Appendix

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APPENDIX A

Appendix A.1 Stoichiometry and Reaction Kinetics

a) Propane oxidative dehydrogenation, *exothermic*

Mechanism of the oxidative conversion of propane to propylene and ethylene process is *free radical*. The reactions are expected to occur in the homogenous pyrolysis and thermal cracking of propane and thermal cracking of propane along with the heterogeneous catalytic oxidative propane in the presence of limited oxygen. Temperature increases, the conversion of propane increases. The two kinds of occurred reactions are shown separately as eq .A.2.1 to eq.A.2.10.

Heterogeneous reactions

$$C_3H_{8(g)} + \frac{1}{2}O_{2(g)} \rightarrow C_3H_{6(g)} + H_2O_{(g)} \qquad \text{(A.1.1)}$$

$C_3H_8 \rightarrow CO, CO_2, CH_4$ and H_2O

Mechanism:

$$C_3H_8 + [O] \rightarrow C_3H_7^* + [OH] \qquad \text{(A.1.2)}$$

The radical mechanism of reaction can be indicated as below

Initiation:

$O_2 \rightarrow O^{2*} + O^{2*}$

$CH_3CH_2CH_3 \rightarrow CH_3CH_2CH_2^* + H^*$

$CH_3CH_2CH_3 \rightarrow CH_3CH_2^* + CH_3^*$

Propagation:

$O^{2*} + CH_3CH_2CH_3 \rightarrow OH^* + CH_3CH_2CH_2^*$

$OH^* + CH_3CH_2CH_3 \rightarrow H_2O + CH_3CH_2CH_2^*$

$CH_3CH_2CH_2^* \rightarrow H_3CHC = CH_2 + H^*$

$CH_3CH_2CH_2^* + CH_3CH_2CH_3 \rightarrow H_3CHC = CH_2 + CH_3CH_2CH_2^* + H_2$

$CH_3CH_2^* \rightarrow H_2C = CH_2 + H^*$

$CH_3CH_2^* + CH_3CH_2CH_3 \rightarrow H_2C = CH_2 + CH_3CH_2CH_2^* + H_2$

$CH_3^* + CH_3CH_2CH_3 \rightarrow CH_4 + H_2C = CH_2$

$H^* + CH_3CH_2CH_3 \rightarrow CH_3CH_2CH_2^* + H_2$

etc.

Termination:

$H^* + OH^* \rightarrow H_2O$

$OH^* + CH_3CH_2CH_2^* \rightarrow H_2O + H_3CHC = CH_2$

$OH^* + CH_3CH_2^* \rightarrow H_2O + H_2C = CH_2$

$H^* + H^* \rightarrow H_2$

$H^* + CH_3CH_2CH_2^* \rightarrow H_2 + H_3CHC = CH_2$

$H^* + CH_3CH_2^* \rightarrow H_2 + H_2C = CH_2$

$H^* + CH_3^* \rightarrow CH_4$

$CH_3CH_2^* + CH_3^* \rightarrow CH_4 + H_2C = CH_2$

$CH_3CH_2^* + \rightarrow H_2C = CH_2 + CH_3CH_2CH_3$

$CH_3^* + CH_3CH_2CH_2^* \rightarrow CH_4 + H_3CHC = CH_2$

etc.

$$2[OH] + 0.5O_2 \rightarrow H_2O + [O] \qquad \text{(A.1.3)}$$

The catalytic reaction is initiated on the catalyst surface by formation of propyl radicals. Homogenous reactions

$$C_3H_{8(g)} \rightarrow C_2H_{4(g)} + CH_{4(g)} \qquad ;\Delta H = +18.7 kcal/mol \qquad \text{(A.1.4)}$$

$$C_3H_{8(g)} \rightarrow C_3H_{6(g)} + H_{2(g)} \qquad ;\Delta H = +30.9 kcal/mol \qquad \text{(A.1.5)}$$

and oxidation of propane will be linear to the amount of oxygen in reactant.

$$C_3H_{8(g)} + \frac{1}{2}O_{2(g)} \rightarrow C_3H_{6(g)} + H_2O_{(g)} \qquad ; \Delta H = -28.3 kcal/mol \qquad \text{(A.1.6)}$$

$$C_3H_{8(g)} + 3.5O_{2(g)} \rightarrow 3CO_{(g)} + 4H_2O_{(g)} \; ; \Delta H = -286.4 kcal/mol \qquad \text{(A.1.7)}$$

$$C_3H_{8(g)} + 5O_{2(g)} \rightarrow 3CO_{2(g)} + 4H_2O_{(g)} \qquad ;\Delta H = -489 kcal/mol \qquad \text{(A.2.8)}$$

$$C_3H_{8(g)} + \frac{1}{2}O_{2(g)} \rightarrow \frac{1}{2}C_2H_{4(g)} + H_2O_{(g)} \qquad ; \Delta H = -33.5 kcal/mol \qquad \text{(A.2.9)}$$

$$CO_{(g)} + 2H_{2(g)} \rightarrow \frac{1}{2}C_2H_{4(g)} + H_2O_{(g)} \qquad ; \Delta H = -25.1 kcal/mol \qquad \text{(A.2.10)}$$

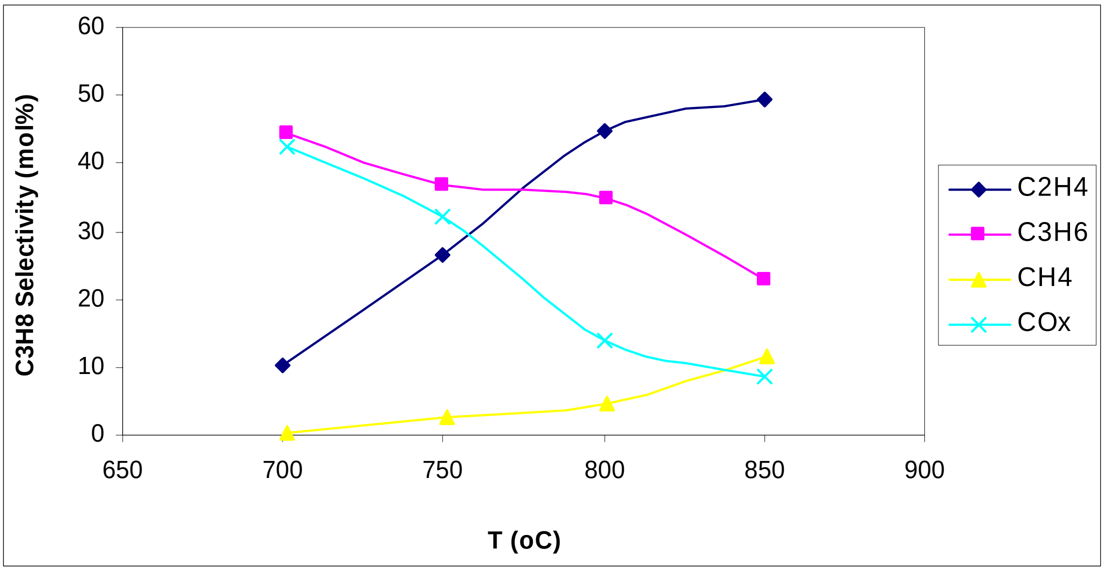

Figure A.1.1. *Effects of temperature on the COx, CH_4, C_2H_4, and C_3H_6 selectivity in oxidative conversion of propane (VH Rane et al, 2003)*

