BASIC DESIGN OF PROPYLENE RECOVERY UNIT

A PROJECT REPORT

Submitted by

MALAVIKA MANOJ
(R670215007)

In partial fulfillment of the requirement for the award of the degree of

MASTER OF TECHNOLOGY
in
CHEMICAL ENGINEERING
with specialization in
PROCESS DESIGN ENGINEERING

Under the guidance of

Mr. Vimal Chikhaliya
Process Design Engineer
Essar Oil Ltd

Ms. Rose Havilah Pulla
Dr. M Gopinath
Assistant Professor
Department of Chemical Engineering

DEPARTMENT OF CHEMICAL ENGINEERING

COLLEGE OF ENGINEERING STUDIES

UNIVERSITY OF PETROLEUM & ENERGY STUDIES

Bidholi campus, Energy Acres,
Dehradun-248007

APRIL – 2017

DECLARATION BY THE SCHOLAR

I hereby declare that this submission is my own and that, to the best of my knowledge and belief, it contains no material previously published or written by another person nor material which has been accepted for the award of any other Degree or Diploma of the University or other Institute of Higher learning, except where due acknowledgement has been made in the text.

Malavika Manoj
R670215007

CERTIFICATE

This is to certify that the thesis entitled "**BASIC DESIGN OF PROPYLENE RECOVERY UNIT"** submitted by **MALAVIKA MANOJ (R670215007)**, to the University of Petroleum and Energy Studies, for the award of the degree of Master of Technology in Chemical Engineering with specialization in Process Design Engineering is a bonafide record of project work carried out by her under our supervision. The results embodied in this project report are based on literature and the research done in Essar Oil Ltd.

Signature of the Internal Guide

Ms. Rose Havilah Pulla
Assistant Professors
Department of Chemical Engineering
UPES
Dehradun - 248007

Signature of the Internal Co-Guide

Dr. M Gopinath
Assistant Professors
Department of Chemical Engineering
UPES
Dehradun - 248007

Signature of the Head of the Department

Date:……………………………………

ESSAR
OIL & GAS

REFINERY DIVISION

CERTIFICATION OF INDUSTRIAL TRAINING

This is to certify that

Ms. Malavika Manoj

a student of

University Of Petroleum & Energy Studies.

has successfully completed the Industrial Training on

Basic Design of Propylene Recovery Unit.

from 1st july, 2016 to 30th January, 2017

under the guidance of

Mr. Vimal Chikhaliya

We wish him/her all the success.

30/1/2017

Head-Training

Project Guide

ACKNOWLEDGEMENT

With a sense of gratitude and respect, I would like to extend my heartfelt acknowledgement to all those who have rendered their help and guidance during the entire period of the dissertation.

I am highly indebted to **The Chemical Engineering Department, University of Petroleum & Energy Studies, Dehradun and Essar Oil Limited, Vadinar, Gujarat,** for giving me this opportunity to pursue an internship and work on a live industrial project.

First of all, I take this opportunity to express my intense feeling of gratitude towards my external guide from Essar Oil Ltd. **Mr. Vimal Chikhaliya** and my internal guides from UPES **Ms. Rose Havilah Pulla** and **Dr. M Gopinath** their suggestions, guidance and constant inspiration at every stage of this project work. Their focusing ability and criticisms have always helped me during my work.

Then, I would like to expresses my profound thanks to **Dr. P Vijay**, Head of Chemical Engineering Department, for his support in doing this project.

It would be difficult, to sum up in words my gratitude towards my parents for being the pillar of strength throughout the work. Finally, great thanks to my department members from college and Vadinar refinery for giving me valuable knowledge, which would help me in future to prove myself as a Master of Technology in Chemical Engineering with Specialization in Process Design Engineering.

Sincere thanks to all those people whose names I might have missed out, for their cooperation in every possible way that helped me to complete this project successfully.

Date : 27th April 2017 MALAVIKA MANOJ

Place: Dehradun

ABSTRACT

Propylene, otherwise called Propene, is one of the most vital petrochemical feedstock created in the business sectors around the world. The by a wide margin biggest share of worldwide propylene yield around 66% is handled into polypropylene (PP). Interest for this plastic is anticipated to increment by 3.7% for every annum until 2021 and will, in this way, rule request improvement in the propylene advertise. Polypropylene is a standout amongst the most adaptable bundling materials. Different applications incorporate filaments, materials, vehicle parts, electric gadgets and family unit merchandise, and substantially more. The growing source for propylene production is from the refineries mainly from FCCU, coker and cracking units. In this thesis, a propylene recovery unit has been simulated based on the data obtained from Essar oil Ltd. The design criterion and objectives for the project were fixed from the findings obtained as a result of literature survey and the product requirement studies. A model for the propylene recovery unit was created in Aspen Hysys by using the Soave-Redlich-Kwong (SRK) as the fluid package. After completing the basic design of Propylene Recovery Unit (PRU) in Hysys, a yield of 99.58% by volume pure propylene was obtained which is 0.08% more than the value acquired from the simulation data received from Engineering India Ltd. (EIL). The study can be further continued by changing the fluid packages or by comparing the study with other licensors.

TABLE OF CONTENTS

DECLARATION BY THE SCHOLAR		
CERTIFICATE		
ACKNOWLEDGEMENT		i
ABSTRACT		ii
TABLE OF CONTENTS		iii
LIST OF TABLES		vi
LIST OF FIGURES		viii
ABBREVIATIONS		ix
1	**INTRODUCTION**	**1**
1.1	Oil refinery…………………………………………………………………………	1
1.2	History of Essar oil ……………………………………………………………....	2
1.2.1	Refinery units in Essar…………………………………………………....	3
1.3	Introduction to propylene………………………………………………………..	4
1.3.1	Stabilization of propylene price…………………………………………	4
1.3.2	Higher production propylene by propane dehydrogenation……………..	5
1.3.3	Regional differences in supply development…………………………....	5
1.3.4	Propylene determining demand for propylene…………………………....	5
1.3.5	Multiple applications of propylene……………………………………	5
1.3.6	General process description of propylene recovery unit………………...	6
1.4	Objectives of the thesis…………………………………………………………	8
2	**LITERATURE SURVEY**	**9**
2.1	Introduction……………………………………………………………………...	9
2.2	Propylene production methods and FCC process rules in propylene demands……	9
2.3	Optimizing The PSA process of P/P using neuro-fuzzy modeling………………...	10
2.4	Modification of a de-ethanisation plant for enhancing propane and propylene recovery …………………………………………………………………………	10
2.5	Propylene/propane separation by vacuum swing adsorption using Cu-BTC spheres…………………………………………………………………………	11

	2.6	Recovery of propylene from refinery off-gas …………………………………..	12
	2.7	New 13X zeolite for propylene/propane separation …………………………	12
	2.8	Pressure swing adsorption process for the separation of nitrogen and propylene …	13
	2.9	Optimal design of cryogenic distillation columns …………………………	14
	2.10	Optimal feed locations and number of trays for distillation columns with multiple feeds……………………………………………………………………..	14
	2.11	Design and economic investigation of shell and tube heat exchangers ……………	15
	2.12	Conclusions………………………………………………………………	15
3	**MATERIALS AND METHODS**		**16**
	3.1	Design basis: unit capacity…………………………………………………….	16
	3.2	Feed and product characteristics……………………………………………	16
		3.2.1 Feed stock characteristics…………………………………………..	16
		3.2.2 Product characteristics……………………………………………	17
	3.3	Feed battery limit conditions……………………………………………….	17
	3.4	Design procedure for distillation columns………………………………….	18
		3.4.1 Major steps to determine number of trays in a column…………………..	18
	3.5	Design procedure for heat exchangers………………………………………..	24
		3.5.1 Major steps in heat exchanger design…………………………….	24
	3.6	Simulation by aspen hysys………………………………………………….	30
	3.7	Controllers in PRU…………………………………………………………….	31
	3.8	Process data sheet in PRU…………………………………………………..	34
	3.9	Heat and mass transfer data sheet in PRU………………………………….	35
4	**RESULTS AND DISCUSSION**		**36**
	4.1	De-propaniser column design………………………………………………	36
	4.2	COS hydrolyser design………………………………………………………..	41
	4.3	Simulated results……………………………………………………………	45
	4.4	Detailed process description for PRU………………………………………	48
	4.5	Control philosophy of PRU…………………………………………………...	50
5	**CONCLUSIONS**		**53**

REFERENCES **54**

APPENDIX 1

(PROCESS DATA SHEET) **57**

APPENDIX 2

(HEAT AND MASS BALANCE DATA SHEET) **82**

APPENDIX 3

(DE- PRIESTER CHART) **107**

LIST OF TABLES

3.1	Design capacity data………………………………………........	16
3.3	Composition of feed in PRU……………………………………........	16
3.4	Feed battery limit condition………………………………………..	18
3.5	List of heat exchangers in PRU…………………………………........	29
4.1	Compositions of feed, distillate and residue in de propaniser column..	37
4.2	Feed temperature determination………………………………………	38
4.3	Dew point determination…………………………………………...	39
4.4	Bubble point determination………………………………………..	39
4.5	The average relative volatility determination………………………	40
4.6	Feed and product properties obtained after simulation………..........	45
A1.1	PDS of de-propaniser column………………………………………	57
A1.2	PDS of de-ethaniser column…………………………………..........	58
A1.3	PDS of P/P splitter…………………………………………………	60
A1.4	PDS of P-101……………………………………………………….	61
A1.5	PDS of P-102……………………………………………………….	63
A1.6	PDS of E-101……………………………………………………….	64
A1.7	PDS of E-102……………………………………………………….	66
A1.8	PDS of E-103……………………………………………………….	67
A1.9	PDS of E-104……………………………………………………….	69
A1.10	PDS of E-105……………………………………………………….	70
A1.11	PDS of E-106……………………………………………………….	71
A1.12	PDS of E-107……………………………………………………….	73
A1.13	PDS of E-108……………………………………………………….	74
A1.14	PDS of E-109……………………………………………………….	76
A1.15	PDS of E-110……………………………………………………….	77
A1.16	PDS of E-111……………………………………………………….	78
A1.17	PDS of E-112……………………………………………………….	80
A2.1	HMB of material stream 1……………………………………........	82

A2.2	HMB of material stream 2……………………………………….……..	83
A2.3	HMB of material stream 3……………………………………….……..	83
A2.4	HMB of material stream 4……………………………………….……..	84
A2.5	HMB of material stream 5……………………………………….……..	85
A2.6	HMB of material stream 6……………………………………….……..	86
A2.7	HMB of material stream 7……………………………………….……..	87
A2.8	HMB of material stream 8……………………………………….……..	88
A2.9	HMB of material stream 9……………………………………….……..	89
A2.10	HMB of material stream 10…………………………………………...	90
A2.11	HMB of material stream 11…………………………………………...	90
A2.12	HMB of material stream 12…………………………………………...	91
A2.13	HMB of material stream 13…………………………………………...	92
A2.14	HMB of material stream 14…………………………………………...	93
A2.15	HMB of material stream 15…………………………………………...	94
A2.16	HMB of material stream 16…………………………………………...	95
A2.17	HMB of material stream 17…………………………………………...	96
A2.18	HMB of material stream 18…………………………………………...	97
A2.19	HMB of material stream 19…………………………………………...	97
A2.20	HMB of material stream 20…………………………………………...	98
A2.21	HMB of material stream 21…………………………………………...	99
A2.22	HMB of material stream 22…………………………………………...	100
A2.23	HMB of material stream 23…………………………………………...	101
A2.24	HMB of material stream 24…………………………………………...	102
A2.25	HMB of material stream 25…………………………………………...	103
A2.26	HMB of material stream 26…………………………………………...	103
A2.27	HMB of material stream 27…………………………………………...	104
A2.28	HMB of material stream 28…………………………………………...	105

LIST OF FIGURES

1.1	Flow diagram of typical petroleum refinery units……………………	1
1.2	Industrial overview of Essar Oil Ltd, Vadinar…………………….....	3
1.3	Applications of propylene in a pie chart………………………..........	6
1.4	Block flow diagram of propylene recovery unit………………..........	7
3.1	Schematic flow diagram of distillation column……………………...	19
3.2	BFD of the main column used in PRU……………………………....	20
3.3	Schematic diagram of heat exchanger………………………….........	25
4.1	BFD of de propaniser column…………………………………….........	36
4.2	COS heat exchanger………………………………………………….....	42
4.3	Simulation diagram of propylene recovery unit after aspen hysys......	47
4.4	Process flow diagram of propylene recovery unit, Vadinar refinery...	52
A.3.1	De-Priester chart for k-values for hydrocarbons, high temperature.....	107

ABBREVIATIONS

EOL	:	Essar Oil Ltd
PSA	:	Pressure Swing Adsorption
P/P	:	Propylene/Propene
FCC	:	Fluidized catalytic cracking
FCCU	:	Fluidized Catalytic Cracking Unit
LPG	:	Liquefied Petroleum Gas
PRU	:	Propylene Recovery Unit
BFD	:	Block Flow Diagram
PDS	:	Process Data Sheet
HMB	:	Heat and Mass balance
MMBOE	:	Million Barrels of Oil Equivalents
SAARC	:	South Asian Association for Regional Cooperation
BPSD	:	Barrels per Stream Day
CBM	:	Coal Bed Methane
PDH	:	Propene Dehydrogenation
TPA	:	Tones per Annum
PR	:	Peng Robinson
SRK	:	Soave-Redlich-Kwong
EOS	:	Equations of State
P	:	Proportional
PI	:	Proportional Integral
PID	:	Proportional Integral Derivative
BP	:	Boiling Point

CHAPTER 1

INTRODUCTION

1.1 OIL REFINERY

An oil refinery is a modern procedure plant where unrefined petroleum is changed over into more valuable items, for example, petrol, kerosene, diesel, fuel oil, asphalt base, heating oil, and liquefied petroleum gas (Gary & Handwerk, 1984; Leffler, 1985). An extensive refinery unit comprises of gigantic funnels and other complex structures used to convey floods of liquids between vast concoction handling units. An oil refinery is viewed as a fundamental piece of the downstream side of the oil business. Contingent on the refinery area, desired products, and financial contemplations, every refinery has its own novel course of action and mix of refining procedures. The general schematic flow diagram of a typical petroleum refinery is given in figure 1.1

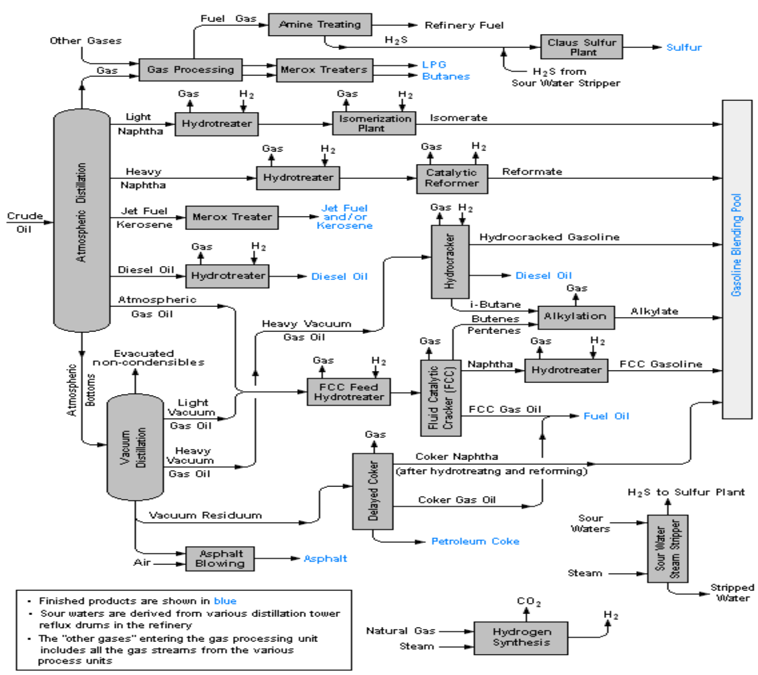

Figure 1.1: Flow diagram of typical petroleum refinery units

1.2 HISTORY OF ESSAR OIL

Essar Oil Ltd. (EOL) was begun in June 1976 underneath the name of Essar Construction Limited and was locked in essentially in center part exercises, and additionally marine developments, pipeline laying, digging and distinctive port-related exercises. In 1984, the corporate wandered extra into various center parts primarily the circle of investigation and improvement, penetrating coastal and seaward oil and gas wells for Indian open division oil investigation firms. The organization's name was then adjusted to Essar Offshore and Exploration Limited in May 1987.

In August 2000, the organization's name was altered to Essar Gujarat Limited, to mirror it to a great degree differed business intrigue. In 1988, the corporate made an underlying open supply for its shares that range unit at present recorded on Bombay stock exchange, national stock exchange of India and two other Indian stock trades. In the 1990s the gathering went into steel making market with its Hazira plant in Gujarat and a pilot plant in Visakhapatnam.

Essar's essential business is inside the power and oil divisions. The greater part of this is regularly dealt with by Essar Energy that is around seventy six percent firmly held by Essar gathering, is these days India's second biggest power era organization inside the non-open area. Its present era capacity is 1,600 MW and is being extended to 8,070 MW. Control era originates from a blend of gas, coal and Liquid fuel based power plants.

Essar Oil might be a completely coordinated oil and open utility of universal scale with a powerful nearness over the natural compound worth chain from investigation and creation to refinement and oil retail. Essar Oil also possesses India's second biggest single site mechanical plant at Vadinar, Gujarat, having a capacity of 20 MMTPA, or 405,000 barrels for each day. Vadinar mechanical plant fuses a multifaceted nature of 11.8 that is among the absolute best all inclusive. The modern plant is equipped for process some of the hardest crudes and in any case creates top notch money related unit IV and V review item. The mechanical plant has been found at a truly focused capex of $12,746/bbl that is concerning the overall normal. Industrial overview of Essar Oil Ltd, Vadinar is given in figure 1.2.

Figure 1.2: Industrial overview of Essar Oil Ltd, Vadinar

1.2.1 REFINERY UNITS IN ESSAR

Crude Distillation Unit (CDU) is used to separate crude into different products by taking boiling point difference into consideration. The capacity of this unit in EOL is about 18MMTPA.

In Vacuum Distillation Unit (VDU), the Atmospheric residue from CDU bottom is used as the feed which further separates the reduced crude oil into medium and heavy distillates under vacuum. This is a very good energy saving process.

The visbreaking unit is converted to Crude Distillation Unit II (CDU II), where the vacuum column functions as both CDU/VDU. This unit is capable of processing ultra-heavy crude on a standalone basis and also provides improved economics. Saturated Gas Unit (SGU) in the refinery is used to separate LPG and stabilized naphtha from unstabilized naphtha. EOL has one of the largest coker units in the world, which has a capacity of 6 MMTPA. The unit is capable of converting bottom of the barrel vacuum residue to valuable products.

Diesel Hydrodesulphurization unit (DHDS) is a medium pressure diesel hydro-Treater which produces Euro IV grade diesel. Diesel Hydrotreater unit (DHDT) is a very high-pressure hydrotreater which is capable of producing Euro V diesel. Vacuum Gas oil Hydro Treater (VGO-HDT) is used to hydrotreat FCC feed to enable the refinery to produce premium quality low Sulfur, high octane products.

Fluidized Catalytic Cracking Unit (FCCU) is the most important unit in any refinery that produces high-value product like LPG, Gasoline, and Diesel from low-value product VGO (vacuum gas oil). This unit helps in improving overall refinery profitability. Naphtha Hydro

Treater /Continuous Catalytic Reformer unit (NHT/CCR) prepares clean feed stroke for CCR and ISOM. CCR Produces reformate and hydrogen, reformate is the key component of gasoline and the H_2 produced is used in DHDS.

Isomerization Unit (ISOM) produces high octane Isomerate. The unit is used to converts Naphtha to Gasoline and also enables an increased production of BS-IV and BS-V grade gasoline. Amine Recovery Unit (ARU) is used to strip H_2S from Rich Amine coming from units DHDS, DCU and CDU and, supplying lean amine back to the unit.

Sour Water Striper (SWS) in the refinery is used to remove H_2S and NH_3 from sour water coming from the units FCCU, DCU, CDU-1 and CDU-2 and supply stripped water to CDU-1. The next unit is Sulfur Recovery Unit (SRU), which converts H_2S present in acid gasses to elemental sulfur and thereby avoids pollution of the environment.

1.3 INTRODUCTION TO PROPYLENE

In 2013, Propylene processed about 85 million tonnes worldwide and became the second most important petrochemical feedstock in the global markets. Uses of propylene incorporate the assembling of plastic polypropylene (PP) and furthermore the creation of critical chemicals, for example, propylene oxide, acrylonitrile, cumene, butyraldehyde, and acrylic corrosive.

The global propylene market is mainly being dominated by some established vendors such as BASF, ExxonMobil Chemical, Dow Chemical, Lyondell Basell Industries, and INEOS.

1.3.1 STABILIZATION OF PROPYLENE PRICES

A huge share of overall propylene yield starts as a by-result of ethylene generation utilizing steam breaking or by reactant splitting in refineries. Accordingly of an expanding utilization of C_2H_6, steam splitting produces a greater amount of ethylene and less of propylene. Be that as it may, in light of an extra across the board utilization of deliberately innovations, the accessibility of propylene worldwide is most likely going to unwind. This advancement is anticipated to direct to an adjustment of expenses for propylene. Overall incomes created with propylene are relied upon to perpetually increment by 5.3% p.a. in the vicinity of 2013 and

2021, and in this way at bottomless lower development rates than inside the past eight year sum.

1.3.2 HIGHER PRODUCTION POTENTIAL BY PROPANE DEHYDROGENATION

The most key of the intentionally advancements for the creation of propylene is the dehydrogenation of propane (PDH innovation). Given an expanding value differential amongst propylene and propane, this innovation is changing into extra beneficial not exclusively in nations with escalated gas assets however conjointly in nations with less gas assets. Both the USA and China can put numerous PDH plants on stream inside what's to come. Creation potential for propylene can rise subsequently.

1.3.3 REGIONAL DIFFERENCES IN SUPPLY DEVELOPMENT

Bolstered by the new PDH plants, gas generation in North America can rise significantly inside the future after the yield fell in the vicinity of 2005 and 2013. The most astounding development rates, then again, we gauge for Eastern Europe, where improvement will be ruled by Russia and the Middle East. Western Europe, in any case, can endure the consolidated impacts of soaring global rating and feeble request improvement.

1.3.4 POLYPROPYLENE DETERMINING DEMAND FOR PROPYLENE

The by a long shot biggest share worldwide for propylene yield is around 66% which are prepared into polypropylene (PP). Interest for this plastic is anticipated to increment by 3.7% p.a. until 2021 and along these lines will rule request improvement in the propylene advertise. Polypropylene is one among the first flexible bundling materials. Elective applications typify strands, materials, vehicle parts, electrical gadgets and family unit merchandise.

1.3.5 MULTIPLE APPLICATIONS OF PROPYLENE

Second biggest deals commercial center for propylene is the creation of propylene oxide, trailed by the assembling of acrylonitrile, butyraldehyde, cumene, and acrylic corrosive. The subordinates of propylene oxide are required as crude materials for the assembling of items like polyurethane (PUR), veneers and glues, polyester pitches, cooling operators, liquid catalyst, and solvents. Furthermore, the generation of the designing plastic acrylonitrile butadiene styrene (ABS), acrylonitrile is additionally used to deliver acrylic strands that are then handled into materials. Butyraldehyde is a middle of the road in the generation of butanol and 2-ethylhexanol. Cumene is basically used to fabricate phenol and **acetone** and is hence conjointly, a precursor for bisphenol A, phenolic tars, caprolactam, and methyl methacrylate.
By keeping the product value, profitability and all the above factors into account Essar oil Ltd is constructing a new PRU unit in their Vadinar refinery. The product from the unit is polymer grade propylene and is expected to have 99.5% by Volume purity. The propylene is being recovered from FCCU and Delayed Coker units, which supplies unsaturated LPG to PRU. The figure 1.3 shows the application of propylene in a pie chart.

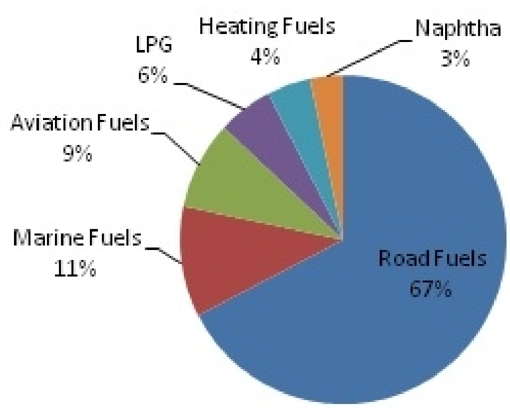

Figure 1.3: Applications of Propylene in a Pie chart

1.3.6 GENERAL PROCESS DESCRIPTION OF PROPYLENE RECOVERY UNIT

The cracked LPG from FCCU and Delayed Coker unit is processed in propylene recovery unit. The LPG is pumped to de propaniser column, from the bottom; the stream is further sent to the LPG storage after cooling. The vapors from the column top are split into two streams.

The major stream is condensed in de propaniser condenser and refluxed back and the other split goes to the COS hydrolyser.

In hydrolyser, COS is converted into H_2S and CO_2 in the presence of water vapors filled with catalyst. The feed to hydrolyzer is preheated before entering the column with hot effluent vapors. Prior to that deaerated water is injected into vapor stream to ensure that minimum 0.2 wt% H_2O of feed is present before entering into the reactor.

Now the cooled product from the hydrolyzer is sent to de ethaniser column, where light components namely C_2 hydrocarbons, H_2S, H_2O, CO, CO_2 etc are separated from C_3 hydrocarbons by the distillation process. The C_3 hydrocarbons are taken from the bottom of the column and used future in the process. The top product is sent as fuel gas to the battery limit.

