CHAPTER IV
RESULTS AND DISCUSSION

4.1 Distillation Column Design

4.1.1 Process Flow Sheets

The feed composition consisted of six compositions, which were separated into three products; C2 and C3 as propane product, iC4, and nC4 and the higher hydrocarbons as LPG product. For three sharp splits of three-main-component mixture, two different sequences calculated from equation 2.10 were possible and for more complex design, scheme 1 and scheme 4 were added as other possible alternative schemes. All the process flow sheet was shown in Figure 4.1.

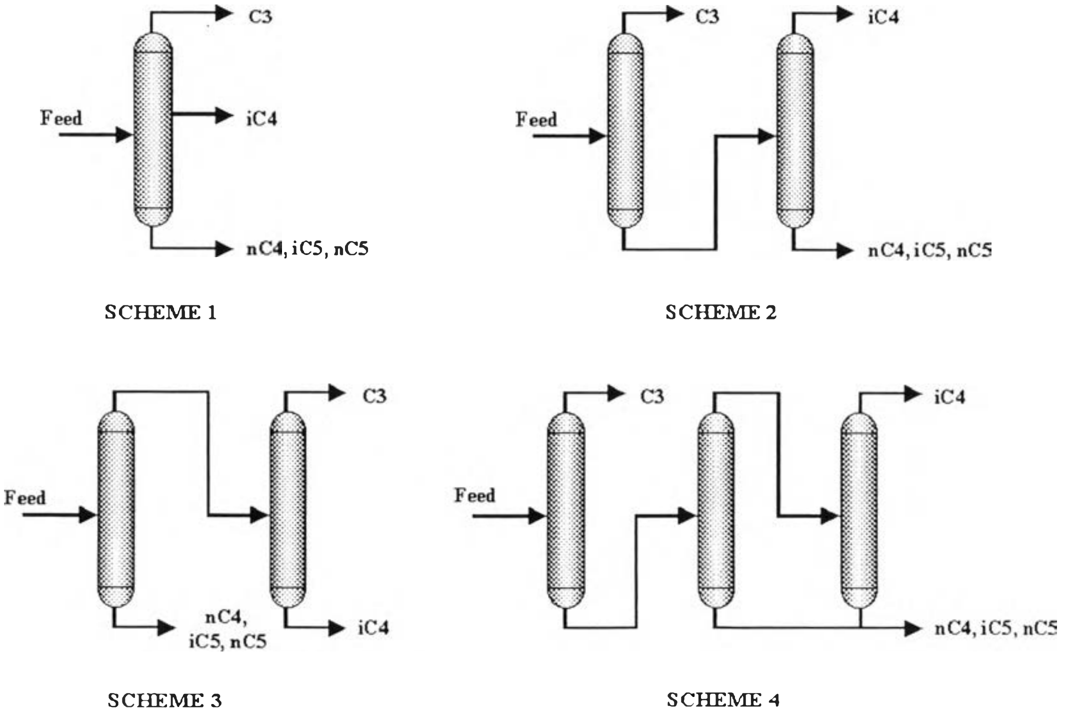

Figure 4.1 Possible column sequences for separating a three-main-component mixture into its pure products

4.1.2 Bubble Point Pressure of Product Stream

According to the algorithm for establishing distillation column pressure, the bubble point pressure of product stream at 110°F were calculated. The selected temperature of 110°F of product distillate depended on cooling water temperature, which was used as a coolant in the overhead condenser. The bubble point pressure of each product stream of all flow sheets is shown in Table 4.1.

4.1.3 Column Operating Pressure and Condenser Type

The column operating pressure and condenser type of all flow sheets were calculated by applying the specified column and condenser pressure drop of 5 psia. The results were shown in Tables 4.2-4.7. From the results, the column, which separated propane as a product, gave a high column pressure around 223-227 psia and the partial pressure was selected for this column. The column separating isobutane as a product gave a column pressure around 91 psia and the total condenser was selected. In scheme 3, the pressure of the feed stream of the second column was increased up to 226-230 psia by installing a pump. From the results in Tables 4.3, 4.5 and 4.7, all scheme have to install pump in order to increase pressure of the feed stream by increasing pressure of the feed 2 up to 230 psia, the feed 4 up to 228 psia and the feed 6 up to 229 psia.