From Figure A.1.1, the favorable temperature for reaction is at 850°C See also Appendix C.1. At this condition, the selectivity of the outlet stream for this experiment is shown as 50%C_2H_4, 25%C_3H_8, 15%CH_4, and 10% CO_X respectively. Then, we can back calculate to obtain the stoichiometry of oxidative dehydrogenation as eq.A.2.11. This reaction will take place in the tube side of this design project.

$$C_3H_{8(g)} + 0.4759O_{2(g)} \rightarrow 0.6946C_2H_{4(g)} + 0.3053CH_{4(g)} + 0.3607C_3H_{6(g)} + 0.2236H_{2(g)} + 0.6945H_2O_{(g)} + 0.1894CO_{(g)} + 0.0339CO_{2(g)} \quad (A.2.11)$$

$; \Delta H = -20 \ \ kcal/mol$

The experimental data for reaction rate is supported by kinetic modeling using Langmuir Huinshelwood (LHHW).

$$r = \frac{k_{LH}K_{O_2}K_{C_3H_8}P_{O_2}P_{C_3H_8}}{(1+K_{O_2}P_{O_2})(1+K_{C_3H_8}P_{C_3H_8})} \quad (A.2.12)$$

b) Propane dehydrogenation, *Endothermic*

Stochiometry

$$C_3H_{8(g)} \rightarrow C_2H_{4(g)} + CH_{4(g)} \qquad ;\Delta H = +18.7kcal/mol \qquad (A.2.13)$$

$$C_3H_{8(g)} \rightarrow C_3H_{6(g)} + H_{2(g)} \qquad ;\Delta H = +30.9kcal/mol \qquad (A.2.14)$$

For the shell side, the reaction takes place only propane dehydrogenation, that we can combine the reactions in eq. A.2.13 and eq.A.2.14. and result in eq. A.2.15.

$$C_3H_{8(g)} \rightarrow 0.002C_2H_{4(g)} + 0.002CH_{4(g)} + 0.998C_3H_{6(g)} + 0.998H_{2(g)} \quad \text{(A.2.15)}$$

$\Delta H = +30.88 \quad kcal/mol$

The experimental data for reaction rate is supported by kinetic modeling using Langmuir Huinshelwood (LHHW).

Dehydrogenation reaction: $C_3H_{8(g)} \Leftrightarrow C_3H_{6(g)} + H_{2(g)}$

$$(-r_1) = \frac{k_1\left(P_{C_3H_8} - \frac{P_{C_3H_6}P_{H_2}}{K_1}\right)}{1 + P_{C_3H_6}K_2} \quad \text{(A.2.16)}$$

$$k_1 = k_{01}\exp\left(\frac{-Ea_1}{R}\left(\frac{1}{T} - \frac{1}{Tm}\right)\right) \quad \text{(A.2.17)}$$

$$K_2 = K_{02}\exp\left(\frac{-\Delta H}{R}\left(\frac{1}{T} - \frac{1}{Tm}\right)\right) \quad \text{(A.2.18)}$$

Cracking reaction: $C_3H_{8(g)} \Leftrightarrow CH_{4(g)} + C_2H_{4(g)}$

$$(-r_2) = k_4P_{C_3H_8} \quad \text{(A.2.19)}$$

$$k_4 = k_{04}\exp\left(\frac{-Ea_4}{R}\left(\frac{1}{T} - \frac{1}{Tm}\right)\right) \quad \text{(A.2.20)}$$

Appendix A.2 Thermodynamic data

Table A.2a. *List of thermodynamics heat data-- Gibbs Energy*

Component name			Gibbs energy of formation of GAS			
Design Systematic	Formula	CAS-Nr.	$G_f = A + BT + CT^2$ (kJ/mol)			
			A	B	C	dGf@298K
Propane	C3H8	74-98-6	-105.603	0.26475	3.3E-05	-23.47
Propylene	C3H6	115-07-1	19.412	0.13685	2.6E-05	62.72
Ethylene	C2H4	74-85-1	51.752	0.049338	1.7E-05	68.12
Methane	CH4	74-82-8	-75.262	0.075925	1.9E-05	-50.84
Hydrogen	H2	1333-74-0	n/a	n/a	n/a	0
Oxygen	O2	7784-44-7	n/a	n/a	n/a	0
Water	H2O	7732-18-5	n/a	n/a	n/a	-228.6
Carbon dioxide	CO2	124-38-9	-393.36	-0.003821	1.3E-06	-394.38
Carbon monoxide	CO	630-08-0	-109.885	-0.092218	1.5E-06	-137.28
ΔG of solid			Gibbs energy of formation of SOLID (kJ/mol)			
			Gf@298K	Af@298K	Sf@298K	
Methyl Diethanolamine	$CH_3N(CH_2CH_2OH)_2$	105-59-9	n/a	n/a	n/a	

Table A.2b. *List of thermodynamics heat data—Saturated liquid density*

Component			Saturated liquid density			
Design Systematic	Formula	CAS-Nr.	$Density = A \cdot B^{-(1-T/T_c)^n}$			
			[g/ml]			
			A	B	n	Tc
Propane	C_3H_8	74-98-6	0.22151	0.27744	0.28700	369.82
Propylene	C_3H_6	115-07-1	0.23314	0.27517	0.30246	364.76
Ethylene	C_2H_4	74-85-1	0.21428	0.28061	0.28571	282.36
Methane	CH_4	74-82-8	0.15998	0.28810	0.27700	190.58
Hydrogen	H_2	1333-74-0	0.03125	0.3473	0.27560	33.18
Oxygen	O_2	7784-44-7	0.43533	0.28772	0.29240	154.58
Water	H_2O	7732-18-5	0.34710	0.27400	0.28571	647.13
Carbon dioxide	CO_2	124-38-9	0.46382	0.26160	0.29030	304.19
Carbon monoxide	CO	630-08-0	0.29818	0.27655	0.29053	132.92
Methyl Diethanolamine	$CH_3N(CH_2CH_2OH)_2$	105-59-9	n/a	n/a	n/a	n/a

A,B,n=regression coefficient for chemical components

T=Temperature, K

T_c=critical temperature, K

Appendix A.3 Input output streams and properties

Feedstocks

a. Propane

Table 3.3.3.1a. *Propane properties*

Steam Name:		Propane			
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
Propane	%wt	>90	90.0	(1)	(1) Http://dixiemail.dixiepipeline.com (2) http://ceh.sric.sri.com
Propylene	%wt	<5	5.0	(1)	
Heavy Ends	%wt	<5	5.0	(1)	
Sulfur	Ppm wt	<123	-	(1)	
Hydrogen sulfide	Ppm wt	<2.0	-	(1)	
Carbonyl sulfide	Ppm wt	<20	-	(1)	
Total			100.0		
Process Conditions and Price					
Temp.	*K*		313		
Press.	Bara		15		
Phase	V/L/S		L		
Price	US$/ton		160.25	(2)	

b. Oxygen

Table3.3.3.1b. *Oxygen properties*

Steam Name:		Oxygen			
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
Oxygen	%wt	>95	95.0	(3)	(3) http://www.indiamart.com/gastek/ (4) http://www-pao.ksc.nasa.gov/ kscpao/nasafact/pdf/SSP.pdf (5) Assume other components except propane do not harmful and influent for the process.
Nitrogen	%wt	<5	5.0	(3)	
Total			100.0		
Process Conditions and Price					
Temp.	*K*		313		
Press.	Bara		15		
Phase	V/L/S		L		
Price	US$/ton		143.4	(4)	

