

Simulation and Optimization of Depropanizer Using Hysys Simulation Package

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ABSTRACT

A new depropanizer is designed for the revamped petrochemical complex PC1 in Basrah. Conventional fractionation column is used to match the design of the existing plant. The feed to the new depropanizer is the bottom product of the revamped deethanizer of the ethylene plant. Hysys package (3.2) is used for the short-cut method, rigorous model and tray sizing. Different variables have been studied such as total number of stages, reflux ratio, feed location and feed temperature. The optimum number of stages is found to be (55) stages and the feed location is at tray 25th from top, with feed temperature of 32°C. The tray layout and sizing is estimated using Hysys, all trays are forced to have the same design so that the column maintains the same diameter throughout its height.

Key Words: Simulation, Propylene, Optimization, Tray layout, Petrochemicals, Distillation.

الخلاصة

تصميم جهاز فصل البروبانات لوحدة الأثيلين المطوّرة في مجمع البتروكيمياويات – البصرة. أختير جهاز فصل تقليدي ليتماشى مع الأجهزة الموجودة حالياً في وحدة الأثيلين. المغذي لهذا الجهاز هو الناتج السفلي لجهاز فصل الأيثان. استخدم برنامج هايسس 3.2 للحصول على أفضل الظروف التشغيلية في موديل الحالة المستقرة واسلوب المحاكاة الصارم المعتمد على النمذجة الحديثة. تم دراسة تأثير عدة متغيرات وهي عدد الصواني، نسبة الراجع، درجة حرارة المغذي وموقع المغذي وتم الحصول على أفضل الظروف التشغيلية، وذلك في ان يكون عدد الصواني الأمثل هو 55 صينية ونسبة الراجع=1 وان يكون المغذي بدرجة حرارة 32 درجة مئوية وموقعه على الصينية 25 من الأعلى. كذلك استخدم البرنامج لاجراء الحسابات الأولية للتصميم باستخدام الطريقة المختصرة لاجاد اقل نسبة راجع واقل عدد صواني بالاضافة الى حساب قطر البرج وتصميم الصواني.

الكلمات الدالة: المحاكاة، البروبلين، الأختيار أمثل، تصميم الصواني، البتروكيمياويات، التقطير

INTRODUCTION

Working from Iraq's significant potentials for a dynamic petrochemical and plastic industry, the downstream petrochemical industries were limited by the country's long isolation from world market. The sector of

the petrochemicals can be revitalized with some investment. Basic infrastructure already exists and can be rehabilitated and expanded.

A revamp study to the ethylene plant at Basrah petrochemical complex PC1 was proposed by Linde ¹ and Al-Azzawi ⁷ for the existing plant to achieve co-cracking of LPG with ethylene in order to produce polymer grade propylene. The net bottom of the revamped deethanizer must be depropanized because of the higher yield of propylene. The top product from the depropanizer (mixed C₃'s) must be sent to selective hydrogenation reactor to increase the yield of propylene. The effluent from the reactor must be sent to propylene/propane fractionator to produce polymer grade propylene, and propane is recycled to the cracking furnace. The depropanizer and propylene fractionator are not constructed because of the low yield of propylene in the existing plant.

In this work, a new depropanizer is designed for the revamped ethylene unit by simulating the data gained from the revamped deethanizer to complete the objective of introducing new feedstock to the existing ethylene plant. The simulation and design are conducted using Hysys package. Hysys incorporates a number of significant features including an innovative separation model that allows accurate simulation and understanding of the liquid and gas separations that occur in oil and gas plants.

Following are the results of a study that is prepared to assess the financial viability of propylene production unit at (PC1). First, the various uses and the demand history for high purity propylene are presented. Next, the financial aspects for the importance of producing propylene in Iraq are discussed. Finally, simulation and design of a new depropanizer for the revamped ethylene plant are conducted. This study includes the optimization and simulation of a conventional distillation with different operating variables.

PROPYLENE

World-wide demand of propylene has been rising steadily over the last 20 years. The consumption of propylene in 2011 was 79 million tons and it is expected to reach 97.5 million tons in 2015 [2], with a majority of the increase in demand occurring in Asia. Propylene prices that were high in 2010 (approximately \$0.6/lb.) have continued to increase and are projected to remain between (\$0.7 – 0.8/lb.) in North_America ². In Europe, propylene went above ethylene for the first time ever during 2010. Since 2001, the price differential of ethylene over propylene has steadily eroded until it reversed in 2010. The price of propylene also has risen above that of ethylene in North East Asia [4, 5, 6, 8].

Propylene is the primary ethylene co-product from steam cracking of hydrocarbons, such as ethane, propane, naphtha and gasoil. Lower molecular weight feedstock (e.g. ethane and propane) yield a higher percentage of ethylene. Heavy molecular weight feedstock like naphtha, LPG and gasoil, are used to obtain more propylene. The propylene produced from an ethylene steam cracker is of sufficient purity to produce polymer grade propylene. In the FCC units, the propylene is produced as a dilute stream in propane. In the case of visbreaking and coking the propylene yields are lower and quality often unacceptable other than for refining fuel. Propylene comes in three grades; polymer grade (99.5% minimum

purity), chemical grade (93 – 94% minimum purity) and refinery grade (60 – 70% purity).

Propylene is the feedstock for many important chemicals and is used to make plastics and fibers as polypropylene (thermoplastic) and acrylic, it is also used to manufacture plenty of consumer products such as; food packaging, table ware, washing machine parts, outdoor furniture, building components, automotive components, pharmaceuticals, cosmetics, skin-care and sun screen products[5]. Worldwide demand for propylene is distributed as follow ¹²:

Polypropylene	52%
Acrylonitrile	12%
Oxo Alcohols	10%
Propylene Oxide	8%
Cumene	6%
Isopropyl Alcohol	4%
Acrylic Acid	3%
Others	5%

The petrochemical industry has been, and always will be, driven by the availability of competitive raw material, first coal then oil and now natural gas. As plastic manufacturers continue to find new uses of polymers, demands for its most basic building blocks, namely ethylene and propylene, will continue to escalate the price of monomers. Another issue is the higher prices of crude oil and petroleum gases in US, Japan, Europe, Canada and Asia impact on the petrochemical industry higher cost of ethylene and propane as well as higher cost for fuel, steam and electricity which raises the cost of producing ethylene and propylene. Great efforts will continue by these countries to seek-out the lowest sources of production [3].

Middle East countries have extensive hydrocarbon resources and geographically well suited to supply these products. These countries are expanding their petrochemical operations in an effort to diversity their industries and strengthen their domestic economics. In terms of future development, the Middle East is at the center of strong global growth ³. Saudi Arabia, Iran, Qatar and most of the Arabian Gulf countries are expanding their petrochemical business, because the industry's center of gravity moves from West to East. Iraq has significant potential to become a major player in both energy exports and petrochemical production. Feedstock's advantages in Iraq should fuel a powerful and continuing upsurge in petrochemical industries to divert Iraq's economy from just exporting oil to industrialization.

Iraq poised to become game-changer for the world's oil markets. Iraq's energy sector holds the key to the country's future prosperity and can make a major contribution to the stability and security of global energy markets. Natural gas can play a much more important role in Iraq's future and a vital first step will be to reduce the amount of gas that is currently flared. Once domestic needs are met, Iraq can provide a cost competitive source of gas supply to US and European market, neighboring countries and to Asia [9].

THE DEPROPANIZER

As noted previously, depropanizer was not constructed in the existing ethylene plant at Basrah PCI because of the low yield of propylene. Even though Lummus[13] (the Construction Company of the petrochemical complex) made a

material and energy balance for the un-constructed yet depropanizer. Their data is simulated using Hysys to ensure the accuracy of the simulation package. The results were pretty good and it is tabulated in appendix A.