Propane –propylene splitter is the next major equipment in the unit. The splitter is built in two units to minimize the difference between height/diameter ratio. The propane is obtained as the bottom product, which after cooling goes to LPG return pool. The top product goes into the H_2S removal.

H_2S removal is effected by H_2S separator. This is filled with ZnO catalyst which reacts with Hydrogen sulphide and the effluent propylene is free of sulfur. The vapor is preheated and resent to the H_2S splitter and bottom product is future taken ahead to the drying unit where it is dried using N_2 and filtered and later stored. The explanation above is from the figure 1.4 showing the block flow diagram of PRU in general

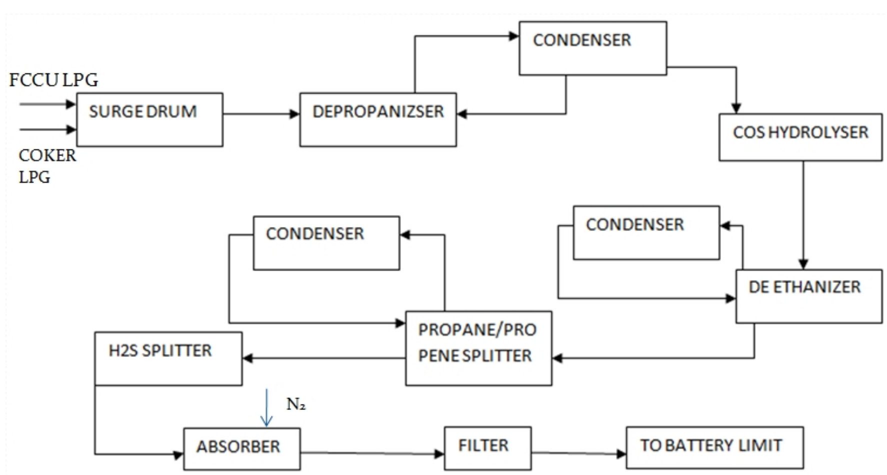

Figure 1.4: Block Flow Diagram of Propylene Recovery Unit

1.4 OBJECTIVES OF THE THESIS

The main objectives of the project are:

Basic design of a Propylene Recovery unit (PRU)

- Simulation of the unit will be done in Aspen Hysys as the first basic step to obtain the product data.
- Calculation of the overall heat and mass balances is done by Aspen Hysys simulation software.
- Process Flow Diagram (PFD) and Process data sheet (PDS) is to be developed.
- Finally, a Control philosophy is to be made for the unit, which would give a clear idea on how the controllers will work (in standard and emergency conditions).

CHAPTER 2

LITERATURE SURVEY

2.1 INTRODUCTION

A literature survey helps in summarizing the current knowledge in the area of investigation, identifying any strengths and weaknesses in previous work, which in turn helps you to identify them in your own research and thus serving to you to spot them in your own analysis and so eliminate the potential weaknesses, whilst bringing to the fore the potential strengths. Additionally, a decent and full literature search can offer the context within which to place your study.

This project is mainly based on the designing of the Propylene Recover Unit. So the survey's topic of interests includes, the procedure for designing columns and heat exchanger, the functioning of the subunits present in the recovery unit and their problems, different technologies used all over the world etc.

2.2 PROPYLENE PRODUCTION METHODS AND FCC PROCESS RULES IN PROPYLENE DEMANDS

Amir, 2008 article on Propylene production methods and FCC process rules in propylene demands states that the markets for propylene as a basic intermediate petrochemical continue to grow at average rates of 4-5% per year and the largest source of propylene supply to the petrochemical markets is steam crackers and the second largest source of propylene supply to petrochemical markets is refinery FCC units. However, the FCC unit is becoming increasingly important as a source of propylene supply to meet future demand growth into the world petrochemical markets. Ethylene growth rates have lagged propylene growth rates and will continue to do so in the future. Similarly, the growth rate of transportation fuels is several times lower than propylene. Currently, steam cracking and refinery operations constitute over 97% of the propylene produced today. Clearly alternative routes to propylene will gain prominence as producers seek to leverage their existing assets and available internal streams to find an optimum solution for meeting the demand for propylene. In this Manuscript all the following methods for increasing of propylene from FCC process is investigated:

- Add additives of ZSM-5 catalyst to FCC catalyst
- Change catalyst of FCC to Propylene mode FCC catalyst as Grace and Davison or Albemarle Company Catalyst that will be considered for propylene purpose production.
- Change the FCC operation condition to HS – FCC process condition.
- Use SUPERFLEX process

2.3 OPTIMIZING THE PSA PROCESS OF P/P USING NEURO-FUZZY MODELING

In optimizing the PSA process of propylene/propane using Neuro-Fuzzy modeling was studied by Mona *et al.,* 2012, Cryogenic distillation is the common method for separating propylene/propane mixtures, but this is highly energy intensive. Some 8-ring silica zeolites, especially pure silica chabazite (SiCHA), are known to show high diffusivity ratio for propylene over propane. In this work, the separation of propylene/propane using pure silica chabazite (SiCHA) in a simple 4-step pressure swing adsorption process is studied. An isothermal isobaric micropore diffusion model was created to simulate this kinetically controlled separation. It is first developed and implemented in the multi-physics software COMSOL to simulate different modes of PSA process. In this study, we present a sequential optimization strategy based on neuro-fuzzy model and genetic algorithm (GA) with synergistic combination of COMSOL simulation model to maximize the purity of propylene and propane productions.

2.4 MODIFICATION OF A DE-ETHANISATION PLANT FOR ENHANCING PROPANE AND PROPYLENE RECOVERY

A study was conducted by Ahmed *et al.,* 2016 on Modification of a de-ethanisation plant for enhancing propane and propylene recovery and solving some operational problems in the Journal of Natural Gas Science and Engineering. In this, the primary goal was to increase the propane and propylene recoveries as well as to overcome some operational problems of this process. In order to accomplish this goal, a change in the process configuration and in some operating conditions was suggested. The simulation tool used in this study to examine the proposed modifications is HYSYS version 8.0 with Peng–Robinson Equation of State (EoS). The validity of simulation is proved by the good correspondence between laboratory and simulation results of the modified plant. The benefits of this study were realized when the proposed modification was applied to the original plant. The results show that the modified plant in operation is capable of recovering 2235 tons/year of propane and propylene more than the original plant. It is also noted that more ethane and ethylene are separated in the modified plant. Furthermore, the modified process provides a solution to some operational problems like increased carryover in the de ethaniser rectifying column. The last part of this work considered the investigation of the maximum feed stream CO_2 concentration at which the plant can operate properly without freezing.

2.5 PROPYLENE/PROPANE SEPARATION BY VACUUM SWING ADSORPTION USING Cu-BTC SPHERES

In the studies carried out by Plaza *et al.,* 2016 on Propylene/propane separation by vacuum swing adsorption using Cu-BTC spheres, Cu-BTC MOF is an appealing candidate to carry out propane/propylene separation by adsorption due to its high adsorption capacity and propylene selectivity. In this work, a brand new sample of Cu-BTC, synthesized and shaped into spheres by the Korean Research Institute of Chemical Technology (KRICT) is studied for this commitment. The sample under evaluation presents the very best propylene specific adsorption capability known up to date for a shaped material (up to 8 mol kg^{-1} of dry adsorbent at 323 K).

A mathematical model is projected that adequately describes the breakthrough in these experiments. This model has conjointly been used to simulate an experimental 5-step VSA cycle formed to produce polymer-grade propylene. Although this cycle presents low recoveries, the valid model is used as a base for VPSA design.

2.6 RECOVERY OF PROPYLENE FROM REFINERY OFF-GAS

The article on Recovery of propylene from refinery off-gas using metal incorporated ethyl cellulose membranes by Susheela *et al.,* 2000 discusses the performance of certain changed ethyl cellulose (EC) films which was explored for the recuperation of propylene from a blend of gas that has an indistinguishable synthesis from the safeguard tail gas (ATG) of a liquid reactant breaking unit of HPCL refinery, Vishakhapatnam. The blend contains C_1–C_5 hydrocarbons and non-hydrocarbons like CO, CO_2, H_2 and N_2. Various metals like silver, ruthenium, palladium and iridium were joined into EC films to roll out improvements in the membrane permeation properties for specifically expanding the propylene flux by encouraging its vehicle. Ethyl cellulose films appeared to have great potential for the business recuperation of propylene from a hydrocarbon rich off-gas blend which contains olefin as a noteworthy constituent. A genuinely high selectivity of 5.1 accompanied by high transport rates was acquired for propylene concerning propane in the multi-component blend. Among the metal-consolidated films, the least complex outcomes were gotten with a silver (5 wt. %)- EC layer which yielded a permeate stream of around 65 mol% propylene. The pervasion of the different hydrocarbon parts introduce in the ATG blend through the metal fused EC layers was likewise broke down. Association of the metal particles with the layer grid has been clarified by portraying the movies with the help of FTIR, WAXD and SEM–EDX procedures that are accessible.

2.7 NEW 13X ZEOLITE FOR PROPYLENE/PROPANE SEPARATION

Campo *et al.,* 2013 has conducted a study on New 13X zeolite for propylene/propane separation by vacuum swing adsorption where adsorption equilibrium isotherms for propane and propylene on a new 13X zeolite from CECA were measured at three temperatures (323, 373, 423 K) up to 5 bar pressure. Maximum capacities of 3.47 and 3.08 mole /kg were found for propylene and propane, respectively. These values are considerably better than those reported in the literature. Adsorption isotherms were fitted with Toth equation. Heats of adsorption were evaluated as a function of loading from experimental data with Van Hoff's equation leading to 53 kJ/mol and 35 kJ/mol for propylene and propane, respectively; values were conjointly obtained from fitting with Toth equation. For fixed bed adsorbers, single and binary breakthrough curves were measured and simulated with the help of a mathematical model. A 5-step VSA cycle was conjointly simulated to produce 99.5% propylene from a mixture of 25:75 propane/propylene; experimental validation confirms a good prediction. Propylene obtained was having a purity of 99.54%, a productivity of 1.46 molC_3H_6/kg_{ads}/h and a recovery of 85%.

2.8 PRESSURE SWING ADSORPTION PROCESS FOR THE SEPARATION OF NITROGEN AND PROPYLENE

Pressure swing adsorption process for the separation of nitrogen and propylene with a MOF adsorbent MIL-100(Fe) by Ana *et al.,* 2013 talks about the recovery of propylene and nitrogen from the stream produced during the resin degassing (70% N_2/30% C_3H_6) within the polypropylene production. This work proposes the utilization of a pressure swing adsorption process with MIL-100(Fe) as an adsorbent to hold out this separation, either for recovering only nitrogen, or for recovering both nitrogen and propylene from the process. On a lab-scale set-up, single and binary breakthrough curves were experimentally determined and simulated to validate the mathematical model. PSA simulation shows that a product purity of 99.9% and a recovery of 81.5% could be achieved for the nitrogen recovery process. For the nitrogen and propylene recovery process, a recovery of 97.4% for nitrogen and 87.6% for propylene has been obtained and the products purities obtained were 99.9% for nitrogen and 97.9% for propylene respectively also the adsorbent productivity values obtained was 0.20 kgN_2/kg_{ads}/h and 0.07 kgC_3H_6/kg_{ads}/h. The overall power consumption raised to 179 W h/kgN_2 or 309 W h/kgC_3H_6 a value that is still below those reported previously in the literature for alternative technologies.

2.9 OPTIMAL DESIGN OF CRYOGENIC DISTILLATION COLUMNS

In studies carried out by Rafael *et al.,* 2014 on optimal design of cryogenic distillation columns with side heat pumps for the propylene/propane separation tell us about propylene/ propane production process. The boiling point of propylene (−47.6 °C) and propane (−42.1 °C) are very similar, so it is very hard to separate them. Plus it's a very high energy-intensive process. To separate this mixture, a conventional columns that operate at high pressure and cryogenic distillation columns that operate at low pressure have to be used; however, these methods are still energy consuming. This work shows the design and optimization of an energy-efficient column, which helps in minimizing the energy consumption in the propylene/propane separator. Conceptual design, superstructure representation, rigorous simulations and mathematical programming techniques are effectively combined to assess all the distillation structures used. Results obtained from the project showed that VRC and distillation columns with heat-integrated stages can scale down the energy consumption between 58 and 75% when compared with conventional distillation at high pressure.

2.10 OPTIMAL FEED LOCATIONS AND NO. OF TRAYS FOR DISTILLATION COLUMNS WITH MULTIPLE FEEDS

In the study, optimal feed locations and a number of trays for distillation columns with multiple feeds done by Jagadisan *et al.,* 1993 uses a MINLP models for finding the optimal locations for the feeds to enter a column and also finds the number of trays required in a column for a specified separation process to take place. Systems with ideal, Soave-Redlich-Kwong equation of state and UNIQUAC thermodynamic models are taken into consideration. The steps in the optimization process automatically determine the order and the location of the feed, no assumptions has to be made. The results obtained, helped to create a frame work that could solve any difficult problems with non-ideal thermodynamics.

2.11 DESIGN AND ECONOMIC INVESTIGATION OF SHELL AND TUBE HEAT EXCHANGERS

This study conducted by Oguz *et al.,* 2014 uses an improved Intelligent Tuned Harmony Search (I-ITHS) algorithm to design a shell and tube heat exchangers. ITHS has an advantage of deciding intensification and diversification processes by adjusting the pitch properly. Their aim was to improve the search capacity of ITHS algorithm by using chaotic sequences instead of uniformly distributed random numbers and also applying alternative search strategies. In order to minimize the total cost of a heat exchanger, baffle spacing, shell diameter, tube outer diameter and number of tube passes are minimized. Results obtained after analysis shows that I-ITHS can be used in optimizing shell and tube heat exchangers.

2.12 CONCLUSIONS

The survey has given an insight on how to move forward with the project, it provided information on different types of feeds used for recovering propylene, the market value of propylene, how the different subunits in Propylene recovery unit work and the challenges faced. A thorough understanding of how a distillation column operates is also acquired from the journals. Few of the papers include aspen techniques and mathematical models which helped in gaining more knowledge on how to approach the simulation part of the project. The design procedure journals have provided a better understanding of how a system can be designed for a multi-component feed.

CHAPTER 3

MATERIALS AND METHODS

3.1 DESIGN BASIS: UNIT CAPACITY

From the findings obtained as a result of literature survey and the product requirement studies the design capacity specifications are fixed and is provided in table 3.1 given below.

Table 3.1: Design capacity data

Propylene Production	100,000 TPA
Design Margin	10% of Design capacity
Recovery of Propylene	95% (minimum)
No. of Stream hours	8000 hrs/Year
Turn Down of Max. Capacity	50% of 110,000 TPA of Propylene production

3.2 FEED AND PRODUCT CHARACTERISTICS

3.2.1 FEEDSTOCK CHARACTERISTICS

The feed for Propylene recovery unit is from FCCU and Delayed Coker units with flow rates 47,390 kg/hr and 7,742.1 kg/hr, which supplies unsaturated LPG and components in the feed is provided in table 3.2.

Table 3.3: Composition of feed in PRU

COMPONENTS	FCCU	COKER	COMBINED
	(wt %)	(wt %)	(wt %)
H_2O	0.044	0	0.0003782926
H_2S	0.00002	0	0.0000001720

C_2H_4	0.001	0	0.0000085976
C_2H_6	0.259	0.03	0.0022689054
C_3H_6	28.47	12.7	0.2626107175
C_3H_8	8.6	40.7	0.1311055204
$N\text{-}C_4H_{10}$	10.01	17.8	0.1110631396
$i\text{-}C_4H_{10}$	10.92	4.6	0.1003464295
$1\text{-}C_4{=}/1\text{-}C_4H_8$	8.38	23.3	0.1047743214
$i\text{-}C_4{=}$	11.34	0	0.0974963249
$C_2\text{-}C_4$	7.54	0	0.0648255988
$t_2\text{-}C_4$	10.99	0	0.0944871791
Butadiene	0.141	0	0.0012122559
C_5H_{12}	3.274	0	0.0281484099
$n\text{-}C_5H_{12}$	0	0.9	0.0012641243
n-Hexane	0.001	0	0.0000085976
Mercaptan	0.00002	0	0.0000001720
Methyl Acetylene	0.000055	0	0.0000004729
Propadiene	0.000005	0	0.0000000430
COS	0.00005	0	0.0000004299
CO	0.00001	0	0.0000000860
CO_2	0.00001	0	0.0000000860
N_2	0.00001	0	0.0000000860
Re-Entry S+RSH-S	0	0.00002	0.0000000281
Caustic	0.000001	0.000001	0.0000000100

3.2.2 PRODUCT CHARACTERISTICS

The major products from the unit are Polymer grade Propylene (with a purity of 99.5%), Lean LPG and Light Gas. The specifications of LPG and Light Gas are to be confirmed by EIL (if IS specification is met or not).

3.3 FEED BATTERY LIMIT CONDITIONS

The table 3.4 given below shows the conditions of the feed coming from the boundaries of the other plant units.

Table 3.4: Feed battery limit condition

	PRESSURE (kg/cm^2 g)	TEMPERATURE (^oC)
Feedstock FCCU LPG from Merichem Unit	18	40
Feedstock Coker LPG from Merichem Unit	18	40

3.4 DESIGN PROCEDURE FOR DISTILLATION COLUMNS

3.4.1 MAJOR STEPS TO DETERMINE NUMBER OF TRAYS IN A COLUMN

Identify the components to be separated

1. Identify light and heavy key
2. Material balance across the column
3. Determine the feed temperature
4. Define minimum Reflux ratio
5. Identify the bubble point and dew point temperature
6. Determine the actual number of trays by finding out minimum and theoretical number of trays

The detailed procedure for the above steps is explained as follows:

- It is convenient to list the feed components and their weight% in the order of the boiling point.
- Find out the individual component flow rates (kg/h) by multiplying weight % to feed flow rate.
- Next step is to find the light and heavy key components. The light key is the component present in the residue in important amounts while components lighter than light key are present in small amounts. Similarly, heavy key is the component present

in distillate in important amounts and the components heavier than heavy key are present in small amounts.

- Now the feed components have to be separated to distillate and residue with respect to the column chosen.

  For example:

  If we assume the column to be De propaniser column, then C_3+ components will be in distillate and C_4+ components will be in the residue.

  If the column is a De ethaniser column, then C_2+ components will be in distillate and C_3+ components will be in the residue.

- The distillate and residue flow rates can be obtained by adding individual component flow rates that belong to the distillate and residue composition.
- In order to find the weight % for components in distillate and residue, the following material balance equations can be used.

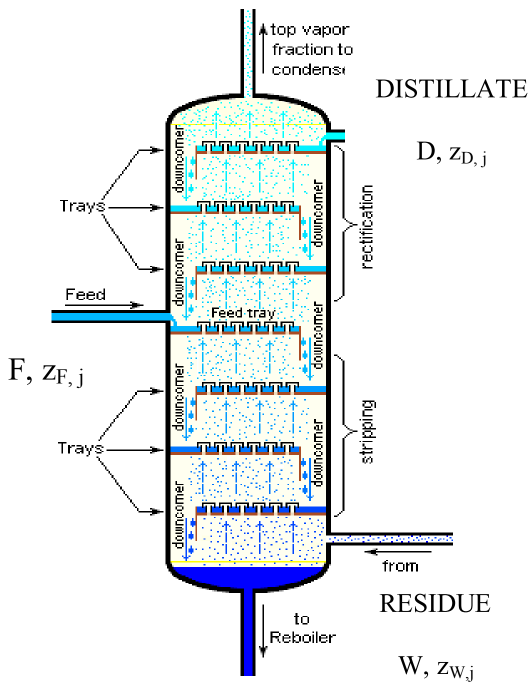

Figure 3.1: Schematic diagram of distillation column

Over all material balance:

$$F = D+W \qquad \ldots 3.1$$

Individual component balance:

$$Fz_{F,j} = Dy_{D,j} + Wx_{W,j} \qquad \ldots 3.2$$

For Distillate:

$$Fz_{F,j} = Dy_{D,j} \qquad \ldots 3.3$$

For residue:

$$Fz_{F,j} = Wx_{W,j} \qquad \ldots 3.4$$

Where,

F = flow rate of feed (kg/h)

$z_{F,j}$ = the weight fraction of j^{th} component in the feed

D = the flow rate of distillate (kg/h)

$y_{D,j}$ = the weight fraction of j^{th} component in the distillate

W = the flow rate of residue (kg/h)

$x_{W,j}$ = the weight fraction of j^{th} component in the residue

To find the weight percent of light and heavy key equation $Fz_{F,j}=Dy_{D,j}+Wx_{W,j}$ has to be used since they are present in both distillate and residue composition. It is important to know the exact weight fraction of the light key and heavier key present in distillate and residue for future calculations.

Example: Given below are the main columns in PRU unit

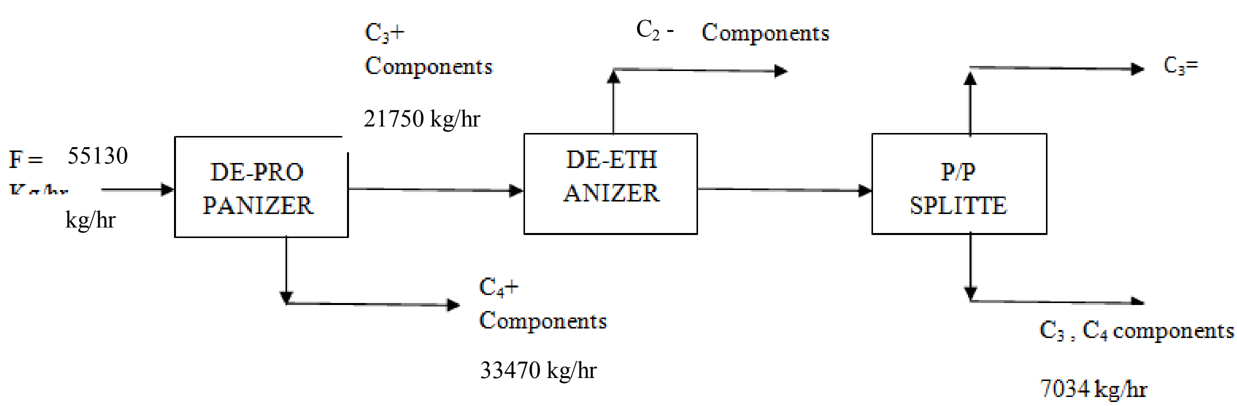

Figure 3.2: BFD of the main columns used in PRU unit

Assuming that the heavy and light components are i-C_4H_8 and C_3H_8, now we need to find out weight percent of these components in De propaniser. Also, assume the flow rate of residual P/P splitter as 7034 kg/h.

Residual flow rate of C_3 component (light key) in P/P splitter= $x_{W,\ C3}$X Feed flow rate

Residual flow rate of C_4 component (heavy key) in P/P splitter

= 7034 - Residual flow rate of C_3 component

$$\% \text{ of split} = \frac{\text{Residual flow rate of } C_4 \text{ in } \frac{P}{P} \text{ splitter}}{\text{Flow rate of feed}} \quad \ldots 3.5$$

$z_{F,\ C4}$ of % of split would give the percentage of weight fraction of i-C_4H_8 in the distillate.
To find the weight fraction in residue equation (2) is used

$$43515\ z_{F,i\text{-}c4=} = 17744\ y_{D,i\text{-}c4=} + 25771\ x_{W,i\text{-}c4=}$$

$$x_{W,i\text{-}c4=} = \frac{(43515\ zF,i-c4=) - (17744\ yD,i-c4=)}{25771}$$

Similarly, C_3H_8 is also found out

- Reflux ratio = $\frac{L}{D}$
- Ratio between Residue and distillate flow rates, $\frac{W}{D}$

INLET TEMPERATURE

- In order to fix the inlet temperature, trial and error method has to be used. The procedure is as follows:
- At first from the data obtained assume an inlet temperature, now with respect to that temperature fill in the table given in table 4.2

In the table calculation, if the total of y_i (concentration in gas) becomes equal to 1 then the assumed temperature is apt for the column, if not keep changing the temperature until or unless you get $\sum y_i$=1

The equation for volatility:

$$\text{Volatility, } \alpha = \frac{m_j}{m_{hk}} \quad \ldots 3.6$$

Where,

m_j = distribution coefficient of j^{th} component

m_{hk}= distribution coefficient of heavy key

MINIMUM REFLUX

Many methods of estimating the minimum reflux have been proposed, Underwood's method [13] is the one that we use here. In this, two equations have to be solved:

$$\sum \frac{\alpha_j z_{jF} F}{\alpha_j - \theta} = F(1-q) \quad \ldots 3.7$$

$$\sum \frac{\alpha_j y_{jD} D}{\alpha_j - \theta} = D(1+R_m) \quad \ldots 3.8$$

Where,

D = the flow rate of distillate (kg/h)

F = flow rate of feed (kg/h)

α_j = relative volatility of j^{th} component

$z_{F,j}$ = the weight fraction of j^{th} component in the feed

y_{jD} = weight fraction of j^{th} component in the distillate

R_m = minimum reflux ratio

q = D/F

ACTUAL REFLUX RATIO

The rule of thumb is:

$R = (1.2\ldots1.5)\, R_{min}$ …3.9

DEW POINT

Assume the dew point temperature and fill in table 4.3 from materials and methods.

Here $\sum y_i/\alpha = m_{hk}$

So the obtained value has to be compared to De-Priester chart (Appendix 3: Image A3.1) for the corresponding heavy key component and find the temperature with respect to it.

BUBBLE POINTS

Assume the bubble point temperature.

According to the table 4.4 in materials and methods, collect the data

Here $m_{hk} = 1/\sum X_i * \alpha$

Similarly like above, compared the obtained value from the above equation to De-Priester chart (Appendix 3: Image A3.1) and find the temperature with respect to it.