Products

a. Ethylene

Table 3.3.3.2a. *Ethylene properties*

Steam Name:		Ethylene			
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
Ethylene	%wt	>95	100.0		(7) http://ceh.sric.sri.com

Heavy Ends	%wt	<5	0.0		
Total			100.0		
Process Conditions and Price					
Temp.	*K*		303		
Press.	Bara		15		
Phase	V/L/S		L		
Price	US$/ton		450	(7)	

b. Propylene

Table 3.3.3.2b.*Propylene properties*

Steam Name:		Propane			
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
Propane	%wt	>95	100.0		(9) http://ceh.sric.sri.com
Heavy Ends	%wt	<5	0.0		
Total			100.0		
Process Conditions and Price					
Temp.	*K*		303		
Press.	Bara		15		
Phase	V/L/S		L		
Price	US$/ton		326	(9)	

c. Light ends gas: Methane/Carbon monoxide/Hydrogen

Table 3.3.3.2c. *$CH_4/CO/H_2$ properties*

Steam Name:		Propane			
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
Methane	%wt		43.0	(11)	(11) Values from mass balance (12) As principle suggested, 25% of propane's price.
Carbon monoxide	%wt		45.9	(11)	
Hydrogen	%wt		11.1	(11)	
Nitrogen	%wt				
Total			100.0		
Process Conditions and Price					
Temp.	*K*		313		
Press.	Bara		15		
Phase	V/L/S		L		
Price	US$/ton		64	(12)	

d. Carbon dioxide

Table 3.3.3.2d. *CO_2 properties*

Steam Name:		Propane			
Comp.	Units	Specification		Notes	Additional Information (also ref.

		Available	Design		note numbers)
Carbon dioxide	%wt	>95	100.0		
Heavy Ends	%wt	<5	0.0		
Total			100.0		
Process Conditions and Price					
Temp.	*K*		298		
Press.	Bara		4.5		
Phase	V/L/S		L		
Price	US$/ton		6.5		

APPENDIX B

Appendix B Binary system comparison

Keep the temperature constant, to find different pressure according to difference liquid mole fraction. Compare the result from experimental data[1] and simulation by Aspen. The results are listed as bellow.

Table B.1 *The comparison between experiment data and results from ASPEN*

Number	T	Experimental data		ASPEN	
		x1	P	x1	p
1	260.93	0.2053	3.447	0.2	3.356
2		0.5557	3.731	0.55	3.634
3	269.54	0.472	4.826	0.475	4.725
4		0.857	5.185	0.85	5.106
5	277.59	0.6554	6.322	0.65	6.253
6		0.8454	6.536	0.85	6.505
7	301.32	0.799	12.169	0.8	12.171
8		0.9	12.348	0.9	12.395
9	310.93	0.1794	13.604	0.175	13.622
10		0.7904	15.241	0.8	15.309
11	330.32	0.756	23.194	0.75	23.152
12		0.874	23.594	0.875	23.654
13	344.26	0.1769	27.49	0.175	27.614
14		0.7901	30.569	0.8	30.736

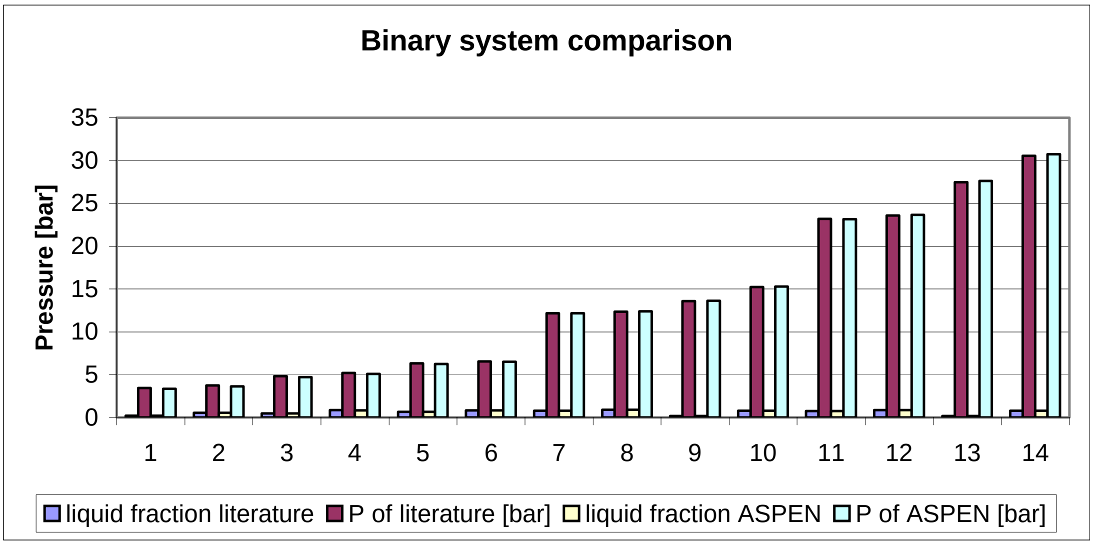

Figure B.1 *The results from ASPEN compared with experimental data*

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[1] Reference: H.Knapp, *Vapor-liquid Equilibria for mixtures of low boiling substances*, Chemistry data series.

Figure B.1 shows that at the same liquid mole fraction, the vapor pressure of the binary system from Aspen is almost the same with experimental data at the same temperature. These verify the correct of Aspen simulation prediction.

APPENDIX C

Appendix C.1 Selection of operating temperature

In the design stage, operating temperature of reaction is imperative to be determined, because it influences to the product selectivity, and then the margins of the process. In view of exergy loss reduction, operating temperature also impact the loss work or loss energy from the temperature difference, thereby the operating temperature should be estimated to optimum both economic point of view and exergy loss reduction.

In this project, the reactions of both, endothermic and exothermic, take place in shell and tube reactor respectively. The temperature of tube side should be higher than that of the shell side. From the shell side, the reaction takes place at the temperature 540^oC in order to achieve the high selectivity and high conversion. So, we will base on this data for the shell side and figure the suitable operating temperature in tube side following data.