Simulating the data gained from the revamped deethanizer is of high importance to complete the objective of introducing new feedstock to the existing ethylene plant in Basrah PC1. The feed comes from the revamped deethanizer, as bottom product, which contains C_3^+ . The feed composition of the depropanizer is given in table 1. The column operates at a fixed pressure of 10 bars to reduce fouling in the bottom section of the tower¹⁰. The feed pressure produces a flash (split between vapor and liquid), depending on its temperature. Heat Q_r is provided at the bottom of the column by low pressure steam, and extracted at its top by condenser Q_c .

The vapor – liquid equilibrium on the tower trays is calculated using Peng Robenson Fluid Package from Hysys Environment. Peng Robenson equation of state becomes one of the most widely used in industry for correlating mixtures containing hydrocarbons¹¹. There are two assumptions for this simulation; first, the state of the process model for simulation is normally considered to be steady-state when the optimal manipulated variables are searched. Second, the model of the process simulator is conducted by simple K-value and a Jacobian Matrix is solved using inside-out method. In the simulator, the controlled variables are the propane mole fraction in the bottom product of the depropanizer is specified to be (0.0078), Reflux ratio, number of trays, feed temperature, propylene composition in distillate and feed location. The major considerations for the simulation are zero mole fraction of isobutane in the top product, and as minimum as possible heat duties of the reboiler and the condenser.

Table (1) Flow rate and composition of the feed stream to the depropanizer.

Component	Flow rate kmole/hr	Mole Fraction
C_2H_6	2.399	0.0165
C_3H_6	77.64	0.5347
C_3H_8	21.9	0.1508
C_3H_4	1.29	0.0089
i- C_4H_{10}	32.09	0.2210
i- C_5H_{12}	4.15	0.0286
n- C_5H_{12}	5.74	0.0395
Total molar rate kmole/hr	145.2	1
Total mass rate kg/hr	9620.81	
Pressure bar	10	

The overhead stream must contain zero isobutane because this component affects the hydrogenation step before propylene-propane splitter. We have to find the minimum reflux ratio and the minimum number of stage to accomplish the separation at total reflux.

Hysys has separate calculations for shortcut method. This calculation has been conducted before steady state simulation of depropanizer. Propylene is assumed to

be the light key-component and propane is the heavy key-component. Hysys asked about other data to be supplied such as feed pressure = 10 bar, feed flow rate = 145.2 kmole/hr, feed temperature 38°C, propylene composition in the bottom product = 0.001 and propane composition in the distillate = 0.21. The results for the short-cut method are shown in Table (2).

With this information it is possible to conduct preliminary simulation based on equilibrium stage model or steady state model in Hysys, Constant pressure operation is assumed.

RESULTS AND DISCUSSION

First consideration is to assume a certain number of stages (50, 55 and 60) and for each assumption the reflux ratio is changed (0.8, 0.9, 1.0, 1.05, 1.1 and 2) and the results are drawn in Figures (1, 2, and 3).

Figure (1), shows the propylene mole fraction in distillate for different reflux ratio for each number of stages. Higher composition can be reached at reflux ratio = 1 and N = 60.

Figure (2), shows the condenser duty for different reflux ratio at each number of stages, as reflux ratio increased, the condenser duty is increased and the condenser duty is higher for N=60 for all reflux ratios.

Figure (3) shows the reboiler duty for different reflux ratio at each number of stages. It shows the same trend as for the condenser duty but with higher values.

From the simulation results at reflux ratio 0.8 & 0.9 for each number of stages, it is noted that a fraction of isobutane appears in the top product which is not preferred, as mentioned before. The isobutane disappears from the top product when reflux ratio increased to 1 and also for the higher values.

Table (2) Results for the short-cut Method.

Component	Mole Frac.Feed	Mole Frac.Distillate	Mole Frac.Bottom
C ₂ H ₆	0.0165	0.0233	0
C ₃ H ₆	0.5347	0.7549	0.001
C ₃ H ₈	0.1508	0.2099	0.0078
C ₃ H ₄	0.0089	0.0119	0.0017
i-C ₄ H ₁₀	0.221	0	0.7564
i-C ₅ H ₁₂	0.028	0	0.0978
n-C ₅ H ₁₂	0.0395	0	0.1353
Flow rate kmole/hr.	145.2	102.8	42.42
Mass flow kg/hr	6935	4338	2598
Heat duty kcal/hr		6.167x10 ⁵	3.184x10 ⁵
Temp..C°	38	6.479	66.11
Pressure bar	10	7.19	8.25
External Reflux Ratio		0.6	
Minimum Reflux Ratio		0.155	
Minimum No. of stages		45	
Optimal feed tray		5-6 from top	

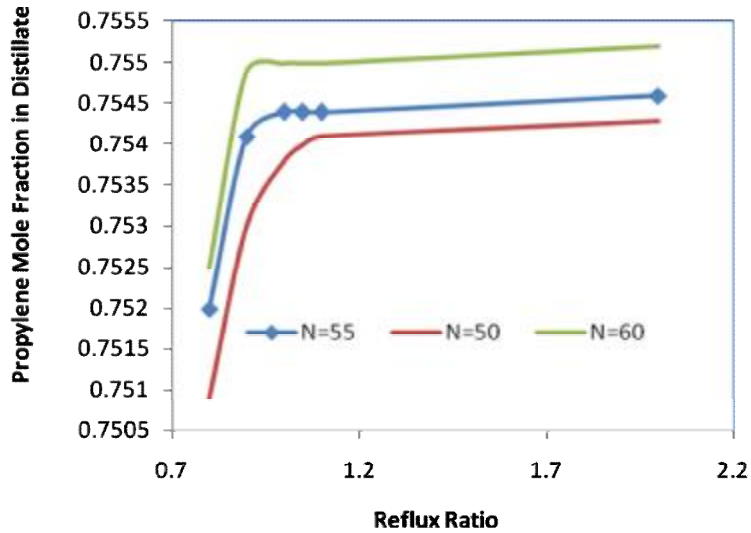


Figure (1) Propylene Profile vs. Reflux Ratio at Different Number of Trays.

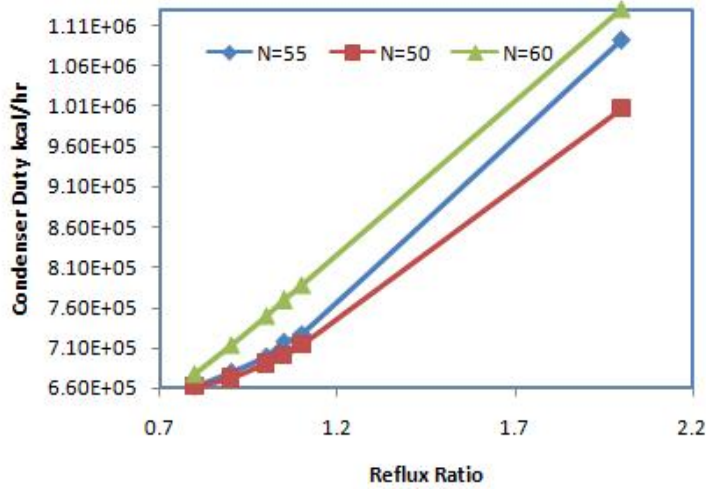


Figure (2) Condenser Duty vs. Reflux Ratio at Different Number of Trays.