MINIMUM NUMBER OF TRAYS

Fenskey equation is not limited to binary mixture and can be applied to the key components to determine the mini number of trays (Robert, 1956)

$$N_{min} + 1 = \frac{\log\left[\left(\frac{y_{lk}}{y_{hk}}\right)_D \left(\frac{X_{HK}}{X_{LK}}\right)_W\right]}{\log\left(\alpha_{D,W}\right)_{ave}} \qquad \ldots 3.10$$

Where,

N_m+1 = Total number of theoretical stages including the reboiler

$Y_{lk,d}$ = Distillate light key component weight fraction

$Y_{hk,d}$ = Distillate heavy key component weight fraction

$x_{lk,d}$ = Residue light key component weight fraction

$x_{hk,d}$ = Residue heavy key component weight fraction

$\left(\alpha_{D,W}\right)_{ave} = (\alpha_{\text{dew pt temp.}} \text{ X } \alpha_{\text{bubble pt temp}})^{0.5}$

Average relative volatility of light and heavy component with respect to the dew and bubble point temperature obtained

THEORETICAL NUMBER OF PLATES

Gilliland related the number of equilibrium stages and the minimum reflux ratio and the no. of equilibrium stages with a plot that was transformed by Eduljee into the relation (Niranjan, 1980):

$$\left(N - N_{min}\right) / \left(N + 1\right) = .75\left[1 - \left(R - R_{min} / R + 1\right)^{0.566}\right] \qquad \ldots 3.11$$

Where,

N = Theoretical plates

N_{min}= Minimum reflux

R = Actual reflux ratio

R_{min} =Minimum reflux ratio

ACTUAL NUMBER OF TRAYS

Assuming the tray efficiency, find the actual number of trays:

$$\text{Actual number of trays} = \frac{\text{Theoretical trays}}{\text{Tray efficiency}} \qquad \ldots 3.12$$

The above design procedure is elaborated through the calculations done in 4.1 of Results and Discussion.

3.5 DESIGN PROCEDURE FOR HEAT EXCHANGERS

3.5.1 MAJOR STEPS IN HEAT EXCHANGER DESIGN:

- Find out the heat transfer area
- Determine the number of tubes
- Calculate bundle diameter and shell diameter
- Calculation of shell and tube side heat transfer coefficients
- Overall Heat transfer coefficient determination
- Calculate heat exchanger pressure drop

The detailed procedure for the above steps is explained as follows:

Shell and tube heat exchangers are generally designed by trial and error calculations. In Kem method following steps is involved for designing a heat exchanger:

1. The properties of the hot and cold fluids used in the heat exchanger have to be found.

2. Energy balance is performed and heat duty (Q) of the exchanger is found out.

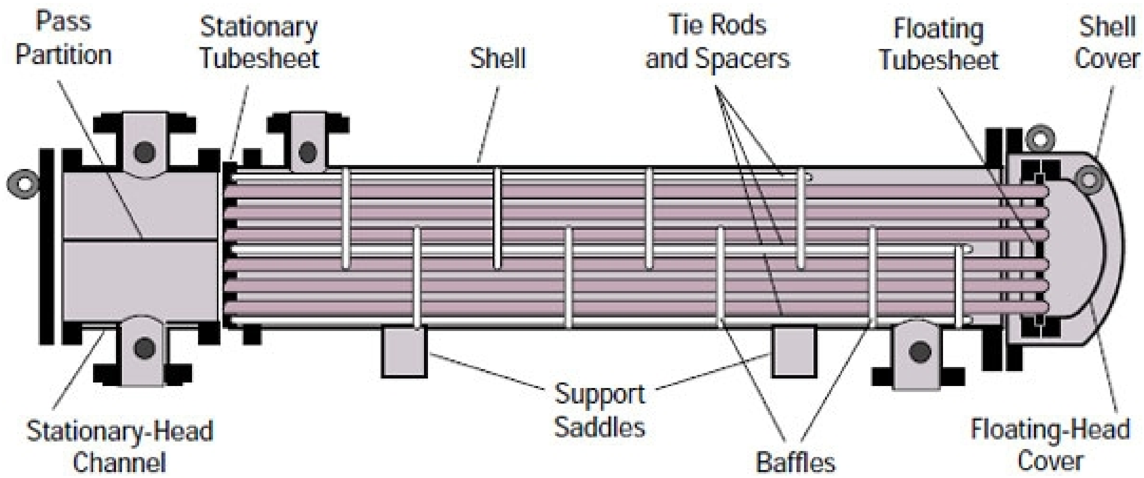

Figure 3.3: Schematic diagram of Heat exchanger

3. Overall heat transfer coefficient (U) value has to be assumed. The value of U for hot and cold fluids can be taken from the books (Kern, 1965; Dutta, 2006).

4. Number of shell and tube passes is then decided. The LMTD and the correction factor FT of the heat exchanger unit is then found out (Kern, 1965; Dutta, 2006). FT normally should be greater than 0.75 for the steady operation of the exchangers. Otherwise, it is required to increase the number of passes to obtain higher FT values.

5. Calculate heat transfer area (A) required from:

$$Q = U\,A\,\Delta T_M \qquad \ldots 3.12$$

   Where,

   Q = Heat transfer per unit time

   U = Over all heat transfer coefficient

   A = Heat transfer area

   ΔT_{M} = the mean temperature difference (LMTD X FT)

6. A tube material is selected and tube diameter (ID= id, OD = od), wall thickness (in terms of BWG or SWG) and tube length (L) are fixed to proceed further in the

calculations. Calculate the number of tubes (N_t) required to provide the heat transfer area (A):

$$N_t = \frac{A}{\pi d_o L} \qquad \text{...3.13}$$

7. Decide the type of shell and tube exchanger (fixed tube sheet, U-tube etc.). Select the tube pitch, determine inside shell diameter (D_s) that can accommodate the calculated number of tubes. Use the standard tube counts table for this purpose. Tube counts are available in standard textbooks (Kern, 1965; Dutta, 2006).

8. On the basis of general guidelines the fluid to shell side or tube side has to be assigned. Select the type of baffle (segmental, doughnut etc.), its size (i.e. percentage cut, 25% baffles are widely used), spacing (B) and number. The baffle spacing is usually chosen to be within 0.2 D_s to D_s.

9. Calculate bundle diameter and shell diameter:

$$D_b = d_o\left(\frac{N_t}{K_t}\right)^{1/n_1} \qquad \text{...3.14}$$

10. The shell side and tube side film heat transfer coefficient is estimated from:

$$j_h = \frac{h\, d_e}{k} \frac{C_p \mu}{k}^{1/3} \frac{\mu}{\mu_w}^{-0.14} \qquad \text{... 3.15}$$

Where,

$\frac{\mu}{\mu_w} = 1$

d_e = Equivalent diameter (Sinnott *et. al.,* 1993)

k = thermal conductivity of the fluid

C_p = Specific heat capacity

h = Heat transfer coefficient (h_o = outside film coefficient; h_i= inside film coefficient)

The overall heat transfer coefficient (U_{cal}) based on the outside tube area is estimated (including dirt factors) from:

$$\frac{1}{U_{cal}} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2k_w} + \frac{d_o}{d_i}\frac{1}{h_{id}} + \frac{d_o}{d_i}\frac{1}{h_i} \quad \text{…3.16}$$

Where,

U_{cal} =Overall coefficient

h_o = outside fluid film coefficient

h_i = Inside fluid film coefficient

h_{od} = outside dirt coefficient

h_{id} = inside dirt coefficient

d_i = tube inside diameter

d_o = tube outside diameter

k_w= thermal conductivity of tube wall material

Select the outside tube (shell side) dirt factor and inside tube (tube side) dirt factor (Dutta, 2006).

11. If

$$0 < \frac{U_{cal} - U}{U} < 30\%$$

Then go to the next step 11 or else go to step 5 and the heat transfer area (A) is re calculated using U_{cal} from step 5. A baffle space of 0.2 D_s is assumed, if the calculated shell side heat transfer coefficient is too low. Now the shell side heat transfer coefficient is recalculated for better result.

12. Calculate the tube-side pressure drop (ΔP_T) by calculating the pressure drop in the straight section of the tube (frictional loss) (ΔP_t) and pressure drop due to return loss (ΔP_{rt}) which is caused because of the change in direction of fluid in a multi-pass exchanger. Total tube side pressure drop:

$$\Delta P_T = \Delta P_t + \Delta P_{rt} \quad \text{…3.17}$$

Or

$$\Delta P_T = N_P\left[8 j_f \frac{L}{d_i}\left(\frac{\mu}{\mu_w}\right)^{-m} + 2.5\right]\frac{\rho u_t^2}{2} \quad \text{…3.18}$$

Where,

ΔP_T = Tube side pressure drop

j_f = friction factor (Sinnott *et. al.,* 1993)

N_P = Number of tube side passes

L = Length of one tube

u_t = Tube side velocity = $\frac{W_t}{\rho A_t}$

W_t = Tube side fluid flow rate

A_t = Tube side area

ρ = Density

13. Calculate shell side pressure drop ΔP_S by calculating pressure drop for flow across the tube bundle (frictional loss) (ΔP_s) and pressure drop due to return loss (ΔP_{rs}). Now the total shell side pressure drop is:

$$\Delta P_S = \Delta P_s + \Delta P_{rs} \quad \text{...3.19}$$

Or

$$\Delta P_S = 8 j_f \frac{D_s}{d_e} \frac{L}{l_B} \frac{\rho u_s^2}{2} \left(\frac{\mu}{\mu_w}\right)^{-m} \quad \text{...3.20}$$

Where

j_f = friction factor (Sinnott *et. al.,* 1993)

L = Length of one tube

l_b = baffle spacing

D_s =Shell diameter

d_e = Equivalent diameter

u_s = shell side velocity = $\frac{W_s}{\rho A_s}$

W_s = shell side fluid flow rate

A_s = shell side area

ρ = Density

If the for the system, the tubes passes are increased or decreased when the tube-side pressure drop exceeds the allowable pressure drop. After this, recalculation starts from step 6.

If the shell-side pressure drop exceeds the allowable pressure drop, go back to step7 and re do the calculations.

The above design procedure is elaborated through the calculations done in 4.2 of Results and Discussion.

Given below are the heat exchangers available in the propylene recovery unit. The detailed data sheet is provided in Appendix 1: Process datasheet.

Table 3.5: List of heat exchangers in PRU

	Heat Exchanger Name	Shell Side Fluid	Tube Side Fluid	Shell Side		Tube Side	
				Inlet Temp (°C)	Outlet Temp (°C)	Inlet Temp (°C)	Outlet Temp (°C)
1	De propaniser Reboiler	LP steam	Hydro Carbon	143	138.2	104.3	106.2
2	De propaniser Cooler	Hydro Carbon	Cooling water	46.5	45.7	31	40
3	COS Hydrolyser Heat exchanger	Hydro Carbon	Hydro carbon	46.5	128.8	150	64
4	COS Reboiler	MP steam	Hydro Carbon	259	235.4	128.8	148.8
5	COS Cooler	Hydro Carbon	Cooling water	64.18	41.76	31	41.05
6	De ethaniser Reboiler	Hydro Carbon	Hot water	68.21	68.2	94	77.35
7	De ethaniser Cooler	Cooling water	Hydro Carbon	31	38.62	57.78	44.8
8	PP Splitter Reboiler	Hot water	Hydro Carbon	59.15	59.18	94	73.65
9	PP Splitter Cooler	Cooling water	Hydro Carbon	31	41.41	48.45	47.78
10	H_2S Separate Heat exchanger	Hydro Carbon	Hydro Carbon	48.45	149.6	180	79.39
11	H_2S Reboiler	MP steam	Hydro Carbon	259	236.4	149.6	179.9
12	H_2S Separate Cooler	Cooling water	Hydro Carbon	31	38.34	79.39	40.36

3.6 SIMULATION BY ASPEN HYSYS

Simulation is the imitation of the operation of a real-world method or system over time (Banks *et. al.,* 2001). Selecting a suitable property package is one among the foremost vital concerns for a successful process simulation. Two of the key factors to taken into consideration are:

- Specific system under consideration
- Operating conditions

The property package is a collection of methods for calculating the properties of the selected components. After you have established a component list, you combine the component list with a property package.

For oil, gas and natural segments, the Peng-Robinson EOS (PR) is for the most part the suggested property package. Hyprotech's upgrades to the present equation of state empower it to be precise for an assortment of frameworks over a decent scope of conditions. It thoroughly settles any single, two-stage or three-stage framework with a high level of effectiveness and dependability, and is material over a decent scope of conditions. The PR condition of state has been improved to yield exact stage harmony computations for frameworks beginning from low-temperature cryogenic systems to high temperature, high-pressure reservoir systems. A similar condition of state acceptably predicts part circulations for overwhelming oil frameworks, watery glycol and CH_3OH frameworks, and corrosive gas/acrid water frameworks, however particular sour water models (Sour PR and Sour SRK) are accessible for more specific treatment.

The Soave-Redlich-Kwong (SRK) condition will offer practically identical outcomes to the PR in a few cases, it has been found that its scope of use is essentially constrained and is not as solid for non-ideal systems. For example, it shouldn't be utilized for systems with CH_3OH or glycols. As a substitute, the PRSV condition of state can likewise be considered. It can deal with an indistinguishable system from the PR condition with equivalent, or higher accuracy, in addition to it is more reasonable for taking care of respectably non-ideal systems.

The upside of the PRSV condition is that not exclusively will it can possibly more precisely foresee the stage conduct of hydrocarbon systems, especially for systems made out of dissimilar components, however it can likewise be reached out to deal with non-ideal system with accuracies that rival traditional activity coefficient models. The sole bargain is the increased computational time and the additional collaboration parameter that is required for the equation (Murdock & James, 1993).

The PR and PRSV conditions of state perform thorough three-stage flash estimations for aqueous systems containing water, CH_3OH or glycols, moreover as a system containing different hydrocarbons or non-hydrocarbons in the second liquid stage. For SRK, water is the exclusively segment that will start an aqueous phase. The PR may likewise be utilized for crude systems that have generally been demonstrated with dual model thermodynamic bundles. These earlier models are suspect for systems with giant amounts of light ends or when approaching critical regions. Likewise, the dual model system winds up in inward irregularities. The restrictive improvements to the PR and SRK techniques permit these EOSs to appropriately represent vacuum conditions and heavy components (an issue with conventional EOS strategies).

However, in my project the fluid package used is SRK because the feed composition has only water and hence it is safe to use this package. The mathematical equations used by SRK are (Murdock & James, 1993):

$$P=\frac{RT}{V-b}-\frac{a}{V(V+b)+b(V-b)} \quad \ldots 3.21$$

$$Z^3-(1-B)\,Z^2+(A-2A-3A^2)\,Z-(AB-B^2-B^3)=0 \quad \ldots 3.22$$

Where,

$$A= aP\,/\,((RT)^2) \quad \ldots 3.23$$

$$B = bP/RT \quad \ldots 3.24$$

$$b=\sum_{i=1}^{n} x_i\left(0.077796\frac{RT_i}{P_i}\right) \quad \ldots 3.25$$

3.7 CONTROLLERS IN PRU

A process has to be controlled tolerably so as to get more uniform and better quality products. Processes are controlled by either manually or automatically controllers. Automatic controllers are proven to be apt for industries where one has to control so many variables. Generally, controllers are used in maintaining temperature, pressure, flow, level etc by retaining the set point provided in a process. There exist different types of controls to control a process such as:

1. Open Loop control: An open loop reaction is controlled by shifting the contribution to a framework and measuring the yield's reaction from the framework. In an open loop control, the controller sets the information incentive to the procedure with no learning of the output variable.

   A typical case of open circle control is the control of traffic in a town. The activity lights change as per an arrangement of foreordained tenets.

2. Feedback control: This control is accomplished by bolstering the procedure output data back to the controller. The controller makes utilization of the present data about the process variable to figure out what move to make to manage the process variable. This is the least difficult and most by and large utilized controller framework.

   The disadvantage of this is that the controller has to wait until disturbances upset the process to respond. There are different types of feedback controllers like:

   - Digital On/Off controller- On/Off controller is one among the first fundamental administrative control. An on-off controller only drives the controlled variable from completely closed to completely open contingent upon the position of the controlled variable with respect to the set point. This controller is a suitable controller, if the deviation from the set point is inside an adequate range and the cycling doesn't destabilize whatever is left of the process.

   - P-Only Controller – This controller will damp out motions from unsettling influences/disturbances and can stop the cycling of the procedure variable. The sole detriment of this controller is that it will have an offset i.e. it will never accomplish the set point yet will run parallel to the set point. The output response is

     $$\mathrm{OP}(t) = K_C E(t) \qquad \ldots 3.26$$

     Where,

     OP (t) = Controller Output

K_c = Proportional gain of the controller

E (t) = Error at the time

- PI Controller – In contrast to P controllers, PI controllers will dampen out oscillations and return the process variables to the set point. Notwithstanding the very certainty that PI control brings about zero error, the fundamental activity of the controller builds increases the time of oscillation and sets aside longer opportunity to time out the process variable. The output of the proportional integral controller is

$$OP(t) = K_C E(t) + \frac{K_C}{T_i} \int E(t) \quad \ldots 3.27$$

Where,

T_i = Integral time

OP (t) = Controller Output

K_c = Proportional gain of the controller

E (t) = Error at the time

- PID Controller- If the response of a PI controller to a disturbance isn't sufficient; the derivative action in a PID controller will cut back the natural period of significantly further. They are faster compared to other controllers. The output is defined as

$$OP(t) = K_C E(t) + \frac{K_C}{T_i} \int E(t) + K_C T_d \frac{dE(t)}{dt} \quad \ldots 3.28$$

Where,

OP (t) = Controller Output

K_c = Proportional gain of the controller

E (t) = Error at the time

T_i = Integral time

T_d = Differential time

3. Feedforward control: This control is employed when the feedback controller fails to control a process variable. Here, process disturbances are measured and compensated

for, without waiting for the disturbance to affect the process. Feedforward control is also useful wherever the final controlled variable can't be measured.

4. Cascade control: This is an often used method for minimizing the disturbances entering a slow process is cascade multi-loop control. Cascade control can also speed up the response of the controller system by reducing the time constant of the process transfer function relating the manipulated variable and process output. In here, the output of the primary control is the set point for the secondary control loop.

STABILITY:

Stability of a system is an essential angle to consider when outlining control plans. A few frameworks have oscillatory reactions, contingent upon its controller tuning parameters. Once a process is agitated with a bounded disturbance or bounded change in the input forcing function, the output typically will respond in one of these three ways:

1. The reaction can oscillate with the diminishing amplitude and in the end achieve steady state and stabilize.
2. The reaction can oscillate with consistent amplitude.
3. The reaction can develop persistently and never achieve steady state

3.8 PROCESS DATA SHEET IN PRU

As the role of information in our world grows, it is increasingly important that people should know how to handle and use the obtained information or data. A datasheet is a complete encyclopedia where data are stored in a particular format. Datasheets can be of different formats depending upon the different departments it is being used. To put it plainly, a datasheet also called a spec sheet is a record that outlines the execution and other specialized qualities of an item, machine, segment, material and so forth. These sufficient details are used by a design engineer to integrate the component into a system. Most of the design errors are caused due to (deliberately or not) overlooking certain specifications in the datasheet. The data sheets can start with an early on page depicting the record, trailed by postings of particular qualities, with more data In this project, a process data sheet has been provided which includes all the information related to all the equipment used in the PRU unit.

The process data sheets obtained after simulating the PRU unit in Aspen Hysys has been provided in Appendix 1: Process datasheet section of the thesis.

3.9 HEAT AND MASS BALANCE DATA SHEET IN PRU

Material and energy balances are critical in an industry. The expanding estimation of energy has made the ventures take a gander at recommendations to diminish energy utilization in handling. Energy balances are utilized as a part of the examination of the different phases of a process, over the entire process and notwithstanding stretching out over the full generation framework from the raw material to the finished products.
Similarly, material balances are basic to the control of processing, especially in the control of yields of the products. In industries, heat and mass balance are documented in a heat and mass balance data sheet (HMB data sheet). In fact, HMB datasheet is a document made by process design engineers while designing a process plant. Sometimes heat and mass balance is not a separate sheet/document but appears along the Process Flow Diagram (PFD). A heat and mass balance sheet represent each and every process stream on the corresponding PFD in terms of the process conditions.

The HMB data obtained after the simulation is provided in Appendix 2: Heat and Mass balance data sheet.

CHAPTER 4

RESULTS AND DISCUSSION

4.1 DE-PROPANISER COLUMN DESIGN

According to the detailed procedure provided in 3.4.2 of Materials and Methods, the number of trays has been found out for a multiple component systems. Similarly, the number of trays in De-propaniser column is also obtained. Given below are the detailed steps used for the manual calculation.

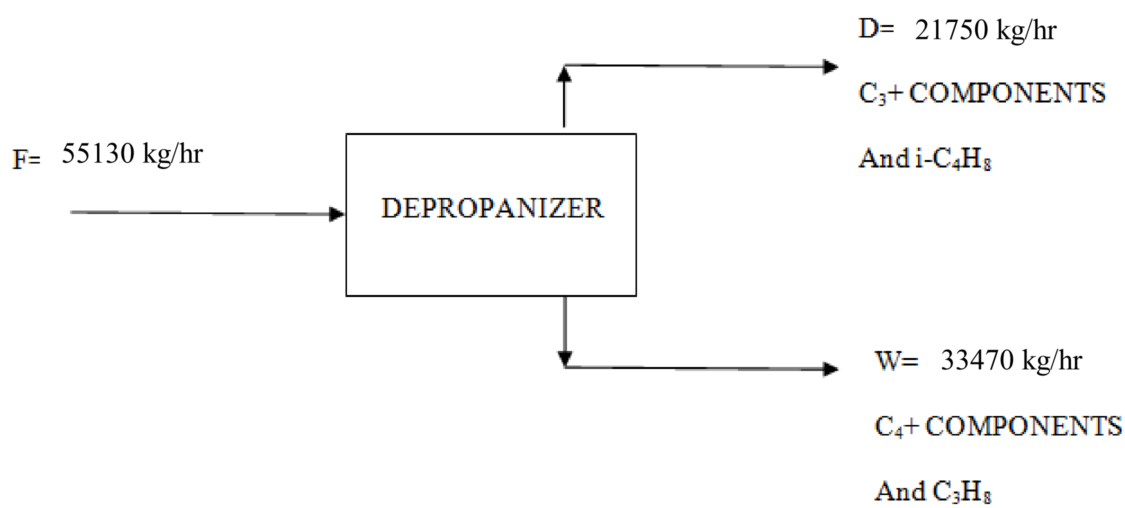

Figure 4.1: BFD of De propaniser column

STEP 1:

1. Identify the components to be separated
2. Identify light and heavy key
3. Material balance across the column

Table4. 1: Compositions of feed, distillate, and residue in De-propaniser column

COMPONENTS	BP (°C)	kg/h W.R.T FLOWRATE	wt FRACTION IN F	SUM OF FLOW RATES (kg/h)	DISTILLATE wt%	RESIDUE wt %
N_2	-195.8	0.004739	0.00000009	21827.4	0.00000022	
CO	-192	0.004739	0.00000009		0.00000022	
C_2H_6	-103.9	125.06273	0.00226891		0.00572963	
C_2H_4	-88.6	0.4739	0.00000860		0.00002171	
CO_2	-78.5	0.004739	0.00000009		0.00000022	
H_2S	-59.6	0.009478	0.00000017		0.00000043	
COS	-50.2	0.023695	0.00000043		0.00000109	
C_3H_6	-48	14475.1797	0.26261072		0.66316642	
C_3H_8(light key)	-42.2	7226.5747	0.13110552		0.03131711	0.19652840
Propadiene	-34	0.0023695	0.00000004		0.00000011	
Methyl Acetylene	-23.21	0.0260645	0.00000047		0.00000119	
i-C_4=(heavy key)	-11.7	5374.026	0.09749632	33292.9	0.08120000	0.10817960
i-C_4H_{10}	-10	5531.1246	0.10034643			0.16613515
1-C_4H_8/1-C_4=	-5	5775.1913	0.10477432			0.17346604
Butadiene	-4.41	66.8199	0.00121226			0.00200703
n-C_4H_{10}	-1	6121.8328	0.11106314			0.18387791
t2-C_4	0.9	5208.161	0.09448718			0.15643448
c2-C_4	3.7	3573.206	0.06482560			0.10732630
C_5H_{12}	27	1551.5486	0.02814841			0.04660296
Mercaptan	34.4	0.00004	0.00000020			0.00000033
n-C_5H_{12}	36	69.6789	0.00126412			0.00209290
n-Hexane	69	0.4739	0.00000860			0.00001423
H_2O	100	20.8516	0.00037829			0.00062631
Caustic	1390	0.000551321	0.00000001			0.00000002

STEP 2: FEED TEMPERATURE DETERMINATION

Assuming the inlet temperature to be 50 °C

Table 4.2: Feed temperature determination

COMPONENTS	Z/100	$m_{j,i}$	α	$Y = \frac{z_f((\frac{W}{D}) + 1)}{1 + (\frac{W}{D_{m_j}})}$
H_2O	0.00000009			
H_2S	0.00000009			
C_2H_4	0.00226891	5.1	8.5	1.7 X10^{-5}
C_2H_6	0.00000860	3.5	5.833	0.00397
C_3H_6	0.00000009	1.45	2.417	0.3227
C_3H_8	0.00000017	1.3	2.167	0.15218
Mercaptan	0.00000043	3.5	5.833	3.5 X10^{-7}
Methyl Acetylene	0.26261072	5.1	8.5	9.1 X10^{-7}
Propadiene	0.13110552	1.45	2.416	5.3 X10^{-8}
COS	0.00000004			
CO	0.00000047			
CO_2	0.09749632			
N_2	0.10034643			
N-C_4H_{10}	0.10477432	0.44	0.73	0.06297
i-C_4H_{10}	0.00121226	0.6	1	0.07168
1-C_4H_8	0.11106314	0.6	1	0.07484
i-C_4=	0.09448718	0.6	1	0.06964
c_2-C_4	0.06482560	0.6	1	0.0463
t_2-C_4	0.02814841	0.6	1	0.06749
Butadiene	0.00000020	0.6	1	0.00087
i-C_5H_{12}	0.00126412	2	3.33	0.04021
n-C_5H_{12}	0.00000860	0.153	0.255	0.00029
n-Hexane	0.00037829	0.068	0.113	9.3 X10^{-7}
Caustic	0.00000001			
Total				0.913

The total of y is found almost equal to, that means the assumed temperature is correct for the column.