From the data in Appendix A.1, Figure A.1.1, show the relationship between product selectivity and temperature. When temperature is changed, the production selectivity will also change. Consequently, the results of reactants and products in chemical reaction estimated will change also. Therefore, to estimate the best operating temperature, the calculations of 3 temperature levels, which are 750^oC, 800^oC, and 850^oC indicated as the equation below.

$$\begin{aligned} C_3H_{8(g)} + 1.255O_{2(g)} \rightarrow\ & 0.195C_2H_{4(g)} + 0.077CH_{4(g)} + 0.566C_3H_{6(g)} + 0.204H_2 + \\ & + 1.556H_2O_{(g)} + 0.719CO_{(g)} + 0.118CO_{2(g)} \end{aligned}$$

; at T = 750 ^oC , $\Delta H = -100\ kcal/mol$ (C.1)

$$\begin{aligned} C_3H_{8(g)} + 1.067O_{2(g)} \rightarrow\ & 0.491C_2H_{4(g)} + 0.074CH_{4(g)} + 0.457C_3H_{6(g)} + 0.012H_2 + \\ & + 1.490H_2O_{(g)} + 0.507CO_{(g)} + 0.068CO_{2(g)} \end{aligned}$$

; at T = 800 ^oC , $\Delta H = -80\ kcal/mol$ (C.2)

$$\begin{aligned} C_3H_{8(g)} + 0.476O_{2(g)} \rightarrow\ & 0.695C_2H_{4(g)} + 0.305CH_{4(g)} + 0.361C_3H_{6(g)} + 0.224H_2 + \\ & + 0.695H_2O_{(g)} + 0.189CO_{(g)} + 0.034CO_{2(g)} \end{aligned}$$

; at T = 850 ^oC , $\Delta H = -20\ kcal/mol$ (C.3)

In fact, the process can operate with the recycle stream of unconverted propane, C_3H_8. So, to compare the whole results of these three operating temperature. The iterative calculation is necessary to figure it out. The results to achieve the design product and compatible with the data from literature show in the Table C.1.1.

Table C.1.1 **The results of iterative calculation for different operating temperature in tube side**

Operating Temperature (°C)	Input (kton/a)		Output (kton/a)						
	C_3H_8	O_2	C_2H_4	C_3H_6	CH_4	H_2	H_2O	CO	CO_2
750	240.87	63.78	81.51	118.5	20.48	3.35	52.36	22.1	6.22
800	236.98	102.27	41.26	158.74	3.61	3.76	80.34	42.31	8.97
850	222.18	40.6	5.72	194.29	1.36	7.46	28.3	20.24	5.22

The price of the substances related shows in the Table C.1.2.

***Table C.1.2** The price of materials in the process*

Substances	C_3H_8	O_2	C_2H_4	C_3H_6	CH_4	H_2	H_2O	CO	CO_2
Price (US$/ton)	160.25	143.4	450	325	2.94	-	-	-	-

Margins of the process expresses as

Margin=Total Value (Products, Wastes *OUT*) - Total Value (Feedstock's, Process Chemicals, *IN*)

And work loss that related to exergy loss mentioned in Chapter 8, eq.8.3.1.15 can be figured out. Consequently, the margins and work loss of different operating temperature for tube side can be estimated as the Table C.1.3

***Table C.1.3.** The margins and Wlost of difference operating temperature.*

Temp.	Margin(US$/a)	Wlost (kcal)
850	31491	10.610
800	21007	47.316
750	26298	37.964

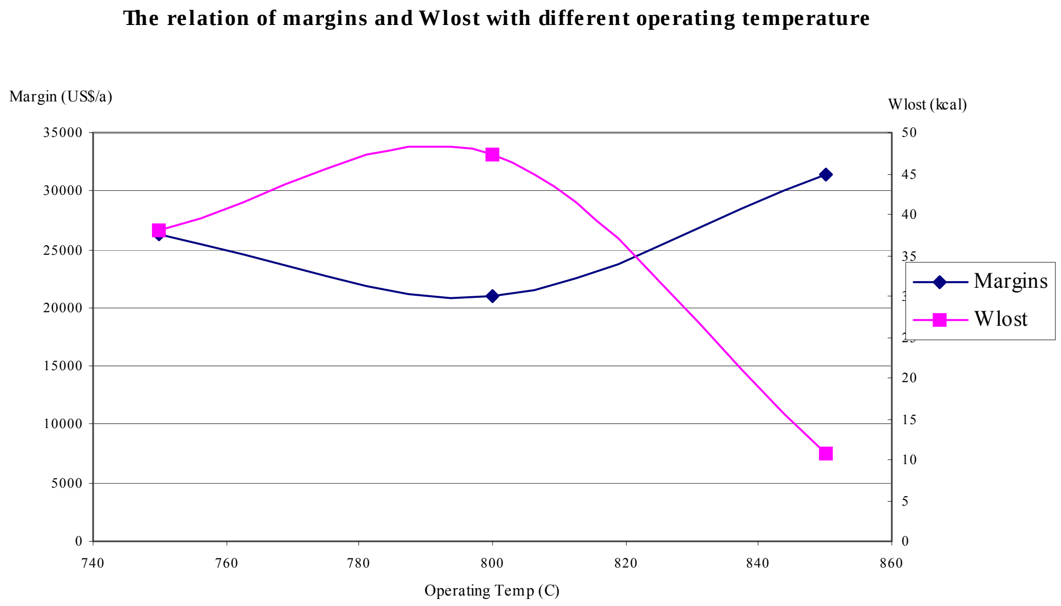

***Figure C.1** The margins and W_{lost} of difference operating temperature.*

Appendix C.2 Criteria and Selections for CO2 Removal

Carbon dioxide is green house gas, and will cause global warming. It is not allowed that High concentration of carbon dioxide present in product. So CO_2 produced in process should be removed.

Solvent chosen

In practice, physical and chemical absorption are both used to remove CO_2. According to higher separation efficiency of chemical absorption, it is applied in design. CO_2 is acid gas and should use amine to absorb it. Normally, methyl ethanolamine (MEA), diethanolamine (DEA) and methyl diethanolamine (MDEA) are chosen. Compare these three amines, MDEA has higher energy efficiency, greater acid gas removal capacity, higher resistance to degradation, smaller equipment size for the new plants and above all much less corrosivity as compared to primary and secondary amines. As a result of following advantages, MDEA is the best choice to remove CO_2 in this process.

The advantage of MDEA in acid gas treating

- MDEA, a tertiary amine, is less basic and can be used in significantly higher concentrations. For identical flows, MDEA has a greater capacity to react with acid gas. The comparison is shown in Table C.2.1.
- MDEA has increased capacity for existing units, decreased capital cost for new units, lower energy costs and higher selectivity than primary or secondary amines. Table C.2.2 summarized actual MDEA operation data.
- In primary treating MDEA rich loading have averaged 0.5 moles of acid gas per mole of MDEA. Reboiler Steam requirements have ranged from 0.67 to 0.85 lbs per gpm of solvent in circulation.
- CO2 selectivity of 50% under 200 psig and higher at lower pressures have also been achieved.
- Solvent concentration between 35 to 50 percent has, been proved successful. Typical concentration between 35 to 50% and pickup rates as high as 0.45 or 0.50 moles acid gas per mole of MDEA significantly increase capacity of existing units and allow equipment to be considerably smaller for new units. Higher concentration and higher pickup rates correspond to lower solvent circulation rates for equivalent capacities, too.
- MDEA also delivers energy savings from reduced reboiler duties (reflux ratio of 0.5 to 1.0).
- Among MEA, DEA, and MDEA, MEA has worst reputation for corrosion related problems. It is well documented in literature, that MEA and DEA form degradation products when reacted with CO2 whereas MDEA does not. Operating MEA, DEA and MDEA plants have demonstrated that corrosion can be minimized under proper operating conditions. However based on plant

experience and laboratory data, relative corrosivity of amines are ranked as follows: MEA >> DEA >> MDEA. Table C.2.4 generates corrosion data for various amine-based solutions.