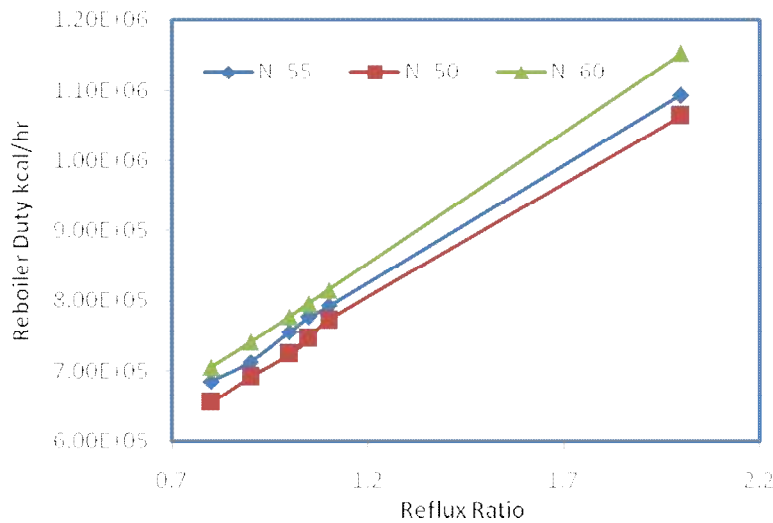


Figure (3) Reboiler Duty vs. Reflux Ratio at Different Number of Trays.

The feed location is studied at each number of stages and different reflux ratio. For number of stages $N=50$, Figure (4) shows the variation of the propylene composition in distillation with reflux ratio. Higher composition can be reached for feed location at stage 15 from the top, but for most reflux ratios a fraction of isobutane appeared, except for $R=1.1$ and higher, hence this location is not preferred. For feed location at 25th stage from top, it is found that only when $R=0.8$, the isobutane appears in the top product, and this is also observed for feed location at 32nd stage and 38th.

Figure (5) and Figure (6) show the heat duty for each feed location for $N=50$ as the feed moves to the top the reboiler, and condenser duties are increased. But at lower reflux ratio they have approximated values. The same has been done for each number of stages and they all show the same trends as for $N=50$, but the results are not shown. From all above, the optimum reflux ratio was found to be $R=1$.

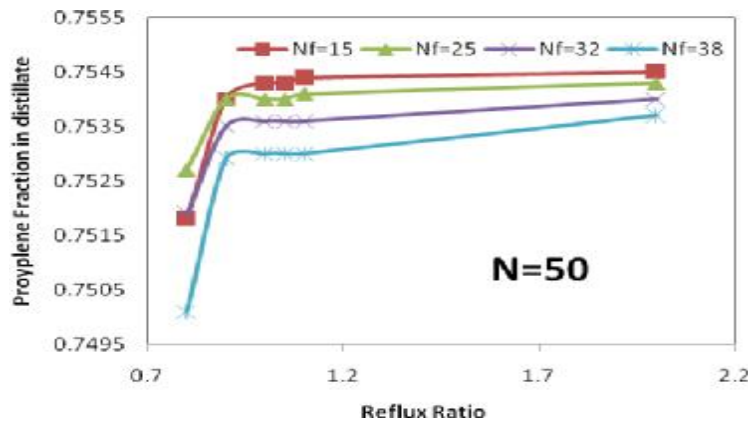


Figure (4) Propylene Composition in Distillate vs. Reflux Ratio at Different Feed Location for $N=50$.

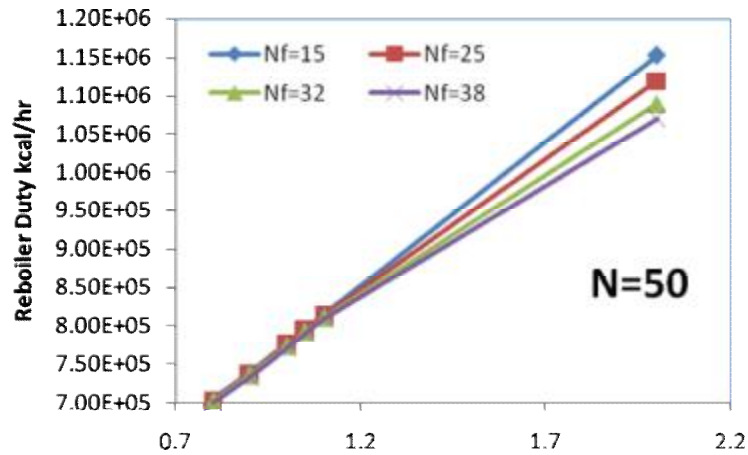


Figure (5) Reboiler Duty vs. Reflux Ratio at Different Feed Location for N=50.

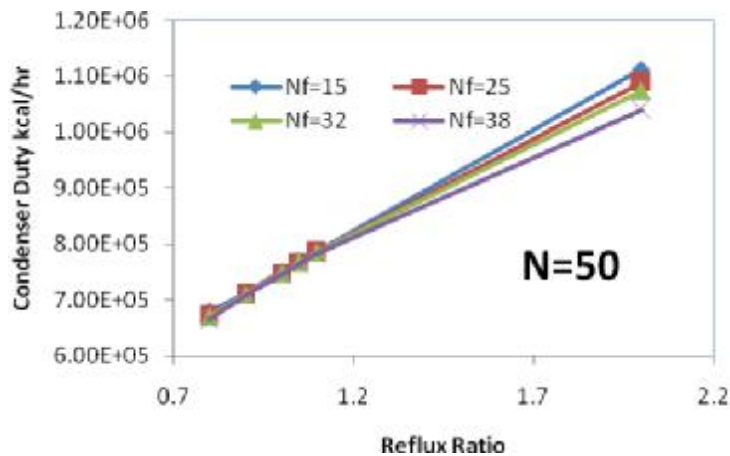


Figure (6) Condenser Duty vs. Reflux Ratio at Different Feed Location at N=50.

For the chosen reflux ratio, a comparison between the numbers of the stages is carried out. For each number of stages with fixed $R=1$, the propylene composition in distillate, the reboiler and condenser duty are plotted against feed location to choose the optimum number of stages and the best feed location. Figures 7 & 8 show the propylene mole fraction in the distillate, the reboiler and condenser duty against feed location at $N=50$. Location at 25th stage from the top gives higher composition which is equal to 0.75, and heat duty $Q_c=7.49 \times 10^5$ kcal/hr and $Q_r=7.76 \times 10^5$ kcal/hr

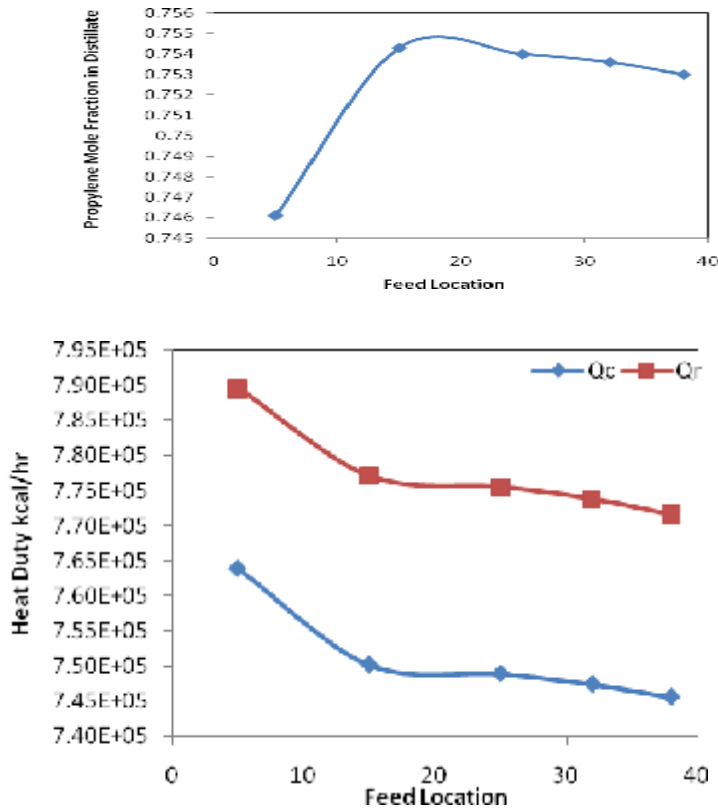


Figure (8) Feed Location vs. Heat Duty for N=50 and R=1

Figure (9 & 10) show the same variables but for N=55. The feed location at stage 25 also gives higher propylene composition equal to 0.7542, $Q_c=7.5 \times 10^5$ kcal/hr and $Q_r=7.74 \times 10^5$ kcal/hr.