4.1.4 Shortcut Design Summary

Selected results are shown for each shortcut column of all feed types in Tables 4.8 to 4.13. Feed tray location, number of theoretical stage, reflux ratios and condenser and reboiler duties have been calculated for a specified operating point at optimum reflux ratio which equals to 1.3 times minimum reflux ratio.

4.2 Distillation Column Sequencing

For all sequences of distillation column, the overall capacity variable of the process proposed as a cost indicator for vapor-liquid equilibrium separation process was calculated and shown in Figure 4.2.

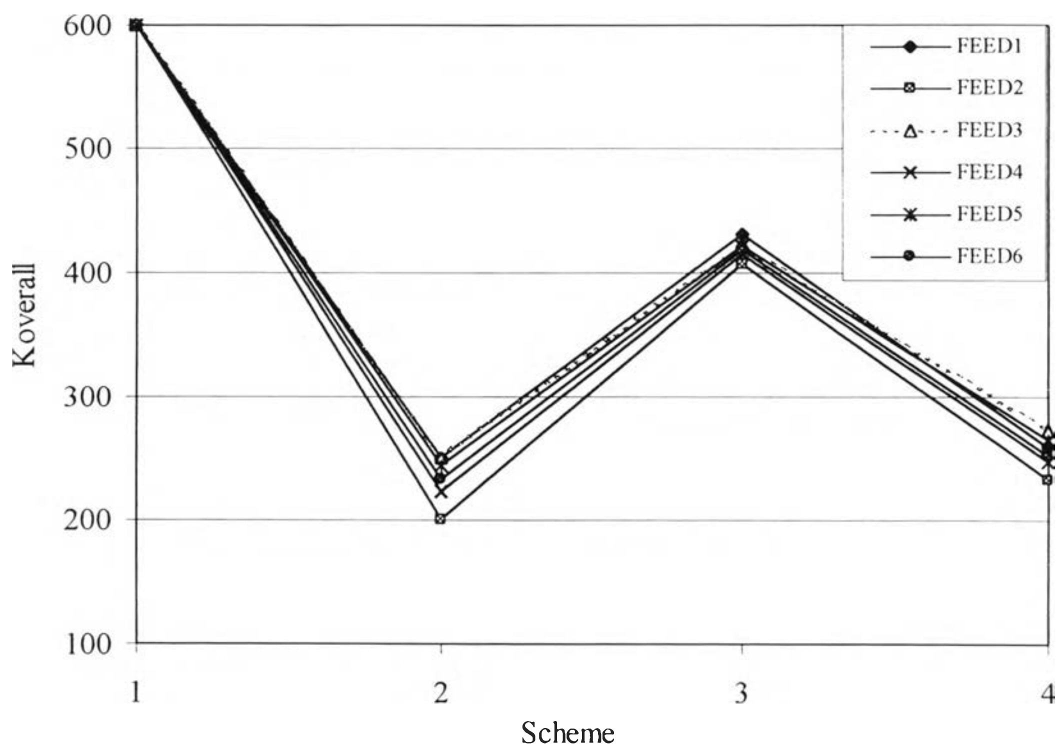

Figure 4.2 Overall capacity variable, K_{ov}, for the design column sequences

From the results, it showed that the best sequence of each feed stream was the scheme 2, which meant this scheme gave the minimum value of the overall boiling capacity. The next ones were schemes 4, 3 and 1, respectively.

4.3 Column Sizing

4.3.1 Column Rigorous Simulation and Feed Stage Correction

In PRO/II, the column sizing module was used with the rigorous column simulation. For developing the rigorous column simulation, the shortcut calculation results were used for specification of the rigorous column and used as the initial estimation values. When the rigorous simulation was developed, some errors from the shortcut calculation was found such as the incorrect values of feed stage location and the incorrect design calculation of the scheme 1 of all feed. The incorrect feed stage location was often occurred in the column separating isobutane from normal butane. The reason for this error came from the close values of the relative volatility, which was approximately 1.5. For the error in the scheme 1, the heat duty of condenser and reboiler in rigorous calculation was so high and impossible to meet this condition in the real operation. The error of the shortcut calculation are a mathematical calculation error due to the non key component calculation. The rigorous column simulation with feed stage correction was shown in Table 4.14.

