow area, at:**		
IDs (in diameter) =	46.1100	in		n (jumlah passes) =	8.0000	
B (bafflespace) = 1/4*IDs=	11.5275	in		Nt (jumlah tube) =	950.0000	

Pt (harga pitch) = 1.8750 in
c (harga pitch - OD) = 0.3750 In

$$as = \frac{IDs \times C' \times B}{Pt \times 144}$$

$$as = \frac{46.1100 \times 0.3750 \times 11.5275}{1.8750 \times 144}$$

as = 0.7382 ft^2

A't (flow area per tube) = 1.5400 in²

$$at = \frac{Nt \times a't}{n \times 144}$$

$$at = \frac{950.0000 \times 1.5400}{8.0000 \times 144}$$

at = 1.2700 ft^2

Menentukan kecepatan massa, Gs:

$$Gs = Ws/As = \frac{70027.0924 \text{ lb/jam}}{0.7382 \text{ ft}^2}$$

= 94856.7712 lb/jam.ft²

Menentukan kecepatan massa, G_t :

Gt = Wt/At

$$= \frac{50454.5178 \text{ lb/jam}}{1.2700 \text{ ft}^2}$$

= 39729.0529 lb/jam.ft²

Menentukan bilangan Reynolds, Re_s :

Untuk Odt 1, De = 1.0800 in
fig 28 Kern 0.1800 ft

Re_s = De x Gs/μ

$$Re_s = \frac{0.1800 \times 94856.7712}{1.2258}$$

Re_s = 13929.0565

dari Kern Fig 28, diperoleh harga J_H = 68.0000

Menentukan bilangan Reynolds, Re_s :

IDt = 1.4000 in
Table 10, Kern 0.2333 ft

Re_t = Dt x Gt/μ =

$$= \frac{0.2333 \times 39729.0529}{1.5354}$$

Re_t = 6037.7147
L/D = 25.7143
dari Kern Fig 24, diperoleh harga J_H = 25.0000

Menentukan koefisien perpindahan pipa lapisan luar, ho:

$$ho = \frac{jH \times K}{De}\left(\frac{C \times \mu}{K}\right)^{\frac{1}{3}}(\phi_s)$$

ho/Φs = 15.3361 Btu/ft².jam(ºF/ft)

$$Tw = tc + \frac{(hio/\phi t)}{(hio/\phi t) + (ho/\phi s)}(Tc - tc)$$

Tw = 107.5950 ºF

Dari *Chemical properties*, μw = 0.50541 cp
1.2231 lb/ft.hr

Φs = (μ/μw) $^{0.14}$= 0.8839

ho = 13.5555 Btu/ft².jam(ºF/ft)

Menentukan koefisien perpindahan pipa lapisan dalam, hi:

$$hi = \frac{jH \times k}{D}\left(\frac{C \times \mu}{k}\right)^{\frac{1}{3}}(\phi_t)$$

hi/Φt = 8.1402 Btu/ft².jam(ºF/ft)

Tw = 107.5950 F
μw = 0.0014 lb/ft.hr

Φt = (μ/μw)$^{0.14}$
Φt = (1.5354/0.0014)$^{0.14}$
= 2.6657

hi = 21.6997 Btu/ft².jam(ºF/ft)

$$hio = hi\left(\frac{IDt}{ODt}\right)$$

$$hio = 21.6997 \times \frac{(0.2333)}{(1.5000/24)}$$

hio = 20.2531 Btu/ft².jam(ºF/ft)

Menentukan Clean Overall Coefficient, Uc:

$$Uc = \frac{hio \times ho}{hio + ho}$$

$$Uc = \frac{20.2531 \times 13.5555}{20.2531 + 13.5555}$$

Menentukan faktor kekotoran, Rd:

$$Rd = \frac{Uc - Udact}{Uc \times Udact}$$

$$Rd = \frac{8.1204 - 71.8341}{8.1204 \times 71.8341} = \mathbf{-0.109}$$

= 8.1204 Btu/ft2.jam(°F/ft)	
Menentukan Perubahan Tekanan	
Shell Side (Aqueous solutions)	**Tube Side (air)**
Re_s = 13929.0565 fs (fig 29, Kern) = 0.0022 ft^2/in^2 Number of crosses = N + 1 = 6 L/B 3.1230 ft Ds = IDs/6 7.6850 ft SG pada Tc = 0.0600 lb/ft^3 $\Delta P_s = \frac{f \times Gs^2 \times Ds \times (N+1)}{5.22 \times 10^{10} \times De \times SG \times \phi_s}$ ΔPs = 0.9534 psi Karena ΔPs < 10 psi, maka alat tersebut layak.	Re_t = 6037.7147 ft (fig 26, Kern) = 0.000036 ft^2/in^2 SG = 48.5997 lb/ft3 $\Delta P_t = \frac{f \times Gt^2 \times L \times n}{5.22 \times 10^{10} \times De \times SG_{sb} \times \phi_t}$ ΔPt = 0.000077162 psi $\Delta P_r = \frac{(4xn)}{SG_{sb}} x \frac{V^2}{2g'} \times \frac{62,5}{144}$ Untuk Gt = 39729.0529 lb/jam ft2 (V2/2g')(62,5/144) = 0.0010 Fig 27, Kern ΔPr = 0.0007 psi ΔPT = ΔPt + ΔPr = 0.0007 psi Karena ΔPT < 10 psi, maka alat tersebut layak.

L3.2 Pehitungan Spesifikasi Alat Utilitas

L3.2.1 Perhitungan Pompa Proses P-01

Fungsi : Mengalirkan air dari Tangki (T-01) menuju MP-01

Jenis : Centrifugal Pump Horizontal

Kondisi Operasi

Tekanan	1	atm
Temperature	30	°c
	303	Kelvin
Laju Alir Massa	17802.1535	Kg/jam
Densitas	782.6364	Kg/m³
	48.8643	lb/ft³
Viskositas	0.9645	Cp
	2.3333	lb/ft.h

Penentuan Ukuran Pipa

● **Kapasitas Pompa, Qf**

mf	=	(1 + f) x ms	
	=	43171.2904	lb/hr
	=	719.5215	lb/min

$$Qf = \frac{m}{\rho}$$

	=	14.7249	ft³/jam
	=	0.2454	ft³/s
	=	110.1578	Gal/min

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

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

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

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

● **Penentuan Spesifikasi Pipa**

Diasumsikan aliran viscous

Di, optimum	=	3.9 x qf^0.45 x ρ^0.13	
	=	3.4363	inch

Suction Pipe

Nominal Pipe Size	=	4	inch	0.3333	ft
OD	=	4.5	inch	0.3750	ft
Schedule Number	40				
ID	=	4.0260	inch	0.3355	ft
Ls	=	3	m	9.8425	ft
A_o	=	12.7296	inch²	0.0884	ft²

Discharge Pipe

Nominal Pipe Size	=	3.5	inch	0.2917	ft
OD	=	4	inch	0.3333	ft
Schedule Number	40				
ID	=	3.5480	inch	0.2957	ft
Ls	=	3	m	9.8425	ft
A_0	=	9.8928	$inch^2$	0.0687	ft^2

Suction Friction Loss

●Kecepatan Linear Fluida, V

Kecepatan Linear V = Qf/A_0

= 2.7762 ft/s

Penentuan Faktor Friksi

●Nre = $\frac{\rho \times v \times ID}{\mu}$

= 70221.2953

Diperoleh F = 0.0031

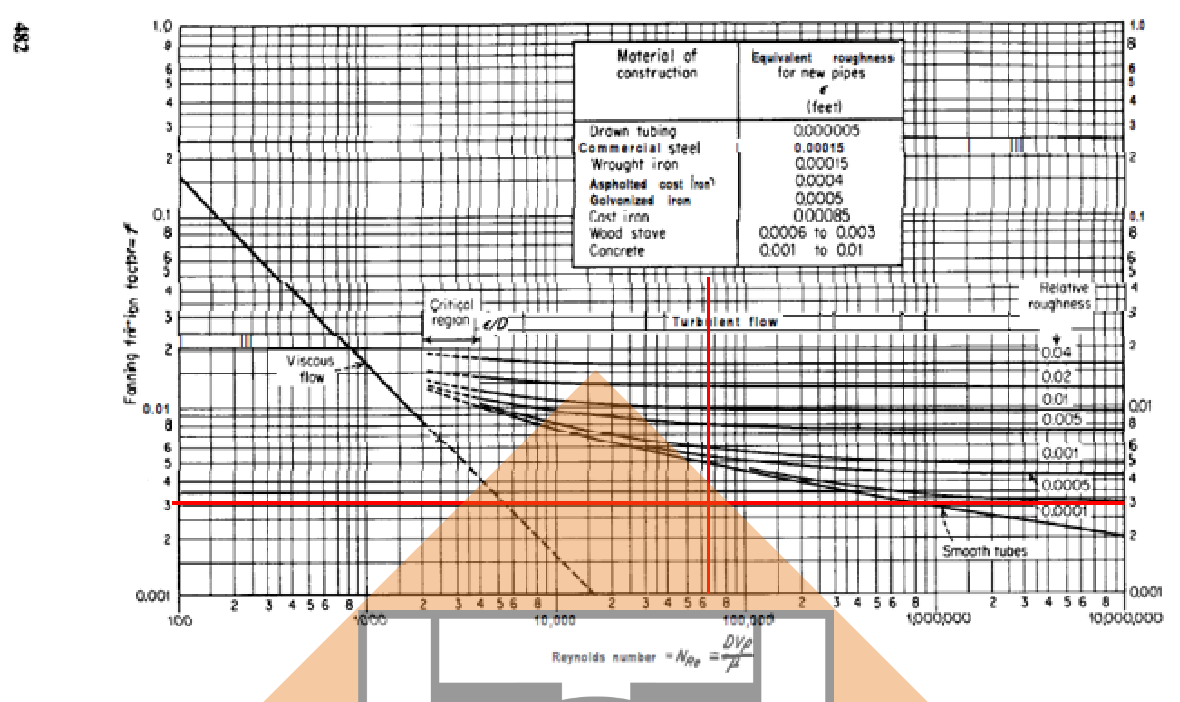


FIGURE 14-1
Fanning friction factors for long straight pipes. [Based on L. F. Moody, *Trans. ASME*, **66**:671–684 (1944).]

GAMBAR L3.7 NILAI FRICTION FACTOR SUCTION

Skin Friction Loss

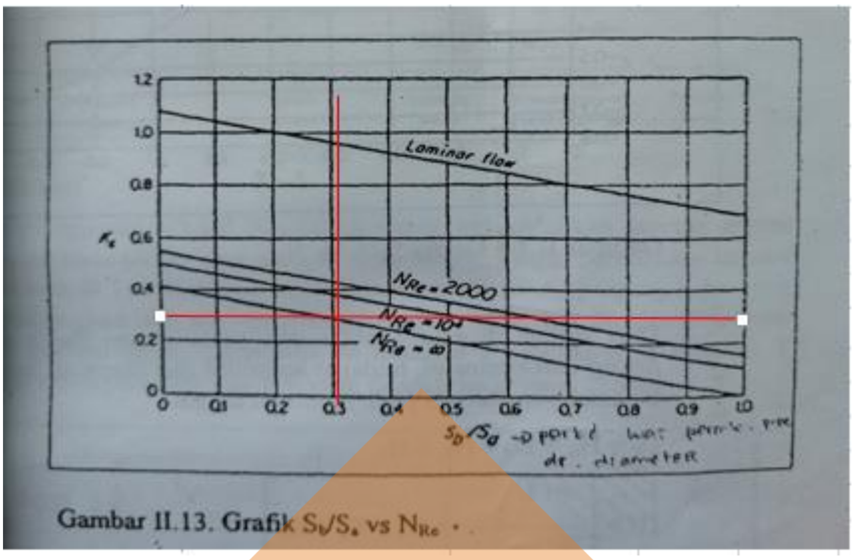
Asumsi

Standard elbow 90º	=	32	buah
Gate Valve	=	7	buah
Tee	=	90	buah
L Fitting	=	71	buah
Le	=	33.6630	ft

$$Hfs = \frac{4 \times f \times Le}{D} \times \frac{V^2}{2gc}$$

= 0.1490 ft. lbf/ lb

Kc = 0.31



Gambar II.13. Grafik S_b/S_a vs N_{Re}

GAMBAR L3.8 GRAFIK S_B/S_A VS N_{RE} SUCTION

A	=	1	
Hfc	=	$\dfrac{Kc}{2\alpha} \times \dfrac{V^2}{2gc}$	
	=	0.0186	ft. lbf/ lb
Kf	=	3	
Hff	=	$Kf \times \dfrac{V^2}{2gc}$	
	=	0.3593	ft. lbf/ lb
Hfsuc	=	Hfs + Hfc + Hff	
	=	0.5269	ft. lbf/ lb
Za	=	13	m
	=	42.6509	ft
Zb	=	0	m
	=	0	ft
Static Suction (Zs)	=	42.6509	ft
Origin Pressure (Pa)	=	1	bar
		0.9870	atm
		14.5050	psi

		2088.7131	lb/ft2		
Pressure Head (Hp)	=	$\frac{Pa}{\rho}$			
	=	42.7451	ft. lbf/ lb		
g/gc	=	1	lbf/lb		
Static Suction Head (Hs)	=	$\frac{g}{gc}$		x	(Za - Zb)
	=	42.6509	ft		
va – vb	=	0	ft/hr		
Velocity head (Hv)	=	0	ft. lbf/lb		
H_{suc}	=	$Hp + Hs + Hv - H_{suction}$			
	=	84.869	ft. lbf/lb		
Pb	=	Hsuc × ρ			
	=	4147.0756	lbf/ft2		
	=	28.7991	psi		
Hp_{uap}	=	$\frac{P_{uap}}{\rho}$			
	=	4.4692	ft.lbf/lb		
Total NPSH	=	$H_{suc} - Hp_{uap}$			
	=	80.4000	ft.lbf/lb		

Discharge Friction Loss

●Kecepatan Linear Fluida,V

Kecepatan Linear V	=	Q_f/A_o	
	=	3.5723	ft/s

Penentuan Faktor Friksi

●Nre	=	$\frac{\rho \times v \times ID}{\mu}$
	=	79629.5396
Diperoleh F	=	0.0031

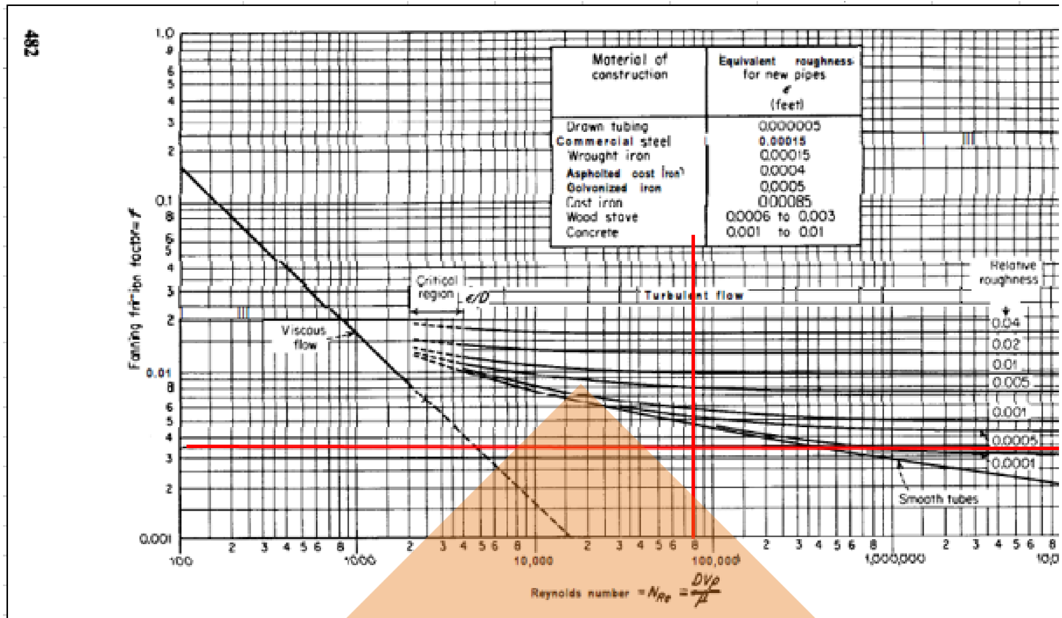


FIGURE 14-1
Fanning friction factors for long straight pipes. [Based on L. F. Moody, *Trans. ASME*, 66:671–684 (1944).]