- MDEA is tertiary amine and therefore carbamate formation with CO2 does not take place in MDEA based system. MEA and DEA form carbamates with CO2. Therefore operation with MDEA is far more stable with no spurious shutdowns over longer periods.

Table C.2.1. *Comparison of amines (MEA, DEA and MDEA)*

SOLVENT	**MEA**	**DEA**	**MDEA**
CONCENTRATION %	15	30	35-50
SOLVENT CIRCULATION GPM	100	100	100
ACID GAS REMOVAL CAPACITY MOL/HR	49.8	58.6	87.5
CAPACITY INCREASE % (MEA BASE =100)	100	118	175

Table C.2.2. *Performance of MDEA*

SOLVENT CONCENTRATION %	35-50
SOLVENT CIRCULATION, GPM	10-1600
RICH MDEA LOADING MOL/MOL	0.50
LEAN MDEA LOADING MOL/MOL	0.01
REBOILER STEAM, #/GPM	0.67-0.85
LEAN MDEA TEMPERATURE ^{O}F	130-160
CO2 SLIP, % CO2 REJECTED	50

Table C.2.3. *Selectivity and capacity of amines*

AMINE	**SELECTIVITY***	**CAPACITY**	
		Mol H_2S/Mol amine	**Mol CO_2/Mol Amine**
MDEA	3.85	0.10	0.12
DEA	2.27	0.09	0.32
MEA	0.89	0.07	0.50

*Selectivity is defined as ratio of (mole percent of H_2S removed to mole percent of H_2S in feed gas) to (mole percent of CO_2 removed to mole percent of CO_2 in feed gas)

Table C.2.4. *Corrosion comparison*

Solvent	**Corrosion Rate MPY**
30% Wt MEA	32

50% Wt DEA	25
15% Wt MEA	13
20% Wt DEA	8
50% Wt MDEA	3

Packed column chosen

Due to one liquid input (MDEA solution) and one vapor input (raw product stream) a vertical absorption column is preferred which can get high exchange efficiency and need less volume. Followed the criteria below, packed column absorber is chosen and material of equipment is carbon steel.

From *Coulson and Richardson, Volume 6*, it mentioned main advantages and disadvantages of plate and packed column which are listed below:

1) Plate columns can be designed to handle a wider range of liquid and gas flow-rates than packed columns.
2) Packed columns are not suitable for very low liquid rates.
3) The efficiency of a plate can be predicted with more certainty than the equivalent term for packing (HETP or HTU).
4) Plate columns can be designed with more assurance than packed columns. There is always some doubt that good liquid distribution can be maintained throughout a packed column under all operating conditions, particularly in large columns.
5) It is easier to make provision for cooling in a plate column; coils can be installed on the plates.
6) It is easier to make provision for the withdrawal of side-streams from plate columns.
7) If the liquid causes fouling, or contains solids, it is easier to make provision for cleaning in a plate column; manways can be installed on the installed on the plates. With small diameter columns it maybe cheaper to use packing and replace the packing when it becomes fouled.
8) For corrosive liquids a packed column will usually be cheaper than the equivalent plate column.
9) The liquid hold-up is appreciably lower in a packed column than a plate column. This can be important when the inventory of toxic or flammable liquids needs to be kept as small as possible for safety reasons.
10) Packed columns are more suitable for handling foaming systems.
11) The pressure drop pre equilibrium stage (HETP) can be lower for packing than plates; and packing should be considered for vacuum columns.
12) Packing should always be considered for small diameter columns, say less than 0.6m, where plates would be difficult to install, and expensive.

Here, MDEA is corrosive for equipment and much pressure drop is not good for the separations followed. As a result, packed column is chosen. Figure C.2.1 shows the scheme of packed absorption column. Actually structured packing has low HETP (typically less than 0.5) and low-pressure drop (around 100Pa/m); however, the cost of structured packing per cubic meter is significantly higher than that of random packing. So, random packed column is preferred. Assume a 99 per cent recovery of the carbon dioxide is reached. In order to improve the liquid distribution characteristics the type of packing is Pall ring, which increases the free area. Ring packings are available in a variety of materials: ceramics, metals, plastics and carbon. Metal and plastics (polypropylene) rings are more efficient than ceramic rings, as it is possible to male the walls thinner. Due to MDEA a kind of amine that is corrosive to plastics, metal (carbon steel) is used in design.

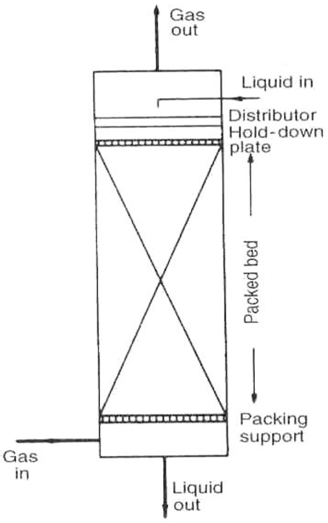

***Figure C.2.1.** The scheme of packed absorption column*

Note: reference from Coulson, Volume 6, Figure11.36

In general, the largest size of packing that is suitable for the size of column should be used, up to 50mm. The reason is that small sizes are appreciably more expensive than the larger sized. However, above 50mm the lower cost pre cubic meter does not normally compensate for the lower mass transfer efficiency. If packing size is too large in a small column it can cause poor liquid distribution. Recommended size ranges are: [Couson, volume 6]

***Table C.2.5.** Recommended size ranges related to packing column diameter*

Column diameter	Use packing size
<0.3 m	<25 mm
0.3 to 0.9 m	25 to 38 mm
>0.9 m	50 to 75 mm

According to the stream flow rate into the unit 700m^3/h approximate, the large range of column diameter and related range of use packing size are chosen.
And Table C.2.6 [Distillation principles and practices] lists typical values of specific area and porosity for several random column packings. Herewith related data is shown in Table C.2.6.

***Table C.2.6.** Design data*

	Size	Metal

Particle	Diameter d_n, mm	Specific Area a m^2/m^3	Porosity ε
Pall rings	10	515	0.92
	20	360	0.93
	25	215	0.94
	35	145	0.94
	50	105	0.95
	80	78	0.96

In order to use the largest packing size suitable for the size of column and get higher specific area, 50mm Pall rings are used. And Onda's method is based on a large amount of data on gas absorption and distillation; with a variety of packings, which included Pall rings. Accordingly, it is used for the calculation of column design.

Appendix C.3 Heat Integration

Heat integration is required to develop an energy-efficient process. In this stage of the synthesis of a flowsheet, the source and the target temperature, Ts and Tt, and power demand of all streams are known. In every plant, it is needed to design an effective Heat Exchanger Network (HEN) by heat integration. It is desired to calculate the Minimum Energy Requirement (MER) before synthesizing the HEN. In MER calculation (usually called as MER targeting), its is obligatory to compute the minimum usage of heating and cooling utilities by exchanging heat between the hot and the cold streams in a process (Seider et al, 2003).

The method which is used in MER targeting is the Temperature Interval (TI) method. The temperature-interval method was applied according to Linnhoff (1987).
After MER targeting, HEN is designed with a unit-by-unit method beginning at the closest-approach temperature difference (the pinch analysis) (Linnhoff et al, 1987). For the minimum utility requirements aver all possible HENs, the minimum approach temperature in heat exchangers, $\Delta T_{\min}$, is 10 oC.