STEP 3: DEW POINT DETERMINATION

Assuming the dew point temperature to be 46 ^oC

Table 4.3: Dew point determination

COMPONENTS	y_i*D/100	$m_{j,i}$ @46 °C	A	y_i*D/D	y_i/ α
C_2H_4	2.17 X10^{-5}	4.4	8.8	2.78 X10^{-5}	3.16 X10^{-6}
C_2H_6	5.73 X10^{-3}	3	6	7.33 X10^{-3}	1.22 X10^{-3}
C_3H_6	0.663	1.2	2.4	0.849	0.354
C_3H_8	0.0313	1.1	2.2	0.0401	0.0182
Methyl Acetylene	1.19 X10^{-6}	4.4	8.8	1.53E-06	1.74 X10^{-7}
Propadiene	1.09 X10^{-7}	1.2	2.4	1.39 X10^{-7}	5.79 X10^{-8}
i-C4=	0.0812	0.5	1	0.104	0.104
H_2S	4.34 X10^{-7}				
COS	1.09E-06				
CO	2.17 X10^{-7}				
CO_2	2.17 X10^{-7}				
N_2	2.17 X10^{-7}				
Total	0.781			0.896	0.476
	D Value				

Here $\sum y_i/\alpha = m_{hk=}$ 0.47

After comparing the value to De-Priester chart (Appendix 3, image A3.1), we obtained the temperature as 44^0 C, which is almost close to the assumed value.

STEP 4: BUBBLE POINT DETERMINATION

Assuming the dew point temperature to be 106°C

Table 4.4: Bubble point determination

COMPONENTS	X_i *W	X_i *W/W	$m_{j,I}$ @106 °C	A	X_i* α
C_3H_8	0.196	0.171	7.2	1.894	0.325
N-C_4H_{10}	0.184	0.161	3.1	0.815	0.131
i-C_4H_{10}	0.166	0.145	3.8	1	.145
1-C_4H_8	0.173	0.152	3.8	1	0.151

i-C_4=	0.108	0.0946	3.8	1	0.094
c2-C_4	0.107	0.0939	3.8	1	0.093
t2-C_4	0.156	0.137	3.8	1	0.136
Butadiene	2.01 X10^{-3}	1.76 X10^{-3}	3.8	1	0.001
i-C_5H_{12}	0.0466	0.0408	1.6	0.421	0.017
n-C_5H_{12}	0.00209	1.83 X10^{-3}	1.3	0.342	0.0006
n-Hexane	1.42 X10^{-5}	1.25 X10^{-5}	0.62	0.163	2.031 X10^{-6}
Mercaptan	3.31 X10^{-7}	2.90 X10^{-7}	16	4.210	1.219 X10^{-6}
H_2O	6.26 X10^{-4}	5.48 X10^{-4}			
Caustic	1.66 X10^{-8}	1.45 X10^{-8}			
Total	1.143	W VALUE			1.098

Here $m_{hk} = 1/\sum X_i * \alpha = 1.098$

After comparing with De-Priester chart (Appendix 3, image A3.1), the obtained temperature is 61 ^{0} C.

STEP 5: NUMBER OF ACTUAL TRAYS

1. Minimum Number of trays determination

   By using Fenskey equation

   We determined the minimum number of trays as 14

Table 4.5: The average relative volatility determination

COMPONENTS	$m_{j,i}$	α @44 °C	$m_{j,i}$	α @61 °C	AVG
C_2H_4	4.2	8.75	7.2	8	8.36
C_2H_6	2.8	5.83	4.8	5.33	5.57
C_3H_6	1.15	2.39	2.2	2.44	2.42
C_3H_8	1.05	2.18	1.9	2.11	2.14
Mercaptan	2.8	5.83	4.8	5.33	5.57
Methyl Acetylene	4.2	8.75	7.2	8	8.36
Propadiene	1.15	2.39	2.2	2.44	2.42
N-C_4H_{10}	0.34	0.71	0.68	0.76	0.73

$i\text{-}C_4H_{10}$	0.48	1	0.9	1	1
$1\text{-}C_4H_8$	0.48	1	0.9	1	1
$i\text{-}C_4$=	0.48	1	0.9	1	1
$c2\text{-}C_4$	0.48	1	0.9	1	1
$t2\text{-}C_4$	0.48	1	0.9	1	1
Butadiene	0.48	1	0.9	1	1
$i\text{-}C_5H_{12}$	1.56	3.25	3	3.33	3.29
$n\text{-}C_5H_{12}$	0.13	0.271	0.26	0.28	0.27
n-Hexane	0.051	0.11	0.26	0.28	0.17

2. Theoretical tray determination

   By using Eduljee's relation:

   The number of theoretical trays obtained = 29 trays

3. Actual tray determination

   Assuming the tray efficiency to be 70% (This may not be the correct efficiency used)

   $$\text{Actual number of trays} = \frac{\text{Theoretical trays}}{\text{Tray efficiency}}$$

   $$=60$$

4.2 COS HYDROLYSER DESIGN

The manual design of the heat exchanger is explained in 3.5.2 of Materials and Methods by taking that into consideration, the COS hydrolyser has been designed. The steps are as follows:

Data provided:

Tube and shell fluid are the same with different temperatures

Use a 1/ 2 Heat exchanger with split head floating head

Duty : 1.987 X 106 kJ/h

FT : 0.92

U : 266.9 kJ/h m^2 ^{o} C

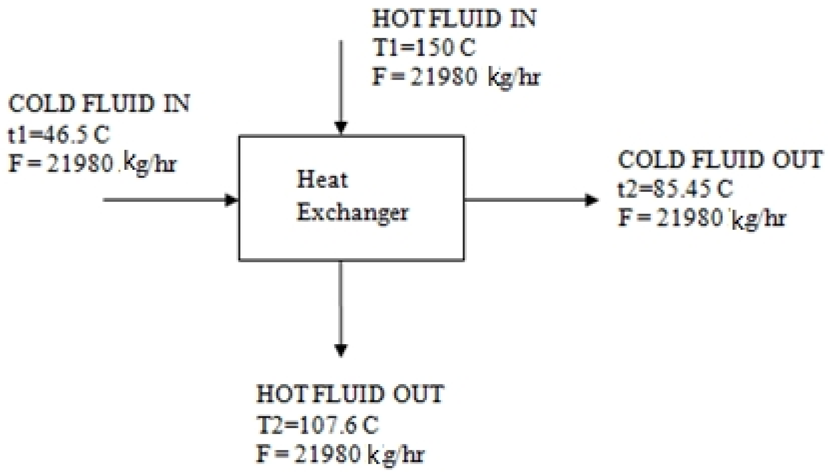

Figure 4.2: COS heat exchanger

Cp	: 1.63 kJ/ kg º C
K_f	: 0.25 W/m º C
Density	: 23.1 kg/m^3
Viscosity	: 0.01 cP
Pitch	: Triangular pitch

Assumptions:

1. Carbon steel tube material is used, K_w =45 W/m º C
2. U =266.9 kJ/ h m^2 º C
3. O.D = 20 mm , I.D =16 mm, L= 5.5 m

LMTD $= \dfrac{(150-85.45)-(107.6-46.5)}{\ln\frac{(150-85.45)}{(107.6-46.5)}}$

$= 62.81$ ºC

Heat transfer Area $= \dfrac{1987000}{266.9 \text{ X } 62.81 \text{X} .92} = 120.7 \text{ m}^2$

Area of 1 tube $= \pi \text{ X } 20 \text{ X } 10^{-3} \text{ X}5.5$

$= 0.345 \text{ m}^2$

No. of tubes $= \frac{\text{Heat transfer Area}}{\text{Area of 1 tube}}$

$= 350$ tubes

Since the tube pass is 2

$K_1 = 0.249$; $n_1 = 2.207$

Bundle dia, D_b $= d_o (\frac{N_t}{K_t})^{1/n_1}$

$= 535$ mm

For split head floating head exchanger, the typical shell clearance is 58mm ([2] page 831, Fig 12.12)

Shell dia, D_s $= 534 + 58$

$= 592$ mm

Tube side

Heat transfer coefficient:

Mean temperature $= \frac{(46.5+85.45)}{2}$

$= 66\ ^{o}C$

u_t $= \frac{21980}{60 \text{ X } 0.046 \text{ X } 23.1}$

$= 451.81$ m/s

To find the value for j_h

Reynolds no., Re $= \frac{\rho u d_i}{\mu}$

$= 16698.53$

$\frac{L}{d_i}$ $= 344$

j_h $= 2.9 \text{ X } 10^{-3}$

Applying the equation from procedure

$h_i = 303.33\ W/m^2\ ^{o}C$

Pressure drop:

$$\Delta P_T = N_P [8 j_f \frac{L}{d_i} (\frac{\mu}{\mu_w})^{-m} + 2.5] \frac{\rho u_t^2}{2}$$

ΔP_T $= 1.2 \text{ X } 10^5$ kPa

Shell side

l_b = 20% of D_s

A_s $= \frac{(P_t - d_O) D_s l_B}{P_t}$

$= 0.014\ m^2$

Considering a triangle pitch

d_e = 14.4 mm ([19]pages 858)

Mean temperature $= \frac{150+107.6}{2}$

$= 128.\ 8\ ^{o}C$

Re $= \frac{\rho u d_i}{\mu}$

= 37680.05

Use baffle segment to be 25%

j_h $= 9.2\ X\ 10^{-3}$

j_f $= 3.4\ X 10^{-3}$

h_s $= 1760\ W/m^2\ ^{o}\ C$

Pressure drop:

$$\Delta P_S = 8 j_f \frac{D_s}{d_e} \frac{L}{l_B} \frac{\rho u_s^2}{2} (\frac{\mu}{\mu_w})^{-m}$$

ΔP_s $= 2.26\ X 10^5$ kPa

Overall Heat transfer

$$\frac{1}{U_{cal}} = \frac{1}{1760} + \frac{20X\ 10^{-3} \ln (\frac{20}{16})}{2\ X\ 45} + \frac{20}{16} \frac{1}{307.33}$$

U_{cal} $= 221.5$ KJ/h $m^2\ ^{o}\ C$

The obtained U_{cal} obeys the condition below

$$0 < \frac{U_{cal} - U}{U} < 30\%$$

4.3 SIMULATED RESULTS

The feed consists of unsaturated LPG from FCCU and delayed Coker units, which after undergoing the process yields 99.58% by volume pure polymer grade propylene.

The expected purity was 99.5% by Volume but the simulation result has shown a 0.08% by volume increase in the purity which is a sign of successes in the designing aspect.
The table given below shows the major product streams coming out of the Propylene Recovery Unit.

Table 4.6: Feed and Product Properties obtained after simulation

	Feed (1)	To LPG Storage Pool (3)	Fuel Gas (15)	To LPG Storage Pool (19)	Propylene (28)
Temperature (^oC)	64	106.43	52.15	59.74	40.04
Pressure (kPa)	1961.33	1892.68	2834.12	2147.65	1667.13
Molar Flow (kgmole/h)	1092.10	577.65	14.09	171.69	332.29
Mass Flow (kg/h)	55115.82	33168.43	544.01	7570.58	13986.00
Liquid Volume Flow (m^3/h)	98.52	55.92	1.15	14.91	26.85
Heat Flow (kJ/h)	-5.7X 107	-3.7 X 10^7	-2.6 X 10^5	-1.8 X 10^7	-2.0 X 10^7
COMPOSITION	wt%	wt%	wt%	w%	wt%
H_2O	0.0004	0	0	0	0
H_2S	0	0	0.000012	0	0
Ethane	0.003795	0	0.297123	0	0
Propane	0.150104	0.000235	0.057457	0.956425	0.00414
CO_2	0	0	0.000006	0	0

Nitrogen	0	0	0.000006	0	0
n-Hexane	0	0	0	0	0
E-Mercaptan	0.000008	0.000015	0	0	0
M-Acetylene	0	0	0	0.000002	0
Propylene	0.315073	0.000059	0.643988	0.037856	0.995859
i-Butane	0.08716	0.163492	0	0.002051	0
n-Butane	0.096472	0.182364	0	0.000038	0
n-Pentane	0.00084	0.001588	0	0	0
1-Butene	0.094279	0.177474	0	0.001279	0
COS	0	0	0	0	0
i-Pentane	0.019664	0.037176	0	0	0
cis2-Butene	0.058311	0.110239	0	0.000004	0
tr2-Butene	0.085019	0.160714	0	0.000034	0
Propadiene	0	0	0	0	0
13-Butadiene	0.00112	0.002111	0	0.000011	0
i-Butene	0.087737	0.164536	0	0.0023	0
NaOH	0	0	0	0	0
Ethylene	0.000018	0	0.001407	0	0

The above results have been obtained from the simulation done in Aspen Hysys. The figure 4.3 shows the Aspen Hysys simulation of the entire PRU unit, which includes three distillation units, two hydrolyser units and rest of the equipment, includes splitters heat exchangers reboilers and coolers.

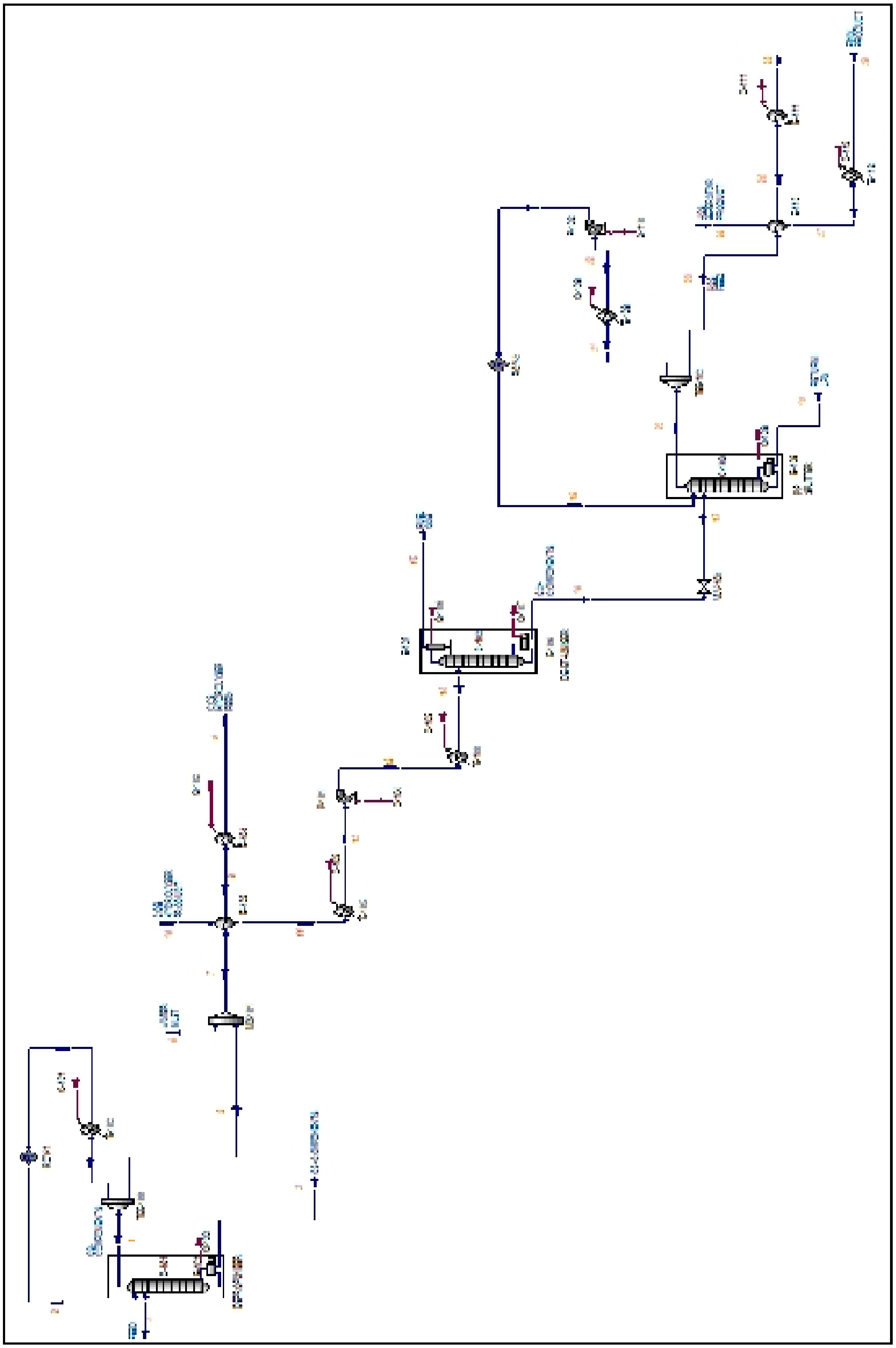

Figure 4.3: Simulation diagram of Propylene Recovery unit in Aspen Hysys

4.4 DETAILED PROCESS DESCRIPTION FOR PRU

DE-PROPANISER COLUMN

The cracked LPG from FCCU and Delayed Coker unit is available at the plant battery limit at a pressure of 18 kg/cm^2 g and temperature equal to 40ºC. The LPG is pumped to de propaniser column 36C-101 under flow control.

The top pressure of the column is maintained at 17.8 Kg/cm^2 g through the pressure controller by controlling the flow of LP steam into 36E-101.

The de propaniser bottom stream at 106.4 ºC is sent to the LPG storage under flow control cascaded with the level controller installed at the bottom of the column. The vapours from the column top are split into two streams and the major stream is condensed in de propaniser condenser 36E-102 by cooling water and refluxed back. The condenser is partially flooded with C_3 components. Then the other split which is stream 5 goes to the COS hydrolyser.

COS HYDROLYSER

In hydrolyser, COS is converted into H_2S and CO_2 in the presence of water vapours at 150 ºC filled with catalyst. The feed to hydrolyzer is preheated in COS preheaters 36E-103 with hot effluent vapours from reactor 36R-101X from 46.5 ºC to 128.8 ºC. Prior to that deaerated water is injected into vapour stream to ensure that minimum 0.2 wt% H_2O of feed is present before entering into the reactor. The C_3 hydrolyser vapours are further heated to about 150 ºC in COS hydrolyser superheater 36E-104 ºC by means of MP stream. Hydrolyser feed is controlled by a temperature controller.

The hot vapours are sent to R-101X where COS is hydrolyzed and after exchanging heat with the reactor feed vapours and attending a temperature of 64.18 ºC are finally condensed in 36E-105, COS hydrolyser condenser. The condensate is sent to condensate drum 36V-101. System pressure is maintained at a constant level by partially flooding 36E-105

The condensed C_3 carbon stream is pumped by 36P-101 through a control valve which is cascaded to the level controller of 36V-101 to de ethaniser column 36C-102.

DE-ETHANISER

The light components namely C_2 hydrocarbons, H_2S, H_2O, CO, CO_2 etc are separated from C_3 hydrocarbons by distillation process in 36C-101. The C_3 hydrocarbons are taken from the bottom of the column and used future in the process. The above light components are purged through the exchanger 36E-107 under a flow control and are sent as fuel gas to the battery limit.

The system pressure is also controlled by varying the duty of 36E-106. Condenser gets flooded up to the desired level to match the duty required which depends on the product reflux rate. Feed is sent to the column under flow control.C_3 hydrocarbon stream is drawn as a bottom product from column 36C-102 under level/ flow controller. The level controller controls the set point of the flow controller.

P/P SPLITTER

Propane –propylene splitter 36C-103 is the next major equipment in the unit. Feed is sent to the column 36C-103 by the flow controller. Propane is obtained as the bottom product from 36C-103, which after cooling goes to LPG return pool via a cascade control, where level and flow controllers are interconnected in a loop.

PP splitter operates at a pressure of around 20-21 kg/cm^2 and is maintained by the pressure controller which intern is controlled by the hot water flow to 36E-108. Heat to the reboiler is supplied by hot water via closed loop hot water system.

The vapour from the top splits into two streams, the major stream goes for H_2S removal and the later into the column as reflux. The reflux to the column is condensed in 36E-109 and then pumped back trough a pump 36P-102 via flow controller.

The column operation is monitored by checking the propane content in propylene distillate by means of an analyzer.

H_2S SEPARATOR

H_2S removal is effected by H_2S separator 36R-102X. This is filled with ZnO catalyst which reacts with Hydrogen Sulphide and the effluent propylene is free of sulfur. The sulfur removal is more effective at about 180ºC and the exact reactor temperature that has to be maintained will be confirmed by the catalyst vendor.

The vapours from 36C-103 are preheated in exchanger 36E-110 and further heated by MP steam in 36E-111 and enters H_2S separator, 36R-102X.

A temperature controller maintains the inlet temperature by controlling MP steam flow to 36E-111. The effluent from 36R-102X are cooled in 36E-110 and further condensed in 36E-112 by cooling water.

In the reactor 36R-102X, with the help of an analyzer, the total sulfur and COS content is continuously checked. Partial flooding of condenser 36E-112 maintains the system pressure at a constant level. 36E-112 gets flooded up to the desired level to match the duty requirement. The propylene is further taken ahead to the drying unit where it is dried using N_2 and filtered and later stored.

The detailed process flow diagram is given in figure 4.4, which has been created from the data obtained. The software used for making the PFD is MS Visio.

4.5 CONTROL PHILOSOPHY OF PRU

DE-PROPANISER

The feed stream to the column 36C-101 is being connected to a flow controller which helps to keep the flow rate intact, here the flow controller detects the difference in the flow rate in stream 1, compares it to the set point that has been provided and maintains the flow according to it. The top product has been connected to a Pressure controller which is controlled by the reboiler 36E-101. When a variation in pressure is detected, the PC sends a signal to the valve which will open or close the flow of LP steam into the reboiler which thereby helps in maintaining the pressure.

To the bottom of the tower, a cascade system is implemented where the level controller is being connected to a flow control, when there is increase or decrease in the level inside the column the level controller gives instructions to the flow controller to open or close the valve attached to the stream 3 so as to maintain the required Liquid level. An additional flow controller has been provided at the reflux stream after 36E-102, this also serves the same purpose that is to control the flow rate of the reflux stream entering back into the column.

COS HYDROLYSER

A temperature controller is implemented after the heat exchanger 36E-104 before entering 36R-101, the temperature controller detects the temperature variation and instructs the control valve to control the flow of MP stream to 36E-104 and henceforth maintaining the temperature.

The effluent coming from the hydrolyzer is further taken to 36V-101, where a level controller is installed, when there is a change in the level inside the column the level controller gives instructions to the flow controller installed in stream 14 and thereby opening or closing the control valve. The control system provided here is cascade controller. The level controller sets the set point for flow controller.

DE-ETHANISER

In de-ethaniser column 36C-102, the feed is sent to the column under flow control. The set point of the flow controller is controlled by the cracked LPG feed from 36C-101. Reflux to 36C-102 is controlled by another flow controller. Flue gas obtained as the top product from 36C-102, which goes to the boundary limit via the flow controller provided at the 15th stream. C_3 hydrocarbon stream is drawn as the bottom product from the column, 36C-102 under level /flow rate control. In here, level control controls the set point of the flow rate controller.

P/P SPLITTER

The C_3 hydrocarbon bottom product from 36C-102 is sent to 36C-103as feed via flow controller. Propane is the bottom product which goes to LPG return pool via the flow controller. The set point for this flow controller is obtained from level controller installed at the bottom of the 36C-103. PP splitter operates at a pressure of 20- 21kg/cm^2 and is maintained by a pressure controller by means of controlling the hot water flow to 36E-108. A flow controller is installed to the reflux of the column to maintain the reflux flow.

H_2S SEPARATOR

In 36R-102X, temperature controller maintains the inlet temperature of the hydrocarbon stream by approximately controlling the MP steam flow to 36E-111. If any variation in the temperature is detected, the controller detects it and increase or decrease the steam flow.