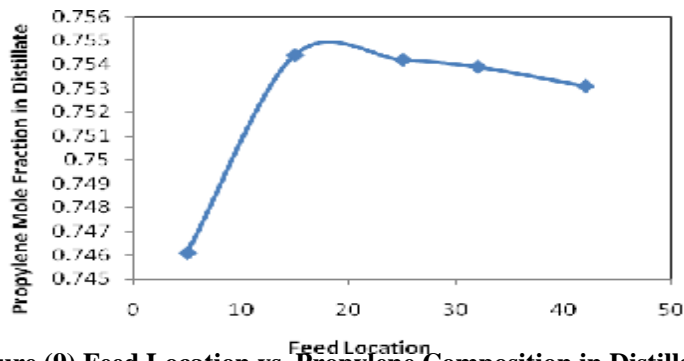


Figure (9) Feed Location vs. Propylene Composition in Distillate at N=55 and R=1.

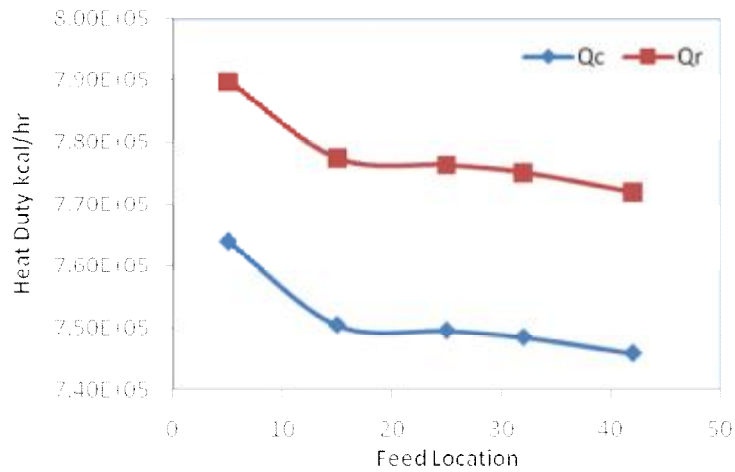


Figure (10) Feed Location vs. Heat Duty for N=55 and R=1.

Figure (11 & 12) show the same variables for N=60. It is found that the optimum location for the feed is 25th from top, which gives propylene composition = 0.7548, $Q_c = 7.48 \times 10^5$ kcal/hr and $Q_r = 7.776 \times 10^5$ kcal/hr.

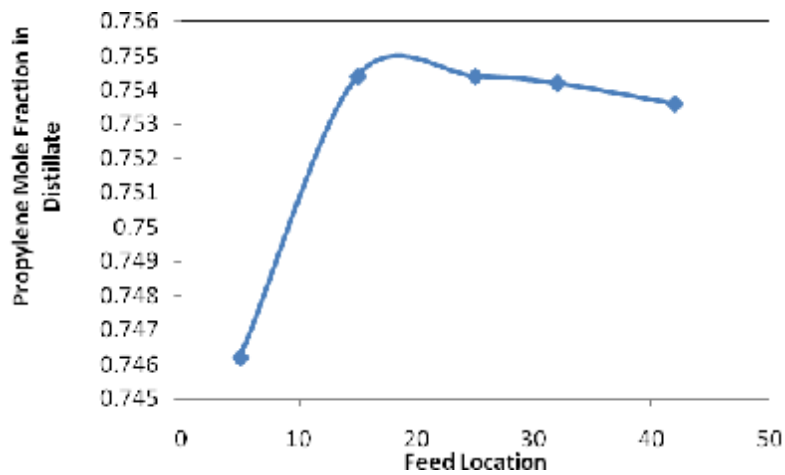


Figure (11) Feed Location vs. Propylene Composition in Distillate for N=60 and R=1.

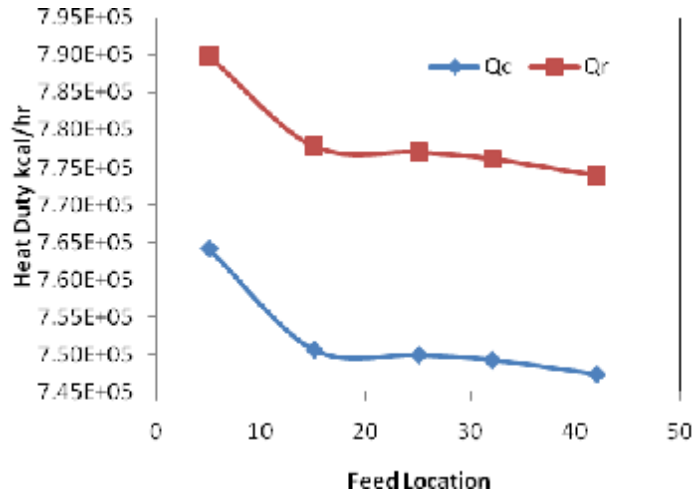


Figure (12) Feed Location vs. Heat Duty for N=60 and R=1.

From all above values, and taking into consideration that isobutane does not exist in the top product, the optimum number of stages is 55, which gives the average values of propylene composition in distillate and the heat duties.

In the above studied cases, the feed temperature was fixed at 30°C. For the chosen configuration, which was found after many trials, the feed temperature has been changed to find its optimum value. The results are shown in table -3.

Table (3) Simulation results for different feed temp. kcal/hr mole fraction.

Feed Temp. C°	f _v	Propylene mole Fraction in Distillate	Isobutane mole Fraction in Distillate	Q _r kcal/hr	Q _c kcal/hr
30	0	0.7542	0	7.5x10 ⁵	7.74x10 ⁵
32	0.0847	0.7540	0	7.244x10 ⁵	7.494x10 ⁵
38	0.5116	0.7465	0.0097	7.6x10 ⁵	4.871x10 ⁵
40	0.7158	0.727	0.0352	7.9x10 ⁵	3.905x10 ⁵

From Table (3) the optimum feed temperature is 32°C. The optimum design variables for the depropanizer are N=55, feed location=25th from top and feed temperature =32°C. Appendix B gives the full design for the chosen column with all specifications and the PFD using Hysys printout results. Table-4 shows the process flow data sheet for the depropanizer which was designed.

To complete the design, Hysys utilities are used for tray sizing. Hysys has a utility program to perform a mechanical design of distillation columns; both tray and packed. Three types of trays are offered; valve tray, sieve tray and bubble cap tray. Valve tray was chosen for the depropanizer. The column diameter is estimated

to be (0.762 m) with maximum flooding of 73%, all the important results for tray sizing are shown in appendix C.

Table (4) Process Flow Data Sheet for the Designed Depropanizer.