GAMBAR L3.9 NILAI FRICTION FACTOR DISCCHARGE

Skin Friction Loss

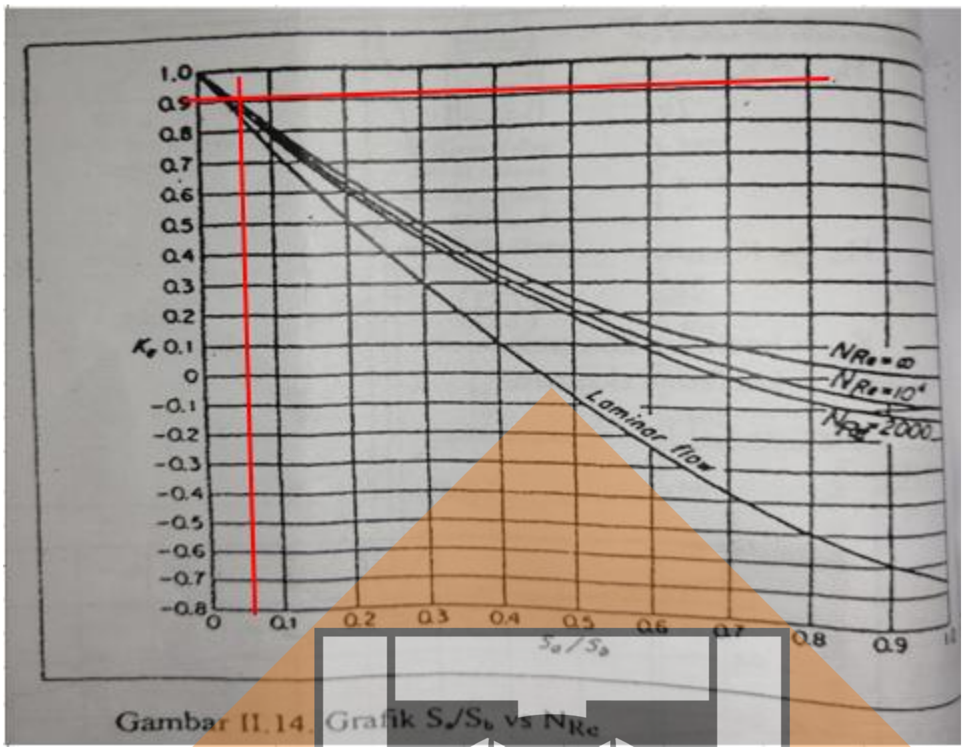
Asumsi

Standard elbow 90º	=	32	buah
Gate Valve	=	7	buah
Tee	=	90	buah
L Fitting	=	64	buah
Le	=	28.7652	ft

$$Hfs = \frac{4 \times f \times Le}{D} \times \frac{V^2}{2gc}$$

$= 0.2392$ ft. lbf/ lb

Ke = 0.9



GAMBAR L3.10 GRAFIK S_B/S_A VS N_{RE} DISCHARGE

α = 1

Hfe $= \dfrac{Ke}{2\alpha} \times \dfrac{V^2}{2gc}$

= 0.0892 ft. lbf/ lb

Kf = 1.8

Hff $= Kf \times \dfrac{V^2}{2gc}$

= 0.3570 ft. lbf/ lb

Hf dis = Hfs + Hfe + Hff

= 0.6854 ft. lbf/ lb

Zc = 0 m

= 0 ft

Zd = 5 m

= 16.4042 ft

Static discharge head (Zdh) = 16.4042 ft

Discharge Pressure (Pd) = 7 bar

= 6.9090 atm

= 101.5347 psi

= 14620.9916 lb/ft2

Pressure Head (Hp) = $\dfrac{Pd}{\rho}$

= 299.2159 ft. lbf/ lb

g/gc = 1 lbf/lb

Static Discharge Head (Hd) = $\dfrac{g}{gc}$ x (Zdh - Zc)

= 16.4042 ft. lbf/ lb

Vc - Vd = 0 ft/hr

Velocity head (Hv) = 0 ft. lbf/lb

H_{dis} = Hp + Hs + Hv - $H_{suction}$

= 314.9347 ft. lbf/lb

Pc = Hdis × ρ

= 15389.0782 lbf/ft2

= 106.8686 psi

Differential pressure = Discharge pressure + Pressure Drop + Suction pressure

= 78.0695 psi

Head untuk QV-01 = 0.000 psi

= 0.000 ft lbf/lb

Total head = Discharge head + Suction head

= 230.0655 ft lbf/lb

Menentukan Daya Pompa (BHP dan Daya Motor DHP)

• Menentukan Daya Fluida (BHP)

BHP = $\dfrac{m_f \text{ x } W_s}{\eta_p}$

Dengan

m_f = 719.5215 lb/min

W_s = 230.0655 ft. lbf/lb [3]

Efisiensi pompa (η_p)	=	68	%

Maka BHP sebagai berikut :

BHP	=	243436.9125	ft. lbf/min
	=	7.3740	HP

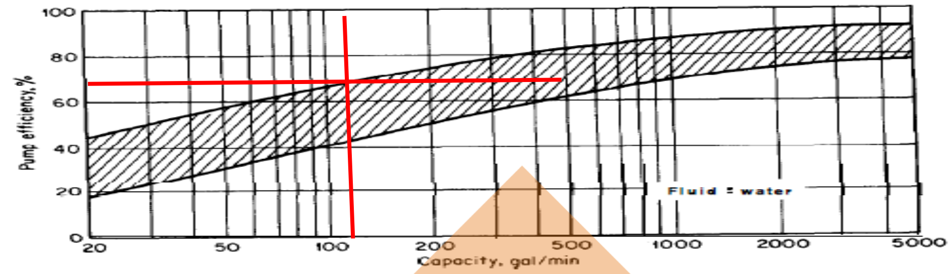


FIGURE 1437
Efficiencies of centrifugal pumps.

520

GAMBAR L3.10 GRAFIK EFISIENSI POMPA

● **Menentukan Daya Motor (MHP)**

Berdasarkan figure 14.38 Peters - Plant Design dengan nilai

BHP	=	7.3740	HP
Efisiensi motor	=	86	%

MATERIALS TRANSFER, HANDLING, AND TREATMENT EQUIPMENT 521

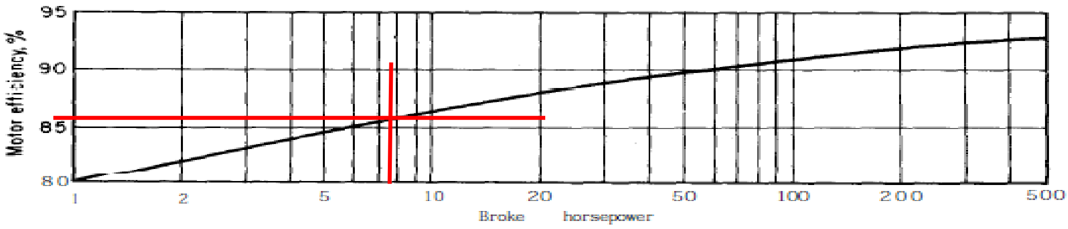


FIGURE 1438
Efficiencies of three-phase motors.

GAMBAR L3.11 GRAFIK EFISIENSI MOTOR

dengan demikian

MHP = $\frac{\text{BHP}}{\text{Efisiensi Motor}}$

= 8.5744 HP

Maka dipilih motor yang berkekuatan = 10.0575 HP

Menentukan NPSH

Untuk mengatasi kavitas

NPSH yang dibutuhkan, NPSHa > NPSHr

NPSH (Net Positive Suction Head) available :

Tabel 7.2 Walas edisi 3 hal 144

NPSH dari	=	Cetrifuge horizontal pump for Chemical			
Maksimum	=	19.7	ft	6.004	m

NPSH (Net Positive Suction Head) required :

Pers. 7.15 Walas edisi 3

$$NPSH_R = \left(\frac{nQ^{0.5}}{S}\right)^{\frac{4}{3}}$$

Keterangan

n, kecepatan putaran	=	3500	rpm
Q, debit	=	3.0000	gpm
S, kecepatan spesifik	=	7900	
single suction			

NPSHr	=	0.7025	ft	0.2141	m

Design OK

Kode Pompa	Fungsi	Jenis	Kapasitas	Efisiensi Pompa	Dimensi	Daya	NPSH A	NPSH R	Jumlah
P-02	Mengalirkan air sebagai medium dari Tangki (T-02) menuju MP-01	Centrifugal Pump Horizontal	116.9141 gpm	70%	Suction : NPS 4 inch Schedule 40 inch Panjang pipa lurus 3 m Standard Elbow 90ºC 32 unit Gate valve 7 unit Tee 90 unit Discharge : NPS 3.5 inch Schedule 40 inch Panjang pipa lurus 3 m Standard Elbow 90ºC 32 unit Gate valve 7 unit Tee 90 unit	10.0575 HP	6.004 m	2.461 m	2 unit, 1 unit standby

Kode Pompa	Fungsi	Jenis	Kapasitas	Efisiensi Pompa	Dimensi	Daya	NPSH A	NPSH R	Jumlah
P-03	Mengalirkan bahan baku untuk masuk ke dalam vaporizer sebagai fase cair	Centrifugal Pump Horizontal	227.0709 gpm	78%	Suction : NPS 6 inch Schedule 40 inch Panjang pipa lurus 3 m Standard Elbow 90ºC 32 unit Gate valve 7 unit Tee 90 unit Discharge : NPS 5 inch Schedule 40 inch Panjang pipa lurus 3 m Standard Elbow 90ºC 32 unit Gate valve 7 unit Tee 90 unit	20.1150 HP	6.004 m	3.831 m	2 unit, 1 unit standby

Kode Pompa	Fungsi	Jenis	Kapasitas	Efisiensi Pompa	Dimensi	Daya	NPSH A	NPSH R	Jumlah
P-04	Mengalirkan buangan dari Flash drum ke tangki pembuangan	Centrifugal Pump Horizontal	150.6451 gpm	78%	Suction : NPS 5 inch Schedule 40 inch Panjang pipa lurus 3 m Standard Elbow 90ºC 32 unit Gate valve 7 unit Tee 90 unit Discharge : NPS 4 inch Schedule 40 inch Panjang pipa lurus 3 m Standard Elbow 90ºC 32 unit Gate valve 7 unit Tee 90 unit	14.7510 HP	6.004 m	2.914 m	2 unit, 1 unit standby

L3.2.2 Perhitungan Pompa Utilitas

Kode Pompa	Fungsi	Jenis	Kapasitas	Efisiensi Pompa	Dimensi	Daya	Jumlah
PU-01	Memompa air dari badan sungai ke screen	Centrifugal Pump Horizontal	127.4320 m^3/jam = 561.0662 gpm	82%	NPS 8 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit Tee 0 unit	2.95 HP	1 unit, 1 unit standby
PU-02	Memompa air dari screen ke reservoir	Centrifugal Pump Horizontal	127.4320 m^3/jam = 561.0662 gpm	82%	NPS 8 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit	2.95 HP	1 unit, 1 unit standby

					Tee 0 unit		
PU-03	Memompa air dari reservoir ke bak pengadukan cepat (BPC)	Centrifugal Pump Horizontal	127.4320 m^3/jam = 561.0662 gpm	82%	NPS 8 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit Tee 0 unit	2.95 HP	1 unit, 1 unit standby
PU-04	Memompa air dari bak pengaduk cepat ke bak pengendap I	Centrifugal Pump Horizontal	127.4320 m^3/jam = 561.0662 gpm	82%	NPS 8 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit Tee 0 unit	2.95 HP	1 unit, 1 unit standby

PU-05	Memompa air dari bak pengendap I ke bak pengendap II	Centrifugal Pump Horizontal	127.4320 m^3/jam = 561.0662 gpm	82%	NPS 8 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit Tee 0 unit	2.95 HP	1 unit, 1 unit standby
PU-06	Memompa air dari bak pengendap II ke tangki filtrasi	Centrifugal Pump Horizontal	127.4320 m^3/jam = 561.0662 gpm	82%	NPS 8 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit Tee 0 unit	2.95 HP	1 unit, 1 unit standby
PU-07	Memompa air dari tangki filtrasi ke bak air bersih	Centrifugal Pump Horizontal	127.4320 m^3/jam	82%	NPS 8 inch Schedule 40 inch	2.95 HP	1 unit, 1 unit standby

			= 561.0662 gpm		Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit Tee 0 unit		
PU-08	Memompa air dari bak air bersih ke tangki demineralisasi	Centrifugal Pump Horizontal	127.4320 m^3/jam = 561.0662 gpm	82%	NPS 8 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit Tee 0 unit	2.95 HP	1 unit, 1 unit standby
PU-09	Memompa air dari bak air bersih ke bak penampung cooling tower	Centrifugal Pump Horizontal	87.2472 m^3/jam = 384.1378 gpm	80%	NPS 8 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit	2.01 HP	1 unit, 1 unit standby

					Globe valve 1 unit Gate valve 1 unit Tee 0 unit		
PU-10	Memompa air dari tangki demineralisasi ke bak umpan boiler	Centrifugal Pump Horizontal	55.8200 m^3/jam = 245.7680 gpm	82%	NPS 6 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit Tee 0 unit	2.01 HP	1 unit, 1 unit standby
PU-11	Memompa air dari bak air bersih ke bak domestik	Centrifugal Pump Horizontal	0.9066 m^3/jam = 3.9916 gpm	42%	NPS 1 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit	0.54 HP	1 unit, 1 unit standby

					Tee 0 unit		
PU-12	Memompa air dari bak air bersih ke bak limbah	Centrifugal Pump Horizontal	94.3583 m^3/jam = 415.4472 gpm	82%	NPS 8 inch Schedule 40 inch Panjang pipa lurus 20 m Standard Elbow 5 unit Globe valve 1 unit Gate valve 1 unit Tee 0 unit	2.01 HP	1 unit, 1 unit standby

LAMPIRAN 4 PERHITUNGAN UTILITAS

L.4.1 Unit Penyedia Air

Kebutuhan air parbik meliputi kebutuhan air domestik, air untuk *steam,* air untuk pendinginan dan lain-lain. Pemasok kebutuhan air berasal dari PT Krakatau Tirta. Dengan beberapa dasar perhitungan sebagai berikut.

L.4.1.1 Kebutuhan Air Pemanas (*Steam)*

Steam dihasilkan dalam sebuah *boiler fire tube* yang menggunakan bahan bakar bio solar, dengan jenis *steam* yang dihasilkan adalah *superheated steam.* Spesifikasi *steam* yang digunakan dalam kegiatan proses pabrik Ethylene ini adalah sebagai berikut (TLV, 2023):

Temperature = 480ºC = 471.30 K

Tekanan (Psteam) = 600 kPa = 15 bar

Entalpi *steam* = 3439.91 Kj/Kg = 1478.88 Btu/lb

- Kebutuhan Steam untuk peralatan pabrik adalam sebagai berikut:

Tabel L4.1 Kebutuham Steam Peralatan Pabrik Ethylene

No	Alat	Kebutuhan (Kg/h)
1	*Vaporizer* (VP-01)	26357.9076
2	*Heater* (H-01)	8865.5864
3	Reaktor (R-01)	117.1964
	Total Kebutuhan Steam	35340.6904

Dengan asumsi factor keamanan dan kehilangan panas dimasing-masing alat, setiap steam dialirkan ke unit proses di lebihkan 10% sehingga :

$$Total + FK\ 10\% = 1.1\ \times 35340.6904 = 38874.7595\,\frac{Kg}{jam}$$

Atau setara dengan 85703.29474 lb/jam.

- Boiler Untuk Mendapatkan Steam

  Boiler untuk mendapatkan steam berjenis *Boiler fire tube* dengan jumlah 1 unit. Penentuan daya *boiler* adalah sebagai berikut (Severn, 2004):

$$BHP = \frac{Ms\ \times (Hv - Hf)}{Cf\ \times (34.5)}$$

Keterangan :

Ms	*= Massa Steam*	*= 85703.29474 lb/jam*
Hv	*= Entalpi vapor pada temperature 480ºC*	*= 1478.8836 Btu/lb*
Hf	*= Entalpi steam pada suhu 30ºC*	*= 54.0602 Btu/lb*
Cf	*= Panas laten penguapan air pada 100ºC*	*= 970.0999 Btu/lb*

Konversi $= 34.5 \frac{\frac{Lb}{h}}{HP}$

$$BHP = \frac{85703.29474 \times (1478.8836 - 54.0602\ )}{970.0999 \times (34.5)} = 3648.5725\ HP$$

Heating surface untuk boiler tiap 1 HP adalah 10 ft³/HP (Severn, 2004). Maka,

$$Heating\ Surface\ Boiler = 10\ \times 3648.5725 = 36485.7254\ ft^3$$

- Kebutuhan Air Untuk Menghasilkan *Steam*

$$Kebutuhan\ Air = \frac{MS\ \times (Hv - Hf)}{Cf}$$

Keterangan :

Ms	*= Massa Steam*	*= 85703.29474 lb/jam*
Hv	*= Entalpi vapor pada temperature 480ºC*	*= 1478.8836 Btu/lb*
Hf	*= Entalpi steam pada suhu 30ºC*	*= 54.0602 Btu/lb*
Cf	*= Panas laten penguapan air pada 100ºC*	*= 970.0999 Btu/lb*

$$Kebutuhan\ Air = \frac{85703.29474\ \times (1478.8836 - 54.0602)}{970.0999}$$

$$= 125875.7528\ lb/jam$$

Atau setara dengan 57096.8669 kg/jam.

Densitas air pada 30 ºC adalah = 1022.8753 kg/m³.

Sehingga debit air (Qa) yang dibutuhkan adalah,

$$Qa = \frac{57096.8669}{1022.8753} = 55.8200\ m^3/jam$$

- Kebutuhan Air Untuk *Make Up Boiler*

  Kehilangan air dikarenakan *Blowdown* dan hilang akibat penguapan diasumsikan sebesar 10%, maka jumlah air yang harus diumpankan sebagai *make up boiler* adalah = 10% x massa air.

$$Volume\ yang\ dibutuhkan = \frac{10\% \times 57096.8669}{1022.8753} = 5.582\ m^3/jam$$

- Kebutuhan Bahan Bakar Untuk *Boiler*

  Bahan bakar yang digunakan untuk *Boiler* adalah bio solar. Dimana,

$$Bio\ solar\ yang\ dibutuhkan = \frac{m\ steam \times (hv - hf)}{\eta \times Hv}$$

Keterangan

m steam	= massa steam = 85703.2947 kg/jam	= 188941.4836 lb/jam
Hv	= Heating value bio solar	= 19200 btu/lb
hv	= Entalpi vapor pada suhu 480ºC	= 1478.8836 btu/lb
hf	= Entalpi steam pada suhu 30ºC	= 54.0602 btu/lb
η	= efisiensi pembakaran	= 85%
ρ bio solar	= densitas bio solar	= 850 kg/m³

$$Massa\ Bio\ solar = \frac{188941.4836 \times (1478.8836 - 54.0602)}{85\% \times 19200}$$

$$= 16495.6029\ lb/jam$$

Atau setara dengan 7482.3564 kg/jam.