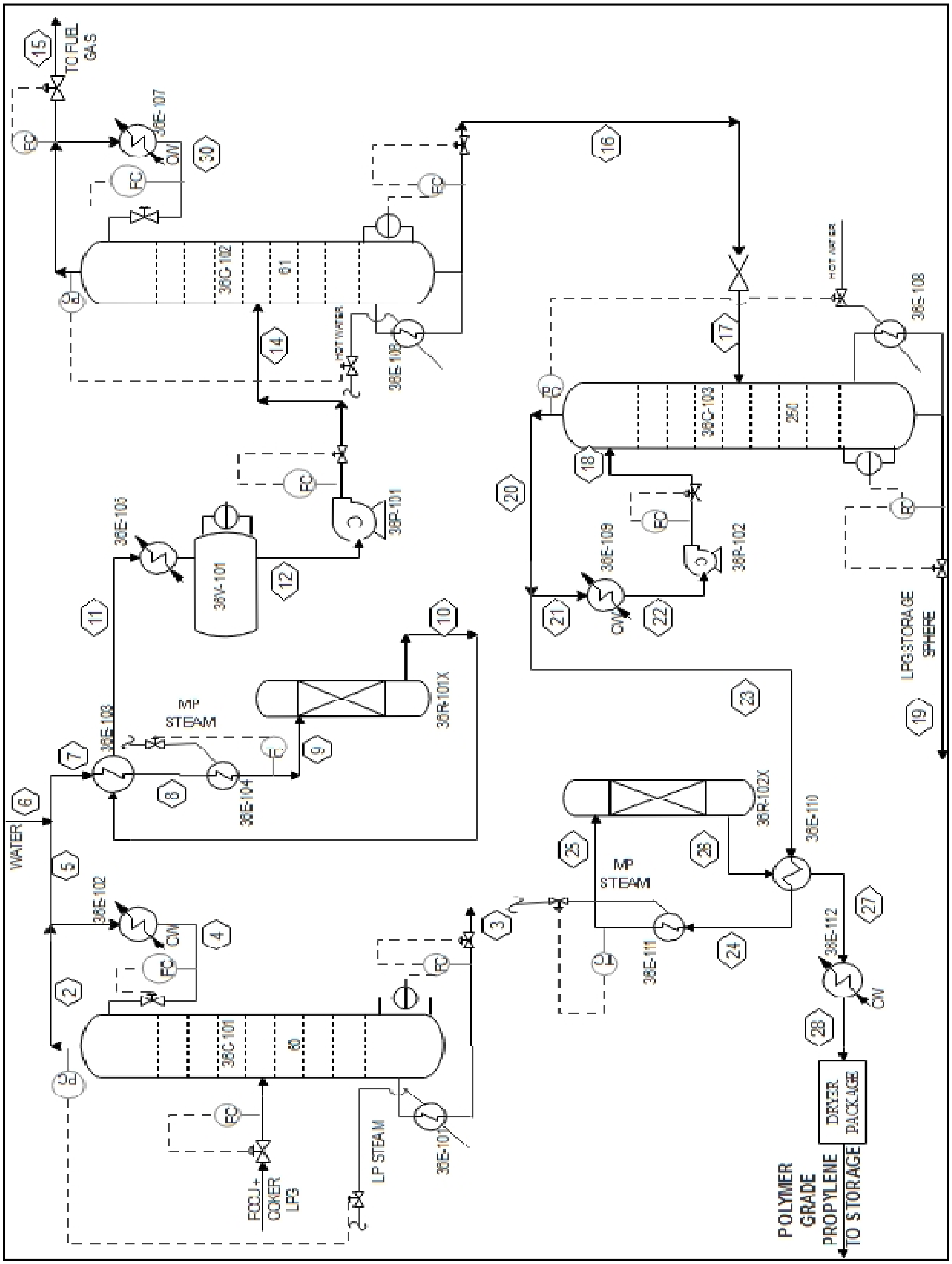

Figure 4.4: Process flow diagram of Propylene Recovery unit, Vadinar refinery

CHAPTER 5

CONCLUSIONS

The overall aim of this project is to achieve an understanding in designing a Propylene unit. EOL aims at commissioning a PRU unit, in order to effectively use the unsaturated naphtha obtained from FCCU and coker units. The propylene market is experiencing a hike and this process could thus economically benefit any industry that is a producer of propylene.

The design criterion and objectives for the project were fixed from the findings obtained as a result of literature survey and the product requirement studies. After completing the basic design of Propylene Recovery Unit (PRU) in Hysys, a yield of 99.58% by volume pure propylene was obtained, whereas originally a yield of 99.5% by volume was expected as per the EIL's technology. Hence it can be concluded that if the conditions used in the simulation are followed while constructing and running the plant, a better result can be expected. The project effectively demonstrates an improvement in the quality of yield by a factor of 0.08% which can prove to be quite substantial in the present industrial scale.

Detailed Process Data Sheets and Heat & Mass Balance Data study have also been done and the data obtained is given in Appendix 1 and Appendix 2.

REFERENCES

Ahmed Bhran, A., & M. Mohamed El-Gharbawy (2016). Modification of a deethanization plant for enhancing propane and propylene recovery and solving some operational problems. Journal of Natural Gas Science and Engineering, **31**, p. 503–514.

Amir Farshi (2008). Propylene production methods and FCC process rules in propylene demands. Research gate.

Ana Ribeiro, **M.**, **C. Marta Campo**, **C. Guler Narin**, **Joao Santos**, **Alexandre Ferreira**, **Jong-San Chang**, **Young Kyu Hwang**, **You-Kyong Seo**, **U-Hwang Lee**, **M. Jose Loureiro**, **E. Alírio Rodrigues** (2013). Pressure swing adsorption process for the separation of nitrogen and propylene with a MOF adsorbent MIL-100(Fe). Separation and purification technology, **110**, p. 101-111.

Banks, J., J. Carson, **B. Nelson, D. Nicol** (2001). Discrete-Event System Simulation. Prentice Hall, p. 3.

Campo, **M.C., A.M. Ribeiro**, **A. Ferreira**, **J.C. Santos**, **C. Lutz**, **J.M. Loureiro**, **A.E. Rodrigues** (2013). New 13X zeolite for propylene/propane separation by vacuum swing adsorption. Separation principle and purification technology, **103**, p. 60-70.

Dutta, B.K (1st ed. 2006). Heat Transfer-Principles and Application, PHI Pvt. Ltd., New Delhi .

Gary, J.H., & **G.E Handwerk** (2nd ed. 1984). Petroleum Refining Technology and Economics. Marcel Dekker.

Jagadisan Viswanathan & **E. Ignacio Grossmann** (1993). Optimal feed locations and number of trays for distillation columns with multiple feeds. Engineering Design Research Center', Carnegie Mellon University. Pittsburgh, Pennsylvania, **32**, p. 2942-2949.

Kern, D. Q. (1965). Process Heat Transfer. McGraw-Hill Book Company, Int.

Leffler, W.L., (2nd ed. 1985). Petroleum refining for the nontechnical person. PennWell Books.

Mona Khalighi, **S. Farooq**, **I.A. Karimi** (2012). Optimizing the PSA process of propylene/propane using Neuro-Fuzzy modeling. 11th International Symposium on Process Systems Engineering, **31**, p. 1336–1340.

Murdock & **W. James**(1993). Fundamental fluid mechanics for the practicing engineer. CRC Press, p. 25–27

Niranjan Kodanda, Design of Distillation column, p.123

Oguz Emrah Turgut, **Mert Sinan Turgut**, **Mustafa urhan Coban** (2014),'Design And Economic Investigation Of Shell And Tube Heat Exchangers Using Improved Intelligent Tuned Harmony Search Algorithm', Department of Mechanical Engineering, Faculty of Engineering, Ege University, Bornova, and Dokuz Eylul University, Tınaztepe,Izmir, Turkey

Plaza, **M.G.,** **A.M. Ribeiro**, **Ferreira**, **J.C. Santos**, **U-Hwang Lee**, **Jong-San Chang**, **J.M. Loureiro**, **A.E. Rodrigues** (2016), 'Propylene/propane separation by vacuum swing adsorption using Cu-BTC spheres', **90**, p.109-119

Rafael Alcantara-Avila, **J.,** **I. Fernando Gomez-Castro**, **J. Gabriel Segovia-Hernandez**, **Ken-Ichiro Sotowa**, **Toshihide Horikawa** (2014). Optimal design of cryogenic distillation columns with side heat pumps for the propylene/propane separation. Chemical Engineering and Processing: Process Intensification, **82**, p. 112-122.

Ray Sinnott & **Gavin Towler** (5th ed. 2009). Chemical Eng. Design, Elsevier.

Robert Perry, H. & **H. Cecil Chilton** (1895-1953). Chemical Eng. Handbook. McGraw Hill.

Robert Treybal, **E.** (1956)**.** Mass transfer operations (3rd ed.), McGraw Hill, p.439

Sinnott, R. K., Coulson & **Richardsons** (1993). Chemical Engineering: Chemical Engineering Design , Butterworth-Heinemann, **6**

Susheela Bai, **S. Sridhar**, **A. A. Khan** (2000). Recovery of propylene from refinery off-gas using metal incorporated ethyl cellulose membranes. Journal of membrane science, **174**, p. 67-79.

Underwood, **A. J. V.** (1948, 1949). Che. Eng. Prog., p.44; 45

APPENDIX 1

PROCESS DATA SHEET

DISTILLATION COLUMN DATA SHEET

Table A1.1: PDS OF De-propaniser column

PROCESS DATA SHEET FOR COLUMN TRAYS					
1	CLIENT:	EOL	DESCRIPTION	Distillation Column	
2	PLANT:	PRU	TAG NO.:	C-101	
3	COLUMN TYPE:	Vertical	SERVICE :	De-propaniser column	
4					
5	COLUMN CHARACTERISTICS				
6	Existing Tower			No	
7	Inside Dia		mm	1500	
8	Type Of Trays In Section			Sieve	
9	Tray Spacing		mm	610	
10	Total Trays In Column:			60	
11	Feed Location At:			30	
12	TRAY CHARACTERISTICS				
13				TOP	BOTTOM
14	Tray Location			2	5
15	Number Of Trays			30	30
16	Tray Spacing		mm	610	610
17	Pressure		$kg/cm^2(g)$	18.53	18.45
18	Liquid Temperature		°C	47.19	103.1
19	Vapor Temperature		°C	47.19	103.1
20	Vapor To Tray(S)				
21		Mass Rate	kg/hr	70410	95890
22		Molal Rate	kgmol/hr	1647	1683
23		Volume Rate	m^3/hr	136	162
24		Absolute Density	kg/ m^3	517.72	591.91
25		Molecular Weight		42.9	57.1
26	Liquid From Tray(S)				

27	Mass Rate	kg/hr	48500	12910
28	Volume Rate	m^3/hr	93.62	217.5
29	Surface Tension	Dynes/cm	4.24	3.45
30	Absolute Density	kg/m^3	462.00	463
31	Foaming Tendency		n/a	
32	Viscosity	cP	n/a	
33	Pressure Drop Limitations	kg/cm^2(g)	16.62	
34	Tray Material		n/a	
35	Tray Thickness	mm	n/a	
36	OPERATING DATA			
37	Feed Rate	kg/hr	55120	
38	Feed Temperature	°C	64	
39	Reflux	kg/hr	48080	
40	Temperature			
41	Top	°C	46.72	
42	Bottom	°C	106.4	
43	Pressure			
44	Top	kg/cm^2(g)	20.33	
45	Bottom	kg/cm^2(g)	20.77	
46	Condenser Duty	Gcal/hr	-3.36	
47	Reboiler Duty	Gcal/hr	5.65	
	Notes:			
PROCESS DATA SHEET FOR CRUDE COLUMN INTERNALS			Document Number 1	
Tag No.: C-101			Page no. :1	

Table A1.2: PDS of De-ethaniser column

PROCESS DATA SHEET FOR COLUMN TRAYS				
1	CLIENT:	EOL	DESCRIPTION	Distillation Column
2	PLANT:	PRU	TAG NO.:	C-102
3	COLUMN TYPE:	Vertical	SERVICE :	De-ethaniser column
4				
5	COLUMN CHARACTERISTICS			
6	Existing Tower			No

7	Inside Dia	mm	1500	
8	Type Of Trays In Section		Sieve	
9	Tray Spacing	mm	610	
10	Total Trays In Column:		61	
11	Feed Location At:		22	
12	TRAY CHARACTERISTICS			
13			TOP	BOTTOM
14	Tray Location		22	2
15	Number Of Trays		22	39
16	Tray Spacing	mm	610	610
17	Pressure	$kg/cm^2(g)$	28.64	28.83
18	Liquid Temperature	°C	67.21	68.13
19	Vapor Temperature	°C	67.21	68.13
20	Vapor To Tray(S)			
21	Mass Rate	kg/hr	41040	51090
22	Molal Rate	kgmol/hr	965.3	1197
23	Volume Rate	m^3/hr	79.27	98.52
24	Absolute Density	$kg/\ m^3$	517.72	518.57
25	Molecular Weight		42.6	42.7
26	Liquid From Tray(S)			
27	Mass Rate	kg/hr	71820	72780
28	Volume Rate	m^3/hr	138.7	140.3
29	Surface Tension	Dynes/cm	2.07	2
30	Absolute Density	kg/m^3	415.00	413
31	Foaming Tendency		n/a	
32	Viscosity	cP	n/a	
33	Pressure Drop Limitations	$kg/cm^2(g)$	0.0054	
34	Tray Material			
35	Tray Thickness	mm	n/a	
36	OPERATING DATA			
37	Feed Rate	kg/hr	22110	
38	Feed Temperature	°C	42.07	
39	Reflux	kg/hr	35400	
40	Temperature			
41	Top	°C	52.15	
42	Bottom	°C	68.33	
43	Pressure			

44	Top	$kg/cm^2(g)$	31.09
45	Bottom	$kg/cm^2(g)$	31.63
46	Condenser Duty	Gcal/hr	-2.14
47	Reboiler Duty	Gcal/hr	2.67
Notes:			
PROCESS DATA SHEET FOR CRUDE COLUMN INTERNALS			Document Number 1
Tag No.: C-102			Page no. :2

Table A1.3: PDS of P/P Splitter

PROCESS DATA SHEET FOR COLUMN TRAYS					
1	CLIENT:	EOL	DESCRIPTION	Distillation Column	
2	PLANT:	PRU	TAG NO.:	C-103	
3	COLUMN TYPE:	Vertical	SERVICE :	P/P Splitter	
4					
5	COLUMN CHARACTERISTICS				
6	Existing Tower			No	
7	Inside Dia		mm	1500	
8	Type Of Trays In Section			Sieve	
9	Tray Spacing		mm	610	
10	Total Trays In Column:			250	
11	Feed Location At:			181	
12	TRAY CHARACTERISTICS				
13				TOP	BOTTOM
14	Tray Location			181	2
15	Number Of Trays			181	69
16	Tray Spacing		mm	610	610
17	Pressure		$kg/cm^2(g)$	20.95	21.28
18	Liquid Temperature		°C	52.08	59.02
19	Vapor Temperature		°C	52.08	59.02
20	Vapor To Tray(S)				
21		Mass Rate	kg/hr	287100	298900
22		Molal Rate	kgmol/hr	6739	6794
23		Volume Rate	m^3/hr	553	587.8

24	Absolute Density	kg/ m^3	519.17	508.51
25	Molecular Weight		42.6	44
26	Liquid From Tray(S)			
27	Mass Rate	kg/hr	290100	306700
28	Volume Rate	m^3/hr	559.2	602.6
29	Surface Tension	Dynes/cm	3.64	3.14
30	Absolute Density	kg/m^3	452.00	508.96
31	Foaming Tendency		n/a	
32	Viscosity	cP	n/a	
33	Pressure Drop Limitations	kg/cm^2(g)	0.0052	
34	Tray Material		n/a	
35	Tray Thickness	mm	n/a	
36	OPERATING DATA			
37	Feed Rate	kg/hr	21560	
38	Feed Temperature	°C	52.87	
39	Reflux	kg/hr	260300	
40	Temperature			
41	Top	°C	48.45	
42	Bottom	°C	59.74	
43	Pressure			
44	Top	kg/cm^2(g)	22.05	
45	Bottom	kg/cm^2(g)	23.57	
46	Condenser Duty	Gcal/hr	17.87	
47	Reboiler Duty	Gcal/hr	18.52	
Notes:				
PROCESS DATA SHEET FOR CRUDE COLUMN INTERNALS			Document Number 1	
Tag No.: C-103			Page no. :3	

PUMPS DATA SHEET

Table A1.4: PDS OF P-101

PUMP DATASHEET					
1	Unit: PRU	Client : EOL	Corrosive due to :		Toxic due to:
2	Item No. :	P-101	Service	: HC pump	PFD:

3	Operating :	1	Standby	:1	Type Of Pump:		Centrifugal
4	PROPERTIES OF LIQUID						
5	Liquid Handled						HC
6	Pumping Temperature					°C	41.76
7	Viscosity At Pumping Temperature					cP	0.0654
8	Vapour Pressure At Pumping Temperature					$kg/cm^2(g)$	15.66
9	Liquid Density At Pumping Temperature					kg/m^3	472.1
10	Presence Of Corrosive / Toxic Components						n/a
11	Solids In Suspension						n/a
12	Pour Point (For Congealing Service)					°C	n/a
13	OPERATING CONDITIONS FOR ONE PUMP						
14	Flow Rate	Normal				m^3/hr	46.83
15		Maximum				m^3/hr	70.25
16		Minimum				m^3/hr	23.42
17	Suction Pressure					$kg/cm^2(g)$	n/a
18	Discharge Pressure					$kg/cm^2(g)$	n/a
19	Differential Pressure					$kg/cm^2(g)$	n/a
20	Differential Head					m	n/a
21	NPSH Available					M	26.48
22	CAPACITY CONTROL FOR VOLUMETRIC PUMPS						
23	Method Of Control						n/a
24	Type Of Control						n/a
25	Control Range						n/a
26	Precision At Minimum Rate						n/a
27	MECHANICAL DATA						
28	Design Pressure					$kg/cm^2(g)$	
30	Design Temperature					°C	44.06
31	Material Of Construction						
32		Casing					
33	IMPELLER/ PISTON/ PLUNGER						
34	Seal Type						
35	Line Rating						
36		Suction					
37		Discharge					
38	Driver Type						Motor
39	STEAM TURBINE DATA						
40	Inlet Pressure (Min./Nor./Max.)					$kg/cm^2(g)$	n/a
41	Inlet Temperature (Min./Nor./Max.)					°C	n/a
42	Design Pressure					$kg/cm^2(g)$	n/a
43	Design Temperature					°C	n/a
44	Exhaust Pressure (Min./Nor./Max.)					$kg/cm^2(g)$	n/a
45	Line Rating						n/a

46	Inlet	n/a
47	Outlet	n/a
	NOTES:	
1	DEC / VENDOR to furnish the balance required data	
2	Suction pressure, Discharge pressure, Differential pressure, Differential head are obtained after detailed hydraulic calculation	
3	Design pressure will depend on the discharge pressure	

0		Issued for detail engineering	
Rev. No.	Date	Purpose	Prepared By

Table A1.5: PDS OF P-102

PUMP DATASHEET							
1	Unit: PRU	Client : EOL		Corrosive due to :			Toxic due to:
2	Item No. :	P-101		Service:	HC pump		PFD:
3	Operating :		1	Standby	:1	Type Of Pump:	Centrifugal

4	PROPERTIES OF LIQUID			
5	Liquid Handled			HC
6	Pumping Temperature		°C	47.91
7	Viscosity At Pumping Temperature		Cp	0.04829
8	Vapour Pressure At Pumping Temperature		$kg/cm^2(g)$	15.66
9	Liquid Density At Pumping Temperature		kg/m^3	463.8
10	Presence Of Corrosive / Toxic Components			n/a
11	Solids In Suspension			n/a
12	Pour Point (For Congealing Service)		°C	n/a
13	OPERATING CONDITIONS FOR ONE PUMP			
14	Flow Rate	Normal	m^3/hr	561.23
15		Maximum	m^3/hr	841.85
16		Minimum	m^3/hr	280.62
17	Suction Pressure		$kg/cm^2(g)$	n/a
18	Discharge Pressure		$kg/cm^2(g)$	n/a
19	Differential Pressure		$kg/cm^2(g)$	n/a
20	Differential Head		m	n/a
21	NPSH Available		m	26.48
22	CAPACITY CONTROL FOR VOLUMETRIC PUMPS			
23	Method Of Control			n/a
24	Type Of Control			n/a
25	Control Range			n/a
26	Precision At Minimum Rate			n/a

27	MECHANICAL DATA		
28	Design Pressure	$kg/cm^2(g)$	
30	Design Temperature	°C	47.91
31	Material Of Construction		
32	Casing		
33	Impeller/ Piston/ Plunger		
34	Seal Type		
35	Line Rating		
36	Suction		
37	Discharge		
38	Driver Type		Motor
39	STEAM TURBINE DATA		
40	Inlet Pressure (Min./Nor./Max.)	$kg/cm^2(g)$	n/a
41	Inlet Temperature (Min./Nor./Max.)	°C	n/a
42	Design Pressure	$kg/cm^2(g)$	n/a
43	Design Temperature	°C	n/a
44	Exhaust Pressure (Min./Nor./Max.)	$kg/cm^2(g)$	n/a
45	Line Rating		n/a
46	Inlet		n/a
47	Outlet		n/a
	NOTES:		
1	DEC / VENDOR to furnish the balance required data		
2	Suction pressure, Discharge pressure, Differential pressure, Differential head are obtained after detailed hydraulic calculation		
3	Design pressure will depend on the discharge pressure		

0		Issued for detail engineering	
Rev. No.	Date	Purpose	Prepared By

HEAT EXCHANGER DATA SHEET

Table A1.6: PDS OF E-101

HEAT EXCHANGER DATA SHEET									
1		Client: EOL			Plant : PRU				
2		Service Name	De propaniser Reboiler						
3		Type :		Horizontal/ ~~Vertical~~	Connected in :	Parallel	1	Series	2

4		Size:					
5		Surface / Unit (m^2) (Eff.)	60.95	Shells/ Unit	1	Surface area / Shell (m^2) (Eff.)	60.95
6		Performance of the Unit					
7		Fluid Allocation		Shell Side		Tube Side	
8		Fluid		LP STEAM		HC (De propaniser bottom)	
9		Fluid Quantity (Total)	kg/h	20000		127900	
10				Inlet	Outlet	Inlet	Outlet
11		Vapor	kg/h	20000	9167	n/a	93780
12		Density	kg/m^3	1.986	1.85	n/a	46.33
13		Molecular Weight	kg/kgmole	18.02	18.02	n/a	57.29
14		Viscosity	cP	0.01366	0.01347	n/a	0.0108
15		Thermal Conductivity	W/m°C	0.02909	0.02758	n/a	0.0250
16		Specific Heat	kJ/kgmole °C	35.28	35.19	n/a	136
17		Liquid	kg/h	n/a	n/a	127900	34480
18		Density	kg/m^3	n/a	n/a	464.20	467.7
19		Viscosity	cP	n/a	n/a	0.0747	0.0747
20		Thermal Conductivity	W/m°C	n/a	n/a	0.0600	0.0608
21		Specific Heat	kJ/kgmole °C	n/a	n/a	196.30	198.3
22		Surface Tension	dyne/cm	n/a	n/a	3.4600	3.567
23		Temperature	°C	143.0	138.2	104.3	106.2
24		Inlet Pressure	$kg/cm^2(g)$	3.8	3.5	18.8	19.3
25		Pressure Drop-Allowable/Calculated	$kg/cm^2(g)$	0.3		0.3	
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a	
27		Heat Exchanged (Normal)	kcal/h	5650000			
28		LMTD	^oC	35.23			
29		Heat Transfer rate-Fouled / Clean	$kJ/m^2 \cdot h^oc$	11010			
30		Notes:					
31	1	Detail sizing to be done during detail engineering.					

32	2	20% overdesign margin to be considered on flow rate and duty.						
A	Rev	Date	Revision description	Ini.	Sign.	Ini.	Ini.	Sign.
B				Prepared by		Checked by	Approved by	
C		Shell & Tube Exchanger PDS		Document Number :3				
D								
E		Tag No: E-101		Document Contains 12 Pages			Page 1	

Table A1.7: PDS OF E-102

HEAT EXCHANGER DATA SHEET										
1		Client: EOL				Plant : PRU				
2		Service Name	De propaniser Cooler							
3		Type :	BHU	Horizontal/ ~~Vertical~~		Connected in :	Parallel	1	Series	2
4		Size:								
5		Surface / Unit (m^2) (Eff.)	60.34	Shells/ Unit	1	Surface area / Shell (m^2) (Eff.)			60.34	
6		Performance of the Unit								
7		Fluid Allocation				Shell Side		Tube Side		
8		Fluid				HC (De propaniser top)		Cooling Water		
9		Fluid Quantity (Total)			kg/h	481840		2000		
10						Inlet	Outlet	Inlet	Outlet	
11		Vapor			kg/h	481840	22270	n/a	n/a	
12		Density			kg/m^3	40.43	39.69	n/a	n/a	
13		Molecular Weight			kg/kgmole	42.63	42.54	n/a	n/a	
14		Viscosity			cP	0.01025	0.01020	n/a	n/a	
15		Thermal Conductivity			W/m°C	0.02189	0.02174	n/a	n/a	
16		Specific Heat			kJ/kgmole °C	91.5	90.48	n/a	n/a	
17		Liquid			kg/h	n/a	25870	2000	2000	
18		Density			kg/m^3	n/a	464.4	1003	996.1	
19		Viscosity			cP	n/a	0.06298	0.7804	0.6514	
20		Thermal Conductivity			W/m°C	n/a	0.09156	0.6196	0.6315	

21		Specific Heat	kJ/kgmole °C	n/a	143.3	79.89	79.88
22		Surface Tension	dyne/cm	n/a	4.369	71.06	69.49
23		Temperature	°C	46.9	45.7	31.0	40.0
24		Inlet Pressure	$kg/cm^2(g)$	18.9	23.7	4.5	4.3
25		Pressure Drop-Allowable/Calculated	$kg/cm^2(g)$	0.29		0.29	
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a	
27		Heat Exchanged (Normal)	kJ/h	79790			
28		LMTD	°C	10.2			
29		Heat Transfer rate-Fouled / Clean	$kJ/m^2 \cdot h^o c$	128.2			
30		Notes:					
31	1	Detail sizing to be done during detail engineering.					
32	2	20% overdesign margin to be considered on flow rate and duty.					
A	Rev	Date	Revision description	Ini.	Sign.	Ini.	Ini. | Sign
B				Prepared by		Checked by	Approved by
C		Shell & Tube Exchanger PDS		Document Number :3			
D		Tag No: E-102		Document Contains 12 Pages			Page 2

Table A1.8: PDS OF E-103

HEAT EXCHANGER DATA SHEET									
1		Client: EOL			Plant : PRU				
2		Service Name	COS Hydrolyser Heat exchanger						
3		Type :	BHU	Horizontal/ ~~Vertical~~	Connected in :	Parallel	1	Series	2
4		Size:							
5		Surface / Unit (m^2) (Eff.)	120	Shells/ Unit	1	Surface area / Shell (m^2) (Eff.)	120		
6		Performance of the Unit							
7		Fluid Allocation			Shell Side		Tube Side		
8		Fluid			HC (De propaniser top)		HC (COS hydrolyser top)		

9		Fluid Quantity (Total)	kg/h	21980		22110	
10				Inlet	Outlet	Inlet	Outlet
11		Vapor	kg/h	21980	21980	22110	22110
12		Density	kg/m^3	40.84	25.74	24.70	34.83
13		Molecular Weight	kg/kgmole	42.4	42.40	42.65	42.65
14		Viscosity	cP	n/a	0.01228	0.01286	0.0105
15		Thermal Conductivity	W/m°C	n/a	0.03000	0.03248	0.0233
16		Specific Heat	kJ/kgmole °C	12260	12260	12260	12260
17		Liquid	kg/h	n/a	n/a	n/a	n/a
18		Density	kg/m^3	n/a	n/a	n/a	n/a
19		Viscosity	cP	n/a	n/a	n/a	n/a
20		Thermal Conductivity	W/m°C	n/a	n/a	n/a	n/a
21		Specific Heat	kJ/kgmole °C	n/a	n/a	n/a	n/a
22		Surface Tension	dyne/cm	n/a	n/a	n/a	n/a
23		Temperature	°C	46.5	128.8	150.0	64.2
24		Inlet Pressure	$kg/cm^2(g)$	18.9	18.5	18.9	18.5
25		Pressure Drop-Allowable/Calculated	$kg/cm^2(g)$	0.4		0.4	
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a	
27		Heat Exchanged (Normal)	kJ/h	2975000			
28		LMTD	^oC	62.81			
29		Heat Transfer rate-Fouled / Clean	$kJ/m^2 \cdot h^oc$	266.9			
30		Notes:					
31	1	Detail sizing to be done during detail engineering.					
32	2	20% overdesign margin to be considered on flow rate and duty.					