Component	Feed	Distillate	Vapor to Condenser	Reflux to Column	Bottom	Vapor from Reboile
C ₂ H ₆	2.399	2.399	4.798	2.399	0	0
C ₃ H ₆	77.64	77.447	154.9	77.447	0.1933	1.754
C ₃ H ₈	21.9	21.568	43.14	21.568	0.3318	2.967
C ₃ H ₄	1.29	1.2749	2.55	1.275	0.0151	0.139
i-C ₄ H ₁₀	32.09	0	0	0	32.09	150.4
i-C ₅ H ₁₂	4.15	0	0	0	4.15	8.174
n-C ₅ H ₁₂	5.74	0	0	0	5.74	9.5
Total						
kmole/hr	145.2	102.689	205.388	102.689	45.42	172.934
kg/hr	6935	4333	8667	4333	2602	10229.3
Mwt. avge.	47.762	42.199	41.635	42.199	61.198	59.15
Temp. C°	32	12.57	12.569	12.569	72.36	72.36
P bar	10	8.5	8.5	8.5	9.5	9.5
Vapor Frac f _v	0.0847	0	1	0	0	1
Total Number of Stages	55					
Feed Location	25 th from top					
Reflux Ratio	1.0					
Reboiler Duty kcal/hr	7.244x10 ⁵					
Condenser Duty kcal/hr	7.494x10 ⁵					

CONCLUSIONS

Based on the process design optimization, the following may be concluded:

1. The short-cut method is important for primary design prior to performing the rigorous tray-to-tray calculations, because the preliminary calculations give an idea of what is reasonable, such as minimum reflux ratio and minimum number of stages. It is found that the minimum reflux ratio is 0.155 and minimum number of stages is 45.
2. The optimum number of stages for the depropanizer is found to be 55 stages. The feed location is at stage 25th from top, and the feed temperature is 32°C. These values are chosen from the simulation results. This configuration gives zero concentration of isobutane in the propylene-propane top product and fairly acceptable heat duty compared to other configurations. The presence of isobutane in top product of the depropanizer affects the catalyst and the reaction of the hydrogenator, to increase the yield of propylene before the propylene propane splitter.

3. Hysys has been tested for the same data of the designed, yet unimplemented, depropanizer of Lummus Company (1976), the results have shown good compatibility with Lummus results.
4. Hysys is also used for tray sizing, and the valve-trays have been chosen. Other efficient packages for the tray design can be used as well, which offer new tray technology, such as Ultra-capacity tray, Superfrac tray, Flexi tray ...etc.

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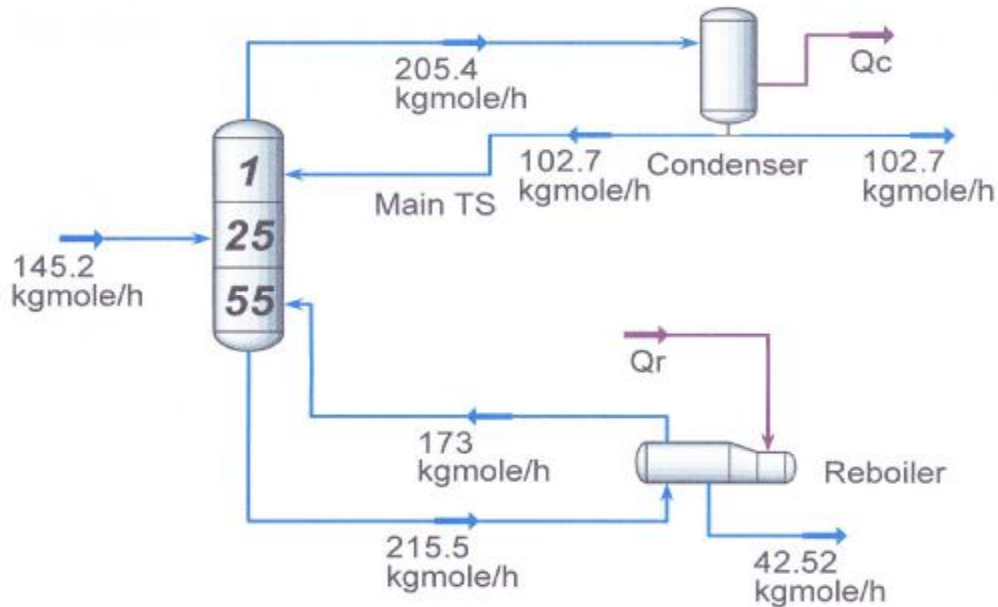
APPENDICES


**Appendix A:
 Composition of the design results for the "Lummus"
 depropanizer (1976) and Hysys simulation results.**


Comp.	X _f	Lummus X _D	Hysys X _D	Lummus X _w	Hysys X _w
C ₂ H ₆	0.0032	0.0045	0.0043	0	0
C ₃ H ₆	0.5920	0.8239	0.8245	0.0052	0.0055
C ₃ H ₈	0.1155	0.1544	0.1532	0.0218	0.02
C ₃ H ₄	0.0099	0.0119	0.0110	0.0054	0.0065


i-C ₄ H ₁₀	0.0037	0.0003	0.0002	0.0125	0.0123
C ₄ H ₈	0.0274	0.0008	0.0011	0.0945	0.0927
C ₄ H ₆	0.1379	0.0039	0.0045	0.4799	0.4727
n-C ₄ H ₁₀	0.0298	0.0003	0.0002	0.1021	0.1034
i-C ₅ H ₁₂	0.0282	0	0	0.0935	0.1007
n-C ₅ H ₁₂	0.0523	0	0	0.1851	0.1862
Molar flow kmol/hr	61.29	42.82	44.11	18.47	17.18
Mass flow kg/hr	2896	1814.28	1872	1104.7	1023
Mwt	47.25	42.37	42.44	59.81	59.55
Temp. C°	31.4	11.7	12.5	72.9	76.65
P bar	8.47	7.19	8.113	8.25	9.103
Reflux Ratio	1.2				
Number of stages	26				
Feed location	4 th from top				
Q _C kcal/hr	Lummus = 3.253x10 ⁵		Hysys = 3.518x10 ⁵		
Q _r kcal/hr	Lummus = 2.768x10 ⁵		Hysys = 2.891x10 ⁵		


Appendix B:
Hysys report documents for the whole design of the depropanizer of N=55, Feed Location 25th from top and Feed Temperature = 32 C°