L.4.1.2 Kebutuhan Media Pendingin *(Cooling Tower)*

Air pendingin yang digunakan memiliki kondisi sebagai berikut :

Temperature masuk = 15ºC

Temperature keluar = 25ºC

Tabel L4.2 Kebutuhan Cooling Tower Pabrik Ethylene

No	Alat	Kebutuhan Cooling Tower (Kg/h)
1	Cooler (C-01)	1921.8838
2	Cooler (C-02)	22885.8115
Total Kebutuhan Cooling Tower		24807.6953

Dengan asumsi factor keamanan dan penyerapan panas dimasing-masing alat, setiap *cooling tower* yang dialirkan ke unit proses dilebihkan 10%, sehingga :

$$Total + FK\ 10\% = 1.1\ \times 24807.6953 = 27288.4649\,\frac{Kg}{jam}$$

Atau setara dengan 60160.1497 lb/jam.

- **Perancangan Menara Pendingin (Cooling Tower)**

Fungsi : Mendinginkan kembali air yang telah digunakan sebagai fluida pendingin pada alat-alat pendingin.

Jenis : *Induced Draft Cooling Tower*

Pola Aliran : *Counter current*, karena pola ini paling umum digunakan dan alirannya lebih efesien

Tabel L4.3 Kebutuhan Cooling Tower Pabrik Ethylene

No	Alat	Kebutuhan Cooling Tower (Kg/h)
Total Kebutuhan Cooling tower		27288.4649

Dengan asumsi factor keamanan dan penyerapan panas dimasing-masing alat, setiap *cooling tower* yang dialirkan ke unit proses dilebihkan 10%, sehingga :

$$Total + FK\ 10\% = 1.1\ \times 27288.4649 = 30017.3114\,\frac{Kg}{jam}$$

- Data

Laju alir massa masuk = 30017.3114 Kg/h

= 60160.1497 lb/h

Densitas bahan (30°C) = 1022.8753 Kg/m³

T air masuk (T1) = 50°C

= 122°F

T air keluar (T2) = 30°C

= 86°F

T reference = 25°C

= 77°F

Laju alir volumetrik air yang masuk menara pendingin (Wc)

$$Laju\ alir\ (Wc) = \frac{Laju\ alir\ massa\ bahan\ masuk}{densitas} = \frac{30017.3114}{1022.8753}$$

$$= 29.3460\ m^3/jam$$

Atau setara dengan 107.4946 gpm

Perhitungan Cooling range

Cooling range = T1-T2

= 36ºF

Perhitungan Approach Temperature

Temperatur bola basa (Tw) = 24ºC

= 75.2ºF

Temperatur Approach = T2 – Tw

= 10.8ºF

Penentuan tinggi menara cooling tower

Tinggi Menara = 7.5 – 9 m (Fig. 12-14, Perry's hal 12-16)

Dipilih, Tinggi Menara = 8 m

Perhitungan Luas menara cooling tower (A)

Kandungan air = 2.7 gal/menit ft2

$$Luas\ Menara = \frac{Wc}{Kandungan\ Air} = \frac{107.4946}{2.7} = 39.8128\ ft^2$$

Diperkirakan effisiensi menara adalah 90%, maka

$$\text{Luas menara sesungguhnya} = \frac{Luas\ Menara}{Efisiensi} = \frac{39.8128}{0.9} = 44.2364\ ft^2$$

Atau setara dengan 4.1096 m^2

Jari - Jari menara (r) = 1.1440 m

Diameter Menara = 2 x r

= 2.2880 m

Menghitung daya fan

Dari figur,12–15, Perry’s, hal 12-17 untuk effisiensi kerja cooling tower 90%, maka diperoleh :

Daya fan = 0.03 HP/ft^2

Daya fan sesungguhnya = daya fan x luas menara sesungguhnya

= 1.3271 HP

Menghitung jumlah air make up (Wm)

Wm = We + Wb + Wd (perry's, pers (12-9) hal 12-17

Dimana :

Wm = Jumlah make up water (lb/jam)

We = Jumlah air yang teruapkan (gpm)

Wb = Jumlah air yang dibuang untuk mengurangi konsentrasi padatan dalam air (blodown), gpm

Wd = Jumlah air yang terbawa dalam aliran uap yang keluar dari cooling tower (drift loss),gpm

Menghitung jumlah air yang menguap (We)

We = 0,0085 x Wc x (Tin - Tout) (pers.12-10 , perry's hal 12-17)

= 32.8933 gpm

Menghitung jumlah air yang terbawa blowdown (Wb)

Diketahui :

Jumlah siklus berkisar antara 3 -5 siklus (perry's hal 12-17)

Siklus = 3

Wb = 1.0000 gpm

Menghitung drift loss (Wd)

Wd = (0.1% – 0.2 %) x Wc

= 6.5787 gpm

Jadi, jumlah make up water (Wm)

Wm = We + Wb + Wd

= 32.8933 + 1.0000 + 6.5787

= 40.4720 gpm

= 9.3086 m^3/jam

= 9521.4959 kg/jam

= 381267.3500 lb/jam

L.4.1.3 Kebutuhan Air untuk *Dowtherm*

Air pendingin yang digunakan memiliki kondisi sebagai berikut :

Temperature masuk = 10ºC

Temperature keluar = 25ºC

Tabel L4.4 Kebutuhan Air umtuk *Dowtherm* Pabrik Ethylene

No	Alat	Kebutuhan Dowtherm (Kg/h)
1	Kondenser Parsial (CS-01)	212971.8203
Total Kebutuhan Cooling Tower		212971.8203

Dengan asumsi factor keamanan dan penyerapan panas dimasing-masing alat, setiap *dowtherm* yang dialirkan ke unit proses dilebihkan 10%, sehingga :

$$Total + FK\ 10\% = 1.1\ \times 212971.8203 = 234269.0023\,\frac{Kg}{jam}$$

Atau setara dengan 516469.4425 lb/jam.

L.4.1.4 Kebutuhan Air Domestik

- Air Sanitasi

  Kebutuhan Air per orang = 100 – 120 L/hari

  Permisalan = 120 L/hari

  Jumlah Karyawan = 146

  Total = 17520 L/hari

- Air Laboratorium

  Kebutuhan Air Lab = 95 -200 L/hari/staff

  Permisalan = 95 L/hari/staff

  Jumlah Staff RnD & QC = 20 staff

  Total = 1900 L/hari

- *Hydrant*

  Kebutuhan = 20 L/jam

  Total = 480 L/hari

$$\text{Total Air Domestik} = 17520 + 1900 + 480 = 19900\ \text{L/hari}$$

Atau setara dengan 19.90 m³/hari atau 0.829 m³/hari atau 848.1341 kg/jam.

Kondisi Air Domestik yang digunakan :

Temperature = 30ºC

ρ = 1022.8753Kg/m³

Massa air domestik = 848.1341Kg/jam

Asumsi factor keamanan sebesar 10%, maka jumlah air domestik yang dibutuhkan adalah:

$$Total + FK\ 10\% = 1.1\ \times 848.1341 = 932.9475\ Kg/jam$$

Maka kebutuhan Air Keseluruhan adalah ;

Tabel L4.5 Kebutuhan Keseluruhan Air Pabrik Ethylene

No	Kebuthan Air	Star Up (Kg/jam)	Kontinyu (Kg/jam)
1	Air Umpan Boiler	57096.8669	-
2	Make Up Boiler	-	5709.6867
3	Air Pendingin	27288.4649	-
4	Make Up Cooling Tower	-	9521.4959
5	Air Pendingin (10˚C)	234269.0023	-
6	Make up air pendingin (10˚C)	-	234269.0023
7	Air Domestik	932.9475	932.9475
Total		319587.2816	250433.1324

Atau setara dengan 269.3158 m³/jam atau 275476.446 kg/jam.

L.4.1.5 Unit Pengolahan Air

Keterangan :

Densitas air suhu (30 degC) = 1022.8753 kg/m³

Volume air bersih yang diperlukan (kontinyu) = 250433.1324 Kg/jam

= 244.8325 m³/jam

Dengan memperhitungkan faktor keamanan 10%, total kebutuhan air

$$Total + FK\ 10\% = 1.1\ \times 244.8325 = 269.3158\frac{\mathrm{m}^3}{jam}$$

Atau setara dengan 275476.4456 Kg/jam.

Asumsi :

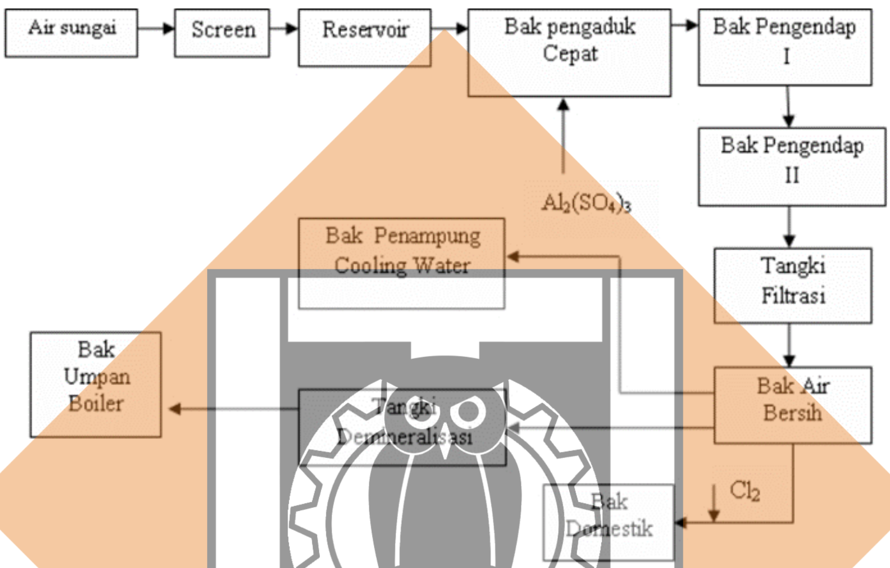
Padatan tertahan di filter pada bak reservoir, bak pengendap I, II, dan tangki filtrasi masing-masing sebesar 5 % dari volume total tiap bak, maka :

Jumlah air sungai yang harus diolah = 330.6491 m^3/jam

= 338212.8076 Kg/jam

= 87357.7597 Gallon

Spesifikasi Peralatan Unit Pengolahan Air



GAMBAR L4.1 UNIT PENGOLAHAN AIR

- **Screen (SC)**

  Fungsi = Menyaring benda-benda pengotor yang berukuran besar seperti plastik,sampah, ranting pohon, daun, dan sebagainya yang terbawa dalam aliran sungai.

  Ukuran screen = 2 m x 3 m

  Ukuran lubang = 1 cm x 1 cm

  Bahan Konstruksi = Besi

- **Reservoir**

  Fungsi = Menampung air sungai yang keluar dari screen

  Bentuk = Empat persegi panjang

  Bahan Konstruksi = Beton

  Jumlah = 1 unit

  Waktu tinggal = 12 jam

Diasumsikan partikel2 pengotor yang tertinggal adalah sebesar 5% dari volume total tiap bak, sehingga

$$Volume\ air\ yang\ harus\ diolah = \frac{Kebutuhan\ air\ yang\ tersedia}{(0.95)^4} = \frac{269.3158}{(0.95)^4}$$

$$= 304.2110\ m^3/jam$$

Volume air yang harus ditampung

Volume air = Volume air yang harus diolah x Waktu tinggal

= 3650.5315 m^3

Volume Bak + FK 20% = 1.2 x Volume air yang harus ditampung

= 4380.6378 m^3

V = P X L X T

Dimana :

P ; L ; T = 03 ; 02 ; 01

Maka :

V = 3T x 2T x T

= $6T^3$

T^3 = V/6

= 730.1063 m^3

T = 9.0046 m

Diperoleh :

P = 3T

= 27.0137 m

L = 2T

= 18.0091 m

- **Bak Pengaduk Cepat (BPC)**

Fungsi = Menggumpalkan partikel-partikel pengotor yang ada di dalam air sungai dengan penambahan koagulan ($Al_2(SO_4)_3$)

Bentuk = Tangki silinder vertikal

Bahan Konstruksi = Stainless Steel SA-240 Grade S Type 304

Jumlah = 1 unit

Waktu tinggal = 20 menit - 1 jam (sumber : powel, water condition for industry)

= 40 menit

= 0.6667 jam

Volume air pada bak pengaduk cepat :

$$Volume\ air = \frac{Kebutuhan\ air\ yang\ tersedia}{(0.95)^3} = \frac{269.3158}{(0.95)^3} = 314.1167\ m^3/jam$$

Volume air yang harus ditampung

Volume air = Volume air yang harus diolah x Waktu tinggal

= 209.4111 m^3

Volume Bak + FK 20% = 1.2 x Volume air yang harus ditampung

= 251.2933 m^3

$$Volume = \frac{1 \text{ x } \pi \text{ x ID2 x 2ID}}{4}$$

$$Diambil = \frac{H}{D} = 2, maka\ H = 2D$$

Maka :

$$Volume = \frac{1 \text{ x } \pi \text{ x D3}}{2}$$

$$D3 = \frac{2 \text{ x Volume}}{\pi} = \frac{2 \text{ x } 251.2933}{3.14} = 160.0594\ m^3$$

D = 5.4295 m

= 213.7603 inch

r = D/2

= 2.7148 m

= 106.8801 inch

H = D x 2

= 10.8590 m

= 427.5205 inch

Kebutuhan koagulan ($Al_2(SO_4)_3$)

Kadar ($Al_2(SO_4)_3$) = 50 – 70 mg/L , Diambil : 60 mg/L = 0.00006 Kg/L

Jumlah air yang harus diolah = 209.4111 m^3/jam

= 209411.1097 L/jam

Kebutuhan ($Al_2(SO_4)_3$) = Jumlah air yang harus diolah x Kadar ($Al_2(SO_4)_3$

= 12.5647 Kg/jam

= 301.5520 Kg/hari

Tebal tangki

$$t = \frac{\text{Pdesign x r}}{\text{f x E} - 0.6 \text{ x Pdesign}} + C$$

Dimana,

P = Tekanan operasi = 14.696 psi

r = jari-jari dalam tangki (inch) = 106.8801 inch

f = tegangan maksimum yang diinginkan (allowable stress) = 18750

E = Efisiensi penyambungan = 0.8, (tipe double welded butt joint tanpa diradiografi)

c = Faktor korosi = 0.0125 inch/tahun

umur tangki diperkirakan 10 tahun, maka c = 0.125 inch

t = Tebal dinding tangki (inch) = 0.2298 inch

Diambil tebal standart = 0.5000 inch

= 0.0127 m

Diamter luar tangka (DO)

DO = Di + (2 x ts)

= 214.7603 inch

= 5.4549 m

Diambil DO standar = 216 inch

- **Bak Pengendap I**

Fungsi = Mengendapkan gumpalan - gumpalan yang lebih besar dari bak pengadukan cepat

Bentuk = Tangki silinder vertikal

Bahan Konstruksi = Stainless Steel SA-240 Grade S Type 304

Jumlah = 1 unit

Waktu tinggal = 2 – 6 jam (Sumber : Powel, Water Condition for Industry)

= 2 jam

Volume air pada bak pengaduk cepat :

$$Volume\ air = \frac{Kebutuhan\ air\ yang\ tersedia}{(0.95)^3} = \frac{269.3158}{(0.95)^3} = 314.1167\ m^3/jam$$

Volume air yang harus ditampung

Volume air = Volume air yang harus diolah x Waktu tinggal

= 628.2333 m^3

Volume Bak + FK 20% = 1.2 x Volume air yang harus ditampung

= 753.8800 m^3

$$Volume = \frac{1 \text{ x } \pi \text{ x ID2 x 2ID}}{4}$$

$$Diambil = \frac{H}{D} = 2, maka\ H = 2D$$

Maka :

$$Volume = \frac{1 \text{ x } \pi \text{ x D3}}{2}$$

$$D3 = \frac{2 \text{ x Volume}}{\pi} = \frac{2 \text{ x } 753.8800}{3.14} 480.1783\ m^3$$

D = 7.8307 m

= 308.2956 inch

r = D/2

= 3.9154 m

= 154.1478 inch

H = D x 2

= 15.6614 m

= 616.5913 inch

Tebal tangki

$$t = \frac{\text{Pdesign x r}}{\text{f x E} - 0.6 \text{ x Pdesign}} + C$$

Dimana,

P = Tekanan operasi = 14.696 psi

r = jari-jari dalam tangki (inch) = 154.1478 inch

f = tegangan maksimum yang diinginkan (allowable stress) = 18750

E = Efisiensi penyambungan = 0.8, (tipe double welded butt joint tanpa diradiografi)

c = Faktor korosi = 0.0125 inch/tahun

umur tangki diperkirakan 10 tahun, maka c = 0.125 inch

t = Tebal dinding tangki (inch) = 0.2761 inch

Diambil tebal standart = 1.0000 inch

= 0.0254 m

Diamter luar tangka (DO)

DO = Di + (2 x ts)

= 308.3464 inch

= 7.8320 m

Diambil DO standar = 228 inch

- **Bak Pengendap II**

Fungsi = Mengendapkan flok-flok yang lebih halus partikel partikelnya, yang tidak dapat terendapkan pada bak sebelumnya

Jenis = Bak persegi empat

Bahan Konstruksi = Beton

Jumlah = 1 unit

Waktu tinggal = 2 – 6 jam (Sumber : Powel, Water Condition for Industry)

= 4 jam

Volume air pada bak pengaduk cepat :

$$Volume\ air = \frac{Kebutuhan\ air\ yang\ tersedia}{(0.95)^2} = \frac{269.3158}{(0.95)^2} = 298.4108\ m^3/jam$$