4.3.2 Column Sizing

Tray sizing was performed for new columns with valve trays. All valve tray calculations use the methods of Glitsch by specifying a stage spacing of 24 inches, a system loading factor of 1 and a flooding factor of 78 percent. A pure transport method was chosen for calculating liquid densities. For sizing calculations, the diameter of each tray is adjusted to meet the flooding factor criterion. The sizing results of the selected design schemes were shown in Table 4.15.

4.4 Column Cost and Profitability Estimation

4.4.1 Capital Investment

From the column sequencing results, the best two schemes of all feed were the schemes 2 and 4. The total capital investment of this two schemes has been calculated in order to show that the column sequencing results were correct and considered as a factor for the investment decision. The total capital investments of the selected scheme of all feed were shown in Table 4.16.

From the results, it confirmed that the scheme 2 is the most economic scheme for isobutane separation compared to the other schemes. The investment costs of the distillation process with different feed streams were directly proportional to the feed capacity and were classified into 2 groups. The first one is the low cost group with 7-9 million dollars and the second one is the high cost group with 13-14 million dollars.

4.4.2 Return on Investment and Net Present Value

The comparison of the profitability of different processes having different capacity have used the return on investment (ROI) and the net present value (NPV) to indicate the profitability. The profitability can not be predicted with absolute accuracy. Risk factors and the degree of uncertainty involved in predicted ROI and NPV plays an important role in determining what return are acceptable. Peters and Timmerhaus (1991) stated that for moderate risk investment, a 20 percent return would be the minimum acceptable return for any type of industrial projects. The annual cost, annual income, annual profit, ROI and NPV value with the schemes 2 of all feed were shown in Table 4.17.

From the results, it showed that all design processes of selected schemes were acceptable for investment according to Peters and Timmerhaus

(1991). The recommended design processes, which gave the maximum net present value, were the designed processes with feed 1 and 5 for the low and high cost groups of investment, respectively. When compared to all design processes, the designed process of feed 5 was recommended.

Table 4.1 Bubble point pressure of distillate product temperature at 110°F

Components	Feed 1			Feed 2			Feed 3		
Top product stream	C2+C3	IC4	C2+C3+IC4	C2+C3	IC4	C2+C3+IC4	C2+C3	IC4	C2+C3+IC4
Flow rate, lbmol/hr	1155.8	386.6	1542.5	1061.6	268.4	1330.0	781.8	299.6	1081.4
Composition									
-Ethane	0.001	0	0	0.009	0	0	0.003	0	0.002
-Propane	0.999	0.003	0.749	0.991	0.003	0.792	0.997	0.003	0.722
-Isobutane	0	0.997	0.250	0	0.997	0.201	0	0.997	0.276
Temperature, F	110	110	110	110	110	110	110	110	110
Bubble point P, psia	217.7	85.6	178.9	221.4	85.6	189.0	218.6	85.6	175.7

Table 4.1 (Cont'd) Bubble point pressure of distillate product temperature at 110°F

Component	Feed 4			Feed 5			Feed 6		
Top product stream	C2+C3	IC4	C2+C3+IC4	C2+C3	IC4	C2+C3+IC4	C2+C3	IC4	C2+C3+IC4
Flow rate, lbmol/hr	2217.1	665.4	2872.5	1937.6	686.3	2623.9	1843.6	567.9	2411.4
Composition									
-Ethane	0.005	0	0.003	0.001	0	0.001	0.006	0	0.005
-Propane	0.995	0.003	0.769	0.999	0.003	0.738	0.994	0.003	0.760
-Isobutane	0	0.997	0.227	0	0.997	0.261	0	0.997	0.235
Temperature, F	110	110	110	110	110	110	110	110	110
Bubble point P, psia	219.5	85.6	183.6	218.0	85.6	177.6	220.2	85.6	183.0