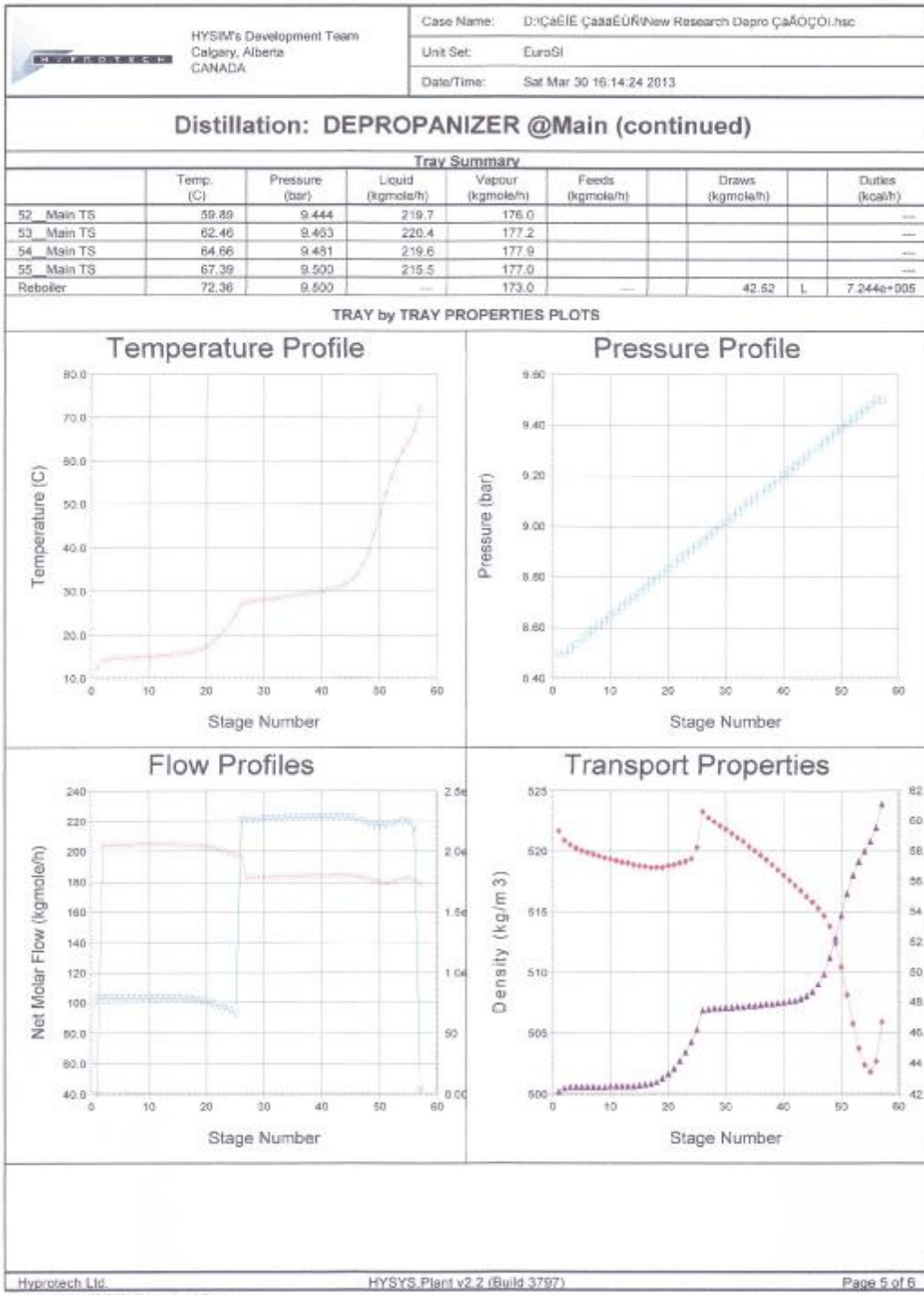


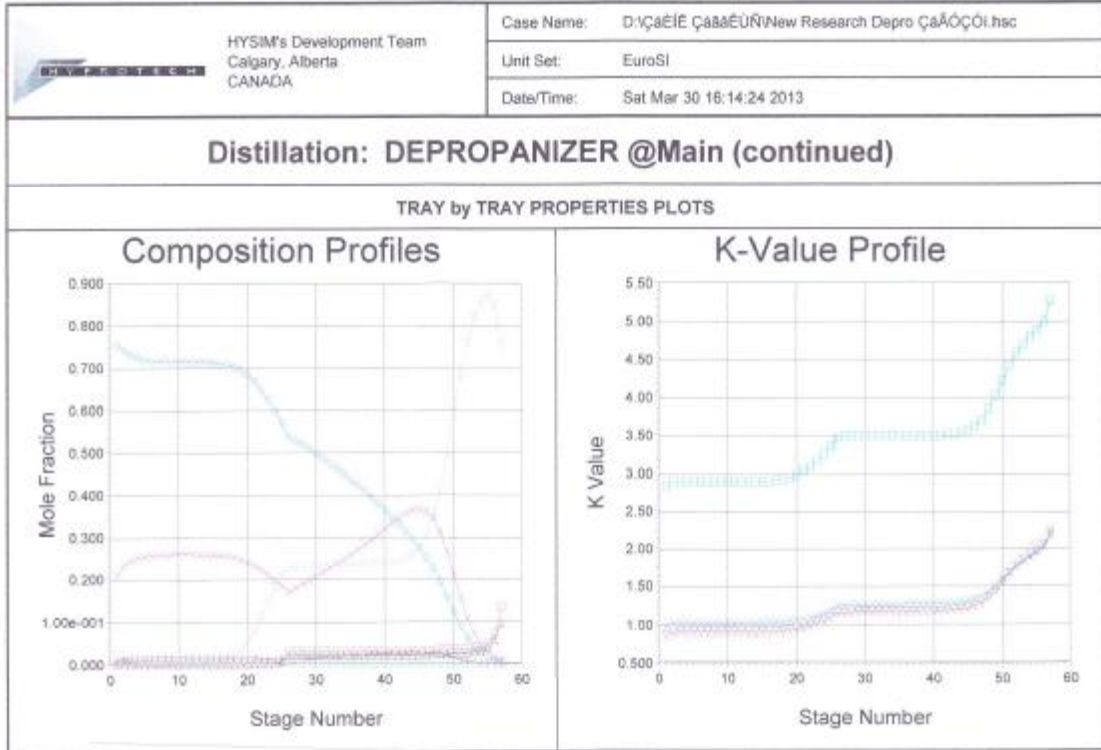
 HYSIM's Development Team Calgary, Alberta CANADA		Case Name: D:\ÇaEİE ÇääâEÜN\New Research Depro ÇääÖÇÖi.hsc Unit Set: EuroSI Date/Time: Sat Mar 30 18:14:24 2013						
Distillation: DEPROPANIZER @Main								
CONNECTIONS								
Inlet Stream								
STREAM NAME	Stage	FROM UNIT OPERATION						
Or Feed	Reboiler 25_Main TS							
Outlet Stream								
STREAM NAME	Stage	TO UNIT OPERATION						
Qc buta propa	Condenser Reboiler Condenser							
SPECS								
Column Specification Parameters								
Reflux Ratio								
Stage:	Condenser	Flow Basis:	Molar Liquid Specification: ---					
Reflux Rate								
Stage:	Condenser	Flow Basis:	Molar Liquid Specification: ---					
Btms Prod Rate								
Stream:	buta	Flow Basis:	Molar					
Distillate Rate								
Stream:	propa	Flow Basis:	Molar					
Comp Fraction								
Stage:	Condenser	Flow Basis:	Mole Fraction Phase: Liquid					
Components:	i-Butane							
Comp Fraction - 2								
Stage:	Reboiler	Flow Basis:	Mole Fraction Phase: Liquid					
Components:	Propane							
Temperature								
Stage:	16_Main TS							
MONITOR								
Specifications Summary								
	Specified Value	Current Value	Wt. Error	Wt. Tol	Abs. Tol	Active	Estimate	Used
Reflux Ratio	1.000	1.000	1.414e-008	1.000e-002	1.000e-002	On	On	On
Reflux Rate	---	102.7 kgmole/h	---	1.000e-002	1.000 kgmole/h	Off	On	Off
Btms Prod Rate	---	42.52 kgmole/h	---	1.000e-002	1.000 kgmole/h	Off	On	Off
Distillate Rate	102.0 kgmole/h	102.7 kgmole/h	6.755e-003	1.000e-002	1.000 kgmole/h	Off	On	Off
Comp Fraction	1.000e-004	9.365e-007	-0.8091	1.000e-002	1.000e-003	Off	On	Off
Comp Fraction - 2	7.800e-003	7.803e-003	1.742e-004	1.000e-002	1.000e-003	On	On	On
Temperature	35.00 C	16.06 C	-3.788e-002	1.000e-002	1.000 C	Off	On	Off
PROFILES								
General Parameters								
Sub-Flow Sheet:	DEPROPANIZER (COL1)	Number of Stages:	55					
Profile Estimates								
	Temperature (C)	Net Liquid (kgmole/h)	Net Vapour (kgmole/h)					
Condenser	12.58	102.8	1.132e-018					
1_Main TS	14.02	103.0	205.6					
2_Main TS	14.43	103.0	205.7					
3_Main TS	14.55	103.0	205.6					
4_Main TS	14.63	103.0	205.8					
5_Main TS	14.81	102.9	205.7					
6_Main TS	15.00	102.7	205.6					
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 HYSYS's Development Team Calgary, Alberta CANADA		Case Name: D:\CASE\Case6\New Research Depro Ca\OC\O\hsc	
		Unit Set: EuroSI	
		Date/Time: Sat Mar 30 16:14:24 2013	
Distillation: DEPROPANIZER @Main (continued)			
Profile Estimates			
	Temperature (C)	Net Liquid (kgmole/h)	Net Vapour (kgmole/h)
7_Main TS	15.25	102.5	205.5
8_Main TS	15.59	102.1	205.2
9_Main TS	16.07	101.5	204.6
10_Main TS	16.74	100.6	204.2
11_Main TS	17.67	99.46	203.4
12_Main TS	18.66	98.11	202.2
13_Main TS	20.30	96.63	200.9
14_Main TS	21.93	94.94	199.4
15_Main TS	23.90	92.48	197.7
16_Main TS	26.42	220.0	195.3
17_Main TS	25.55	220.4	177.6
18_Main TS	15.59	101.4	204.7
19_Main TS	17.58	100.5	204.1
20_Main TS	18.52	99.36	203.2
21_Main TS	19.72	98.02	202.0
22_Main TS	27.15	220.6	178.0
23_Main TS	27.36	220.8	178.2
24_Main TS	27.54	221.0	178.4
25_Main TS	27.73	221.2	178.5
26_Main TS	27.91	221.3	178.7
27_Main TS	25.10	221.5	178.9
28_Main TS	25.30	221.6	179.0
29_Main TS	25.50	221.8	179.2
30_Main TS	25.70	221.9	179.3
31_Main TS	25.90	222.1	179.5
32_Main TS	29.11	222.2	179.5
33_Main TS	29.33	222.4	179.9
34_Main TS	29.55	222.5	179.0
35_Main TS	29.77	222.7	180.1
36_Main TS	30.00	222.9	180.3
37_Main TS	30.24	223.0	180.4
38_Main TS	30.49	223.2	180.5
39_Main TS	30.75	223.3	180.7
40_Main TS	31.05	223.4	180.9
41_Main TS	31.41	223.5	181.0
42_Main TS	31.87	223.4	181.0
43_Main TS	32.32	223.1	181.0
44_Main TS	33.50	222.5	180.7
45_Main TS	35.00	221.5	180.1
46_Main TS	37.26	220.1	179.1
47_Main TS	40.47	218.5	177.6
48_Main TS	44.57	217.4	176.1
49_Main TS	49.17	217.2	175.0
50_Main TS	53.67	217.9	174.5
51_Main TS	57.55	219.1	175.4
52_Main TS	60.60	220.2	176.6
53_Main TS	62.93	220.6	177.5
54_Main TS	64.87	219.9	178.4
55_Main TS	67.59	215.7	177.4
Reboiler	72.49	42.43	173.3
PROPERTIES			
Properties : Feed			
	Overall	Vapour Phase	Liquid Phase
Vapour/Phase Fraction	0.0847	0.0847	0.9153
Temperature: (C)	32.00	32.00	32.00
Pressure: (bar)	10.00	10.00	10.00
Molar Flow (kgmole/h)	145.2	12.30	132.9
Mass Flow (kg/h)	6935	540.2	6395
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 HYSYS's Development Team Calgary, Alberta CANADA	Case Name:	D:\Çalışmalar\New Research Depro ÇAĞUÇOI.hsc		
	Unit Set:	EuroSI		
	Date/Time:	Sat Mar 30 16:14:24 2013		
Distillation: DEPROPANIZER @Main (continued)				
Properties : Feed				
	Overall	Vapour Phase	Liquid Phase	
Liquid Volume Flow (m ³ /h)	12.92	1.041	11.87	
Molar Enthalpy (kcal/kgmole)	-1.428e+004	-5603	-1.508e+004	