Volume air yang harus ditampung

Volume air = Volume air yang harus diolah x Waktu tinggal

= 1193.6433 m^3

Volume Bak + FK 20% = 1.2 x Volume air yang harus ditampung

= 1432.3720 m^3

Volume = P x L x T

Kedalaman bak (T) = 10 - 20 ft (Powel, water conditioning for industry) diambil :

Kedalaman bak (T) = 10 ft

= 3.0480 m

P/ L = 1 – 2.5 (Powel, water conditioning for industry), diambil :

P/L = 2

P = 2L

A = V/T

$= 469.9383 \ m^2$

A $= P \times L$

$= 2L \times L$

$= 2L^2$

L $= (A/2)^{0.5}$

$= 15.3287$ m

P $= 2L$

$= 30.6574$ m

T $= 3.0480$ m

- **Tangki Filtrasi (Sand Filter)**

Fungsi = Untuk menyaring partikel partikel halus yang tersisa dalam air yang berasal dari bak pengendap II

Bentuk = Tangki silinder vertikal

Bahan Konstruksi = Stainless Steel SA-240 Grade S Type 304

Jumlah = 1 unit

Media Penyaring = Pasir dan Kerikil

Laju alir volumetrik air sungai yang akan disaring,:

$$Laju\ alir\ volumetrik\ air = \frac{Kebutuhan\ air\ yang\ tersedia}{(0.95)^1} = \frac{269.3158}{(0.95)^1}$$

$$= 283.4903\ m^3/jam$$

Atau setara dengan 1248.1510 gpm

Kecepatan filtrasi = 30 – 60 gpm/ft^2 (Sumber : Powel, Water Condition for Industry)

Diambil :

Kecepatan filtrasi = 60 gpm/ft^2

Luas penampang saringan (A)

$$A = \frac{Laju\ alir\ sungai\ yang\ disaring}{Kecepatan\ filtrasi} = \frac{1248.1510}{60} = 20.8025\ ft^2$$

Atau setara dengan 1.9326 m^2

$= 0.1933$ m^3

Diameter Permukaaan Tangki (D)

$$Volume = \frac{1 \text{ x } \pi \text{ x ID}^2 \text{ x 2ID}}{4}$$

$$Diambil = \frac{\text{H}}{D} = 2, maka\ H = 2D$$

Maka :

$$Volume = \frac{1 \text{ x } \pi \text{ x D3}}{2}$$

$$D3 = \frac{2 \text{ x Volume}}{\pi} = \frac{2 \text{ x 0.1933}}{3.14} = 0.1231\ m^3$$

D = 0.4974 m

= 19.5844 inch

r = D/2

= 0.2487 m

= 9.7922 inch

H = D x 2

= 0.9949 m

= 39.1688 inch

Tebal tangki (ts)

Komposisi ketebalan lapisan media penyaring :

Tinggi tumpukan pasir = 12-20 inch

Diambil, 15 inch

Tinggi tumpukan kerikil = 21-40 inch

Diambil, 30 inch

Maka :

Tinggi lapisan media penyaring (tebal bed) = 45 inch

= 1.143 m

Diamsusikan : tebal head = 1/3 x tinggi tangki

Maka :

Tinggi tangki = 0.3810 m

= 15.0000 inch

Tebal tangki

$$t = \frac{\text{Pdesign x r}}{\text{f x E} - 0.6 \text{ x Pdesign}} + C$$

Dimana,

P = Tekanan operasi = 14.696 psi

r = jari-jari dalam tangki (inch) = 9.7922 inch

f = tegangan maksimum yang diinginkan (allowable stress) = 18750

E = Efisiensi penyambungan = 0.8, (tipe double welded butt joint tanpa diradiografi)

c = Faktor korosi = 0.0125 inch/tahun

umur tangki diperkirakan 10 tahun, maka c = 0.125 inch

t = Tebal dinding tangki (inch) = 0.1346 inch

Diambil tebal standart = 0.1875 inch

= 0.0048 m

Diamter luar tangki (DO)

DO = Di + (2 x ts)

= 19.5939 inch

= 0.4977 m

Diambil DO standar = 20 inch

- **Bak Penampung Air Bersih**

Fungsi = Menampung air yang keluar dari bak filtrasi

Bentuk = Empat persegi panjang

Bahan Konstruksi = Beton

Jumlah = 1 unit

Waktu tinggal = 6 jam

= 360 menit

Volume air pada tangki penampung air bersih :

$$Volume\ air\ yang\ harus\ diolah = \frac{Kebutuhan\ air\ yang\ tersedia}{(0.95)^0} = \frac{269.3158}{(0.95)^0}$$

$$= 269.3158\ m^3/jam$$

Volume air yang harus ditampung

Volume air = Volume air yang harus diolah x Waktu tinggal

= 1615.8947 m^3

Volume Bak + FK 20% = 1.2 x Volume air yang harus ditampung

= 1939.0736 m^3

V = P X L X T

Dimana :

P ; L ; T = 03 ; 02 ; 01

Maka :

V = 3T x 2T x T

= $6T^3$

T^3 = V/6

= 323.1789 m^3

T = 6.8625 m

Diperoleh :

P = 3T

= 20.5874 m

L = 2T

= 13.7250 m

T = 6.8625 m

- **Tangki Demineralisasi *(Ion Exchanger)***

Fungsi = Untuk menghilangkan kesadahan air dan kandungan mineral dalam air dengan menggunakan resin penukar ion

Bentuk = Tangki silinder vertikal

Bahan Konstruksi = Stainless Steel SA-240 Grade S Type 304

Jumlah = 1 unit

Media = Resin sintesis

Tipe Resin = *Mixed cation and strong base anion (Chemical equivalent mixture)*

Air yang Diproses = *Make up boiler*

Dari tabel 19-7, Perry's, hal 19-41 diperoleh :

Kecepatan alir maksimum (v) = 40 m/jam

= 16 gpm

Diambil :

Kecepatan alir = 20 m/jam

Digunakan tinggi bed = 1.2 m

= 3.9370 ft

= 47.2435 inch

Laju alir air yang didemineralisasi = 5709.6867 kg/jam

= 5.7097 m^3/jam

Dengan memperhitungkan faktor keamanan sebesar 10%, maka laju alir yang akan didemineralisasi, yaitu :

Laju alir air yang dimineralisasi = 1.1 x laju alir air yg didemineralisasi

= 6.2807 m^3/jam

Luas permukaan tangki (A)

A = Q/V

= 0.3140 m^2

= 0.0314 m^3

Diameter Permukaaan Tangki (D)

$$Volume = \frac{1 \text{ x } \pi \text{ x } ID^2 \text{ x } 2ID}{4}$$

$$Diambil = \frac{H}{D} = 2, maka\ H = 2D$$

Maka :

$$Volume = \frac{1 \text{ x } \pi \text{ x } D3}{2}$$

$$D3 = \frac{2 \text{ x Volume}}{\pi} = \frac{2 \text{ x } 0.0314}{3.14} = 0.0200\ m^3$$

D = 0.2715 m

= 10.6871 inch

r = D/2

= 0.1357 m

= 5.3435inch

H = D x 2

= 0.5429 m

= 21.3741 inch

Tinggi tangki keseluruhan

Diasumsikan :

Tinggi resin = 1.2 m

Tinggi bed = 1/2 x tinggi tangki

Maka :

Tinggi tangki keseluruhan = 2 x tinggi resin

= 2.4 m

= 94.4882 inch

Tebal tangka (ts)

$$t = \frac{\text{Pdesign x r}}{\text{f x E} - 0.6 \text{ x Pdesign}} + C$$

Dimana,

P = Tekanan operasi = 14.696 psi

r = jari-jari dalam tangki (inch) = 0.1357inch

f = tegangan maksimum yang diinginkan (allowable stress) = 18750

E = Efisiensi penyambungan = 0.8, (tipe double welded butt joint tanpa diradiografi)

c = Faktor korosi = 0.0125 inch/tahun

umur tangki diperkirakan 10 tahun, maka c = 0.125 inch

t = Tebal dinding tangki (inch) = 0.1251inch

Diambil tebal standart = 0.1875 inch

= 0.0048 m

Diamter luar tangki (DO)

DO = Di + (2 x ts)

= 11.0621 inch

= 0.2810 m

Diambil DO standar = 12 inch

- **Bak Umpan Boiler**

Fungsi = Menampung air sungai yang keluar dari screen

Bentuk = Empat persegi panjang

Bahan Konstruksi = Beton

Jumlah = 1 unit

Waktu tinggal = 1 jam

Volume air yang harus ditampung adalah volume air yang diperlukan untuk membuat steam

$$Volume\ air\ yang\ ditampung = \frac{Kebutuhan\ air\ boiler}{Densitas\ air} = \frac{57096.8669}{1022.8753}$$

$$= 55.8200\ m^3/jam$$

Volume bak penampung dengan faktor keamanan sebesar 20%, maka :

Volume Bak + FK 20% = 1.2 x Volume air yang harus ditampung

= 66.9840 m^3

V = P X L X T

Dimana :

P ; L ; T = 03 ; 02 ; 01

Maka :

V = 3T x 2T x T

= $6T^3$

T^3 = V/6

= 11.1640 m^3

T = 2.2350 m

Diperoleh :

P = 3T

= 6.7049 m

L = 2T

= 4.4700 m

T = 2.2350 m

- **Bak Umpan Cooling Tower**

Fungsi = Untuk menampung air sebagai feed cooling tower

Bentuk = Empat persegi panjang

Bahan Konstruksi = Beton

Jumlah = 1 unit

Waktu tinggal = 1 jam

Volume air yang harus ditampung adalah volume air yang diperlukan untuk membuat steam

$$Volume\ air\ yang\ ditampung = \frac{Kebutuhan\ air\ pendingin}{Densitas\ air} = \frac{9521.4959}{1022.8753}$$

$$= 9.3086\ m^3/jam$$

Volume bak penampung dengan faktor keamanan sebesar 20%, maka :

Volume Bak + FK 20% = 1.2 x Volume air yang harus ditampung

= 11.1703 m^3

V = P X L X T

Dimana :

P ; L ; T = 03 ; 02 ; 01

Maka :

V = 3T x 2T x T

= $6T^3$

T^3 = V/6

= 1.8617 m^3

T = 1.2302 m

Diperoleh :

P = 3T

= 3.6906 m

L = 2T

= 2.4604 m

T = 1.2302 m

- **Bak Penampung Air Domestik**

Fungsi = Untuk menampung kondensat steam untuk feed boiler

Bentuk = Empat persegi panjang

Bahan Konstruksi = Beton

Jumlah = 1 unit

Waktu tinggal = 24 jam

Volume air yang harus ditampung adalah volume air yang diperlukan untuk membuat steam

$$Volume\ air\ yang\ ditampung = \frac{Kebutuhan\ air\ domestik}{Densitas\ air} = \frac{932.9475}{1022.8753}$$

$$= 0.9121\ m^3/jam$$

Volume bak penampung dengan faktor keamanan sebesar 20%, maka :

Volume Bak + FK 20% = 1.2 x Volume air yang harus ditampung

= 1.0945 m^3

V = P X L X T

Dimana :

P ; L ; T = 03 ; 02 ; 01

Maka :

V = 3T x 2T x T

= $6T^3$

T^3 = V/6

= 0.1824 m^3

T = 0.5671 m

Diperoleh :

P = 3T

= 1.7014 m

L = 2T

= 1.1343 m

T = 0.5671 m

- **Bak Penampung Limbah**

Fungsi = Untuk menampung air limbah

Bentuk = Empat persegi panjang

Bahan Konstruksi = Beton

Jumlah = 1 unit

Waktu tinggal = 12 jam

Volume limbah yang ditampung = Limbah yang dihasilkan x waktu tinggal

= 5.8232 m^3/jam x 12 jam

= 69.8784 m^3

Volume bak penampung dengan faktor keamanan sebesar 20%, maka :

Volume Bak + FK 20% = 1.2 x Volume air yang harus ditampung

= 83.8541 m^3

V = P X L X T

Dimana :

P ; L ; T = 03 ; 02 ; 01

Maka :

V $= 3T \times 2T \times T$

$= 6T^3$

$T^3 = V/6$

$= 13.9757\ m^3$

T = 2.4087 m

Diperoleh :

P = 3T

= 7.2262 m

L = 2T

= 4.8175 m

T = 2.4087 m

L4.2 Perhitungan Daya Pabrik

L.4.2.1 Daya Proses

Berikut rincian detail daya proses :

Tabel L4.6 Daya Proses Pabrik Ethylene

Nama Alat/Kode Alat	Fungsi	Daya (HP)
Pompa Utilitas-01 (PU-01)	Memompa air dari badan sungai ke screen	2.95
Pompa Utilitas-02 (PU-02)	Memompa air dari screen ke reservoir	2.95
Pompa Utilitas-03 (PU-03)	Memompa air dari reservoir ke bak pengadukan cepat (BPC)	2.95
Pompa Utilitas-04 (PU-04)	Memompa air dari bak pengaduk cepat ke bak pengendap I	2.95
Pompa Utilitas-05 (PU-05)	Memompa air dari bak pengendap I ke bak pengendap II	2.95

Nama Alat/Kode Alat	Fungsi	Daya (HP)
Pompa Utilitas-06 (PU-06)	Memompa air dari bak pengendap II ke tangki filtrasi	2.95
Pompa Utilitas-07 (PU-07)	Memompa air dari tangki filtrasi ke bak air bersih	2.95
Pompa Utilitas-08 (PU-08)	Memompa air dari bak air bersih ke tangki demineralisasi	2.95
Pompa Utilitas-09 (PU-09)	Memompa air dari bak bersih ke bak penampung cooling tower	2.01
Pompa Utilitas-10 (PU-10)	Memompa air dari tangki demineralisasi ke bak umpan boiler	2.01
Pompa Utilitas-11 (PU-11)	Memompa air dari bak air bersih ke bak domestik	0.54
Pompa Utilitas-12 (PU-12)	Memompa air dari bak air bersih ke bak limbah	2.01
Fan cooling tower	Mendinginkan fluida panas menjadi fluida dingin dengan prinsip "dehumineralisasi"	1.33
Total		31.50

L.4.2.2 Daya Penunjang

- Peralatan bengkel

  Dalam suatu pabrik diperlukan fasilitas pemeliharaan dan perbaikan pabrik.

  Estimasi daya listrik yang dibutuhkan = 50 Kw/hari
- Instrumentasi

  Alat instrumentasi yang digunakan berupa alat control dan alat pendeteksi.

  Estimasi daya listrik yang diburuhkan = 15 Kw/hari
- Penerangan lampu jalan, pendingin ruangan dan perkantoran

  Alat penerangan yang dibutuhkan untuk pabrik, kantor dan lingkungan sekitar pabrik.

  Estimasi daya listrik yang dibutuhkan = 45 Kw/hari
- Alat kantor seperti computer, intercom, pengeras suara dan lainnya

  Estimasi daya listrik yang dibutuhkan = 30 Kw/hari

$$Total\ Daya\ Penunjang = 50 + 15 + 45 + 30 = 140\ KW/hari$$

Atau setara dengan 7.8226 HP/jam.

Kebutuhan Listrik Keseluruhan = Daya Proses + Daya Utilitas + Daya Penunjang

= 8086.9810 + 31.4971 + 7.8226

= 8126.3007 $\frac{HP}{jam}$

Asumsi factor keamanan sebesar 10%, maka jumlah air domestik yang dibutuhkan adalah:

$$Total + FK\ 10\% = 1.2\ \times 8126.3007 = 9751.5609 \frac{HP}{jam}$$

Atau setara dengan 7271.7389 KWH.

L.4.3 Perhitungan Bahan Bakar

L.4.3.1 Penyediaan Bahan Bakar

Untuk menjamin kontinuitas produksi dan kinerja perusahaan, disediakan 2 unit generator yang dilengkapi dengan Uninterupted Power System (UPS) yang akan menjalankan generator 7 detik setelah pemadaman terjadi.

Bahan bakar = Solar

Kapasitas = 1600 kW/unit

Jumlah = 2 unit

Heating value untuk solar = 19200 btu/lb

Densitas solar = 850 kg/m^3

- Penyediaan solar untuk generator

  Diasumsikan :

  Efisiensi pembakaran solar pada generator = 85%

  Terjadi pemadaman listrik selama = 1 jam/hari

  Generator yang digunakan 8000 kW (5 buah) = 7271.7389 kWH

  = 24812195.0051 btu/jam

$$m\ solar = \frac{Kebutuhan\ listrik}{h\ x\ Hv}$$

Keterangan :

Hv = *Heating value solar* = *19200 btu/lb*

h = *Efisiensi pembakaran* = *85%*

$$m\ solar = \frac{24812195.0051}{19200 \times 85\%} = 1520.3551\ lb/jam$$

Atau setara dengan 689.5034 kg/jam.

Diasumsikan pemadaman listrik 1 jam/ hari = 15858.5791 kg/jam.

Total kebutuhan solar dengan faktor keamanan 20% adalah :

$$Kebutuhan\ solar = 1.2\ \times 15858.5791 = 19030.2950 \frac{kg}{hari}$$

$$Volume = \frac{Kebutuhan\ solar}{Densitas\ solar} = \frac{19030.2950}{850} = 22.3886\ m^3/hari$$

Atau setara dengan 939.8766 liter/hari.

- Penyediaan solar untuk boiler

  Kebutuhan solar untuk boiler = 7482.3564 kg/jam

  Kebutuhan solar untuk satu hari = 179576.5539 kg/hari

  Total Kebutuhan Bahan Bakar solar = 195435.1331 kg/hari

$$Volume = \frac{Kebutuhan\ solar}{Densitas\ solar} = \frac{195435.1331}{850} = 229.9237\ m^3/hari$$

Atau setara dengan 229923.6860 liter/hari.