A	Rev	Date	Revision description	Ini.	Sign.	Ini.	Ini.	Sign.
B				Prepared by		Checked by	Approved by	
C		Shell & Tube Exchanger PDS		Document Number :3				
D								
E		Tag No: E-103		Document Contains 12 Pages			Page 3	

Table A1.9: PDS OF E-104

HEAT EXCHANGER DATA SHEET									
1		Client: EOL			Plant : PRU				
2		Service Name	COS Reboiler						
3		Type :	BHU	Horizontal/ ~~Vertical~~	Connected in :	Parallel	1	Series	2
4		Size:							
5		Surface / Unit (m^2) (Eff.)	60.32	Shells/ Unit	1	Surface area / Shell (m^2) (Eff.)	60.32		
6		Performance of the Unit							
7		Fluid Allocation		Shell Side		Tube Side			
8		Fluid		MP STEAM		HC (De propaniser top)			
9		Fluid Quantity (Total)	kg/h	20000		21980			
10				Inlet	Outlet	Inlet	Outlet		
11		Vapor	kg/h	20000	2000	21980	21980		
12		Density	kg/m^3	5.231	5.37	25.74	23.46		
13		Molecular Weight	kg/kgmole	18.02	18.02	42.40	42.4		
14		Viscosity	cP	0.01841	0.01737	0.0123	0.0128		
15		Thermal Conductivity	W/m°C	0.04053	0.03837	0.0300	0.0322		
16		Specific Heat	kJ/kgmole °C	37.32	37.21	90.64	92.75		
17		Liquid	kg/h	n/a	n/a	n/a	n/a		
18		Density	kg/m^3	n/a	n/a	n/a	n/a		
19		Viscosity	cP	n/a	n/a	n/a	n/a		
20		Thermal Conductivity	W/m°C	n/a	n/a	n/a	n/a		
21		Specific Heat	kJ/kgmole °C	n/a	n/a	n/a	n/a		
22		Surface Tension	dyne/cm	n/a	n/a	n/a	n/a		
23		Temperature	°C	259.0	235.4	128.8	148.8		
24		Inlet Pressure	kg/cm^2(g)	12.6	12.3	18.5	18.2		
25		Pressure Drop-Allowable/Calculated	kg/cm^2(g)	0.29		0.29			
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a			
27		Heat Exchanged (Normal)	kJ/h	962300					

28		LMTD °C		108.39				
29		Heat Transfer rate- Fouled / Clean $kJ/m^2 \cdot h^oc$		148.2				
30		Notes:						
31	1	Detail sizing to be done during detail engineering.						
32	2	20% overdesign margin to be considered on flow rate and duty.						
A	Rev	Date	Revision description	Ini.	Sign.	Ini.	Ini.	Sign.
B				Prepared by		Checked by	Approved by	
C		Shell & Tube Exchanger PDS		Document Number :3				
D		Tag No: E-104		Document Contains 12 Pages			Page 4	

Table A1.10: PDS OF E-105

HEAT EXCHANGER DATA SHEET											
1		Client: EOL				Plant : PRU					
2		Service Name	COS Cooler								
3		Type :	BHU	Horizontal/ ~~Vertical~~		Connected in :	Parallel	1	Series	2	
5		Surface / Unit (m^2) (Eff.)		60.44	1	Surface area / Shell (m^2) (Eff.)			60.44		
6		Performance of the Unit									
7		Fluid Allocation				Shell Side		Tube Side			
8		Fluid				HC (Hydrolyser bottom)		Cooling Water			
9		Fluid Quantity (Total) kg/h				22110		170000			
10						Inlet	Outlet	Inlet	Outlet		
11		Vapor kg/h				22110	22110	n/a	n/a		
12		Density kg/m^3				34.83	472.10	n/a	n/a		
13		Molecular Weight kg/kgmole				42.65	42.65	n/a	n/a		
14		Viscosity cP				0.01056	0.06540	n/a	n/a		
15		Thermal Conductivity W/m°C				0.02336	0.09387	n/a	n/a		
16		Specific Heat kJ/kgmole °C				87.79	137.80	n/a	n/a		
17		Liquid kg/h				n/a	n/a	170000	170000		

18		Density	kg/m^3	n/a	n/a	1003	995.3
19		Viscosity	cP	n/a	n/a	0.780	0.6386
20		Thermal Conductivity	W/m°C	n/a	n/a	0.619	0.6328
21		Specific Heat	kJ/kgmole °C	n/a	n/a	79.89	79.88
22		Surface Tension	dyne/cm	n/a	n/a	71.06	69.31
23		Temperature	°C	64.18	41.76	31.00	41.05
24		Inlet Pressure	$kg/cm^2(g)$	19.90	19.58	5.95	5.63
25		Pressure Drop-Allowable/Calculated	$kg/cm^2(g)$	0.31		0.31	
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a	
27		Heat Exchanged (Normal)	kJ/h	7573000			
28		LMTD	^oC	16.16			
29		Heat Transfer rate-Fouled / Clean	$kJ/m^2{\cdot}h^oc$	7769			
30		Notes:					
31	1	Detail sizing to be done during detail engineering.					
32	2	20% overdesign margin to be considered on flow rate and duty.					

A	0			Ini.	Sign.	Ini.	Ini.	Sign
B	Rev	Date	Revision description	Prepared by		Checked by	Approved by	
C		Shell & Tube Exchanger PDS		Document Number :3				
D		Tag No: E-105		Document Contains 12 Pages			Page 5	

Table A1.11: PDS OF E-106

HEAT EXCHANGER DATA SHEET									
1		Client: EOL			Plant : PRU				
2		Service Name	De ethaniser Reboiler						
3		Type :	BHU	Horizontal/ ~~Vertical~~	Connected in :	Parallel	1	Series	2
4		Size:							

5		Surface / Unit (m^2) (Eff.)	120.6	shell/ units	1	Surface area / Shell (m^2) (Eff.)	60.44
6		Performance of the Unit					
7		Fluid Allocation		Tube Side		Shell Side	
8		Fluid		HC (De ethaniser bottom)		Hot water	
9		Fluid Quantity (Total)	kg/h	72830		150000	
10				Inlet	Outlet	Inlet	Outlet
11		Vapor	kg/h	n/a	51190	n/a	n/a
12		Density	kg/m^3	n/a	69.48	n/a	n/a
13		Molecular Weight	kg/kgmole	n/a	42.75	n/a	n/a
14		Viscosity	cP	n/a	0.01181	n/a	n/a
15		Thermal Conductivity	W/m°C	n/a	0.02610	n/a	n/a
16		Specific Heat	kJ/kgmole °C	n/a	126.20	n/a	n/a
17		Liquid	kg/h	72830	21640	150000	150000
18		Density	kg/m^3	413.1	413.1	953	966.6
19		Viscosity	cP	0.04729	0.04783	0.2976	0.363
20		Thermal Conductivity	W/m°C	0.07809	0.0792	0.6780	0.6679
21		Specific Heat	kJ/kgmole °C	195	195.1	80.79	80.33
22		Surface Tension	dyne/cm	2.004	2.013	59.74	62.83
23		Temperature	°C	68.21	68.20	94.00	77.35
24		Inlet Pressure	kg/cm^2(g)	31.63	31.47	3.55	3.23
25		Pressure Drop-Allowable/Calculated	kg/cm^2(g)	0.32		0.15	
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a	
27		Heat Exchanged (Normal)	kJ/h	11170000			
28		LMTD	°C	16.05			
29		Heat Transfer rate-Fouled / Clean	kJ/$m^2 \cdot h$°c	5768			
30		Notes:					
31	1	Detail sizing to be done during detail engineering.					
32	2	20% overdesign margin to be considered on flow rate and duty.					

A	R e	Date	Revision description	Ini.	Sign.	Ini.	Ini.	Sign.
B				Prepared by		Checked	Approved by	

	v				by	
C		Shell & Tube Exchanger PDS		Document Number :3		
D						
E		Tag No: E-106		Document Contains 12 Pages		Page 6

Table A1.12: PDS OF E-107

HEAT EXCHANGER DATA SHEET										
1		Client: EOL				Plant : PRU				
2		Service Name	De-ethaniser Cooler							
3		Type :	BHU	Horizontal/ ~~Vertical~~		Connected in :	Parallel	1	Series	2
4		Size:								
5		Surface / Unit (m^2) (Eff.)	60.32	Shells/ Unit	1	Surface area / Shell (m^2) (Eff.)			60.32	
6		Performance of the Unit								
7		Fluid Allocation				Shell Side			Tube Side	
8		Fluid				Cooling Water			HC (De ethaniser top)	
9		Fluid Quantity (Total)	kg/h			270200			34710	
10						Inlet	Outlet		Inlet	Outlet
11		Vapor	kg/h			n/a	n/a		34710	2173
12		Density	kg/m^3			n/a	n/a		58.51	58.93
13		Molecular Weight	kg/kgmole			n/a	n/a		38.61	38.73
14		Viscosity	cP			n/a	n/a		0.0116	0.011
15		Thermal Conductivity	W/m°C			n/a	n/a		0.0256	0.025
16		Specific Heat	kJ/kgmole °C			n/a	n/a		97.31	96.74
17		Liquid	kg/h			270200	270200		n/a	32540
18		Density	kg/m^3			1003	995.3		n/a	424.2
19		Viscosity	cP			0.7804	0.6686		n/a	0.047
20		Thermal Conductivity	W/m°C			0.6196	0.6298		n/a	0.084
21		Specific Heat	kJ/kgmole °C			79.89	79.88		n/a	154.4
22		Surface Tension	dyne/cm			71.06	69.31		n/a	3.117

23		Temperature	°C	31.00	38.62	57.78	44.80
24		Inlet Pressure	$kg/cm^2(g)$	5.95	5.64	31.31	31.00
25		Pressure Drop- Allowable/Calculated	$kg/cm^2(g)$	0.31		0.31	
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a	
27		Heat Exchanged (Normal)	kJ/h	9121000			
28		LMTD	°C	16.33			
29		Heat Transfer rate- Fouled / Clean	$kJ/m^2 \cdot h^oc$	9259			
30		Notes:					
31	1	Detail sizing to be done during detail engineering.					
A	R	Date	Revision description	Ini.	Sign.	Ini.	Ini. / Sign.
C B	e v			Prepared by		Checked by	Approved by
C		Shell & Tube Exchanger PDS		Document Number :3			
D		Tag No: E-107		Document Contains 12 Pages		Page 7	

Table A1.13: PDS OF E-108

HEAT EXCHANGER DATA SHEET									
1		Client: EOL			Plant : PRU				
2		Service Name	PP Splitter Reboiler						
3		Type :	BHU	Horizontal/ ~~Vertical~~	Connected in:	Parallel	1	Series	2
4		Size:							
5		Surface / Unit (m^2) (Eff.)	60.31	Shells/ Unit	1	Surface area / Shell (m^2) (Eff.)		60.31	
6		Performance of the Unit							
7		Fluid Allocation			Shell Side		Tube Side		
8		Fluid			Hot Water		HC (De Ethaniser Bottom)		
9		Fluid Quantity (Total)	kg/h		850000		306800		
10					Inlet	Outlet	Inlet	Outlet	

11		Vapor	kg/h	n/a	298100	n/a	n/a
12		Density	kg/m^3	n/a	48.66	n/a	n/a
13		Molecular Weight	kg/kgmole	n/a	44.09	n/a	n/a
14		Viscosity	cP	n/a	0.01051	n/a	n/a
15		Thermal Conductivity	W/m°C	n/a	0.02424	n/a	n/a
16		Specific Heat	kJ/kgmole °C	n/a	109.4	n/a	n/a
17		Liquid	kg/h	850000	8676	306800	306800
18		Density	kg/m^3	431.50	431.9	953	969.6
19		Viscosity	cP	0.0706	0.07078	0	0.3811
20		Thermal Conductivity	W/m°C	0.0754	0.07529	0.6780	0.6652
21		Specific Heat	kJ/kgmole °C	167.60	167.5	81	80.25
22		Surface Tension	dyne/cm	3.14	3.161	60	63.5
23		Temperature	°C	59.15	59.18	94.00	73.65
24		Inlet Pressure	$kg/cm^2(g)$	23.30	23.18	3.55	3.23
25		Pressure Drop-Allowable/Calculated	$kg/cm^2(g)$	0.31		0.15	
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a	
27		Heat Exchanged (Normal)	kJ/h	77320000			
28		LMTD	°C	23.2			
29		Heat Transfer rate-Fouled / Clean	$kJ/m^2 \cdot h^oc$	55260			
30		Notes:					
31	1	Detail sizing to be done during detail engineering.					
32	2	20% overdesign margin to be considered on flow rate and duty.					

A	Rev	Date	Revision description	Ini.	Sign	Ini.	Ini.	Sign.
B				Prepared by		Checked by	Approved by	
C		Shell & Tube Exchanger PDS		Document Number :3				
D								
E		Tag No: E-108		Document Contains 12 Pages			Page 8	

Table A1.14: PDS OF E-109

HEAT EXCHANGER DATA SHEET								
1		Client: EOL		Plant : PRU				
2		Service Name	PP Splitter Cooler					
3		Type :	BHU	Horizontal/ ~~Vertical~~	Connected in :	Parallel 1	Series 2	
4		Size:						
5		Surface / Unit (m^2) (Eff.)	60.34	Shells/ Unit 1	Surface area / Shell (m^2) (Eff.)		60.34	
6			Performance of the Unit					
7		Fluid Allocation		Shell Side		Tube Side		
8		Fluid		Cooling Water		HC (Splitter top)		
9		Fluid Quantity (Total)	kg/h	1621000		260300		
10				Inlet	Outlet	Inlet	Outlet	
11		Vapor	kg/h	360200	n/a	260300	260300	
12		Density	kg/m^3	43.83	n/a	43.85	459.7	
13		Molecular Weight	kg/kgmole	42.08	n/a	42.08	42.08	
14		Viscosity	Cp	0.01043	n/a	n/a	n/a	
15		Thermal Conductivity	W/m°C	0.02199	n/a	n/a	n/a	
16		Specific Heat	kJ/kgmole °C	90.28	n/a	90.30	143.2	
17		Liquid	kg/h	115.2	1621000	n/a	n/a	
18		Density	kg/m^3	0	995	n/a	n/a	
19		Viscosity	cP	0.0478	0.6344	n/a	n/a	
20		Thermal Conductivity	W/m°C	144.2000	0.6332	n/a	n/a	
21		Specific Heat	kJ/kgmole °C	144.20	79.88	n/a	n/a	
22		Surface Tension	dyne/cm	3.95	69.25	n/a	n/a	
23		Temperature	°C	31.00	41.41	48.45	47.78	
24		Inlet Pressure	kg/cm^2(g)	5.95	5.64	22.05	21.74	
25		Pressure Drop-Allowable/Calculated	kg/cm^2(g)	0.31		0.31		
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a		
27		Heat Exchanged (Normal)	kJ/h	74770000				
28		LMTD	°C	11.21				

29		Heat Transfer rate-Fouled / Clean	kJ/m^2·h°c	110500					
30		Notes:							
31	1	Detail sizing to be done during detail engineering.							
32	2	20% overdesign margin to be considered on flow rate and duty.							
A	Rev	Date	Revision description	Ini.	Sign.	Ini.	Ini.	Sign.	
B				Prepared by		Checked by	Approved by		
C		Shell & Tube Exchanger PDS		Document Number :3					
D									
E		Tag No: E-109		Document Contains 12 Pages			Page 9		

Table A1.15: PDS OF E-110

HEAT EXCHANGER DATA SHEET s									
1		Client: EOL			Plant : PRU				
2		Service Name	H_2S Separate Heat exchanger						
3		Type :	BHU	Horizontal/ ~~Vertical~~	Connected in :	Parallel	1	Series	2
4		Size:							
5		Surface / Unit (m^2) (Eff.)	120	Shells/ Unit	1	Surface area / Shell (m^2) (Eff.)		120	
6		Performance of the Unit							
7		Fluid Allocation			Shell Side		Tube Side		
8		Fluid			HC (P/P splitter top)		HC (H2S Separate top)		
9		Fluid Quantity (Total) kg/h			13990		13990		
10					Inlet	Outlet	Inlet	Outlet	
11		Vapor kg/h			13990	13990	13990	13990	
12		Density kg/m^3			43.83	25.88	24.09	34.32	
13		Molecular Weight kg/kgmole			42.08	42.08	42.09	42.09	
14		Viscosity cP			0.01043	0.01302	0.01390	0.0111	
15		Thermal Conductivity W/m°C			0.02199	0.03200	0.03549	0.0246	
16		Specific Heat kJ/kgmole			90.28	89.44	92.90	84.62	

			°C				
17		Liquid	kg/h	n/a	n/a	n/a	n/a
18		Density	kg/m^3	n/a	n/a	n/a	n/a
19		Viscosity	Cp	n/a	n/a	n/a	n/a
20		Thermal Conductivity	W/m°C	n/a	n/a	n/a	n/a
21		Specific Heat	kJ/kgmole °C	n/a	n/a	n/a	n/a
22		Surface Tension	dyne/cm	n/a	n/a	n/a	n/a
23		Temperature	°C	48.45	149.60	180.00	79.39
24		Inlet Pressure	kg/cm^2(g)	22.05372	21.51609	22.0	21.4
25		Pressure Drop-Allowable/Calculated	kg/cm^2(g)	0.5379571		0.53795	
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a	
27		Heat Exchanged (Normal)	kJ/h	2929000			
28		LMTD	^{o}C	29.8			
29		Heat Transfer rate-Fouled / Clean	kJ/m^2·h^{o}c	1584			
32		Notes:					
33	1	Detail sizing to be done during detail engineering.					

A	Rev	Date	Revision description	Ini.	Sign.	Ini.	Ini.	Sign.
B				Prepared by		Checked by	Approved by	
C		Shell & Tube Exchanger PDS		Document Number: 3				
D								
E		Tag No: E-110		Document Contains 12 Pages			Page 10	

Table A1.16: PDS OF E-111

HEAT EXCHANGER DATA SHEET									
1		Client: EOL			Plant : PRU				
2		Service Name	H_2S Reboiler						
3		Type :	BHU	Horizontal/ ~~Vertical~~	Connected in :	Parallel	1	Series	2
4		Size:							

5		Surface / Unit (m^2) (Eff.)	60.3	Shells/ Unit	1	Surface area / Shell (m^2) (Eff.)	60.3
6		Performance of the Unit					
7		Fluid Allocation		Shell Side		Tube Side	
8		Fluid		MP Steam		HC (PP Splitter Top)	
9		Fluid Quantity (Total)	kg/h	20000		13990	
10				Inlet	Outlet	Inlet	Outlet
11		Vapor	kg/h	20000	2000	13990	13990
12		Density	kg/m^3	5.231	5.37	25.74	23.46
13		Molecular Weight	kg/kgmole	18.02	18.02	42.40	42.4
14		Viscosity	cP	0.01841	0.01737	0.0123	0.01282
15		Thermal Conductivity	W/m°C	0.04053	0.03837	0.0300	0.03228
16		Specific Heat	kJ/kgmole °C	37.32	37.21	89.44	92.63
17		Liquid	kg/h	n/a	n/a	n/a	n/a
18		Density	kg/m^3	n/a	n/a	n/a	n/a
19		Viscosity	cP	n/a	n/a	n/a	n/a
20		Thermal Conductivity	W/m°C	n/a	n/a	n/a	n/a
21		Specific Heat	kJ/kgmole °C	n/a	n/a	n/a	n/a
22		Surface Tension	dyne/cm	n/a	n/a	n/a	n/a
23		Temperature	°C	259.0	236.4	149.6	179.9
24		Inlet Pressure	kg/cm^2(g)	12.6	12.3	18.5	18.2
25		Pressure Drop- Allowable/Calculated	kg/cm^2(g)	0.29		0.29	
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a	
27		Heat Exchanged (Normal)	kJ/h	920500			
28		LMTD	^oC	81.51			
29		Heat Transfer rate- Fouled / Clean	kJ/$m^2 \cdot h^oc$	187.2			
30		Notes:					
31	1	Detail sizing to be done during detail engineering.					
32	2	20% overdesign margin to be considered on flow rate and duty.					

A	Re	Date	Revision description	Ini.	Sign.	Ini.	Ini.	Sign.
B				Prepared by		Checked	Approved by	

	v				by	
C		Shell & Tube Exchanger PDS		Document Number :3		
D		Tag No: E-111		Document Contains 12 Pages		Page 11

Table A1.17: PDS OF E-112

HEAT EXCHANGER DATA SHEET										
1		Client: EOL				Plant : PRU				
2		Service Name	H_2S Separate Cooler							
3		Type :	BHU	Horizontal/ ~~Vertical~~		Connected in :	Parallel	1	Series	2
4		Size:								
5		Surface / Unit (m^2) (Eff.)	60.32	Shells/ Unit	1	Surface area / Shell (m^2) (Eff.)		60.32		
6		Performance of the Unit								
7		Fluid Allocation				Shell Side		Tube Side		
8		Fluid				Cooling Water		HC (Separate bottom)		
9		Fluid Quantity (Total)	kg/h			162100		13990		
10						Inlet	Outlet	Inlet	Outlet	
11		Vapor	kg/h			n/a	n/a	13990	13990	
12		Density	kg/m^3			n/a	n/a	34.32	480.2	
13		Molecular Weight	kg/kgmole			n/a	n/a	42.09	42.09	
14		Viscosity	cP			n/a	n/a	0.0111	0.0551	
15		Thermal Conductivity	W/m°C			n/a	n/a	0.0246	0.0995	
16		Specific Heat	kJ/kgmole °C			n/a	n/a	84.62	132.4	
17		Liquid	kg/h			1621000	1621000	n/a	n/a	
18		Density	kg/m^3			1003	997.4	n/a	n/a	
19		Viscosity	Cp			0.7804	0.6725	n/a	n/a	
20		Thermal Conductivity	W/m°C			0.6196	0.6294	n/a	n/a	
21		Specific Heat	kJ/kgmole °C			79.89	79.88	n/a	n/a	
22		Surface Tension	dyne/cm			71.06	69.78	n/a	n/a	

23		Temperature	°C	31.00	38.34	79.39	40.36
24		Inlet Pressure	$kg/cm^2(g)$	5.95	5.64	21.42	21.10
25		Pressure Drop-Allowable/Calculated	$kg/cm^2(g)$	0.31		0.31	
26		Fouling Resistance	m^2 h °C/kcal	n/a		n/a	
27		Heat Exchanged (Normal)	kJ/h	5272000			
28		LMTD	^oC	21.445			
29		Heat Transfer rate- Fouled / Clean	$kJ/m^2 \cdot h^oc$	4688			
30		Notes:					
31	1	Detail sizing to be done during detail engineering.					
A	Rev	Date	Revision description	Ini.	Sign.	Ini.	Ini. / Sign.
B				Prepared by		Checked by	Approved by
C		Shell & Tube Exchanger PDS		Document Number :3			
D							
E		Tag No: E-112		Document Contains 12 Pages		Page 12	

APPENDIX 2

HEAT AND MASS BALANCE DATA SHEET

Table A2.1: HMB of material stream 1

	CONDITIONS				
			Over all	Liquid Phase	
	Vapour / Phase Fraction		0.0000	1.0000	
	Temperature	^{o}C	64.00	64.00	
	Pressure	kPa	1961	1961	
	Molar Flow	kgmole/h	1092	1092	
	Mass Flow	kg/h	$5.512X10^{4}$	$5.512X10^{4}$	
	Std Ideal Liquid Volume Flow	m^3/h	98.53	98.53	
	Molar Enthalpy	kJ/kgmole	$-5.302\ X10^{4}$	$-5.302\ X10^{4}$	
	Molar Entropy	kJ/kgmole^{o}C	87.55	87.55	
	Heat Flow	kJ/h	$-5.790X10^{7}$	$-5.790X10^{7}$	
	Liquid Volume Flow @Std Cond	m^3/h	97.47	97.47	
	PROPERTIES				
			Over all	Liquid Phase	
	Molecular Weight		50.47	50.47	
	Molar Density	kgmole/m^3	9.781	9.781	
	Mass Density	kg/m^3	493.6	493.6	
	Act. Volume Flow	m^3/h	111.7	111.7	
	Mass Enthalpy	kJ/kg	-1051	-1051	
	Std. Gas Flow	STD_m^3/h	$2.582X10^{4}$	$2.582X10^{4}$	
	Z Factor		$7.154X10^{-2}$	$7.154X10^{-2}$	
	Watson K		13.68	13.68	
	C_p/C_v		1.555	1.555	
	Surface Tension	dyne/cm	5.191	5.191	
	Thermal Conductivity	W/m-K	$7.979X10^{-2}$	$7.979X10^{-2}$	
	Viscosity	cP	$8.495X10^{-2}$	$8.495X10^{-2}$	
	Reid VP at 37.8 C	kPa	838.1	838.1	
	True VP at 37.8 C	kPa	881.1	881.1	