Mass Enthalpy (kcal/kg)	-298.9	-127.6	-313.4	
Molar Entropy (kJ/kgmole-C)	66.57	104.9	63.02	
Mass Entropy (kJ/kg-C)	1.394	2.390	1.310	
Heat Flow (kcal/h)	-2.073e+008	-6.894e+004	-2.004e+006	
Molar Density (kgmole/m ³)	3.790	0.4734	10.76	
Mass Density (kg/m ³)	181.0	20.79	519.1	
Std Liquid Mass Density (kg/m ³)	542.4	524.1	544.0	
Molar Heat Capacity (kJ/kgmole-C)	123.7	78.72	127.9	
Mass Heat Capacity (kJ/kg-C)	2.591	1.793	2.658	
Thermal Conductivity (W/m-K)	---	1.890e-002	9.633e-002	
Viscosity (cP)	---	9.101e-003	0.1006	
Surface Tension (dyne/cm)	---	---	7.212	
Molecular Weight	47.76	43.91	48.12	
Z Factor	---	0.8326	3.654e-002	
Properties : propa				
	Overall	Vapour Phase	Liquid Phase	
Vapour/Phase Fraction	0.0000	0.0000	1.0000	
Temperature: (C)	12.57	12.57	12.57	
Pressure: (bar)	8.500	8.500	8.500	
Molar Flow (kgmole/h)	102.7	0.0000	102.7	
Mass Flow (kg/h)	4333	0.0000	4333	
Liquid Volume Flow (m ³ /h)	8.414	0.0000	8.414	
Molar Enthalpy (kcal/kgmole)	-5498	-2223	-5499	
Mass Enthalpy (kcal/kg)	-130.3	-53.40	-130.3	
Molar Entropy (kJ/kgmole-C)	38.26	94.23	38.26	
Mass Entropy (kJ/kg-C)	0.9066	2.263	0.9066	
Heat Flow (kcal/h)	-5.647e+005	0.0000	-5.647e+005	
Molar Density (kgmole/m ³)	12.36	0.4228	12.36	
Mass Density (kg/m ³)	521.7	17.60	521.7	
Std Liquid Mass Density (kg/m ³)	517.9	513.2	517.9	
Molar Heat Capacity (kJ/kgmole-C)	110.4	69.84	110.4	
Mass Heat Capacity (kJ/kg-C)	2.615	1.677	2.615	
Thermal Conductivity (W/m-K)	0.1122	1.714e-002	0.1122	
Viscosity (cP)	8.683e-002	8.585e-003	8.683e-002	
Surface Tension (dyne/cm)	8.175	---	8.175	
Molecular Weight	42.20	41.64	42.20	
Z Factor	2.894e-002	0.8483	2.894e-002	
Properties : buta				
	Overall	Vapour Phase	Liquid Phase	
Vapour/Phase Fraction	0.0000	0.0000	1.0000	
Temperature: (C)	72.36	72.36	72.36	
Pressure: (bar)	9.500	9.500	9.500	
Molar Flow (kgmole/h)	42.52	0.0000	42.52	
Mass Flow (kg/h)	2602	0.0000	2602	
Liquid Volume Flow (m ³ /h)	4.502	0.0000	4.502	
Molar Enthalpy (kcal/kgmole)	-3.607e+004	-3.117e+004	-3.607e+004	
Mass Enthalpy (kcal/kg)	-589.3	-527.1	-589.3	
Molar Entropy (kJ/kgmole-C)	107.8	156.7	107.8	
Mass Entropy (kJ/kg-C)	1.762	2.650	1.762	
Heat Flow (kcal/h)	-1.534e+006	0.0000	-1.534e+006	
Molar Density (kgmole/m ³)	8.266	0.4093	8.266	
Mass Density (kg/m ³)	505.8	24.21	505.9	
Std Liquid Mass Density (kg/m ³)	580.4	570.1	580.4	
Molar Heat Capacity (kJ/kgmole-C)	172.9	122.3	172.9	
Mass Heat Capacity (kJ/kg-C)	2.825	2.088	2.825	
Thermal Conductivity (W/m-K)	7.222e-002	2.114e-002	7.222e-002	
Viscosity (cP)	0.1152	9.204e-003	0.1152	
Surface Tension (dyne/cm)	6.092	---	6.092	
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 HYSIM's Development Team Calgary, Alberta CANADA		Case Name: D:\ÇaEİE ÇâââEÜÑNew Research Depro ÇâÄÇÖİ.hsc Unit Set: EuroSI Date/Time: Sat Mar 30 16:14:24 2013	
Distillation: DEPROPANIZER @Main (continued)			
Properties : buta			
	Overall	Vapour Phase	Liquid Phase
Molecular Weight	61.20	59.14	61.20
Z Factor	4.001e-002	0.8079	4.001e-002
SUMMARY			
Tray Summary			
Flow Basis:	Molar		Reflux Ratio: 1.003
	Temp. (C)	Pressure (bar)	Liquid (kgmole/h)
			Vapour (kgmole/h)
			Feeds (kgmole/h)
			Draws (kgmole/h)
			Duties (kcal/h)
Condenser	12.57	8.500	102.7
1 Main TS	14.00	8.500	102.9
2 Main TS	14.39	8.519	103.0
3 Main TS	14.55	8.537	103.0
4 Main TS	14.66	8.556	103.0
5 Main TS	14.75	8.574	103.0
6 Main TS	14.84	8.593	103.0
7 Main TS	14.92	8.611	103.0
8 Main TS	15.01	8.630	103.1
9 Main TS	15.09	8.648	103.1
10 Main TS	15.18	8.667	103.1
11 Main TS	15.27	8.685	103.1
12 Main TS	15.38	8.704	103.0
13 Main TS	15.49	8.722	103.0
14 Main TS	15.63	8.741	102.9
15 Main TS	15.81	8.759	102.7
16 Main TS	16.06	8.778	102.5
17 Main TS	16.40	8.796	102.0
18 Main TS	16.89	8.815	101.4
19 Main TS	17.58	8.833	100.5
20 Main TS	18.52	8.852	99.36
21 Main TS	19.72	8.870	98.02
22 Main TS	21.14	8.889	96.57
23 Main TS	22.74	8.907	94.92
24 Main TS	24.57	8.926	92.50
25 Main TS	27.13	8.944	220.8
26 Main TS	27.58	8.963	221.2
27 Main TS	27.85	8.981	221.4
28 Main TS	28.04	9.000	221.6
29 Main TS	28.22	9.019	221.7
30 Main TS	28.40	9.037	221.9
31 Main TS	28.58	9.056	222.0
32 Main TS	28.76	9.074	222.2
33 Main TS	28.94	9.093	222.3
34 Main TS	29.13	9.111	222.5
35 Main TS	29.33	9.130	222.6
36 Main TS	29.53	9.148	222.7
37 Main TS	29.74	9.167	222.9
38 Main TS	29.96	9.185	223.0
39 Main TS	30.20	9.204	223.1
40 Main TS	30.46	9.222	223.2
41 Main TS	30.76	9.241	223.3
42 Main TS	31.19	9.259	223.2
43 Main TS	31.77	9.278	222.9
44 Main TS	32.63	9.296	222.4
45 Main TS	33.97	9.315	221.4
46 Main TS	36.04	9.333	220.0
47 Main TS	39.06	9.352	218.4
48 Main TS	43.07	9.370	217.2
49 Main TS	47.72	9.389	216.7
50 Main TS	52.42	9.407	217.3
51 Main TS	56.57	9.426	218.5
			145.2
			M