- Penyediaan solar untuk kendaraan

  Kebutuhan solar 3 mobil operasional yaitu sebesar = 60 liter/hari = 0.06 m3/hari = 51 kg/hari

  Total Bahan Bakar Solar yang Dibutuhkan

  Bahan bakar solar yang dibutuhkan adalah

  Untuk generator = 22388.5823 liter/hari

  Untuk boiler = 229923.6860 liter/hari

  Untuk kendaraan = 60.0000 liter/hari

  Total Kebutuhan Bahan Bakar solar = 252372.2683 liter/hari

  = 211344.7122 kg/hari

L.4.3.2 Perancangan Tangki Bahan Bakar

Fungsi	Untuk menyimpan solar sebagai total bahan bakar yang dibutuhkan selama proses (generator,boiler,kendaraan)
Bentuk	Silinder berbentuk tegak berdasarkan rata beratap *flasnged* and *dished heads*
Bahan	Carbon steel SA -283 grade C
Jumlah	1 Unit

T Operasi	30	degC	303	K
P Operasi	1	atm		
ρ	850	Kg/m^3		

Menghitung Volume Tangki

Penggunaan selama 1 hari	=	211344.7122	kg/hari
Total selama 7 hari	=	1479412.9853	Kg/7 hari
Volume bio solar	=	$\frac{\text{m kebutuhan}}{\text{Densitas}}$	
	=	1740.4859	m3
Rencana tangki yang digunakan	=	1	unit
Volume tangki	=	1740.4859	m3
Faktor keamanan 20 %	=	20%	
Volum tangki (VT)	=	2088.5830	m3
	=	551740.9812	gal

Menentukan Ukuran Tangki

Diasumsikan

H	=	2 D	
Volume total	=	$1/4 \times \pi \times ID^2 \times H$	
1805.0338	=	$1/4 \times \pi \times ID^2 \times H$	
ID^3	=	1329.6333	m
ID	=	10.9962	m
	=	432.9226	Inch

H	=	2	D
H	=	21.9925	m
	=	865.8452	inch

Menentukan Tebal Dinding Tangki

Bahan Tangki = SS SA 167, Grade 3, Type 304

Tinggi Cairan Dalam Tangki	=	80%	Tinggi Tangki
	=	16.7560	m

Tekanan Design Tangki

P Operasi	=	1	atm
g	=	9.8	m/s^2
P Hidrostatik	=	ρ x g x h	
	=	146557.8012	kg/ms^2
	=	146557.8012	N/m^2
	=	1.4464	atm
	=	21.2565	psi
P Desain	=	2.9357	atm
	=	43.1430	psi

$$\text{Tebal Tangki (ts)} = \frac{(P \times r)}{(f \times E - 0.6 \times P)} + C$$

Persamaan 14.34, Brownell and Young, hal 275

ts = Tebal Tangki	=			
P = Tekanan Design	=	43.1430	Psi	
r = Jari -Jari Tangki	=	216.4613	inch	
f = Max Allowable Stress	=	18750	Psi	
E = Jenis sambungan (Hal 254)	=	0.8		
C = Faktor Korosi	=	0.15		Petters Hal 542

Umur alat	=	10	Tahun
ts	=	0.77	Inch

ts standar	=	1	Inch	*Brownell & Young, tabel 5.7 Hal 91*

LAMPIRAN 5 ANALISIS EKONOMI

L.5.1 Ketetapan Analisa Ekonomi

Ketetapan yang digunakan dalam perhitungan Analisis Ekonomi Pabrik Ethylene adalah sebagai berikut sumber: (Keuangan, 2024) :

1.	Konstruksi	=	2025		
2.	Produksi komersial	=	2027		
3.	Operasi per tahun	=	330	hari	
4.	Shut down	=	35	hari	
5.	Umur teknis pabrik	=	10	tahun	
6.	Nilai tukar rupiah	=	15623	rupiah	
7.	Suku bunga bank	=	10%	per tahun	*
8.	Kenaikan harga bahan baku dan produk	=	10%	per tahun	
9.	*Salvage Value* (Nilai Rongsok)	=	10%	DFCI tanpa harga tanah	

Tabel L5.1 Suku Bunga Dasar Kredit Bank Umum Konvensional di Indonesia

No	Nama Bank	Kredit Korporasi	Kredit Ritel	Kredit Mikro	Kredit Konsumsi	
					KPR	Non KPR
1	PT BANK RAKYAT INDONESIA (PERSER0), Tbk	8.00	8.25	14.00	7.25	8.75
2	PT BANK MANDIRI (PERSERO), Tbk	8.05	8.30	11.30	7.30	8.80
3	PT BANK NEGARA INDONESIA (PERSERO), Tbk	8.05	8.30	-	7.30	8.80
4	PT BANK DANAMON INDONESIA, Tbk	8.50	9.00	-	8.00	9.25
5	PT BANK PERMATA, Tbk	8.50	9.00	-	0.00	8.75
6	PT BANK CENTRAL ASIA, Tbk	7.90	8.10	-	7.20	5.96
7	PT BANK MAYBANK INDONESIA, Tbk	8.00	9.00	-	8.00	8.75
8	PT BANK PAN INDONESIA, Tbk	8.75	8.55	15.58	8.05	8.33

No	Nama Bank	Kredit Korporasi	Kredit Ritel	Kredit Mikro	Kredit Konsumsi	
					KPR	Non KPR
9	PT BANK CIMB NIAGA, Tbk	8.00	8.75	-	7.30	8.50
10	PT BANK UOB INDONESIA	8.50	9.40	-	8.20	-
11	PT BANK OCBC NISP, Tbk	8.25	8.75	-	8.00	9.25
12	CITIBANK, N.A	5.75	-	-	-	-
13	JP MORGAN CHASE BANK,N.A	7.22	-	-	-	-
14	BANK OF AMERICA, N.A	7.00	-	-	-	-
15	PT BANK CHINA CONSTRUCTION BANK INDONESIA, Tbk	7.27	7.73	-	7.58	8.86
16	PT BANK ARTHA GRAHA INTERNASIONAL, Tbk	10.70	10.19	-	11.62	15.09
17	MUFG BANK, LTD	6.64	-	-	-	-
18	PT BANK DBS INDONESIA	7.41	9.56	-	7.00	-
19	PT BANK RESONA PERDANIA	7.77	-	-	-	-
20	PT BANK MIZUHO INDONESIA	7.25	-	-	-	-
21	STANDARD CHARTERED BANK	7.94	-	-	-	-
22	PT BANK CAPITAL INDONESIA, Tbk	12.00	12.00	12.00	12.50	12.50
23	PT BANK BNP PARIBAS INDONESIA	6.51	-	-	-	-
24	PT BANK ANZ INDONESIA	6.26	-	-	-	-
25	DEUTSCHE BANK AG	7.25	-	-	-	-

No	Nama Bank	Kredit Korporasi	Kredit Ritel	Kredit Mikro	Kredit Konsumsi	
					KPR	Non KPR
26	BANK OF CHINA HONGKONG LIMITED	6.48	6.48	-	-	-
27	PT BANK BUMI ARTA, Tbk	7.74	7.86	12.88	7.31	13.10
28	PT BANK HSBC INDONESIA	6.75	8.75	-	7.88	-
29	PT BANK JTRUST INSONESIA, Tbk	12.05	12.75	-	11.8 5	14.75
30	PT BANK MAYAPADA INTERNATIONAL, Tbk	10.60	12.20	14.20	11.9 0	12.00
31	PT BPD JAWA BARAT DAN BANTEN, Tbk	7.21	9.74	0.93	9.22	9.26
32	PT BPD DKI	8.50	9.00	10.50	8.50	9.25
33	PT BPD DAERAH ISTIMEWA YOGYAKARTA	5.52	5.50	5.43	5.20	6.01
34	PT BPD JAWA TENGAH	7.80	7.86	9.39	7.61	10.34
35	PT BPD JAWA TIMUR, Tbk	7.67	8.19	13.34	8.21	9.57
36	PT BPD JAMBI	9.93	8.14	8.05	7.84	7.87
37	PT BPD SUMATERA UTARA	9.61	9.95	11.50	9.88	11.09
38	PT BANK NAGARI	8.71	10.23	12.37	9.63	10.48
39	PT BPD SUMATERA SELATAN DAN BANGKA BELITUNG	8.60	9.25	11.60	8.85	10.35
40	PT BPD LAMPUNG	9.29	9.29	9.29	9.29	9.29
41	PT BPD KALIMANTAN SELATAN	9.75	9.95	10.57	9.80	9.94
42	PT BPD KALIMANTAN BARAT	5.62	6.63	6.49	6.54	7.89

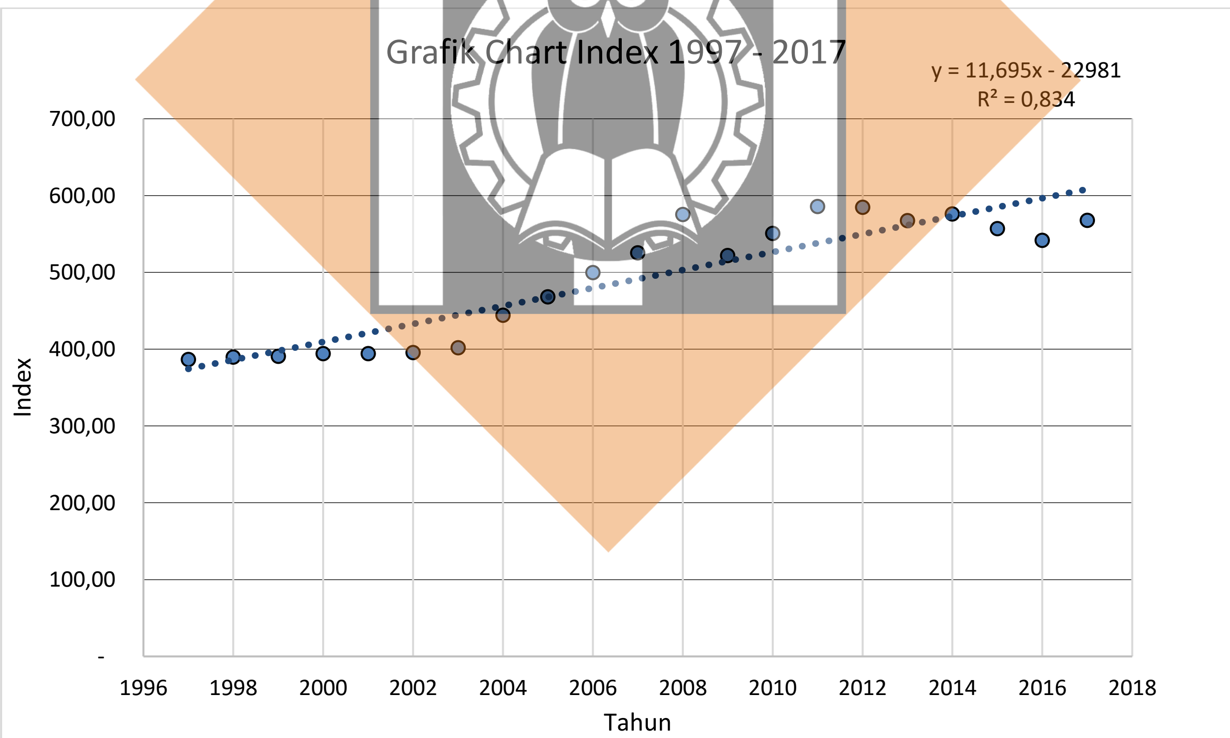
No	Nama Bank	Kredit Korporasi	Kredit Ritel	Kredit Mikro	Kredit Konsumsi	
					KPR	Non KPR
43	PT BPD KALIMANTAN TIMUR DAN KALIMANTAN UTARA	7.34	7.19	7.19	7.34	7.05
44	PT BPD KALIMANTAN TENGAH	7.50	10.50	9.37	8.32	8.81
45	PT BPD SULAWESI SELATAN DAN SULAWESI BARAT	8.40	8.25	8.59	8.25	10.20

Cost Index yangdigunakan dalam perhitungan Analisis Ekonomi adalah sebagai berikut, (Cheresources, 2022):

Tabel L5.2 Tabel Cost Index

Tahun	Index
1997	386.5
1998	389.5
1999	390.6
2000	394.1
2001	394.3
2002	395.6
2003	401.7
2004	444.2
2005	468.2
2006	499.6
2007	525.4
2008	575.4
2009	521.9
2010	550.8
2011	585.7
2012	584.6
2013	567.3

2014	576.1
2015	556.8
2016	541.7
2017	567.5
2018	619.5
2019	631.2
2020	642.9
2021	654.6
2022	666.3
2023	678.0
2024	689.7
2025	701.4
2026	713.1
2027	724.8



L.5.2 Perhitungan Gaji Karyawan

Perkiraan perhitungan gaji karyan bersumber dari UMK daerah Banten (Kompas, 2022) dan Peraturan Menteri Keuangan (PMK) Nomor 250/PMK.03/2008 (Keungan, 2008).

UMK Cilegon	=	Rp4,815,108	
Pendidikan			
SMA	=	8%	Rp 385,209
SMK	=	10%	Rp 481,511
D3	=	15%	Rp 722,266
S1	=	20%	Rp 963,022
S2	=	35%	Rp 1,685,288
Leader	=	45%	Rp 2,166,799
Spv	=	50%	Rp 2,407,554
Shift	=	5%	Rp 240,755

Tabel L5.3 Gaji Karyawan

No.	Jabatan	Jumlah	Jenjang Pendidikan (minimum)	Gaji Bulanan	Total
1	Dewan Komisaris	2	S1	Rp 20,223,454	Rp 40,446,907
2	Direktur	1	S1	Rp 24,075,540	Rp 24,075,540
3	Manager	5	S2	Rp 13,723,058	Rp 68,615,289
4	Supervisor	13	S1	Rp 10,593,238	Rp 137,712,089
5	Sekretaris Direktur	1	S2	Rp 6,500,396	Rp 6,500,396
6	Sekretaris Manager	4	S1	Rp 5,778,130	Rp 23,112,518
	Karyawan *Shift*				
Manager Produksi & Supply Chain					
7	**Bagian Proses**				
	- Leader	2	S1	Rp 6,741,151	Rp 13,482,302
	- Pelaksana	10	SMK	Rp 5,537,374	Rp 55,373,742
8	***Quality Assurance***				
	- Leader	2	S1	Rp 6,741,151	Rp 13,482,302
	- Pelaksana	6	D3	Rp 5,778,130	Rp 34,668,778
9	***Quality Control***				
	- Leader	2	S1	Rp 6,741,151	Rp 13,482,302
	- Pelaksana	8	D3	Rp 5,778,130	Rp 46,225,037
Manager Teknik					
10	**Utilitas**				
	- Leader	2	S1	Rp 6,741,151	Rp 13,482,302

No.	Jabatan	Jumlah	Jenjang Pendidikan (minimum)	Gaji Bulanan	Total
	- Pelaksana	6	D3	Rp 5,778,130	Rp 34,668,778
11	***Engineering***				
	- Leader	2	S1	Rp 6,741,151	Rp 13,482,302
	- Pelaksana	10	D3	Rp 5,778,130	Rp 57,781,296
12	**Keamanan**				
	- Leader	2	D3	Rp 6,018,885	Rp 18,056,655
	- Pelaksana	8	SLTA	Rp 5,200,317	Rp 41,602,533
Karyawan Non *Shift*					
Manager Produksi & Supply Chain					
13	**R&D**				
	- Leader	1	S1	Rp 6,259,640	Rp 6,259,640
	- Pelaksana	5	S1	Rp 5,296,619	Rp 26,483,094
14	**Suppply Chain**				
	- Leader	1	S1	Rp 5,778,130	Rp 5,778,130
	- Pelaksana	2	S1	Rp 5,296,619	Rp 10,593,238
15	**Gudang**				
	- Leader	1	S1	Rp 5,778,130	Rp 5,778,130
	- Pelaksana	4	D3	Rp 5,055,863	Rp 20,223,454
Manager Sales & Marketing					
16	**Bagian Distribusi**				
	- Leader	1	S1	Rp 5,778,130	Rp 5,778,130
	- Pelaksana	4	SMK	Rp 4,815,108	Rp 19,260,432
17	**Bagian Pemasaran**				
	- Leader	1	S1	Rp 5,778,130	Rp 5,778,130
	- Pelaksana	4	D3	Rp 5,055,863	Rp 20,223,454
Manager SDM dan Umum					
18	**Bagian Kesehatan**				
	- Dokter	1	S1	Rp 5,778,130	Rp 5,778,130
	- Perawat	3	S1	Rp 5,296,619	Rp 15,889,856
19	**Bagian HSE *(Health Safety Environment)***				
	- Leader	2	S1	Rp 5,778,130	Rp 11,556,259
	- Pelaksana	4	D3	Rp 5,055,863	Rp 20,223,454
20	**Bagian HRD**				
	- Leader	2	S1	Rp 5,778,130	Rp 11,556,259
	- Pelaksana	5	S1	Rp 5,296,619	Rp 26,483,094
21	**Bagian Rumah Tangga**				
	- Cleaning Service	6	SLTA	Rp 4,237,295	Rp 25,423,770
	- Kebersihan Taman	3	SLTA	Rp 4,237,295	Rp 12,711,885

No.	Jabatan	Jumlah	Jenjang Pendidikan (minimum)	Gaji Bulanan	Total
	- Sopir Perusahaan	3	SLTA	Rp 4,959,561	Rp 14,878,684
Manajer Keuangan					
22	Akunting	2	S1	Rp 5,296,619	Rp 10,593,238
23	Perpajakan	2	S1	Rp 5,296,619	Rp 10,593,238
24	Bagian Pengadaaan	2	S2	Rp 6,018,885	Rp 12,037,770
TOTAL		146			Rp 960,132,535

Gaji / tahun (a)	=	Rp 11,521,590,422
Tunjangan Hari Raya (1 bulan gaji pokok)	=	Rp 703,005,768
Tunjangan makan dan transport (5% a)	=	Rp 576,079,521
Tunjangan Kesehatan (2,5% a)	=	Rp 288,039,761
Total Gaji Per Tahun	=	Rp 13,088,715,472

Kemudian ada kenaikan gaji karyakan tiap tahun sekitar 10%.