Step by steps of MER targeting for heat integration, which has been followed according the Temperature Interval (TI) method (Linnhoff, 1987). And Seader (2003) are presented shortly. First, "Hot and Cold Composite" Curve is constructed in one graph. Then, "Hot and Composite" Curve plus 'the pinch point' is constructed. After that, calculation for Tpinch by making 'Cascade table' is described as follows.
The first step in pinch analysis is to determine the pinch point by using Temperature Interval (TI) method. An interval is an imaginary boundary to make our works easier. We defined a 10oC for the minimum temperature difference, where interval is $\frac{1}{2}\Delta T_{\min}(5^{o}C)$ below the hot stream and $\frac{1}{2}\Delta T_{\min}(5^{o}C)$ above cold stream.

The procedure of TI method is usually applied for the streams with constant the heat-capacity flowrate (FCp). Meanwhile, in the alkenes plant; there are some phase changes involved. Phase changes (latent heats) are counted into TI method formalism simply by assuming 1oC temperature change at the temperature of the phase change; i.e., if the heat corresponding to the phase change is $F\Delta H_v$, it can be wrote

$$FC_p\left(1\right) = F\Delta H_v$$

where F and Cp are the fictitious values.

For the case of mixtures, where a plot of enthalpy versus temperature is curved, we merely linearize the graph and select fictitious FCp values that have the same hat duty.

Thus, phase changes simply increase the number of temperature interval considered (Douglas, 1988).

In the Alkenes plant, heat integration is applied for pre-condition and reaction section. The four hot steams and three cold streams in precondition and reaction systems can be found in Table C.2.1.

The heating and cooling requirement in separation section is not included in heat integration task. The temperature of condenser in separation section is very low due to cryogenic distillation. Therefore, special design methods to reduce energy loss and utilities cost is applied in the separation section as described below:

1. The light gas column T301.

   The Coolant used in condenser E301B is hydrogen expanded, which obtain from the F-T process, from 20 bar to 5 bar. This is because the overhead gas temperature is very cold, around –130 C. and expanded H_2 can be used as refrigerant. The temperature of H_2 expanded is around –250 C.
2. The ethylene column; T302.

   Overhead stream as <306> is condensed by E302B, and reflux stream as stream<307> can be used for purity adjust. The condenser of this column uses part of ethylene product as refrigerant itself, by passing through expansion valve. This is because at the top column, the temperature is around –40 C. Ethylene after expansion valve will be reach to –70 C that can be used to condense the overhead gas.
3. The last distillation column, T303.

   This overhead gas is condensed by E303B, which use part of propylene product as heat pump process referring to Appendix C.5. After heat pump, propylene will be used as the heating media for reboiler, E303A.

Table C.3.1 *The stream conditions and properties for precondition amd reaction section*

No	Stream	Hot/Cold	F (kg/hr)	F (kg/s)	Cp (J/kg-K)	Temp(in) (°C)	Temp(out)(°C)	FCp (kJ/°C/s)
1	102	Hot	39022.39	10.8	2637.2	540	25	28.59
3	205	Hot	32755.24	9.1	2761.2	850	25	25.12
4	104	Hot	39022.39	10.8	2736.9	227	55	29.67
5	208	Hot	27875.32	7.7	2307.3	308	30	17.87
13	001	Cold	29644	8.2	3509.6	25	540	28.90
6	312	Cold	35392.96	9.8	2805.9	43	540	27.59
2	203	Cold	32756.43	9.1	3254.7	540	850	29.61

After making the cascade of temperature intervals, which is shown in Figure C.3.1 and Enthalpy Differences for Temperature Intervals in Table C.3.2, the pinch point is acquired quickly.

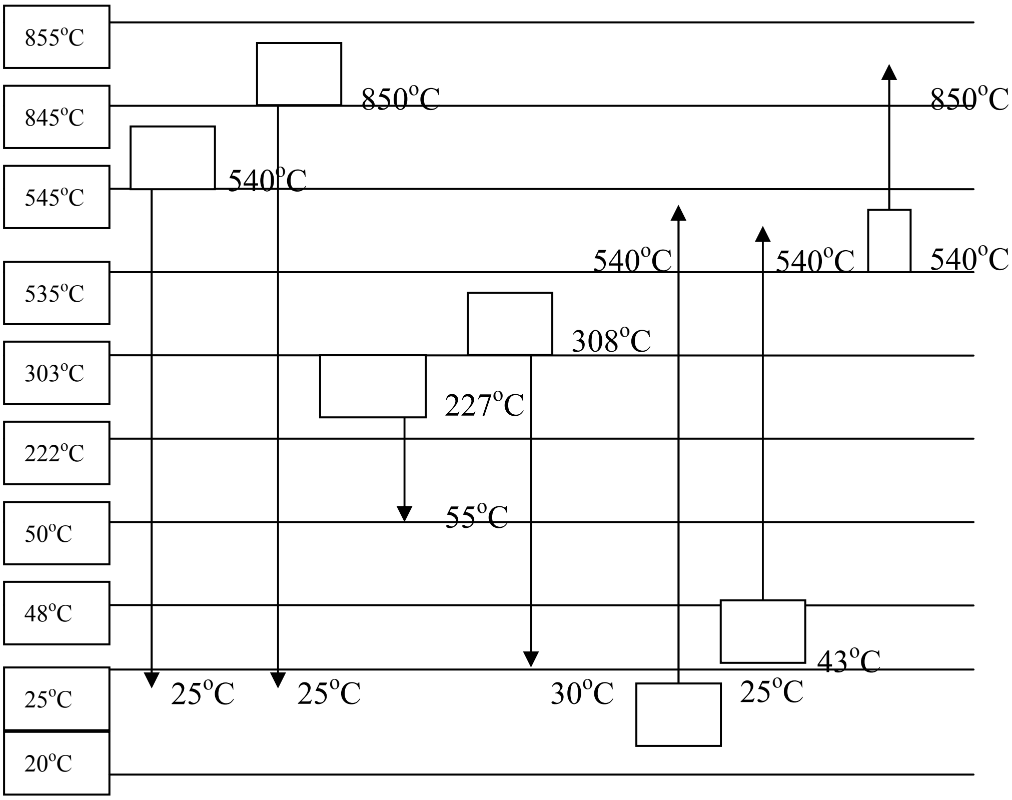

Figure C.3.1 *Temperature-Interval (TI) Method*

Table C.3.2 *Enthalpy Differences for Temperature Intervals*

	Interval i	$T_i - T_{i-1}$	$\sum FC_{P,hot} - \sum FC_{P,cold}$	$\Delta H (kJ/s)$	+ or -	Accumulate	From hot utilities
855°C							2602
845°C	1	10	-30	-296	Deficit	-296	2305
545°C	2	300	-4	-1347	Deficit	-1644	958
535°C	3	10	-31	-314	Deficit	-1957	644
303°C	4	232	-3	-644	Deficit	-2602	0
222°C	5	81	15	1222	Surplus	-1379	1222
50°C	6	172	45	7698	Surplus	6319	8920
48°C	7	2	15	30	Surplus	6349	8950
30°C	8	18	43	768	Surplus	7117	9718
25°C	9	5	54	269	Deficit	7385	9987
20°C	10	5	0	0	Surplus	7385	9987

The Cascade table of temperature intervals including energy balance is presented in Figure C.3.2.