Appendix C

 HYSIM's Development Team Calgary, Alberta CANADA		Case Name: D:\Ç&ÉÉ Ç&â&É&Ö&R\New Research Depro Ç&â&Ö&Ç&Ö&L.hsc Unit Set: EuroSI Date/Time: Sat Mar 30 16:06:08 2013	
Tray Sizing: Tray Sizing-1			
SETUP			
Tray Section:	Main TS @COL1	Liquid Draw:	0.00 % Sieve Tray Flooding Method: Minimum Csb
Section_1			
Section Start	1 Main TS		
Section End	55 Main TS		
Internals	Valve		
Mode	Design		
Active	OFF		
Status	Complete		
Design Limit	Flooding		
Limiting Stage	25 Main TS		
SPECIFICATIONS			
Section_1			
Section Start	1 Main TS		
Section End	55 Main TS		
Internals	Valve		
Mode	Design		
Number of Flow Paths	---		
Section Diameter (m)	---		
Tray for Properties	---		
Tray Spacing (mm)	609.6		
Tray Thickness (mm)	3.175		
Foaming Factor	1.000		
Max Delta P (ht of liq)	152.4 mm		
Max Flooding (%)	85.00		
Packing Correlation	---		
HETP (m)	---		
Packing Type	---		
TRAY INTERNALS			
Section_1			
Section Start	1 Main TS		
Section End	55 Main TS		
Internals	Valve		
Sieve Hole Pitch (mm)	---		
Sieve Hole Diameter (mm)	---		
Valve Mat's Density (kg/m3)	8220		
Valve Mat's Thickness (mm)	1.524		
Hole Area (% of AA)	15.30		
Valve Orifice Type	Straight		
Valve Design Manual	Gätsch		
Bubble Cap Slot Height (mm)	---		
Side Weir Type	Straight		
Weir Height (mm)	50.80		
Max Weir Loading (m3/h-m)	89.42		
Downcomer Type	Vertical		
Downcomer Clearance (mm)	38.10		
Max DC Backup (%)	50.00		
Side DC Top Width (mm)	---		
Side DC Bottom Width (mm)	---		
Centre DC Top Width (mm)	---		
Centre DC Bottom Width (mm)	---		
O.C. DC Top Width (mm)	---		
O.C. DC Bottom Width (mm)	---		
O.S. DC Top Width (mm)	---		
O.S. DC Bottom Width (mm)	---		
TRAY RESULTS			
Section_1			
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 HYSIM's Development Team Calgary, Alberta CANADA		Case Name: D:\Ç&ÉÉ Ç&â&ÉÜ\New Research Depro Ç&â&ÖÇ&Öi.hsc Unit Set: EuroSI Date/Time: Sat Mar 30 16:06:08 2013
Tray Sizing: Tray Sizing-1 (continued)		
Section Start		1 Main TS
Section End		55 Main TS
Internals		Valve
Section Diameter (m)		0.7620
Max Flooding (%)		73.02
X-Sectional Area (m ²)		0.4560
Section Height (m)		33.53
Section DeltaP (bar)		0.2671
NFP		1
Flow Length (mm)		444.5
Flow Width (mm)		716.3
Max DC Backup (%)		35.37
Max Weir Load (m ³ /h-m)		41.48
Max DP/ Tray (bar)		5.631e-003
Tray Spacing (mm)		609.6
Total Weir Length (mm)		616.9
Weir Height (mm)		50.90
Active Area (m ²)		0.3184
DC Clearance (mm)		36.10
DC Area (m ²)		6.863e-002
Side Weir Length (m)		0.6189
Hole Area (m ²)		4.871e-002
Estimated # of Holes/Valves		42
Relief Area (m ²)		0.0000
Relief - S (mm)		---
Relief - A (mm)		---
Relief - B (mm)		---
Side DC Top Width (mm)		158.6
Side DC Btm Width (mm)		158.6
Side DC Top Length (m)		0.6189
Side DC Btm Length (m)		0.6189
Side DC Top Area (m ²)		6.863e-002
Side DC Btm Area (m ²)		6.863e-002
Centre DC Top Width (mm)		0.0000
Centre DC Btm Width (mm)		0.0000
Centre DC Top Length (m)		0.0000
Centre DC Btm Length (m)		0.0000
Centre DC Top Area (m ²)		0.0000
Centre DC Btm Area (m ²)		0.0000
O.C. DC Top Width (mm)		0.0000
O.C. DC Btm Width (mm)		0.0000
O.C. DC Top Length (m)		0.0000
O.C. DC Btm Length (m)		0.0000
O.C. DC Top Area (m ²)		0.0000
O.C. DC Btm Area (m ²)		0.0000
O.S. DC Top Width (mm)		0.0000
O.S. DC Btm Width (mm)		0.0000
O.S. DC Top Length (m)		0.0000
O.S. DC Btm Length (m)		0.0000
O.S. DC Top Area (m ²)		0.0000
O.S. DC Btm Area (m ²)		0.0000
PACKED RESULTS		
Section_1		
Section Start		1 Main TS
Section End		55 Main TS
Internals		Valve
Section Diameter (m)		0.7620
Max Flooding (%)		73.02
X-Sectional Area (m ²)		0.4560
Section Height (m)		33.53
Section DeltaP (bar)		0.2671
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