Tabel L5.4 Kenaikan Gaji Karyawan dalam 10 Tahun

Tahun Ke -	Total Gaji Per Tahun
1	Rp 13,088,715,472
2	Rp 14,397,587,019
3	Rp 15,837,345,721
4	Rp 17,421,080,293
5	Rp 19,163,188,323
6	Rp 21,079,507,155
7	Rp 23,187,457,870
8	Rp 25,506,203,657
9	Rp 28,056,824,023
10	Rp 30,862,506,426

L.5.3 Harga Alat

L.5.3.1 Daftar Harga Alat-Alat

Rumus yang digunakan

;

H2	=	H1 x (I2/I1) x $(K2/K1)^{exp}$		

Dimana ;

H2	=	Harga sekarang		
H1	=	Harga yang diketahui		
I2	=	Index yang akan dicari (2026)	=	762.0
I1	=	Index yang diketahui (2014)	=	576.1
K2	=	Kapasitas sekarang		
K1	=	Kapasitas yang diketahui		
exp	=	eksponen (perry Tabel 9-50 hal 9-69)		
1 m3	=	264.2007026 gallon		
1 $	=	15623 Rupiah		
m3	=	2204.62 pound		

L.5.3.2 Harga Peralatan Utama

*Merujuk pada situs *www.matche.com* dan *www.alibaba.com*

No.	Kode	Nama Alat	Bahan Material	Spesifikasi		Jumlah	Kapasitas		Harga Satuan ($)	Exp	Harga Total ($)	Harga Satuan (Rp)	Harga Total (Rp)
1	T-01	Tangki Etanol 98%	Stainless Steel SA-167 Tipe 304	327.82	m^3	6	86609.51	gal	295,200	0.57	1,771,200	6,100,113,028	36,600,678,166
2	T-01	Tangki Air	Stainless Steel SA-167 Tipe 304	232.94	m^3	3	61543.48	gal	236,400	0.57	709,200	4,885,049,864	14,655,149,591
3	MP-01	Mixer	Stainless steel SA-167 grade 11 Tipe 316	52.98	m^3	1	13998.24	gal	2,400	0.57	2,400	49,594,415	49,594,415
4	T-03	Tangki Etilen	Stainless Steel SA 204 Grade A	51.35	m^3	1	13565.99	gal	2,300	0.57	2,300	47,527,981	47,527,981
5	R-01	Reaktor-01	SS SA 167, Grade 3, type 304	483.15	m^3	1	127649.41	gal	2,370,000	0.56	2,370,000	48,974,484,673	48,974,484,673
6	C-01	Cooler-01	SS, SA-167 Grade 11 Type 316	196.32	ft^2	1	196.32	ft^2	25,500	0.6	25,500	526,940,658	526,940,658
7	C-02	Cooler-02	SS, SA-167 Grade 11 Type 316	2209.08	ft^2	1	2209.08	ft^2	29,500	0.6	29,500	609,598,016	609,598,016

No.	Kode	Nama Alat	Bahan Material	Spesifikasi		Jumlah	Kapasitas		Harga Satuan ($)	Exp	Harga Total ($)	Harga Satuan (Rp)	Harga Total (Rp)
8	H-01	Heater-01	SS, SA-167 Grade 11 Type 316	1226.50	ft^2	1	1226.50	ft^2	115,900	0.6	115,900	2,394,996,951	2,394,996,951
9	V-01	Vaporizer	SS, SA-167 Grade 11 Type 316	809.57	ft^2	1	809.57	ft^2	70,200	0.6	70,200	1,450,636,635	1,450,636,635
10	K-01	Kompresor-01 (air, centrifugal, 125 psi	Carbon Steel (SA-285)	7000.00	HP	1	7000.00	HP	4,466,600	0.69	4,466,600	92,299,338,920	92,299,338,920
11	K-01	Kompresor-02	Carbon Steel (SA-285)	750.00	HP	1	750.00	HP	585,100	0.69	585,100	12,090,705,056	12,090,705,056
12	FD-01	Flash Drum-01	Stainless steel SA 240 grade M Tipe 316	160.11	m^3	1	352991.01	pound	1,680,400	0.62	1,680,400	34,724,356,137	34,724,356,137
13	BW-01	Blower-01	Carbon Steel (SA-285)	0.90	HP	1	0.90	HP	53,500	0.59	53,500	1,105,542,165	1,105,542,165
14	BW-	Blower-02	Carbon Steel (SA-285)	0.19	HP	1	0.19	HP	53,500	0.59	53,500	1,105,542,165	1,105,542,165
15	P-01	Pompa-01 Centrifugal Pump	SS, SA-167 Grade 11 Type 316	3.44	inch	1	3.44	inch	10,300	0.33	10,300	212,842,697	212,842,697

No.	Kode	Nama Alat	Bahan Material	Spesifikasi		Jumlah	Kapasitas		Harga Satuan ($)	Exp	Harga Total ($)	Harga Satuan (Rp)	Harga Total (Rp)
		Horizontal 2 unit,1 stage											
16	P-02	Pompa-02	SS, SA-167 Grade 11 Type 316	3.65	inch	1	3.65	inch	10,300	0.33	10,300	212,842,697	212,842,697
17	P-03	Pompa-03	SS, SA-167 Grade 11 Type 316	4.85	inch	1	4.85	inch	12,600	0.33	12,600	260,370,678	260,370,678
18	P-04	Pompa-04	SS, SA-167 Grade 11 Type 316	4.10	inch	1	4.10	inch	12,600	0.33	12,600	260,370,678	260,370,678
												Total	**138,843,687,535**

Biaya Peralatan Utama		Rp 247,581,518,276
Biaya Pengangkutan + Asuransi	10%	Rp 24,758,151,828
Biaya Administrasi pelabuhan	5%	Rp 12,379,075,914
SubTotal		Rp 284,718,746,017
Bea Masuk	10%	Rp 28,471,874,602
TOTAL		Rp 313,190,620,619

L.5.3.3 Harga Peralatan Penunjang

*Merujuk pada situs *www.matche.com* dan *www.alibaba.com*

No.	Kode	Nama Alat	Bahan Material	Spesifikasi		jumlah	Kapasitas		Harga Satuan ($)	Exp	Harga Total ($)	Harga Satuan (Rp)	Harga Total (Rp)
1	S-01	Screen (SC)	Besi	9.00	ft^2	1	2377.81	gal	194,900	0.92	194,900	4,027,479,773	4,027,479,773
2	BP-01	Reservoir	Beton	1688.30	m^3	1	446048.75	gal	39,800	-	39,800	621,795,400	621,795,400
3	T-04	Bak pengadukan cepat (BPC)	Stainless Steel SA-240 Grade S Type 304	96.85	m^3	1	25587.39	gal	184,000	0.57	184,000	3,802,238,472	3,802,238,472
4	T-05	Bak pengendap I	Stainless Steel SA-240 Grade S Type 304	290.54	m^3	1	76762.16	gal	10,700	0.57	10,700	221,108,433	221,108,433
5	BP-02	Bak Pengendap II	Beton	0.07	m^3	1	19.68	gal	123,900	-	123,900	1,935,689,700	1,935,689,700
6	T-06	Tangki Filtrasi (sand flter)	Stainless Steel SA-240 Grade S Type 304	747.32	m^3	3	747.32	inch	19,300	0.57	57,900	398,821,753	1,196,465,258
7	BP-03	Bak penampung air bersih	Beton	0.31	m^3	1	0.31	inch	6,500	-	6,500	101,549,500	101,549,500

No.	Kode	Nama Alat	Bahan Material	Spesifikasi		jumlah	Kapasitas		Harga Satuan ($)	Exp	Harga Total ($)	Harga Satuan (Rp)	Harga Total (Rp)
8	T-07	Tangki Demineralisasi (ion exchanger)	Stainless Steel SA-240 Grade S Type 304	0.03	m^3	3	0.03	inch	19,300	0.57	57,900	398,821,753	1,196,465,258
9	BP-04	Bak umpan boiler	Beton	66.98	m^3	1	66.98	inch	6,500	-	6,500	101,549,500	101,549,500
10	BP-05	Bak umpan cooling tower	Beton	104.70	m^3	1	104.70	lb/h	242,400	-	242,400	3,787,015,200	3,787,015,200
11	BP-06	Bak penampung air domestik	Beton	1.09	m^3	1	287.42	gal	240,000	-	240,000	3,749,520,000	3,749,520,000
12	BP-07	Bak Penampung Limbah	Beton	83.85	m^3	1	22154.31	gal	150,000	-	150,000	2,343,450,000	2,343,450,000
13	PU-01	Pompa Utilitas-01	SS, SA-167 Grade 11 Type 316	8	in	1	8	in	23,200	0.33	23,200	479,412,677	479,412,677
14	PU-02	Pompa Utilitas-02	SS, SA-167 Grade 11 Type 316	8	in	1	8	in	23,200	0.33	23,200	479,412,677	479,412,677
15	PU-03	Pompa Utilitas-03	SS, SA-167 Grade 11 Type 316	8	in	1	8	in	23,200	0.33	23,200	479,412,677	479,412,677

No.	Kode	Nama Alat	Bahan Material	Spesifikasi		jumlah	Kapasitas		Harga Satuan ($)	Exp	Harga Total ($)	Harga Satuan (Rp)	Harga Total (Rp)
16	PU-04	Pompa Utilitas-04	SS, SA-167 Grade 11 Type 316	8	in	1	8	in	23,200	0.33	23,200	479,412,677	479,412,677
17	PU-05	Pompa Utilitas-05	SS, SA-167 Grade 11 Type 316	8	in	1	8	in	23,200	0.33	23,200	479,412,677	479,412,677
18	PU-06	Pompa Utilitas-06	SS, SA-167 Grade 11 Type 316	8	in	1	8	in	23,200	0.33	23,200	479,412,677	479,412,677
19	PU-07	Pompa Utilitas-07	SS, SA-167 Grade 11 Type 316	8	in	1	8	in	23,200	0.33	23,200	479,412,677	479,412,677
20	PU-08	Pompa Utilitas-08	SS, SA-167 Grade 11 Type 316	8	in	1	8	in	23,200	0.33	23,200	479,412,677	479,412,677
21	PU-09	Pompa Utilitas-09	SS, SA-167 Grade 11 Type 316	8	in	1	8	in	23,200	0.33	23,200	479,412,677	479,412,677
22	PU-10	Pompa Utilitas-10	SS, SA-167 Grade 11 Type 316	6	in	1	6	in	19,500	0.33	19,500	402,954,621	402,954,621

No.	Kode	Nama Alat	Bahan Material	Spesifikasi		jumlah	Kapasitas		Harga Satuan ($)	Exp	Harga Total ($)	Harga Satuan (Rp)	Harga Total (Rp)
23	PU-11	Pompa Utilitas-11	SS, SA-167 Grade 11 Type 316	1	in	1	1	in	6,500	0.33	6,500	134,318,207	134,318,207
24	PU-12	Pompa Utilitas-12	SS, SA-167 Grade 11 Type 316	8	in	1	8	in	23,200	0.33	23,200	479,412,677	479,412,677
25	T-08	Tangki Bahan Bakar	Carbon steel SA - 283 grade C	1805	m3	1	476891	gal	123,900	0.57	123,900	2,560,311,667	2,560,311,667
26	G-01	Generator		1600	kW	5	1600	kW	4,750	-	23,750	371,046,250	1,855,231,250
27	TE-01	Cooling Tower		281	m3	1	74314	gal	10,000	-	10,000	156,230,000	156,230,000
28	B-01	Boiler		213	m3	1	56287	gal	10,000	-	10,000	156,230,000	156,230,000
29	JJ-01	Fan Cooling Tower		13	HP	1	13	HP	3,000	-	3,000	46,869,000	46,869,000

No.	Kode	Nama Alat	Bahan Material	Spesifikasi		jumlah	Kapasitas		Harga Satuan ($)	Exp	Harga Total ($)	Harga Satuan (Rp)	Harga Total (Rp)
30		Dowtherm		516469	lb/h	1	516469	lb/h	400	-	400	6,249,200	6,249,200
31		Kendaraaan		464536	lb/h	4	464536	lb/h	10,000	-	40,000	624,920,000	2,499,680,000
32		Forklift	LTMG EPA Carretilla Elevadora with cushion tire 5 Ton	-	-	10	-	-	4,500	-	45,000	703,035,000	7,030,350,000
33		Chiller				1			750		750	11,717,250	11,717,250
34		Refrigerant				1			5		5	78,115	78,115
35		Truk operasional/mobil tangki	Cargo van truck	-	-	5	-	-	12,000	-	60,000	937,380,000	4,686,900,000
												Total	47,425,572,575

* Dari matche harga alat merupakan harga FOB dan sudah pada kapasitas yang dikehendaki sehingga K1=K2

Biaya Peralatan Utama		Rp 47,425,572,575
Biaya Pengangkutan + Asuransi	10%	Rp 4,742,557,257
Biaya Administrasi pelabuhan	5%	Rp 2,371,278,629
SubTotal		Rp 54,539,408,461
Bea Masuk	10%	Rp 5,453,940,846
Total		Rp 59,993,349,307
Total Alat Utama & Alat penunjang	=	**Rp 373,183,969,926**
Total Alat utama & penunjang, tanpa kendaraan	=	Rp 358,967,039,926

L.5.4 Total Modal Investasi

L.5.5 Struktur Permodalan

Yang dapat dijaminkan	=		DFCI =	Rp	846,599,167,401
Jika bank memberikan pinjaman sebesar	=	75 %	DFCI =	Rp	634,949,375,551
			=	Rp	635,000,000,000

Besar pinjaman dari bank yang diambil sebesar	=	Rp	635,000,000,000
Modal sendiri (TCI - Pinjaman Bank)	=	Rp	11,331,367,511,573

Sehingga komposisi permodalan adalah ;

Modal sendiri = (TCI - pinjaman bank)/TCI x 100%	=	**94,7%**
Pinjaman Bank = pinjaman bank/TCI x 100 %	=	**5.3%**

- Jangka waktu pinjaman 5 tahun
- *Grace Priode* 1 tahun
- Bunga bank per tahun (diasumsikan tetap selama 5 tahun) = 9.95%

Menentukan Bunga Pinjaman dan Sisa Pinjaman

Tahun	Pokok Pinjaman	Angsuran Pokok	Bunga	Jumlah	Sisa
0	Rp 635,000,000,000	-	Rp 63,182,500,000	Rp 63,182,500,000	Rp 635,000,000,000
1	Rp 635,000,000,000	Rp 158,750,000,000	Rp 63,182,500,000	Rp 221,932,500,000	Rp 476,250,000,000
2	Rp 476,250,000,000	Rp 158,750,000,000	Rp 47,386,875,000	Rp 206,136,875,000	Rp 317,500,000,000
3	Rp 317,500,000,000	Rp 158,750,000,000	Rp 31,591,250,000	Rp 190,341,250,000	Rp 158,750,000,000
4	Rp 158,750,000,000	Rp 158,750,000,000	Rp 15,795,625,000	Rp 174,545,625,000	Rp-

		Rp 635,000,000,000	Rp 221,138,750,000	Rp 856,138,750,000	
			Rp 856,138,750,000		

L.5.6 Produk dan Bahan Baku

L.5.6.1 Hasil Penjualan Produk Per Tahun

Hasil produksi	=		85000000.00	kg/th
Harga produk	=	Rp	274,483.61	/kg
Hasil penjualan produk per tahun	=	Rp	23,331,106,986,000.00	

harga paling rendah source : www.alibaba.com	
Tahun	Harga ($)
2023	12.00
2024	13.20
2025	14.52
2026	15.97
2027	17.57

Tabel L5.5 Total Penjualan Produk dalam 10 Tahun

Tahun	Kapasitas Produksi	Hasil penjualan produksi (Total Sales)
1	80%	Rp 18,664,885,588,800
2	90%	Rp 23,097,795,916,140
3	100%	Rp 28,230,639,453,060
4	100%	Rp 31,053,703,398,366
5	100%	Rp 34,159,073,738,203
6	100%	Rp 37,574,981,112,023
7	100%	Rp 41,332,479,223,225
8	100%	Rp 45,465,727,145,548

9	100%	Rp 50,012,299,860,102
10	100%	Rp 55,013,529,846,113

Keterangan : Terjadi kenaikan harga produk sebesar 10 % pertahun

L.5.6.2 Perhitungan Biaya Bahan Baku dan Bahan Penunjang

a. Persediaan Bahan Baku

Tahun Pertama			
Bahan Baku	**Kebutuhan (Kg/jam)**	**Harga (Rp/Kg)**	**Biaya**
Etanol 96%(kg)	17802.1535	Rp18,299	Rp 2,814,566,051,679
Katalis ZSM-5 (kg)	161880.7022	Rp22,874	Rp 3,702,799,983
Total Biaya Bahan Baku		a	Rp 2,818,268,851,662