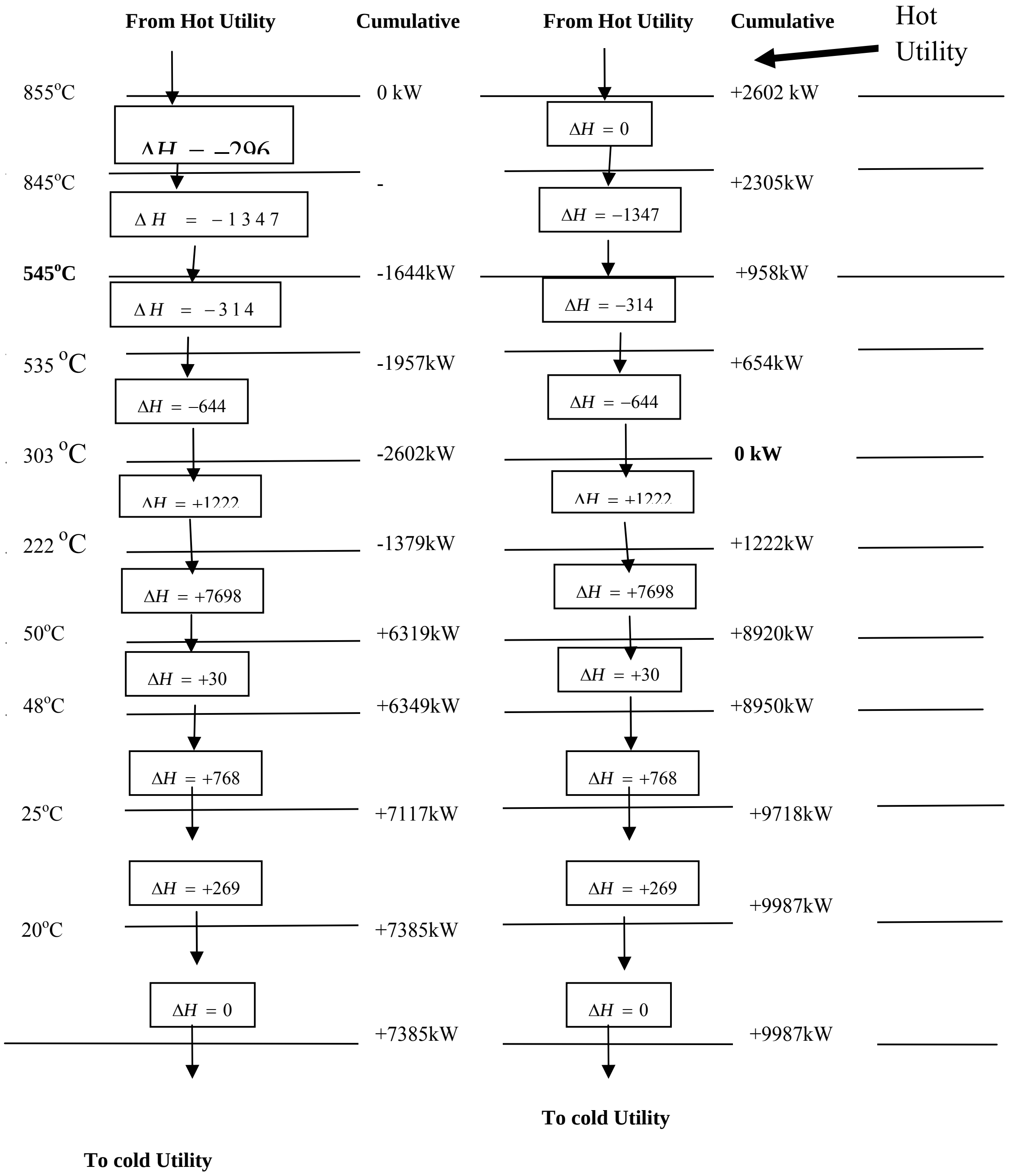

***Figure C.3.2** Cascade of temperature intervals, energy balances*

From the Cascade table above, we found that T_{pinch} is at $T_{interval} = 303^{o}C$. In the other word, $T_{pinch} = 308^{o}C$ for Hot Streams ($5^{o}C$ above $T_{interval}$)

$T_{pinch} = 298^{o}C$ for Cold Streams ($5^{o}C$ below $T_{interval}$).

- The Utility Consumption for Heating and Cooling

As guidance, we should choose the 'minimal' energy as efficient as possible. For Example, There are two hot utilities, which perform the same quantities of energy. In this case we have to choose the utilities at lower pressure first in order to prevent inefficient usage of energy.

Heat Exchanger Network (Matching Diagram, HEN)

These are the basic rules of 'pinch design method' (Linhoff et al, 1987):

-. Problem is divided at the pinch, and designing each part separately. It is not allowed to transfer heat across the pinch.

-. The design starts at the pinch and moves away

-. In case the streams are immediately adjacent to the pinch, the following constraints should be obeyed:

$$\boxed{\begin{array}{l} mC_{P_{HOT}} \le mC_{P_{COLD}} \ \ (above\ pinch) \\ mC_{P_{HOT}} \ge mC_{P_{COLD}} \ \ (below\ pinch) \end{array}}$$

-. Exchanger loads should be maximized

-. Suppliying external heating can be one above the pinch and external cooling can be supplied only below the pinch. To maintain minimum utilities, no energy is permitted to flow across the pinch.