Table 4.2 Column operating pressure and condenser type of feed 1

Column Specifications	Scheme 1	Scheme 2		Scheme 3		Scheme 4		
	Col. 1	Col. 1	Col. 2	Col. 1	Col. 2	Col. 1	Col. 2	Col. 3
Top stage P. psia	223.0	223.0	91.0	184.0	223.0	223.0	94.0	91.0
Feed stage P, psia	225.5	225.5	93.5	186.5	225.5	225.5	96.5	93.5
Bottom stage P, psia	228.0	228.0	96.0	189.0	228.0	228.0	99.0	96.0
Condenser type	partial	partial	total	total	partial	partial	total	total

Table 4.3 Column operating pressure and condenser type of feed 2

Column Specifications	Scheme 1	Scheme 2		Scheme 3		Scheme 4		
	Col. 1	Col. 1	Col. 2	Col. 1	Col. 2	Col. 1	Col. 2	Col. 3
Top stage P, psia	227.0	227.0	91.0	194.0	227.0	227.0	94.0	91.0
Feed stage P, psia	229.5	229.5	93.5	196.5	229.5	229.5	96.5	93.5
Bottom stage P, psia	232.0	232.0	96.0	199.0	232.0	232.0	99.0	96.0
Condenser type	partial	partial	total	total	partial	partial	total	total

Table 4.4 Column operating pressure and condenser type of feed 3

Column Specifications	Scheme 1	Scheme 2		Scheme 3		Scheme 4		
	Col. 1	Col. 1	Col. 2	Col. 1	Col. 2	Col. 1	Col. 2	Col. 3
Top stage P, psia	224.0	224.0	91.0	181.0	224.0	224.0	94.0	91.0
Feed stage P, psia	226.5	226.5	93.5	183.5	226.5	226.5	96.5	93.5
Bottom stage P, psia	229.0	229.0	96.0	186.0	229.0	229.0	99.0	96.0
Condenser type	partial	partial	total	total	partial	partial	total	total

Table 4.5 Column operating pressure and condenser type of feed 4

Column Specifications	Scheme 1	Scheme 2		Scheme 3		Scheme 4		
	Col. 1	Col. 1	Col. 2	Col. 1	Col. 2	Col. 1	Col. 2	Col. 3
Top stage P, psia	225.0	225.0	91.0	189.0	225.0	225.0	94.0	91.0
Feed stage P, psia	227.5	227.5	93.5	191.5	227.5	227.5	96.5	93.5
Bottom stage P, psia	230.0	230.0	96.0	194.0	230.0	230.0	99.0	96.0
Condenser type	partial	partial	total	total	partial	partial	total	total

Table 4.6 Column operating pressure and condenser type of feed 5

Column Specifications	Scheme 1	Scheme 2		Scheme 3		Scheme 4		
	Col. 1	Col. 1	Col. 2	Col. 1	Col. 2	Col. 1	Col. 2	Col. 3
Top stage P, psia	223.0	223.0	91.0	183.0	223.0	223.0	94.0	91.0
Feed stage P, psia	225.5	225.5	93.5	185.5	225.5	225.5	96.5	93.5
Bottom stage P, psia	228.0	228.0	96.0	188.0	228.0	228.0	99.0	96.0
Condenser type	partial	partial	total	total	total	partial	total	total

Table 4.7 Column operating pressure and condenser type of feed 6

Column Specifications	Scheme 1	Scheme 2		Scheme 3		Scheme 4		
	Col. 1	Col. 1	Col. 2	Col. 1	Col. 2	Col. 1	Col. 2	Col. 3
Top stage P, psia	226.0	226.0	91.0	188.0	226.0	226.0	94.0	91.0
Feed stage P, psia	228.5	228.5	93.5	190.5	228.5	228.5	96.5	93.5
Bottom stage P, psia	231.0	231.0	96.0	193.0	231.0	231.0	99.0	96.0
Condenser type	partial	partial	total	total	partial	partial	total	total