Table A2.2: HMB of material stream 2

CONDITIONS				
		Over all	Vapour Phase	Liquid Phase
Vapour / Phase Fraction		0.0045	0.0045	0.9955
Temperature	^{o}C	45.23	45.23	45.23
Pressure	kPa	1824	1824	1824
Molar Flow	kgmole/h	1128	5.050	1123
Mass Flow	kg/h	$4.808X10^{4}$	214.1	$4.787\ X10^{4}$
Std Ideal Liquid Volume Flow	m^3/h	93.37	0.4166	92.95
Molar Enthalpy	kJ/kgmole	$-3.351X10^{4}$	$-1.939X10^{4}$	$-3.357\ X10^{4}$
Molar Entropy	$kJ/kgmole^{o}C$	56.44	94.74	56.27
Heat Flow	kJ/h	$-3.780X10^{7}$	$-9.790\ X10^{4}$	$-3.770\ X10^{4}$
Liquid Volume Flow @Std Cond	m^3/h	92.92	0.4140	92.51
PROPERTIES				
		Over all	Vapour Phase	Liquid Phase
Molecular Weight		42.62	42.39	42.62
Molar Density	$kgmole/m^3$	10.41	0.9330	10.91
Mass Density	kg/m^3	443.7	39.55	465.0
Act. Volume Flow	m^3/h	108.4	5.413	102.9
Mass Enthalpy	kJ/kg	-786.2	-457.3	-787.7
Std. Gas Flow	STD_m^3/h	$2.667X10^{4}$	119.4	$2.655X10^{4}$
Z Factor		---	0.7386	$6.316X10^{-2}$
Watson K		14.38	14.41	14.38
C_p/C_v		1.110	1.418	1.641
Surface Tension	dyne/cm	4.426	---	4.426
Thermal Conductivity	W/m-K	---	$2.171X10^{-2}$	$9.202\ X10^{-2}$
Viscosity	Cp	---	$1.019X10^{-2}$	$6.279\ X10^{-2}$
True VP at 37.8 C	kPa	1543	1579	1543

Table A2.3: HMB of material stream 3

CONDITIONS				
		Over all	Liquid Phase	
Vapour / Phase Fraction		0.0000	1.0000	
Temperature	^{o}C	106.4	106.4	

Pressure	kPa	1893	1893	
Molar Flow	kgmole/h	577.7	577.7	
Mass Flow	kg/h	$3.317X10^{4}$	$3.317\ X10^{4}$	
Std Ideal Liquid Volume Flow	m^3/h	55.93	55.93	
Molar Enthalpy	kJ/kgmole	$-6.476X10^{4}$	$-6.476X10^{4}$	
Molar Entropy	kJ/kgmole^{o}C	119.3	119.3	
Heat Flow	kJ/h	$-3.741X10^{7}$	$-3.741X10^{7}$	
Liquid Volume Flow @Std Cond	m^3/h	55.58	55.58	
PROPERTIES				
		Over all	Liquid Phase	
Molecular Weight		57.42	57.42	
Molar Density	kgmole/m^3	8.042	8.042	
Mass Density	kg/m^3	461.8	461.8	
Act. Volume Flow	m^3/h	71.83	71.83	
Mass Enthalpy	kJ/kg	-1128	-1128	
Std. Gas Flow	STD_m^3/h	$1.366X10^{4}$	$1.366X10^{4}$	
Z Factor		$7.457X10^{-2}$	$7.457X10^{-2}$	
Watson K		13.24	13.24	
C_p/C_v		1.513	1.513	
Surface Tension	dyne/cm	3.340	3.340	
Thermal Conductivity	W/m-K	$5.940\ X10^{-2}$	$5.940X10^{-2}$	
Viscosity	cP	$7.36X10^{-2}$	$7.36X10^{-2}$	
True VP at 37.8 C	kPa	395.1	395.1	

Table A2.4: HMB of material stream 4

CONDITIONS				
		Over all	Vapour Phase	
Vapour / Phase Fraction		1.0000	1.0000	
Temperature	^{o}C	46.72	46.72	
Pressure	kPa	1853	1853	
Molar Flow	kgmole/h	1643	1643	
Mass Flow	kg/h	$7.003X10^{4}$	$7.003X10^{4}$	
Std Ideal Liquid Volume Flow	m^3/h	136.0	136.0	
Molar Enthalpy	kJ/kgmole	$-2.110X10^{4}$	$-2.110X10^{4}$	
Molar Entropy	kJ/kgmole^{o}C	95.48	95.48	

	Heat Flow	kJ/h	$-3.466X10^{7}$	$-3.466X10^{7}$	
	Liquid Volume Flow @Std Cond	m^3/h	135.3	135.3	
	PROPERTIES				
			Over all	Vapour Phase	
	Molecular Weight		42.63	42.63	
	Molar Density	$kgmole/m^3$	0.9495	0.9495	
	Mass Density	kg/m^3	40.48	40.48	
	Act. Volume Flow	m^3/h	1730	1730	
	Mass Enthalpy	kJ/kg	-494.9	-494.9	
	Std. Gas Flow	STD_m^3/h	$3.884X10^{4}$	$3.884\ X10^{4}$	
	Z Factor		0.7340	0.7340	
	Watson K		14.38	14.38	
	C_p/C_v		1.424	1.424	
	Surface Tension	dyne/cm	---	---	
	Thermal Conductivity	W/m-K	$2.189\ X10^{-2}$	$2.189X10^{-2}$	
	Viscosity	cP	$1.025\ X10^{-2}$	$1.025X10^{-2}$	
	Reid VP at 37.8 C	kPa	1529	1529	
	True VP at 37.8 C	kPa	1542	1542	

Table A2.5: HMB of material stream 5

	CONDITIONS				
			Over all	Vapour Phase	
	Vapour / Phase Fraction		1.0000	1.0000	
	Temperature	^oC	46.72	46.72	
	Pressure	kPa	1853	1853	
	Molar Flow	kgmole/h	513.5	513.5	
	Mass Flow	kg/h	$2.189X10^{4}$	$2.189X10^{4}$	
	Std Ideal Liquid Volume Flow	m^3/h	42.50	42.50	
	Molar Enthalpy	kJ/kgmole	$-2.110X10^{4}$	$-2.110X10^{4}$	
	Molar Entropy	$kJ/kgmole^oC$	95.48	95.48	
	Heat Flow	kJ/h	$-1.083X10^{7}$	$-1.083X10^{7}$	
	Liquid Volume Flow @Std Cond	m^3/h	42.30	42.30	
	PROPERTIES				

		Over all	Vapour Phase	
Molecular Weight		42.63	42.63	
Molar Density	kgmole/m^3	0.9495	0.9495	
Mass Density	kg/m^3	40.48	40.48	
Act. Volume Flow	m^3/h	540.8	540.8	
Mass Enthalpy	kJ/kg	-494.9	-494.9	
Std. Gas Flow	STD_m^3/h	1.214 X10^4	1.214 X10^4	
Act. Liquid. Flow	m^3/s	---	---	
Z Factor		0.7340	0.7340	
Watson K		14.38	14.38	
Cp/Cv		1.424	1.424	
Surface Tension	dyne/cm	---	---	
Thermal Conductivity	W/m-K	2.19X10^{-2}	2.19X10^{-2}	
Viscosity	cP	1.02X10^{-2}	1.025X10^{-2}	
Reid VP at 37.8 C	kPa	1529	1529	
True VP at 37.8 C	kPa	1542	1542	

Table A2.6: HMB of material stream 6

CONDITIONS				
		Over all	Aqueous Phase	
Vapour / Phase Fraction		0.0000	1.0000	
Temperature	^oC	105.0	105.0	
Pressure	kPa	1863	1863	
Molar Flow	kgmole/h	4.874	4.874	
Mass Flow	kg/h	87.80	87.80	
Std Ideal Liquid Volume Flow	m^3/h	8.798X10^{-2}	8.798X10^{-2}	
Molar Enthalpy	kJ/kgmole	-2.799X10^5	-2.799X10^5	
Molar Entropy	kJ/kgmole^oC	72.32	72.32	
Heat Flow	kJ/h	-1.364X10^6	-1.364X10^6	
Liquid Volume Flow @Std Cond	m^3/h	8.652X10^{-2}	8.652X10^{-2}	
PROPERTIES				
		Over all	Aqueous Phase	
Molecular Weight		18.02	18.02	
Molar Density	kgmole/m^3	52.43	52.43	
Mass Density	kg/m^3	944.5	944.5	
Act. Volume	m^3/h	9.296X10^{-2}	9.296X10^{-2}	

Flow			
Mass Enthalpy kJ/kg	$-1.554X10^{4}$	$-1.554X10^{4}$	
Act. Gas Flow ACT_m^3/h	---	---	
Std. Gas Flow STD_m^3/h	115.2	115.2	
Z Factor	$1.13X10^{-2}$	$1.13X10^{-2}$	
Watson K	---	---	
C_p/C_v	1.171	1.171	
Thermal Conductivity W/m-K	0.6826	0.6826	
Viscosity cP	0.2650	0.2650	
Reid VP at 37.8 C kPa	6.442	6.442	
True VP at 37.8 C kPa	6.442	6.442	

Table A2.7: HMB of material stream 7

CONDITIONS				
	Over all	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour / Phase Fraction	0.9842	0.9842	0.0108	0.0050
Temperature ^oC	46.50	46.50	46.50	46.50
Pressure kPa	1853	1853	1853	1853
Molar Flow kgmole/h	518.3	510.2	5.600	2.575
Mass Flow kg/h	$2.198\ X10^{4}$	$2.169\ X10^{4}$	239.4	46.38
Std Ideal Liquid Volume Flow m^3/h	42.59	42.08	0.4643	4.647e-002
Molar Enthalpy kJ/kgmole	$-2.353X10^{4}$	$-2.207X10^{4}$	$-3.637X10^{4}$	$-2.846X10^{5}$
Molar Entropy kJ/kgmole^{o}C	95.30	95.89	58.47	58.83
Heat Flow kJ/h	$-1.22X10^{7}$	$-1.126X10^{7}$	$-2.037X10^{5}$	$-7.327X10^{5}$
Liquid Volume Flow @Std Cond m^3/h	42.31	41.84	0.4622	$4.57X10^{-2}$
PROPERTIES				
	Over all	Vapour Phase	Liquid Phase	Aqueous Phase
Molecular Weight	42.40	42.52	42.75	18.02
Molar Density kgmole/m^3	0.9631	0.9489	10.84	55.04
Mass Density kg/m^3	40.84	40.35	463.4	991.5
Act. Volume Flow m^3/h	538.2	537.6	0.5167	$4.678X10^{-2}$

	Mass Enthalpy	kJ/kg	-555.0	-519.1	-850.7	-1.580×10^{4}
	Act. Gas Flow	ACT_m^3/h	537.6	537.6	---	---
	Std. Gas Flow	STD_m^3/h	1.226×10^{4}	1.206×10^{4}	132.4	60.87
	Z Factor		---	0.7350	6.434×10^{-2}	1.267×10^{-2}
	Watson K		14.38	14.38	14.37	9.714
	C_p/C_v		1.411	1.423	1.635	1.153
	Surface Tension	dyne/cm	---	---	4.371	68.35
	Thermal Conductivity	W/m-K	---	2.187×10^{-2}	9.122×10^{-2}	0.6393
	Viscosity	cP	---	1.023×10^{-2}	6.311×10^{-2}	0.5783
	Reid VP at 37.8 C	kPa	1529	1530	1510	6.446
	True VP at 37.8 C	kPa	1543	1543	1521	6.446

Table A2.6: HMB of material stream 8

	CONDITIONS				
			Over all	Vapour Phase	
	Vapour / Phase Fraction		1.0000	1.0000	
	Temperature	^{o}C	128.8	128.8	
	Pressure	kPa	1814	1814	
	Molar Flow	kgmole/h	518.3	518.3	
	Mass Flow	kg/h	2.198×10^{4}	2.198×10^{4}	
	Std Ideal Liquid Volume Flow	m^3/h	42.59	42.59	
	Molar Enthalpy	kJ/kgmole	-1.586×10^{4}	-1.586×10^{4}	
	Molar Entropy	$kJ/kgmole^{o}C$	116.9	116.9	
	Heat Flow	kJ/h	-8.223×10^{6}	-8.223×10^{6}	
	Liquid Volume Flow @Std Cond	m^3/h	42.31	42.31	
	PROPERTIES				
			Over all	Vapour Phase	
	Molecular Weight		42.40	42.40	
	Molar Density	$kgmole/m^3$	0.6071	0.6071	
	Mass Density	kg/m^3	25.74	25.74	
	Act. Volume Flow	m^3/h	853.8	853.8	
	Mass Enthalpy	kJ/kg	-374.1	-374.1	
	Act. Gas Flow	ACT_m^3/h	853.8	853.8	
	Std. Gas Flow	STD_m^3/h	1.226×10^{4}	1.226×10^{4}	
	Z Factor		0.8943	0.8943	
	Watson K		14.38	14.38	

C_p/C_v		1.179	1.179	
Surface Tension	dyne/cm	---	---	
Thermal Conductivity	W/m-K	$3.04X10^{-2}$	$3.04\ X10^{2}$	
Viscosity	cP	$1.23X10^{-2}$	$1.23X10^{-2}$	
Reid VP at 37.8 C	kPa	1529	1529	
True VP at 37.8 C	kPa	1543	1543	

Table A2.9: HMB of material stream 9

CONDITIONS				
		Over all	Vapour Phase	
Vapour / Phase Fraction		1.0000	1.0000	
Temperature	^oC	148.9	148.9	
Pressure	kPa	1795	1795	
Molar Flow	kgmole/h	518.3	518.3	
Mass Flow	kg/h	$2.198\ X10^{4}$	$2.198\ X10^{4}$	
Std Ideal Liquid Volume Flow	m^3/h	42.59	42.59	
Molar Enthalpy	kJ/kgmole	$-1.401X10^{4}$	$-1.401X10^{4}$	
Molar Entropy	$kJ/kgmole^oC$	121.5	121.5	
Heat Flow	kJ/h	$-7.26X10^{6}$	$-7.26\ X10^{6}$	
Liquid Volume Flow @Std Cond	m^3/h	42.31	42.31	
PROPERTIES				
		Over all	Vapour Phase	
Molecular Weight		42.40	42.40	
Molar Density	$kgmole/m^3$	0.5602	0.5602	
Mass Density	kg/m^3	23.75	23.75	
Act. Volume Flow	m^3/h	925.3	925.3	
Mass Enthalpy	kJ/kg	-330.3	-330.3	
Act. Gas Flow	ACT_m^3/h	925.3	925.3	
Std. Gas Flow	STD_m^3/h	$1.226\ X10^{4}$	$1.226\ X10^{4}$	
Z Factor		0.9130	0.9130	
Watson K		14.38	14.38	
C_p/C_v		1.160	1.160	
Surface Tension	dyne/cm	---	---	
Thermal Conductivity	W/m-K	$3.23X10^{-2}$	$3.23\ X10^{-2}$	
Viscosity	cP	$1.28\ X10^{-2}$	$1.28\ X10^{-2}$	

Table A2.10: HMB of material stream 10

CONDITIONS			
	Over all	Vapour Phase	
Vapour / Phase Fraction	1.0000	1.0000	
Temperature °C	150.0	150.0	
Pressure kPa	1853	1853	
Molar Flow kgmole/h	518.3	518.3	
Mass Flow kg/h	2.21×10^4	2.21×10^4	
Std Ideal Liquid Volume Flow m^3/h	42.94	42.94	
Molar Enthalpy kJ/kgmole	-1.16×10^4	-1.16×10^4	
Molar Entropy kJ/kgmole°C	120.5	120.5	
Heat Flow kJ/h	-6.028×10^6	-6.028×10^6	
Liquid Volume Flow @Std Cond m^3/h	42.74	42.74	
PROPERTIES			
	Over all	Vapour Phase	
Molecular Weight	42.65	42.65	
Molar Density kgmole/ m^3	0.5790	0.5790	
Mass Density kg/ m^3	24.70	24.70	
Act. Volume Flow m^3/h	895.2	895.2	
Mass Enthalpy kJ/kg	-272.7	-272.7	
Act. Gas Flow ACT_m^3/h	895.2	895.2	
Std. Gas Flow STD_m^3/h	1.226×10^4	1.226×10^4	
Z Factor	0.9099	0.9099	
Watson K	14.38	14.38	
C_p/C_v	1.161	1.161	
Surface Tension dyne/cm	---	---	
Thermal Conductivity W/m-K	3.2×10^{-2}	3.2×10^{-2}	
Viscosity cP	1.28×10^{-2}	1.28×10^{-2}	
Reid VP at 37.8 C kPa	1529	1529	
True VP at 37.8 C kPa	1536	1536	

Table A2.11: HMB of material stream 11

CONDITIONS			
	Over all	Vapour Phase	
Vapour / Phase Fraction	1.0000	1.0000	

Temperature	^oC	64.18	64.18	
Pressure	kPa	1814	1814	
Molar Flow	kgmole/h	518.3	518.3	
Mass Flow	kg/h	$2.211 X10^4$	$2.211 X10^4$	
Std Ideal Liquid Volume Flow	m^3/h	42.94	42.94	
Molar Enthalpy	kJ/kgmole	$-1.93 X10^4$	$-1.93 X10^4$	
Molar Entropy	$kJ/kgmole^oC$	100.4	100.4	
Heat Flow	kJ/h	$-1.00 X10^7$	$-1.00 X10^7$	
Liquid Volume Flow @Std Cond	m^3/h	42.74	42.74	
PROPERTIES				
		Over all	Vapour Phase	
Molecular Weight		42.65	42.65	
Molar Density	kgmole/ m^3	0.8166	0.8166	
Mass Density	kg/ m^3	34.83	34.83	
Act. Volume Flow	m^3/h	634.8	634.8	
Mass Enthalpy	kJ/kg	-452.4	-452.4	
Act. Gas Flow	ACT_ m^3/h	634.8	634.8	
Std. Gas Flow	STD_ m^3/h	$1.226 X10^4$	$1.226 X10^4$	
Z Factor		0.7921	0.7921	
Watson K		14.38	14.38	
C_p/C_v		1.309	1.309	
Surface Tension	dyne/cm	---	---	
Thermal Conductivity	W/m-K	$2.33 X10^{-2}$	$2.33 X10^{-2}$	
Viscosity	cP	$1.06X10^{-2}$	$1.06X10^{-2}$	
Reid VP at 37.8 C	kPa	1529	1529	
True VP at 37.8 C	kPa	1536	1536	

Table A2.12: HMB of material stream 12

CONDITIONS				
		Over all	Liquid Phase	
Vapour / Phase Fraction		0.0000	1.0000	
Temperature	^oC	41.76	41.76	
Pressure	kPa	1795	1795	
Molar Flow	kgmole/h	518.3	518.3	
Mass Flow	kg/h	$2.211 X10^4$	$2.211 X10^4$	

Std Ideal Liquid Volume Flow m^3/h	42.94	42.94	
Molar Enthalpy kJ/kgmole	$-3.391X10^4$	$-3.391X10^4$	
Molar Entropy kJ/kgmole°C	54.72	54.72	
Heat Flow kJ/h	$-1.758X10^7$	$-1.758X10^7$	
Liquid Volume Flow @Std Cond m^3/h	42.74	42.74	
PROPERTIES			
	Over all	Liquid Phase	
Molecular Weight	42.65	42.65	
Molar Density kgmole/ m^3	11.07	11.07	
Mass Density kg/ m^3	472.1	472.1	
Act. Volume Flow m^3/h	46.83	46.83	
Mass Enthalpy kJ/kg	-795.0	-795.0	
Act. Gas Flow ACT_ m^3/h	---	---	
Std. Gas Flow STD_ m^3/h	$1.226 X10^4$	$1.226 X10^4$	
Z Factor	$6.192X10^{-2}$	$6.192X10^{-2}$	
Watson K	14.38	14.38	
Surface Tension dyne/cm	4.763	4.763	
Thermal Conductivity W/m-K	$9.38X10^{-2}$	$9.38 X10^{-2}$	
Viscosity cP	$6.540X10^4$	$6.540X10^{-2}$	
True VP at 37.8 C kPa	1536	1536	

Table A2.13: HMB of material stream 13

CONDITIONS			
	Over all	Liquid Phase	
Vapour / Phase Fraction	0.0000	1.0000	
Temperature °C	44.06	44.06	
Pressure kPa	3285	3285	
Molar Flow kgmole/h	518.3	518.3	
Mass Flow kg/h	2.211	2.211e+004	
Std Ideal Liquid Volume Flow m^3/h	42.94	42.94	
Molar Enthalpy kJ/kgmole	$-3.373X10^4$	$-3.373X10^4$	
Molar Entropy kJ/kgmole°C	54.82	54.82	
Heat Flow kJ/h	$-1.748X10^7$	$-1.748X10^7$	
Liquid Volume Flow @Std Cond m^3/h	42.74	42.74	
PROPERTIES			

		Over all	Liquid Phase	
Molecular Weight		42.65	42.65	
Molar Density	kgmole/m^3	11.13	11.13	
Mass Density	kg/m^3	474.6	474.6	
Act. Volume Flow	m^3/h	46.58	46.58	
Mass Enthalpy	kJ/kg	-790.7	-790.7	
Act. Gas Flow	ACT_m^3/h	---	---	
Std. Gas Flow	STD_m^3/h	1.226 $X10^4$	1.226 $X10^4$	
Z Factor		0.1119	0.1119	
Watson K		14.38	14.38	
Cp/Cv		1.581	1.581	
Surface Tension	dyne/cm	4.506	4.506	
Thermal Conductivity	W/m-K	$9.24X10^{-2}$	9.24 $X10^{-2}$	
Viscosity	cP	6.43 $X10^{-2}$	6.43 $X10^{-2}$	
Reid VP at 37.8 C	kPa	1529	1529	
True VP at 37.8 C	kPa	1536	1536	

Table A2.14: HMB of material stream 14

CONDITIONS				
		Over all	Liquid Phase	
Vapour / Phase Fraction		0.0000	1.0000	
Temperature	^oC	42.07	42.07	
Pressure	kPa	3285	3285	
Molar Flow	kgmole/h	518.3	518.3	
Mass Flow	kg/h	2.211 $X10^4$	2.211 $X10^4$	
Std Ideal Liquid Volume Flow	m^3/h	42.93	42.93	
Molar Enthalpy	kJ/kgmole	$-3.399X10^4$	$-3.399X10^4$	
Molar Entropy	kJ/kgmole^{o}C	53.99	53.99	
Heat Flow	kJ/h	$-1.762X10^7$	$-1.762X10^7$	
Liquid Volume Flow @Std Cond	m^3/h	42.73	42.73	
PROPERTIES				
		Over all	Liquid Phase	
Molecular Weight		42.65	42.65	
Molar Density	kgmole/m^3	11.22	11.22	

Mass Density	kg/m^3	478.4	478.4	
Act. Volume Flow	m^3/h	46.21	46.21	
Mass Enthalpy	kJ/kg	-796.9	-796.9	
Std. Gas Flow	STD_m^3/h	$1.225\ X10^4$	$1.225\ X10^4$	
Z Factor		0.1118	0.1118	
Watson K		14.38	14.38	
C_p/C_v		1.578	1.578	
Surface Tension	dyne/cm	4.729	4.729	
Thermal Conductivity	W/m-K	$9.36\ X10^{-2}$	$9.369X10^{-2}$	
Viscosity	cP	$6.580X10^{-2}$	$6.58\ X10^{-2}$	
Reid VP at 37.8 C	kPa	1529	1529	
True VP at 37.8 C	kPa	1536	1536	

Table A2.15: HMB of material stream 15

CONDITIONS				
		Over All	Vapour Phase	
Vapour / Phase Fraction		1.0000	1.0000	
Temperature:	^oC	52.15	52.15	
Pressure:	kPa	2834	2834	
Molar Flow	kgmole/h	14.09	14.09	
Mass Flow	kg/h	544.0	544.0	
Std Ideal Liquid Volume Flow	m^3/h	1.159	1.159	
Molar Enthalpy	kJ/kgmole	$-1.914X10^4$	$-1.914X10^4$	
Molar Entropy	$kJ/kgmole^oC$	103.4	103.4	
Heat Flow	kJ/h	$-2.698X10^5$	$-2.698X10^5$	
Liquid Volume Flow @Std Cond	m^3/h	1.121	1.121	
PROPERTIES				
		Over all	Vapour Phase	
Molecular Weight		38.61	38.61	
Molar Density	$kgmole/m^3$	1.591	1.591	
Mass Density	kg/m^3	61.43	61.43	
Act. Volume Flow	m^3/h	8.855	8.855	
Mass Enthalpy	kJ/kg	-495.9	-495.9	
Act. Gas Flow	ACT_m^3/h	8.855	8.855	
Std. Gas Flow	STD_m^3/h	333.2	333.2	

Z Factor		0.6585	0.6585	
Watson K		15.45	15.45	
C_p/C_v		1.740	1.740	
Surface Tension	dyne/cm	---	---	
Thermal Conductivity	W/m-K	$2.52X10^{-2}$	$2.52\ X10^{-2}$	
Viscosity	cP	$1.15\ X10^{-2}$	$1.15\ X10^{-2}$	
Reid VP at 37.8 C	kPa	2331	2331	
True VP at 37.8 C	kPa	2498	2498	