Total pembelian bahan baku tahun pertama = Rp 2,818,268,851,662

Tabel L5.6 Biaya Bahan Baku dalam 10 Tahun

Tahun	Kapasitas Produksi	Biaya Bahan Baku
1	80%	Rp 2,254,615,081,330
2	90%	Rp 2,790,086,163,146
3	100%	Rp 3,410,105,310,511
4	100%	Rp 3,751,115,841,563
5	100%	Rp 4,126,227,425,719
6	100%	Rp 4,538,850,168,291
7	100%	Rp 4,992,735,185,120
8	100%	Rp 5,492,008,703,632
9	100%	Rp 6,041,209,573,995
10	100%	Rp 6,645,330,531,395

Keterangan : Terjadi kenaikan harga bahan baku sebesar 10 % pertahun

b. Bahan Penunjang

Tahun Pertama			
Komponen	**Kebutuhan**	**Harga/satuan**	**Biaya**
1. Solar (lt/jam)	249563.888	Rp 9,955.88	Rp 819,927,279,655
2. Listrik PLN (kwh)	9751.561	Rp 1,115.00	Rp 86,114,083,695
Abonemen/bulan	3.0000	Rp 312,049,947.71	Rp 3,744,599,373
3. Air Keseluruhan	275476.4456	Rp 4,831.53	Rp 57,978
4. Downtherm	234269.0023	Rp 91,494.54	Rp 1,097,934
5. Resin Ion Exchanger	5.7097	Rp 22,873.63	Rp 274,484
Total persediaan bahan penunjang	B		**Rp 909,787,393,119**

Tabel L5.7 Biaya Penunjang dalam 10 Tahun

Tahun	**Kapasitas Produksi**	**Bio Solar**	**Listrik *variabel cost***	**Abonemen**	**Air Keseluruhan**	***Downtherm***	**Resin Ion Exchanger**
1	80%	655,941,823,724	68,891,266,956	3,744,599,373	46,383	878,348	219,587
2	90%	811,728,006,859	85,252,942,858	4,119,059,310	57,399	1,086,955	271,739
3	100%	992,112,008,383	104,198,041,271	4,530,965,241	70,154	1,328,501	332,125
4	100%	1,091,323,209,221	114,617,845,398	4,984,061,765	77,169	1,461,351	365,338
5	100%	1,200,455,530,144	126,079,629,938	5,482,467,941	84,886	1,607,486	401,871
6	100%	1,320,501,083,158	138,687,592,931	6,030,714,735	93,375	1,768,234	442,059
7	100%	1,452,551,191,474	152,556,352,224	6,633,786,209	102,712	1,945,058	486,264
8	100%	1,597,806,310,621	167,811,987,447	7,297,164,830	112,983	2,139,564	534,891
9	100%	1,757,586,941,683	184,593,186,192	8,026,881,313	124,282	2,353,520	588,380

Tahun	Kapasitas Produksi	Bio Solar	Listrik *variabel cost*	Abonemen	Air Keseluruhan	*Downtherm*	Resin Ion Exchanger
10	100%	1,933,345,635,852	203,052,504,811	8,829,569,444	136,710	2,588,872	647,218

Keterangan : Terjadi kenaikan biaya penunjang sebesar 10 % pertahun

L.5.7 Salvage Value (Nilai Aset) dan Depresiasi

L.5.7.1 Salvage Value (Nilai Aset)

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

a. Kendaraan (mobil dan forklift)

= 10% x Rp 7,030,350,000

= Rp 703,035,000

b. DFCI selain kendaraan, bangunan, dan tanah

= 10% x Rp 531,044,723,590

= Rp. 53,104,472,359

c. Bangunan

= 10% x Rp 22,391,038,196

= Rp. 2,239,103,820

Catatan: Tanah tidak didepresiasi. Pada akhir tahun ke-10 harga tanah tetap : Rp 37,500,000,000

Sehingga total nilai salvage value yang akan diperhitungkan pada akhir tahun ke-10 : Rp 93,546,611,179

L.5.7.2 Depresiasi

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

Rumus

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

a. 8 tahun atau 12.5% / tahun untuk kendaraan

b. 16 Tahun atau 6.25% /tahun untuk mesin-mesin industri kimia

HP = Harga perolehan

NS = Nilai sisa

c. 20 tahun atau 5% / tahun untuk bangunan n = taksir umur kegunaan

d. 8 tahun atau 12.5% / tahun untuk IFCI tanpa salvage value (amortisasi)

Tabel L5.8 Depresiasi

Tahun	Kendaraan	DFCI tanpa tanah, bangunan &kendaraan	Bangunan	Nilai depresiasi IFCI	Jumlah Nilai Depresiasi
1	Rp 790,914,375	Rp 29,871,265,702	Rp 1,007,596,719	Rp 46,986,253,791	Rp 78,656,030,587
2	Rp 790,914,375	Rp 29,871,265,702	Rp 1,007,596,719	Rp 46,986,253,791	Rp 78,656,030,587
3	Rp 790,914,375	Rp 29,871,265,702	Rp 1,007,596,719	Rp 46,986,253,791	Rp 78,656,030,587
4	Rp 790,914,375	Rp 29,871,265,702	Rp 1,007,596,719	Rp 46,986,253,791	Rp 78,656,030,587
5	Rp 790,914,375	Rp 29,871,265,702	Rp 1,007,596,719	Rp 46,986,253,791	Rp 78,656,030,587
6	Rp 790,914,375	Rp 29,871,265,702	Rp 1,007,596,719	Rp 46,986,253,791	Rp 78,656,030,587
7	Rp 790,914,375	Rp 29,871,265,702	Rp 1,007,596,719	Rp 46,986,253,791	Rp 78,656,030,587
8	Rp 790,914,375	Rp 29,871,265,702	Rp 1,007,596,719	Rp 46,986,253,791	Rp 78,656,030,587
9	-	Rp 29,871,265,702	Rp 1,007,596,719	-	Rp 30,878,862,421
10	-	Rp 29,871,265,702	Rp 1,007,596,719	-	Rp 30,878,862,421
Total					Rp 691,005,969,534
11	-	Rp 29,871,265,702	Rp 1,007,596,719	-	Rp 30,878,862,421
12	-	Rp 29,871,265,702	Rp 1,007,596,719	-	Rp 30,878,862,421

Tahun	Kendaraan	DFCI tanpa tanah, bangunan &kendaraan	Bangunan	Nilai depresiasi IFCI	Jumlah Nilai Depresiasi
13	-	Rp 29,871,265,702	Rp 1,007,596,719	-	Rp 30,878,862,421
14	-	Rp 29,871,265,702	Rp 1,007,596,719	-	Rp 30,878,862,421
15	-	Rp 29,871,265,702	Rp 1,007,596,719	-	Rp 30,878,862,421
16	-	Rp 29,871,265,702	Rp 1,007,596,719	-	Rp 30,878,862,421
17	-	-	Rp 1,007,596,719	-	Rp 1,007,596,719
18	-	-	Rp 1,007,596,719	-	Rp 1,007,596,719
19	-	-	Rp 1,007,596,719	-	Rp 1,007,596,719
20	-	-	Rp 1,007,596,719	-	Rp 1,007,596,719
Total					Rp 189,303,561,400

L.5.8 Perhitungan Biaya Produksi Total (TPC)

Keterangan : biaya produksi dihitung per tahun operasi pabrik (330 hari)

Pajak Bumi dan Bangunan

Nilai Kena Pajak PBB = 20% x (Harga Tanah dan Bangunan)

= Rp 11,978,207,639

Besarnya PBB = 0,5% x Nilai Kena Pajak

= Rp 59,891,038

Tabel L5.9 Perhitungan Biaya Produksi Total dalam 10 Tahun

TAHUN				I		II	
KAPASITAS PRODUKSI				80%		90%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
A.	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
a.	Biaya Bahan Baku				Rp 2,254,615,081,330		Rp 2,790,086,163,146
b.	Gaji Karyawan			Rp 13,088,715,472		Rp 14,397,587,019	
c.	Biaya Pemeliharaan dan Perbaikan (kenaikan 5% per tahun)	2%	DFCI	Rp 16,931,983,348		Rp 17,778,582,515	
d.	Biaya Royalti dan Paten	0,5%	TS		Rp 93,324,427,944		Rp 115,488,979,581
e.	Biaya Laboratorium	0,5%	BB		Rp 11,273,075,407		Rp 13,950,430,816
f.	Biaya pengemasan produk	0,5%	BB		Rp 11,273,075,407		Rp 13,950,430,816
g.	Biaya sarana penunjang			Rp 3,744,599,373	Rp 724,833,090,680	Rp 4,119,059,310	Rp 896,980,949,717
h.	Biaya Start Up			Rp 9,841,271,432,647		Rp 10,825,398,575,912	
	Total Biaya Manufacturing Langsung (DMC)			**Rp 9,875,036,730,839**	**Rp 3,095,318,750,767**	**Rp 10,861,693,804,756**	**Rp 3,830,456,954,075**
	Biaya Plant Overhead	20%	(b+c)	**Rp 6,004,139,764**		**Rp 6,435,233,907**	

TAHUN				I		II	
KAPASITAS PRODUKSI				80%		90%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
	Biaya Manufacturing Tetap (FMC)						
2.	**Depresiasi**			Rp 78,656,030,587		Rp 78,656,030,587	
3.	Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th	0,1%		Rp 59,891,038		Rp 65,880,142	
a.	Biaya asuransi (kenaikan 10 %) pertahun	0,5%	DFCI	Rp 4,232,995,837		Rp 4,656,295,421	
	Total Biaya Manufacturing Tetap (FMC)			**Rp 82,948,917,462**		**Rp 83,378,206,149**	
B.	**Pengeluaran Umum (General Expenses)**						
a.	Biaya administrasi	5%	b	Rp 654,435,774		Rp 719,879,351	
b.	Biaya distribusi dan penjualan	10%	f		Rp 1,127,307,541		Rp 1,395,043,082
c.	Bunga Bank + Cicilan Pokok			Rp 221,932,500,000		Rp 206,136,875,000	
	Total Pengeluaran Umum			**Rp 222,586,935,774**	**Rp 1,127,307,541**	**Rp 206,856,754,351**	**Rp 1,395,043,082**
Total Biaya				**Rp 10,186,576,723,839**	**Rp 3,096,446,058,308**	**Rp 11,158,363,999,163**	**Rp 3,831,851,997,156**
Total Biaya Produksi (TPC)				**Rp 13,283,022,782,147**		**Rp 14,990,215,996,319**	

TAHUN				III		IV	
KAPASITAS PRODUKSI				100%		100%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
A	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
a.	Biaya Bahan Baku				Rp 3,410,105,310,511		Rp 3,751,115,841,563
b.	Gaji Karyawan			Rp 15,837,345,721		Rp 17,421,080,293	
c.	Biaya Pemeliharaan dan Perbaikan (kenaikan 5% per tahun)	2%	DFCI	Rp 18,667,511,641		Rp 19,600,887,223	
d.	Biaya Royalti dan Paten	0,5%	TS		Rp 141,153,197,265		Rp 155,268,516,992
e.	Biaya Laboratorium	0,5%	BB		Rp 17,050,526,553		Rp 18,755,579,208
f.	Biaya pengemasan produk	0,5%	BB		Rp 17,050,526,553		Rp 18,755,579,208
g.	Biaya sarana penunjang			Rp 4,530,965,241	Rp 1,096,310,049,654	Rp 4,984,061,765	Rp 1,205,941,054,619
h.	Biaya Start Up			Rp 11,907,938,433,503		Rp 13,098,732,276,853	
	Total Biaya Manufacturing Langsung (DMC)			**Rp 11,946,974,256,106**	**Rp 4,681,669,610,536**	**Rp 13,140,738,306,134**	**Rp 5,149,836,571,589**
	Biaya Plant Overhead	20%	(b+c)	**Rp 6,900,971,472**		**Rp 7,404,393,503**	

TAHUN				III		IV	
KAPASITAS PRODUKSI				100%		100%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
	Biaya Manufacturing Tetap (FMC)						
2.	**Depresiasi**			Rp7 8,656,030,587		Rp 78,656,030,587	
3.	Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th	0,1%		Rp 72,468,156		Rp 79,714,972	
a.	Biaya asuransi (kenaikan 10 %) pertahun	0,5%	DFCI	Rp5,121,924,963		Rp 5,634,117,459	
	Total Biaya Manufacturing Tetap (FMC)			**Rp 83,850,423,706**		**Rp 84,369,863,017**	
B.	**Pengeluaran Umum (General Expenses)**						
a.	Biaya administrasi	5%	b	Rp 791,867,286		Rp 871,054,015	
b.	Biaya distribusi dan penjualan	10%	f		Rp 1,705,052,655		Rp 1,875,557,921
c.	Bunga Bank + Cicilan Pokok			Rp 190,341,250,000		Rp 174,545,625,000	
	Total Pengeluaran Umum			**Rp 191,133,117,286**	**Rp 1,705,052,655**	**Rp 175,416,679,015**	**Rp 1,875,557,921**
Total Biaya				**Rp 12,228,858,768,570**	**Rp 4,683,374,663,191**	**Rp 13,407,929,241,670**	**Rp 5,151,712,129,510**
Total Biaya Produksi (TPC)				**Rp 16,912,233,431,761**		**Rp 18,559,641,371,180**	

TAHUN				V		VI	
KAPASITAS PRODUKSI				100%		100%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
A.	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
a.	Biaya Bahan Baku				Rp 4,126,227,425,719		Rp 4,538,850,168,291
b.	Gaji Karyawan			Rp 19,163,188,323		Rp 21,079,507,155	
c.	Biaya Pemeliharaan dan Perbaikan (kenaikan 5% per tahun)	2%	DFCI	Rp 20,580,931,584		Rp 21,609,978,164	
d.	Biaya Royalti dan Paten	0,5%	TS		Rp 170,795,368,691		Rp 187,874,905,560
e.	Biaya Laboratorium	0,5%	BB		Rp 20,631,137,129		Rp 22,694,250,841
f.	Biaya pengemasan produk	0,5%	BB		Rp 20,631,137,129		Rp 22,694,250,841
g.	Biaya sarana penunjang			Rp 5,482,467,941	Rp 1,326,535,160,081	Rp 6,030,714,735	Rp 1,459,188,676,089
h.	Biaya Start Up			Rp 14,408,605,504,538		Rp 15,849,466,054,992	
	Total Biaya Manufacturing Langsung (DMC)			**Rp 14,453,832,092,387**	**Rp 5,664,820,228,748**	**Rp 15,898,186,255,046**	**Rp 6,231,302,251,623**
	Biaya Plant Overhead	20%	(b+c)	**Rp 7,948,823,981**		**Rp 8,537,897,064**	

TAHUN				V		VI	
KAPASITAS PRODUKSI				100%		100%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
	Biaya Manufacturing Tetap (FMC)						
2.	**Depresiasi**			Rp 78,656,030,587		Rp 78,656,030,587	
3.	Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th	0,1%		Rp 87,686,469		Rp 96,455,116	
a.	Biaya asuransi (kenaikan 10 %) pertahun	0,5%	DFCI	Rp 6,197,529,205		Rp 6,817,282,125	
	Total Biaya Manufacturing Tetap (FMC)			**Rp 84,941,246,260**		**Rp 85,569,767,828**	
B.	**Pengeluaran Umum (General Expenses)**						
a.	Biaya administrasi	5%	b	Rp 958,159,416		Rp 1,053,975,358	
b.	Biaya distribusi dan penjualan	10%	f		Rp 2,063,113,713		Rp 2,269,425,084
c.	Bunga Bank + Cicilan Pokok						
	Total Pengeluaran Umum			**Rp 958,159,416**	**Rp 2,063,113,713**	**Rp 1,053,975,358**	**Rp 2,269,425,084**
Total Biaya				**Rp 14,547,680,322,045**	**Rp 5,666,883,342,461**	**Rp 15,993,347,895,296**	**Rp 6,233,571,676,707**
Total Biaya Produksi (TPC)				**Rp 20,214,563,664,506**		**Rp 22,226,919,572,003**	

TAHUN				VII		VIII	
KAPASITAS PRODUKSI				100%		100%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
A	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
a.	Biaya Bahan Baku				Rp 4,992,735,185,120		Rp 5,492,008,703,632
b.	Gaji Karyawan			Rp 23,187,457,870		Rp 25,506,203,657	
c.	Biaya Pemeliharaan dan Perbaikan (kenaikan 5% per tahun)	2%	DFCI	Rp 22,690,477,072		Rp 23,825,000,925	
d.	Biaya Royalti dan Paten	0,5%	TS		Rp 206,662,396,116		Rp 227,328,635,728
e.	Biaya Laboratorium	0,5%	BB		Rp 24,963,675,926		Rp 27,460,043,518
f.	Biaya pengemasan produk	0,5%	BB		Rp 24,963,675,926		Rp 27,460,043,518
g.	Biaya sarana penunjang			Rp 6,633,786,209	Rp 1,605,107,543,698	Rp 7,297,164,830	Rp 1,765,618,298,068
h.	Biaya Start Up			Rp 17,434,412,660,491		Rp 19,177,853,926,541	
	Total Biaya Manufacturing Langsung (DMC)			**Rp 17,486,924,381,643**	**Rp 6,854,432,476,785**	**Rp 19,234,482,295,953**	**Rp 7,539,875,724,464**
	Biaya Plant Overhead	20%	(b+c)	**Rp 9,175,586,988**		**Rp 9,866,240,917**	