Table 4.8 Shortcut design summary of feed 1

Scheme	Unit	Total tray	Feed tray	N/N_{min}	R/R_{min}	Reflux	Duty(MM Btu/hr)	
							Condenser	Reboiler
1	Col.1	96	-	-	-	-	-	-
2	Col.1	41	16	1.987	1.3	2.073	-13.73	25.43
	Col.2	102	39	1.801	1.3	7.965	-26.88	24.55
3	Col.1	112	65	1.905	1.3	2.886	-39.06	42.56
	Col.2	47	28	1.984	1.3	2.030	-13.41	21.32
4	Col.1	43	20	1.984	1.3	2.083	-13.77	25.46
	Col.2	65	5	1.844	1.3	5.180	-21.88	19.70
	Col.3	78	47	1.847	1.3	4.837	-17.50	17.42

Table 4.9 Shortcut design summary of feed 2

Scheme	Unit	Total tray	Feed tray	N/N_{min}	R/R_{min}	Reflux	Duty(MM Btu/hr)	
							Condenser	Reboiler
1	Col.1	89	-	-	-	-	-	-
2	Col.1	44	16	1.998	1.3	1.931	-11.67	22.43
	Col.2	85	42	1.804	1.3	8.008	-18.75	17.07
3	Col.1	114	71	1.925	1.3	2.520	-29.84	33.44
	Col.2	43	19	1.999	1.3	1.917	-11.58	18.56
4	Col.1	43	20	1.998	1.3	1.934	-11.69	22.44
	Col.2	65	5	1.844	1.3	5.200	-15.29	13.71
	Col.3	80	50	1.846	1.3	4.858	-12.19	12.14

Table 4.10 Shortcut design summary of feed 3

Scheme	Unit	Total tray	Feed tray	N/N_{min}	R/R_{min}	Reflux	Duty(MM Btu/hr)	
							Condenser	Reboiler
1	Col.1	102	-	-	-	-	-	-
2	Col.1	42	21	1.983	1.3	2.102	-9.39	18.11
	Col.2	94	71	1.804	1.3	7.801	-20.43	18.68
3	Col.1	112	65	1.899	1.3	3.000	-28.47	31.48
	Col.2	41	17	1.985	1.3	2.091	-9.35	14.83
4	Col.1	43	20	1.982	1.3	2.102	-9.395	18.11
	Col.2	65	5	1.846	1.3	5.086	-16.61	14.96
	Col.3	78	48	1.848	1.3	4.814	-13.50	13.44

Table 4.11 Shortcut design summary of feed 4

Scheme	Unit	Total tray	Feed tray	N/N_{min}	R/R_{min}	Reflux	Duty(MM Btu/hr)	
							Condenser	Reboiler
2	Col.1	42	19	1.992	1.3	2.008	-25.41	47.84
	Col.2	92	59	1.802	1.3	7.986	-45.65	41.63
3	Col.1	113	68	1.913	1.3	2.728	-69.04	76.14
	Col.2	43	18	1.995	1.3	1.968	-24.92	39.79
4	Col.1	43	20	1.990	1.3	2.011	-25.44	47.87
	Col.2	65	5	1.844	1.3	5.188	-37.16	33.40
	Col.3	79	49	1.847	1.3	4.846	-29.70	29.56

Table 4.12 Shortcut design summary of feed 5

Scheme	Unit	Total tray	Feed tray	N/N_{min}	R/R_{min}	Reflux	Duty(MM Btu/hr)	
							Condenser	Reboiler
2	Col.1	42	21	1.983	1.3	2.098	-23.27	43.66
	Col.2	94	69	1.803	1.3	7.892	-47.30	43.23
3	Col.1	112	65	1.901	1.3	2.951	-67.82	74.34
	Col.2	41	17	1.989	1.3	2.045	-22.69	36.06
4	Col.1	43	20	1.983	1.3	2.099	-23.28	43.67
	Col.2	65	5	1.845	1.3	5.139	-38.49	34.68
	Col.3	78	47	1.848	1.3	4.826	-30.99	30.85