Table A2.16: HMB of material stream 16

CONDITIONS				
		Over all	Liquid Phase	
Vapour / Phase Fraction		0.0000	1.0000	
Temperature	^{o}C	68.33	68.33	
Pressure	kPa	2883	2883	
Molar Flow	kgmole/h	504.2	504.2	
Mass Flow	kg/h	$2.156\ X10^{4}$	$2.156\ X10^{4}$	
Std Ideal Liquid Volume Flow	m^3/h	41.77	41.77	
Molar Enthalpy	kJ/kgmole	$-3.002X10^{4}$	$-3.002X10^{4}$	
Molar Entropy	kJ/kgmole^{o}C	65.80	65.80	
Heat Flow	kJ/h	$-1.514X10^{7}$	$-1.514X10^{7}$	
Liquid Volume Flow @Std Cond	m^3/h	41.62	41.62	
PROPERTIES				
		Over all	Liquid Phase	
Molecular Weight		42.77	42.77	
Molar Density	kgmole/m^3	9.649	9.649	
Mass Density	kg/m^3	412.7	412.7	
Act. Volume Flow	m^3/h	52.26	52.26	
Mass Enthalpy	kJ/kg	-702.0	-702.0	
Act. Gas Flow	ACT_m^3/h	---	---	
Std. Gas Flow	STD_m^3/h	$1.192\ X10^{4}$	$1.192\ X10^{4}$	
Z Factor		0.1052	0.1052	
Watson K		14.35	14.35	
C_p/C_v		1.771	1.771	
Surface Tension	dyne/cm	1.992	1.992	
Thermal	W/m-K	$7.80\ X10^{-2}$	$7.80\ X10^{-2}$	

Conductivity				
Viscosity	cP	$4.71X10^{-2}$	$4.71 X10^{-2}$	
Reid VP at 37.8 C	kPa	1510	1510	

Table A2.17: HMB of material stream 17

CONDITIONS				
		Over all	Vapour Phase	Liquid Phase
Vapour / Phase Fraction		0.2098	0.2098	0.7902
Temperature	^oC	52.87	52.87	52.87
Pressure	kPa	2108	2108	2108
Molar Flow	kgmole/h	504.2	105.8	398.4
Mass Flow	kg/h	$2.156 X10^4$	4518	$1.705 X10^4$
Std Ideal Liquid Volume Flow	m^3/h	41.77	8.750	33.02
Molar Enthalpy	kJ/kgmole	$-3.002X10^4$	$-1.868X10^4$	$-3.303X10^4$
Molar Entropy	kJ/kgmole^{o}C	66.25	92.93	59.16
Heat Flow	kJ/h	$-1.514X10^7$	$-1.976X10^6$	$-1.316X10^7$
Liquid Volume Flow @Std Cond	m^3/h	41.62	8.717	32.90
PROPERTIES				
		Over all	Vapour Phase	Liquid Phase
Molecular Weight		42.77	42.72	42.78
Molar Density	kgmole/m^3	3.769	1.103	10.52
Mass Density	kg/m^3	161.2	47.10	450.2
Act. Volume Flow	m^3/h	133.8	95.92	37.86
Mass Enthalpy	kJ/kg	-702.0	-437.4	-772.1
Act. Gas Flow	ACT_m^3/h	95.92	95.92	---
Std. Gas Flow	STD_m^3/h	$1.192 X10^4$	2501	9421
Z Factor		---	0.7054	$7.392X10^{-2}$
Watson K		14.35	14.35	14.36
C_p/C_v		1.078	1.492	1.655
Surface Tension	dyne/cm	3.578	---	3.578
Thermal Conductivity	W/m-K	---	$2.292X10^{-2}$	$8.732X10^{-2}$
Viscosity	cP	---	$1.055X10^{-2}$	$5.802X10^{-2}$

Table A2.18: HMB of material stream 18

CONDITIONS				
		Over all	Liquid Phase	
Vapour / Phase Fraction		0.0000	1.0000	
Temperature	^oC	47.91	47.91	
Pressure	kPa	2020	2020	
Molar Flow	kgmole/h	6186	6186	
Mass Flow	kg/h	$2.603 X10^5$	$2.603 X10^5$	
Std Ideal Liquid Volume Flow	m^3/h	499.7	499.7	
Molar Enthalpy	kJ/kgmole	7602	7602	
Molar Entropy	kJ/kgmole^oC	27.24	27.24	
Heat Flow	kJ/h	$4.703 X10^7$	$4.703 X10^7$	
Liquid Volume Flow @Std Cond	m^3/h	497.5	497.5	
PROPERTIES				
		Over all	Liquid Phase	
Molecular Weight		42.08	42.08	
Molar Density	kgmole/m^3	11.03	11.03	
Mass Density	kg/m^3	464.0	464.0	
Act. Volume Flow	m^3/h	561.0	561.0	
Mass Enthalpy	kJ/kg	180.7	180.7	
Act. Gas Flow	ACT_m^3/h	---	---	
Std. Gas Flow	STD_m^3/h	$1.463 X10^5$	$1.463 X10^5$	
Z Factor		$6.862X10^{-2}$	$6.862X10^{-2}$	
Watson K		14.18	14.18	
C_p/C_v		1.679	1.679	
Surface Tension	dyne/cm	4.008	4.008	
Thermal Conductivity	W/m-K	$9.497X10^{-2}$	$9.497X10^{-2}$	
Viscosity	cP	$4.83 X10^{-2}$	$4.83 X10^{-2}$	
True VP at 37.8 C	kPa	1583	1583	

Table A2.19: HMB of material stream 19

CONDITIONS				
		Over all	Liquid Phase	
Vapour / Phase Fraction		0.0000	1.0000	
Temperature	^oC	59.74	59.74	

Pressure kPa	2148	2148	
Molar Flow kgmole/h	171.7	171.7	
Mass Flow kg/h	7571	7571	
Std Ideal Liquid Volume Flow m^3/h	14.91	14.91	
Molar Enthalpy kJ/kgmole	$-1.102X10^5$	$-1.102X10^5$	
Molar Entropy kJ/kgmole°C	104.1	104.1	
Heat Flow kJ/h	$-1.893X10^7$	$-1.893X10^7$	
Liquid Volume Flow @Std Cond m^3/h	14.88	14.88	
PROPERTIES			
	Over all	Liquid Phase	
Molecular Weight	44.09	44.09	
Molar Density kgmole/ m^3	9.755	9.755	
Mass Density kg/ m^3	430.1	430.1	
Act. Volume Flow m^3/h	17.60	17.60	
Mass Enthalpy kJ/kg	-2500	-2500	
Act. Gas Flow ACT_m^3/h	---	---	
Std. Gas Flow STD_m^3/h	4060	4060	
Z Factor	$7.954X10^{-2}$	$7.954X10^{-2}$	
Watson K	14.67	14.67	
C_p/C_v	1.632	1.632	
Surface Tension dyne/cm	3.083	3.083	
Thermal Conductivity W/m-K	$7.5 X10^{-2}$	$7.5 X10^{-2}$	
Viscosity cP	$7.01 X10^{-2}$	$7.01 X10^{-2}$	
True VP at 37.8 C kPa	1325	1325	

Table A2.20: HMB of material stream 20

CONDITIONS			
	Overall	Vapour Phase	
Vapour / Phase Fraction	1.0000	1.0000	
Temperature °C	48.45	48.45	
Pressure kPa	2010	2010	
Molar Flow kgmole/h	6519	6519	
Mass Flow kg/h	$2.743 X10^5$	$2.743 X10^5$	
Std Ideal Liquid Volume Flow m^3/h	526.6	526.6	

Molar Enthalpy	kJ/kgmole	$1.968\ X10^4$	$1.968\ X10^4$	
Molar Entropy	kJ/kgmole°C	64.85	64.85	
Heat Flow	kJ/h	$1.283\ X10^8$	$1.283\ X10^8$	
Liquid Volume Flow @Std Cond	m^3/h	524.2	524.2	
PROPERTIES				
		Over all	Vapour Phase	
Molecular Weight		42.08	42.08	
Molar Density	$kgmole/m^3$	1.042	1.042	
Mass Density	kg/m^3	43.83	43.83	
Act. Volume Flow	m^3/h	6258	6258	
Mass Enthalpy	kJ/kg	467.8	467.8	
Act. Gas Flow	ACT_m^3/h	6258	6258	
Std. Gas Flow	STD_m^3/h	$1.541X10^5$	$1.541\ X10^5$	
Z Factor		0.7218	0.7218	
Watson K		14.18	14.18	
C_p/C_v		1.478	1.478	
Surface Tension	dyne/cm	--	---	
Thermal Conductivity	W/m-K	$2.199X10^{-2}$	$2.199X10^{-2}$	
Viscosity	cP	$1.043X10^{-2}$	$1.043X10^{-2}$	
True VP at 37.8 C	kPa	1583	1583	

Table A2.21: HMB of material stream 21

CONDITIONS				
Vapour / Phase Fraction		1.0000	1.0000	
Temperature	°C	48.45	48.45	
Pressure	kPa	2010	2010	
Molar Flow	kgmole/h	6186	6186	
Mass Flow	kg/h	$2.603\ X10^5$	$2.603\ X10^5$	
Std Ideal Liquid Volume Flow	m^3/h	499.7	499.7	
Molar Enthalpy	kJ/kgmole	$1.968\ X10^4$	$1.968\ X10^4$	
Molar Entropy	kJ/kgmole°C	64.85	64.85	
Heat Flow	kJ/h	$1.218\ X10^8$	$1.218\ X10^8$	
Liquid Volume Flow @Std Cond	m^3/h	497.5	497.5	

PROPERTIES			
	Over all	Vapour Phase	
Molecular Weight	42.08	42.08	
Molar Density kgmole/m^3	1.042	1.042	
Mass Density kg/m^3	43.83	43.83	
Act. Volume Flow m^3/h	5939	5939	
Mass Enthalpy kJ/kg	467.8	467.8	
Act. Gas Flow ACT_m^3/h	5939	5939	
Std. Gas Flow STD_m^3/h	1.463 $X10^5$	1.463 $X10^5$	
Z Factor	0.7218	0.7218	
Watson K	14.18	14.18	
C_p/C_v	1.478	1.478	
Surface Tension dyne/cm	---	---	
Thermal Conductivity W/m-K	$2.19X10^{-2}$	2.19 $X10^{-2}$	
Viscosity cP	1.04 $X10^4$	1.04 $X10^{-2}$	
Reid VP at 37.8 C kPa	1583	1583	
True VP at 37.8 C kPa	1583	1583	

Table A2.22: HMB of material stream 22

CONDITIONSS			
	Over all	Liquid Phase	
Vapour / Phase Fraction	0.0000	1.0000	
Temperature ^oC	47.91	47.91	
Pressure kPa	1991	1991	
Molar Flow kgmole/h	6186	6186	
Mass Flow kg/h	2.603 $X10^5$	2.603 $X10^5$	
Std Ideal Liquid Volume Flow m^3/h	499.7	499.7	
Molar Enthalpy kJ/kgmole	7598	7598	
Molar Entropy kJ/kgmole^oC	27.28	27.28	
Heat Flow kJ/h	4.700 $X10^7$	4.700 $X10^7$	
Liquid Volume Flow @Std Cond m^3/h	497.5	497.5	
PROPERTIES			
	Over all	Liquid Phase	
Molecular Weight	42.08	42.08	
Molar Density kgmole/m^3	11.02	11.02	

Mass Density	kg/m^3	463.8	463.8	
Act. Volume Flow	m^3/h	561.2	561.2	
Mass Enthalpy	kJ/kg	180.6	180.6	
Act. Gas Flow	ACT_m^3/h	---	---	
Std. Gas Flow	STD_m^3/h	1.463×10^5	1.463×10^5	
Z Factor		6.76×10^{-2}	6.76×10^{-2}	
Watson K		14.18	14.18	
C_p/C_v		1.681	1.681	
Surface Tension	dyne/cm	4.007	4.007	
Thermal Conductivity	W/m-K	9.497×10^{-2}	9.497×10^{-2}	
Viscosity	cP	4.829×10^{-2}	4.829×10^{-2}	
Reid VP at 37.8 C	kPa	1583	1583	
True VP at 37.8 C	kPa	1583	1583	

Table A2.23: HMB of material stream 23

CONDITIONS				
		Over all	Vapour Phase	
Vapour / Phase Fraction		1.0000	1.0000	
Temperature	^oC	48.45	48.45	
Pressure	kPa	2010	2010	
Molar Flow	kgmole/h	332.5	332.5	
Mass Flow	kg/h	1.399×10^4	1.399×10^4	
Std Ideal Liquid Volume Flow	m^3/h	26.85	26.85	
Molar Enthalpy	kJ/kgmole	1.968×10^4	1.968×10^4	
Molar Entropy	kJ/kgmole^oC	64.85	64.85	
Heat Flow	kJ/h	6.544×10^6	6.544×10^6	
Liquid Volume Flow @Std Cond	m^3/h	26.74	26.74	
PROPERTIES				
		Over all	Vapour Phase	
Molecular Weight		42.08	42.08	
Molar Density	$kgmole/m^3$	1.042	1.042	
Mass Density	kg/m^3	43.83	43.83	
Act. Volume Flow	m^3/h	319.2	319.2	
Mass Enthalpy	kJ/kg	467.8	467.8	
Act. Gas Flow	ACT_m^3/h	319.2	319.2	
Std. Gas Flow	STD_m^3/h	7861	7861	

Z Factor	0.7218	0.7218	
Watson K	14.18	14.18	
C_p/C_v	1.478	1.478	
Surface Tension dyne/cm	---	---	
Thermal Conductivity W/m-K	$2.19X10^{-2}$	$2.19X10^{-2}$	
Viscosity cP	$1.04X10^{-2}$	$1.04\ X10^{-2}$	

Table A2.24: HMB of material stream 24

CONDITIONS			
	Over all	Vapour Phase	
Vapour / Phase Fraction	1.0000	1.0000	
Temperature ^oC	149.6	149.6	
Pressure kPa	1961	1961	
Molar Flow kgmole/h	332.5	332.5	
Mass Flow kg/h	$1.399\ X10^{4}$	$1.399\ X10^{4}$	
Std Ideal Liquid Volume Flow m^3/h	26.85	26.85	
Molar Enthalpy kJ/kgmole	$2.849\ X10^{4}$	$2.849\ X10^{4}$	
Molar Entropy kJ/kgmole^oC	88.84	88.84	
Heat Flow kJ/h	$9.473\ X10^{6}$	$9.473\ X10^{6}$	
Liquid Volume Flow @Std Cond m^3/h	26.74	26.74	
PROPERTIES			
	Over all	Vapour Phase	
Molecular Weight	42.08	42.08	
Molar Density kgmole/m^3	0.6149	0.6149	
Mass Density kg/m^3	25.88	25.88	
Act. Volume Flow m^3/h	540.6	540.6	
Mass Enthalpy kJ/kg	677.1	677.1	
Act. Gas Flow ACT_m^3/h	540.6	540.6	
Std. Gas Flow STD_m^3/h	7861	7861	
Z Factor	0.9074	0.9074	
Watson K	14.18	14.18	
C_p/C_v	1.173	1.173	
Surface Tension dyne/cm	---	---	
Thermal Conductivity W/m-K	$3.2\ X10^{-2}$	$3.2\ X10^{-2}$	
Viscosity cP	$1.30X10^{-2}$	$1.30X10^{-2}$	

Table A2.25: HMB of material stream 25

CONDITIONS				
		Over all	Vapour Phase	
Vapour / Phase Fraction		1.0000	1.0000	
Temperature	^{o}C	179.7	179.7	
Pressure	kPa	1910	1910	
Molar Flow	kgmole/h	332.5	332.5	
Mass Flow	kg/h	$1.399 X10^{4}$	$1.399 X10^{4}$	
Std Ideal Liquid Volume Flow	m^3/h	26.85	26.85	
Molar Enthalpy	kJ/kgmole	$3.126 X10^{4}$	$3.126 X10^{4}$	
Molar Entropy	kJ/kgmole^{o}C	95.36	95.36	
Heat Flow	kJ/h	$1.039 X10^{7}$	$1.039 X10^{7}$	
Liquid Volume Flow @Std Cond	m^3/h	26.74	26.74	
PROPERTIES				
		Over all	Vapour Phase	
Molecular Weight		42.08	42.08	
Molar Density	kgmole/m^3	0.5450	0.5450	
Mass Density	kg/m^3	22.93	22.93	
Act. Volume Flow	m^3/h	610.0	610.0	
Mass Enthalpy	kJ/kg	742.9	742.9	
Act. Gas Flow	ACT_m^3/h	610.0	610.0	
Std. Gas Flow	STD_m^3/h	7861	7861	
Z Factor		0.9309	0.9309	
Watson K		14.18	14.18	
C_p/C_v		1.149	1.149	
Surface Tension	dyne/cm	---	---	
Thermal Conductivity	W/m-K	$3.53 X10^{-2}$	$3.53X10^{-2}$	
Viscosity	cP	$1.38 X10^{-2}$	$1.38 X10^{-2}$	
Reid VP at 37.8 C	kPa	1583	1583	

Table A2.26: HMB of material stream 26

CONDITIONS				
		Over all	Vapour Phase	
Vapour / Phase Fraction		1.0000	1.0000	
Temperature	^{o}C	180.0	180.0	

	Pressure kPa	2001	2001	
	Molar Flow kgmole/h	332.3	332.3	
	Mass Flow kg/h	1.399×10^4	1.399×10^4	
	Std Ideal Liquid Volume Flow m^3/h	26.85	26.85	
	Molar Enthalpy kJ/kgmole	3.074×10^4	3.074×10^4	
	Molar Entropy kJ/kgmole°C	95.49	95.49	
	Heat Flow kJ/h	1.02×10^7	1.02×10^7	
	Liquid Volume Flow @Std Cond m^3/h	26.73	26.73	
	PROPERTIES			
		Over all	Vapour Phase	
	Molecular Weight	42.09	42.09	
	Molar Density $kgmole/m^3$	0.5723	0.5723	
	Mass Density kg/m^3	24.09	24.09	
	Act. Volume Flow m^3/h	580.6	580.6	
	Mass Enthalpy kJ/kg	730.4	730.4	
	Act. Gas Flow ACT_m^3/h	580.6	580.6	
	Std. Gas Flow STD_m^3/h	7857	7857	
	Z Factor	0.9278	0.9278	
	Watson K	14.18	14.18	
	C_p/C_v	1.152	1.152	
	Surface Tension dyne/cm	---	---	
	Thermal Conductivity W/m-K	3.54×10^{-2}	3.54×10^{-2}	
	Viscosity cP	1.39×10^{-2}	1.39×10^{-2}	
	Reid VP at 37.8 C kPa	1582	1582	
	True VP at 37.8 C kPa	1582	1582	

Table A2.27: HMB of material stream 27

	CONDITIONS			
		Over all	Vapour Phase	
	Vapour / Phase Fraction	1.0000	1.0000	
	Temperature °C	79.39	79.39	
	Pressure kPa	1952	1952	
	Molar Flow kgmole/h	332.3	332.3	
	Mass Flow kg/h	1.399×10^4	1.399×10^4	
	Std Ideal Liquid Volume Flow m^3/h	26.85	26.85	

Molar Enthalpy	kJ/kgmole	2.193×10^4	2.193×10^4	
Molar Entropy	kJ/kgmole°C	73.72	73.72	
Heat Flow	kJ/h	7.286×10^6	7.286×10^6	
Liquid Volume Flow @Std Cond	m^3/h	26.73	26.73	
PROPERTIES				
		Over all	Vapour Phase	
Molecular Weight		42.09	42.09	
Molar Density	$kgmole/m^3$	0.8154	0.8154	
Mass Density	kg/m^3	34.32	34.32	
Act. Volume Flow	m^3/h	407.5	407.5	
Mass Enthalpy	kJ/kg	521.0	521.0	
Act. Gas Flow	ACT_m^3/h	407.5	407.5	
Std. Gas Flow	STD_m^3/h	7857	7857	
Z Factor		0.8165	0.8165	
Watson K		14.18	14.18	
C_p/C_v		1.286	1.286	
Surface Tension	dyne/cm	---	---	
Thermal Conductivity	W/m-K	2.46×10^{-2}	2.46×10^{-2}	
Viscosity	cP	1.11×10^{-2}	1.11×10^{-2}	
Reid VP at 37.8 C	kPa	1582	1582	
True VP at 37.8 C	kPa	1582	1582	

Table A2.28: HMB of material stream 28

CONDITIONS				
		Over All	Vapour Phase	Liquid Phase
Vapour / Phase Fraction		0.0015	0.0015	0.9985
Temperature	°C	40.05	40.05	40.05
Pressure	kPa	1667	1667	1667
Molar Flow	kgmole/h	332.3	0.4894	331.8
Mass Flow	kg/h	1.399×10^4	20.60	1.397×10^4
Std Ideal Liquid Volume Flow	m^3/h	26.85	3.95×10^{-2}	26.81
Molar Enthalpy	kJ/kgmole	6062	1.905×10^4	6043
Molar Entropy	kJ/kgmole°C	24.65	65.93	24.59
Heat Flow	kJ/h	2.014×10^6	9321	2.005×10^6
Liquid Volume Flow @Std Cond	m^3/h	26.73	3.937×10^{-2}	26.69
PROPERTIES				

		Over all	Vapour Phase	Liquid Phase
Molecular Weight		42.09	42.09	42.09
Molar Density	kgmole/m^3	11.19	0.8439	11.39
Mass Density	kg/m^3	470.9	35.52	479.6
Act. Volume Flow	m^3/h	29.70	0.5799	29.12
Mass Enthalpy	kJ/kg	144.0	452.5	143.6
Act. Gas Flow	ACT_m^3/h	0.5799	0.5799	---
Std. Gas Flow	STD_m^3/h	7857	11.57	7845
Z Factor		---	0.7586	$5.618X10^{-2}$
Watson K		14.18	14.18	14.18
C_p/C_v		1.188	1.398	1.665
Surface Tension	dyne/cm	4.901	---	4.901
Thermal Conductivity	W/m-K	---	$2.05\ X10^{-2}$	$9.97\ X10^{-2}$
Viscosity	cP	---	$9.91\ X10^{-2}$	$5.53\ X10^{-2}$
Reid VP at 37.8 C	kPa	1582	1582	1582
True VP at 37.8 C	kPa	1582	1583	1582

APPENDIX 3

DE-PRIESTER CHART

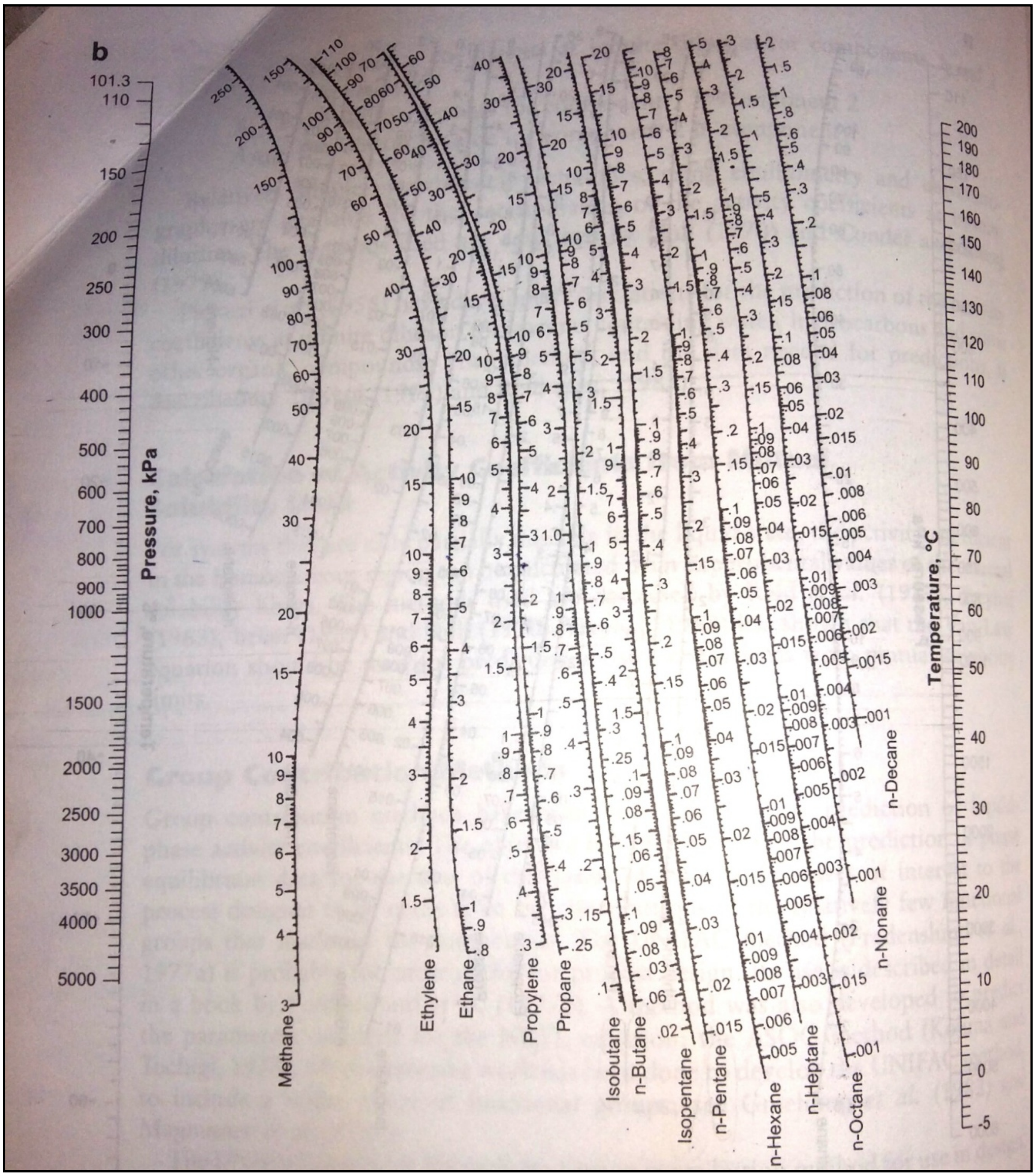

Figure A3.1: De-Priester chart for K-values for hydrocarbons, high temperature