TAHUN				VII		VIII	
KAPASITAS PRODUKSI				100%		100%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
	Biaya Manufacturing Tetap (FMC)						
2.	**Depresiasi**			Rp 78,656,030,587		Rp 78,656,030,587	
3.	Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th	0,1%		Rp 106,100,628		Rp 116,710,690	
a.	Biaya asuransi (kenaikan 10 %) pertahun	0,5%	DFCI	Rp 7,499,010,338		Rp 8,248,911,372	
	Total Biaya Manufacturing Tetap (FMC)			**Rp 86,261,141,552**		**Rp 87,021,652,649**	
B.	**Pengeluaran Umum (General Expenses)**						
a.	Biaya administrasi	5%	b	Rp 1,159,372,894		Rp 1,275,310,183	
b.	Biaya distribusi dan penjualan	10%	f		Rp 2,496,367,593		Rp 2,746,004,352
c.	Bunga Bank + Cicilan Pokok						
	Total Pengeluaran Umum			**Rp 1,159,372,894**	**Rp 2,496,367,593**	**Rp 1,275,310,183**	**Rp 2,746,004,352**
Total Biaya				**Rp 17,583,520,483,077**	**Rp 6,856,928,844,378**	**Rp 19,332,645,499,701**	**Rp 7,542,621,728,816**
Total Biaya Produksi (TPC)				**Rp 24,440,449,327,455**		**Rp 26,875,267,228,517**	

	TAHUN			IX		X	
	KAPASITAS PRODUKSI			100%		100%	
	BIAYA PRODUKSI (PRODUCT COST)			Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
A.	**Biaya Manufacturing (Manufacturing Cost)**						
1.	**Biaya Manufacturing Langsung (DMC)**						
a.	Biaya Bahan Baku				Rp 6,041,209,573,995		Rp 6,645,330,531,395
b.	Gaji Karyawan			Rp 28,056,824,023		Rp 30,862,506,426	
c.	Biaya Pemeliharaan dan Perbaikan (kenaikan 5% per tahun)	2%	DFCI	Rp 25,016,250,972		Rp 26,267,063,520	
d.	Biaya Royalti dan Paten	0,5%	TS		Rp 250,061,499,301		Rp 138,231,653,773
e.	Biaya Laboratorium	0,5%	BB		Rp 30,206,047,870		Rp 33,226,652,657
f.	Biaya pengemasan produk	0,5%	BB		Rp 30,206,047,870		Rp 33,226,652,657
g.	Biaya sarana penunjang			Rp 8,026,881,313	Rp 1,942,180,127,875	Rp 8,829,569,444	Rp 2,136,398,140,662
h.	Biaya Start Up			Rp 21,095,639,319,195		Rp 23,205,203,251,114	
	Total Biaya Manufacturing Langsung (DMC)			**Rp 21,156,739,275,502**	**Rp 8,293,863,296,910**	**Rp 23,271,162,390,504**	**Rp 8,986,413,631,143**
	Biaya Plant Overhead	20%	(b+c)	**Rp 10,614,614,999**		**Rp 11,425,913,989**	

TAHUN				IX		X	
KAPASITAS PRODUKSI				100%		100%	
BIAYA PRODUKSI (PRODUCT COST)				Fixed Cost	Variable Cost	Fixed Cost	Variable Cost
	Biaya Manufacturing Tetap (FMC)						
2.	**Depresiasi**			Rp 30,878,862,421		Rp 30,878,862,421	
3.	Pajak Bumi dan Bangunan diperkirakan 0.1 % x (tanah + bangunan),kenaikan 10 % /th	0,1%		Rp 128,381,759		Rp 141,219,935	
a.	Biaya asuransi (kenaikan 10 %) pertahun	0,5%	DFCI	Rp 9,073,802,509		Rp 9,981,182,760	
	Total Biaya Manufacturing Tetap (FMC)			**Rp 40,081,046,689**		**Rp 41,001,265,116**	
B.	**Pengeluaran Umum (General Expenses)**						
a.	Biaya administrasi	5%	b	Rp 1,402,841,201		Rp 1,543,125,321	
b.	Biaya distribusi dan penjualan	10%	f		Rp 3,020,604,787		Rp 3,322,665,266
c.	Bunga Bank + Cicilan Pokok						
	Total Pengeluaran Umum			**Rp 1,402,841,201**	**Rp 3,020,604,787**	**Rp 1,543,125,321**	**Rp 3,322,665,266**
Total Biaya				**Rp 21,208,837,778,391**	**Rp 8,296,883,901,697**	**Rp 23,325,132,694,930**	**Rp 8,989,736,296,409**
Total Biaya Produksi (TPC)				**Rp 29,505,721,680,089**		**Rp 32,314,868,991,339**	

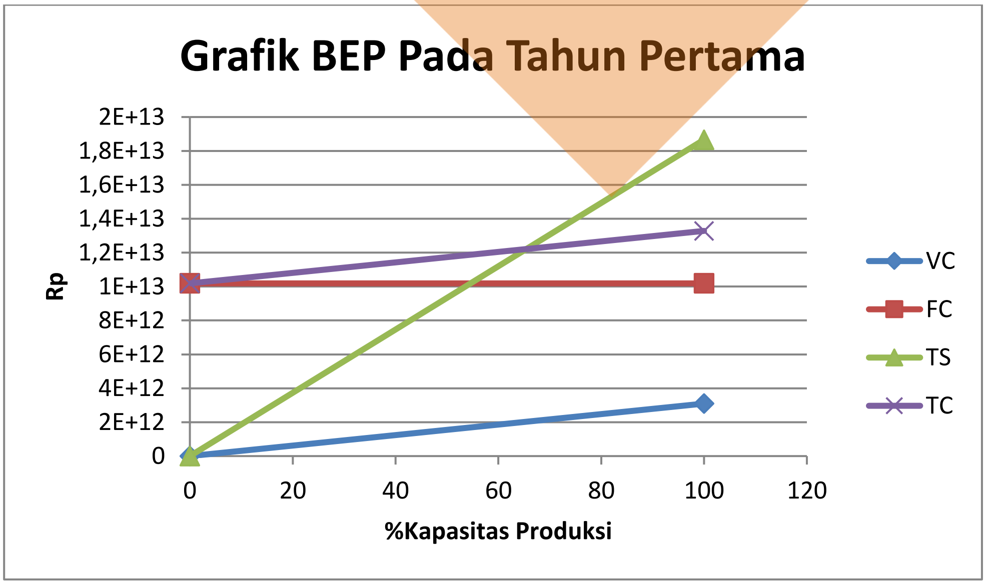
L.5.9 *Break Even Point* (BEP)

Rumus Umum :

$$\text{BEP} = \frac{Total\ Fixed\ Cost}{(Total\ Sales - Total\ Variable\ Cost)}\ x\ 100\%$$

Tabel L5.10 Nilai *Break Even Point* (BEP)

Tahun	Total			BEP
	Fixed Cost (Rp)	Variabel Cost (Rp)	Penjualan (Rp)	(%)
1	Rp 10,186,576,723,839	Rp 3,096,446,058,308	Rp 18,664,885,588,800	65.43
2	Rp 11,158,363,999,163	Rp 3,831,851,997,156	Rp 23,097,795,916,140	57.92
3	Rp 12,228,858,768,570	Rp 4,683,374,663,191	Rp 28,230,639,453,060	51.93
4	Rp 13,407,929,241,670	Rp 5,151,712,129,510	Rp 31,053,703,398,366	51.76
5	Rp 14,547,680,322,045	Rp 5,666,883,342,461	Rp 34,159,073,738,203	51.06
6	Rp 15,993,347,895,296	Rp 6,233,571,676,707	Rp 37,574,981,112,023	51.03
7	Rp 17,583,520,483,077	Rp 6,856,928,844,378	Rp 41,332,479,223,225	51.00
8	Rp 19,332,645,499,701	Rp 7,542,621,728,816	Rp 45,465,727,145,548	50.98
9	Rp 21,208,837,778,391	Rp 8,296,883,901,697	Rp 50,012,299,860,102	50.84
10	Rp 23,325,132,694,930	Rp 8,989,736,296,409	Rp 55,013,529,846,113	50.68



Keterangan :

VC = *Variable Cost*

FC = *Fixed Cost*

TS = Total *Sales*

TC = Total *Cost*

L.5.10 Laba Rugi dan Pajak

Perhitungan laba rugi dan pajak dihitung berdasarkan UU No. 36 Tahun 2008, sebagai berikut:

1 Badan usaha yang memiliki pendapatan bruto sampai dengan 4.8 M/Tahun, dikenakan tariff pajak PPh final sebesar 1% dari seluruh pendapatan bruto

2 Bdan usaha yang memiliki pendapatan bruto antara 4.8 M/Tahun sampai dengan 50 M/tahun dikenakan 2 tarif pajak perhitungan sebagai berikut:

a 12.5% x penghasilan pajak yang mendapat fasilitas (pendapatan bruto antara 4.8 M sampai dengan 50 M)

b 25% x penghasilan pajak yang tidak mendapatkan fasilitas (pendapatan bruto antara 4.8 M sampai dengan 50 M)

3 Bdan usaha yang memiliki pendapatan bruto diatas 50 M/Tahun, dikenakan tariff pajak PPh 25% x laba sebelum pajak.

Tabel mengenai laba rugi dan pajak dapat dilihat pada tabel sebagai berikut:

Tabel L5.11 Laba Rugi dan Pajak dalam 10 Tahun

Tahun	Penjualan	Pengeluaran	Laba sebelum pajak	PPH 25%	Laba setelah pajak
1	Rp 18,664,885,588,800	Rp 13,283,022,782,147	Rp 5,381,862,806,653	Rp 1,345,465,701,663	Rp 4,036,397,104,990
2	Rp 23,097,795,916,140	Rp 14,990,215,996,319	Rp 8,107,579,919,821	Rp 2,026,894,979,955	Rp 6,080,684,939,865
3	Rp 28,230,639,453,060	Rp 16,912,233,431,761	Rp 11,318,406,021,299	Rp 2,829,601,505,325	Rp 8,488,804,515,974
4	Rp 31,053,703,398,366	Rp 18,559,641,371,180	Rp 12,494,062,027,186	Rp 3,123,515,506,797	Rp 9,370,546,520,390
5	Rp 34,159,073,738,203	Rp 20,214,563,664,506	Rp 13,944,510,073,697	Rp 3,486,127,518,424	Rp 10,458,382,555,273
6	Rp 37,574,981,112,023	Rp 22,226,919,572,003	Rp 15,348,061,540,020	Rp 3,837,015,385,005	Rp 11,511,046,155,015
7	Rp 41,332,479,223,225	Rp 24,440,449,327,455	Rp 16,892,029,895,771	Rp 4,223,007,473,943	Rp 12,669,022,421,828
8	Rp 45,465,727,145,548	Rp 26,875,267,228,517	Rp 18,590,459,917,031	Rp 4,647,614,979,258	Rp 13,942,844,937,773
9	Rp 50,012,299,860,102	Rp 29,505,721,680,089	Rp 20,506,578,180,014	Rp 5,126,644,545,003	Rp 15,379,933,635,010
10	Rp 55,013,529,846,113	Rp 32,314,868,991,339	Rp 22,698,660,854,774	Rp 5,674,665,213,693	Rp 17,023,995,641,080

Jumlah nominal aliran masuk = Laba setelah pajak + depresiasi + *salvage value*

Tabel L5.12 Jumlah Nominal Aliran Masuk

Tahun	Laba setelah pajak	Depresiasi	Salvage value+ tanah	Cash in Nominal
1	Rp 4,036,397,104,990	Rp 78,656,030,587	Rp-	Rp 4,115,053,135,576
2	Rp 6,080,684,939,865	Rp 78,656,030,587	Rp-	Rp 6,159,340,970,452
3	Rp 8,488,804,515,974	Rp 78,656,030,587	Rp-	Rp 8,567,460,546,561
4	Rp 9,370,546,520,390	Rp 78,656,030,587	Rp-	Rp 9,449,202,550,976
5	Rp 10,458,382,555,273	Rp 78,656,030,587	Rp 703,035,000	Rp 10,537,741,620,859
6	Rp 11,511,046,155,015	Rp 78,656,030,587	Rp-	Rp 11,589,702,185,602
7	Rp 12,669,022,421,828	Rp 78,656,030,587	Rp-	Rp 12,747,678,452,415
8	Rp 13,942,844,937,773	Rp 78,656,030,587	Rp-	Rp 14,021,500,968,360
9	Rp 15,379,933,635,010	Rp 30,878,862,421	Rp-	Rp 15,410,812,497,431
10	Rp 17,023,995,641,080	Rp 30,878,862,421	Rp 93,546,611,179	Rp 17,148,421,114,679

L.5.11 *Minimum Payback Period* (MPP)

Jangka waktu minimum oengembalian investasi modal sebagai berikut:

Keterangan : Suku bunga tahunan 7.90%

Tabel L5.13 ***Minimum Payback Period*** **(MPP)**

Tahun	NCF nominal (Rp)	Faktor Discount 1/(1+0.0995)^n	NCF PV (Rp)	Akumulasi
0	-Rp 11,966,367,511,573	1.00	-Rp 11,966,367,511,573	-Rp 11,966,367,511,573
1	Rp 4,115,053,135,576	0.93	Rp 3,813,765,649,283	-Rp 8,152,601,862,290
2	Rp 6,159,340,970,452	0.86	Rp 5,290,434,687,021	-Rp 2,862,167,175,269
3	Rp 8,567,460,546,561	0.80	Rp 6,820,053,462,604	Rp 3,957,886,287,335
4	Rp 9,449,202,550,976	0.74	Rp 6,971,229,490,468	Rp 10,929,115,777,803
5	Rp 10,537,741,620,859	0.68	Rp 7,205,105,135,064	Rp 18,134,220,912,868
6	Rp 11,589,702,185,602	0.63	Rp 7,344,184,999,458	Rp 25,478,405,912,326
7	Rp 12,747,678,452,415	0.59	Rp 7,486,537,210,500	Rp 32,964,943,122,826
8	Rp 14,021,500,968,360	0.54	Rp 7,631,729,149,753	Rp 40,596,672,272,579
9	Rp 15,410,812,497,431	0.50	Rp 7,773,785,163,537	Rp 48,370,457,436,115
10	Rp 17,148,421,114,679	0.47	Rp 8,016,959,564,511	Rp 56,387,417,000,627
Total			**Rp 56,387,417,000,627**	

Dengan demikian dapat dihitung MPP sebagai berikut:

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

MPP = 2.43 Tahun

MPP = 2 Tahun 6 Bulan

L.5.12 *Internal Rate of Return*

Keterangan:

Net Cash Flow (NCF) sesudah pajak = (Penjualan – Pengeluaran – Pajak) + Depresiasi

Tabel L5.14 *Internal Rate of Return* (IRR)

Tahun	Net Cash Flow	Bunga 7,90% 1/(1+I)^n	Present Value	Bunga 57.82% 1/(1+I)^n	Present Value	Bunga 25% 1/(1+I)^n	Present Value
0	-Rp11,966,367,511,573	1.00	-Rp11,966,367,511,572.700	1.00	-Rp11,966,367,511,572.700	1.00	-Rp11,925,272,914,658.200
1	Rp4,115,053,135,576	0.93	Rp3,813,765,649,282.980	0.63	Rp2,607,494,003,541.700	0.80	Rp3,429,024,498,176.100
2	Rp6,159,340,970,452	0.86	Rp5,290,434,687,020.940	0.40	Rp2,473,033,437,541.070	0.64	Rp4,069,152,300,340.680
3	Rp8,567,460,546,561	0.80	Rp6,820,053,462,604.170	0.25	Rp2,179,694,990,118.780	0.51	Rp4,504,287,326,486.360
4	Rp9,449,202,550,976	0.74	Rp6,971,229,490,467.880	0.16	Rp1,523,304,209,821.730	0.41	Rp3,974,011,188,329.360
5	Rp10,537,741,620,859	0.68	Rp7,205,105,135,064.440	0.10	Rp1,076,432,744,884.950	0.33	Rp3,544,190,858,958.710
6	Rp11,589,702,185,602	0.63	Rp7,344,184,999,458.190	0.06	Rp750,169,671,157.216	0.26	Rp3,118,412,532,221.690
7	Rp12,747,678,452,415	0.59	Rp7,486,537,210,500.060	0.04	Rp522,836,858,983.676	0.21	Rp2,743,994,581,565.610
8	Rp14,021,500,968,360	0.54	Rp7,631,729,149,752.680	0.03	Rp364,399,249,518.749	0.17	Rp2,414,556,631,839.760

9	Rp15,410,812,497,431	0.50	Rp7,773,785,163,536.630	0.02	Rp253,779,405,071.203	0.13	Rp2,123,086,672,622.870
10	Rp17,148,421,114,679	0.47	Rp8,016,959,564,511.160	0.01	Rp178,938,111,455.942	0.11	Rp1,889,418,237,313.230
Total			**Rp 56,387,417,000,626.500**		**-Rp36,284,829,477.640**		**Rp18,951,265,823,528.500**

Dengan demikian IRR dapat dihitung sebagai berikut:

$$IRR = \%IRR_{(+)} + \left\{\left(\frac{\text{NCF PV}_{(+)}}{\text{NCF PV}_{(+)} + \text{NCF PV}_{(-)}}\right) \text{x} \left(\%IRR_{(-)} - \%\text{IRR}_{(+)}\right)\right\}$$

$$IRR = 25\% + \left\{\left(\frac{Rp18{,}951{,}265{,}823{,}528.500}{Rp18{,}951{,}265{,}823{,}528.500 \ + \ -\text{Rp}36{,}284{,}829{,}477.640}\right) \text{x} \ (57.82\% - 25\%)\right\}$$

IRR = 57.88%

Pada perhitungan IRR dengan cara interpolasi didapatkan bunga sebesar 57.82 %. Perhitungan ini dipastikan menggunakan goalseek pada microsoft excel, sehingga nilai nol dapat diketahui bunganya, dari hasil goalseek untuk nilai nol yaitu 57.82%. Sehingga data interpolasi valid.

Berdasarkan perhitungan IRR didapatkan nilai sebesar 57.88%, dengan bunga pinjaman sebesar 7.90% serta nilai NCPV bertanda positif, sehingga dapat disimpulkan proyek investasi pabrik Ethylene ini layak untuk didirikan atau *Feasible.*