Table 4.13 Shortcut design summary of feed 6

Scheme	Unit	Total tray	Feed tray	N/N_{min}	R/R_{min}	Reflux	Duty(MM Btu/hr)	
							Condenser	Reboiler
2	Col.1	42	20	1.990	1.3	2.022	-21.24	40.71
	Col.2	95	71	1.803	1.3	7.905	-39.21	35.77
3	Col.1	113	68	1.911	1.3	2.757	-58.62	65.22
	Col.2	43	18	1.993	1.3	1.990	-20.92	33.39
4	Col.1	43	20	1.989	1.3	2.023	-21.25	40.72
	Col.2	65	5	1.845	1.3	5.140	-31.89	28.67
	Col.3	79	50	1.847	1.3	4.836	-25.70	25.58

Table 4.14 Column rigorous simulation with feed stage correction

Feed	Scheme	Unit	Total tray	Feed tray	Reflux	Duty(MM Btu/hr)	
						Condenser	Reboiler
1	2	Col.1	41	16	2.013	-13.38	25.07
		Col.2	102	54	7.730	-26.43	24.10
	4	Col.1	43	20	1.870	-12.43	24.12
		Col.2	65	14	5.124	-21.83	19.66
		Col.3	78	40	5.955	-21.05	20.98
2	2	Col.1	44	16	1.893	-11.50	22.24
		Col.2	85	42	8.675	-20.33	18.66
	4	Col.1	43	20	1.707	-10.36	21.11
		Col.2	65	13	5.199	-15.40	13.83
		Col.3	80	50	5.872	-14.44	14.39
3	2	Col.1	42	21	2.052	-9.20	17.92
		Col.2	94	56	8.394	-22.02	20.27
	4	Col.1	43	20	2.019	-9.06	17.78
		Col.2	65	14	5.350	-17.48	15.84
		Col.3	78	48	5.212	-14.57	14.50
4	2	Col.1	42	19	1.845	-23.45	45.90
		Col.2	92	52	8.545	-48.96	44.94
5	2	Col.1	42	21	2.007	-22.38	42.82
		Col.2	94	56	8.457	-50.80	46.67
6	2	Col.1	42	20	1.868	-19.73	39.17
		Col.2	94	55	8.441	-41.97	38.55

Table 4.15 Column sizing summary of the selected schemes

Feed	Scheme	Unit	Total trays	Tray Dia. (in)	Tray spacing (in)	Tray type
1	2	Col.1 Col.2	41 102	102 96	24	Valve
	4	Col.1 Col.2 Col.3	43 65 78	96 90 90	24	Valve
2	2	Col.1 Col.2	44 85	96 84	24	Valve
	4	Col.1 Col.2 Col.3	43 65 80	90 72 72	24	Valve
3	2	Col.1 Col.2	42 94	84 90	24	Valve
	4	Col.1 Col.2 Col.3	43 65 78	84 78 72	24	Valve
4	2	Col.1 Col.2	42 92	138 126	24	Valve
5	2	Col.1 Col.2	42 94	132 132	24	Valve
6	2	Col.1 Col.2	42 95	126 120	24	Valve

Table 4.16 Total purchased equipment and capital investment cost of all selected schemes.

Feed	Scheme	Total purchased equipment cost, $	Total capital investment, $
1	2	1,755,749	9,832,192
	4	2,015,708	11,287,966
2	2	1,320,305	7,393,706
	4	1,533,898	8,589,829
3	2	1,390,081	7,784,454
	4	1,521,558	8,520,725
4	2	2,527,008	14,151,242
5	2	2,634,984	14,755,909
6	2	2,355,788	13,192,413

Tables 4.17 Rate of return on investment and net present value of all selected schemes.

Feed	Scheme	Annual cost, $	Annual income, $	Annual profit, $	%ROI	NPV, $
1	2	97,044,685	101,101,109	3,560,206	22.4	11,739,211
2	2	79,242,636	81,870,458	2,254,671	18.7	6,598,322
3	2	68,916,560	72,046,900	2,737,468	21.8	8,880,087
4	2	176,313,724	182,970,823	5,942,904	26.0	20,889,295
5	2	166,034,457	173,148,265	6,369,096	26.8	22,727,066
6	2	148,113,091	153,917,870	5,138,974	24.1	17,172,875