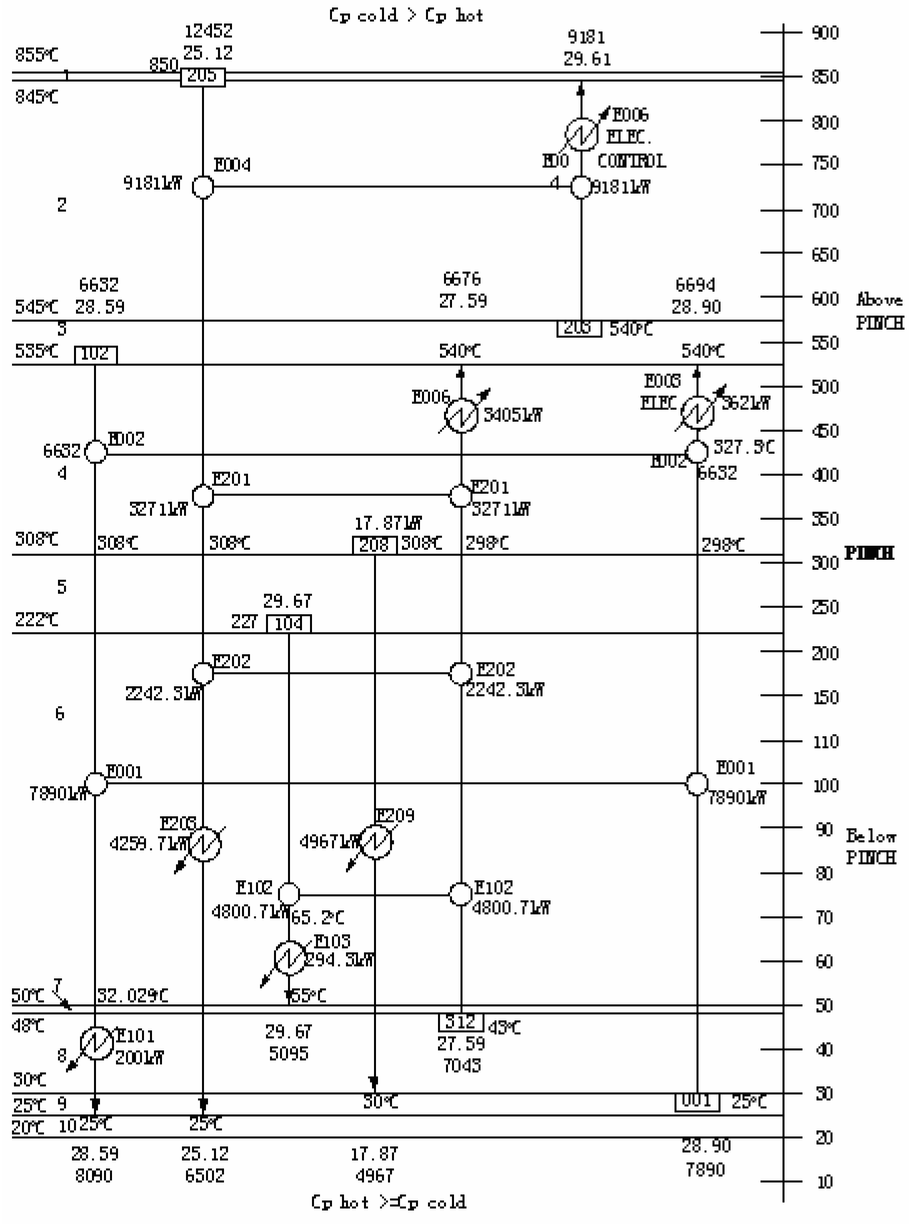

Figure C.3.3 Pinch Design (Heat Exchanger Network)

The list of heat exchangers after heat integration can be found in Chapter 11, economic part and in equipment specification (Appendix E.8).

- ***Summary of heat integraton***

Heat integration ends up with the minimum usage of hot and cold utilities when exchanging heat between the hot and cold utilities in the process. As mentioned before, the direct result of heat integration is the design of a heat exchanger network (HEN). From the calculation in heat integration we obtained a new flowsheet with the HEN. The new HEN is expected to reduce the usage of hot and cold utilities, which reduce the annualized cost for utilities. In this CPD project, after the heat integration is applied, the usage of hot and cold utilities decreases significantly. However, the heat integration has an impact in the investment cost. Most of liquids in the Alkenes plant are in gaseous phase, which have very low heat transfer coefficient (h). Typical overall heat transfer coefficient (U) for gas-gas heat exchanger is 10-50 ($W/m^2.^oC$). Consequently, the heat exchange area becomes very large and will results the very expensive heat exchanger investment costs.

After conducting Heat integration procedure and making the new flowsheet, the CPD project goes to economic assessment. The economic evaluation or profitability analysis shows that the investment for the process with heat integration has significant increasing due to the cost of heat exchanger with large heat exchange area. Some large heat exchangers require very expensive purchased cost and result a high investment cost.

In this CPD report, two flowsheets are presented. One flowsheet is the process 'with' heat integration and the other flowsheet is the process 'without' heat integration. Both flowsheets; the process 'with' and 'without' heat integration are presented together with mass and heat balance and the economic evaluation by profitably analysis. Finally, after the economic evaluation, the profitability analysis shows that the alkenes process 'without' heat integration is more profitable and more robust than the alkenes process 'with' heat integration due to the high investment in heat exchanger equipments.

Appendix C.4 Options and selections of Heat exchanger

Selection is the process in which the designer selects a particular type of heat exchanger for a given application from a variety of heat exchangers. There are a number of alternatives for selecting heat transfer equipment, but only one among them is the best for the given set of conditions.

Selection Criteria

Selection criteria are many, but primary criteria are type of fluids to be handled, operating pressures and temperatures, heat duty, and cost. Fluids involved in heat transfer can be characterized by temperature, pressure, phase, physical properties, toxicity, corrosivity, and fouling tendency. Operating conditions for heat exchangers vary over a very wide range, and a broad spectrum of demands is imposed for their design and performance. All of these must be considered when assessing the type of unit to be used. When selecting a heat exchanger for a given duty, the following points must be considered:

- Materials of construction
- Operating pressure and temperature, temperature program, and temperature driving force
- Flow rates
- Flow arrangements
- Performance parameters i.e. thermal effectiveness and pressure drops
- Fouling tendency
- Types and phases of fluids
- Maintenance, inspection, cleaning, extension, and repair possibilities
- Overall economy
- Fabrication techniques
- Intended applications

Materials of Construction

For reliable and continuous use, the construction materials for pressure vessels and heat exchangers should have a well-defined corrosion rate in the service environments. Furthermore, the material should exhibit strength to withstand the operating temperature and pressure. Shell and tube heat exchangers can be manufactured in virtually any materials that may be required for corrosion resistance, e.g., from nonmetals like glass, Teflon, and graphite to exotic metals like titanium, zirconium, tantalum, etc. compact heat exchangers with extended surfaces are mostly manufactured from any metal that has drawability, formability, and malleability. Heat exchanger types like plate heat exchangers normally require a material that can be pressed or welded.

Operating Pressure and Temperature

Pressure

The design pressure is important to determine the thickness of the pressure retaining components. The higher the pressure, the greater will be the required thickness of the pressure-retaining membranes and the more advantage there is to placing the high-pressure fluid on the tubeside. The pressure level of the fluids has a significant effect on the type of unit selected.

- At low pressures, the vapor-phase volumetric flow rate is high and low allowable pressure drops may require a design that maximizes the area available for flow, such as crossflow or split flow with multiple nozzles.
- At high pressures, the vapor-phase volumetric flow rates are lower and allowable pressure drops are greater. These lead to more compact units.
- In general, higher heat transfer rates are obtained by placing the low-pressure gas on the outside of tubular surfaces.
- Operating pressures of the gasketed plate heat exchangers and spiral plate heat exchangers are limited because of the difficulty in pressing the required plate thickness, and by the gasketed materials in the case of PHEs. The floating nature of floating-head shell and tube heat exchangers and lamella heat exchangers limits the operating pressure.

Temperature

Design Temperature

This parameter is important as it indicates whether the material at the desired temperature can withstand the operating pressure and various loads imposed on the components. For low-temperature and cryogenic applications toughness is prime requirement, and for high-temperature applications the material has to exhibit creep resistance.

Temperature Program

Temperature program in both a single pass and multipass shell and tube heat exchanger decides (1) the mean metal temperatures of various components like shell, tube bundle, and tubesheet, and (2) the possibility of temperature cross. The mean metal temperatures affect the integrity and capability of heat exchangers and thermal stresses induced in various components.

Temperature Driving Force

The effective temperature driving force is a measure of the actual potential for heat transfer that exists at the design conditions. With a counterflow arrangement, the effective temperature difference is defined by the log mean temperature difference (LMTD). For flow arrangements other than counterflow arrangement, the LMTD must be corrected by a correction factor, F. The F factor can be determined analytically for each flow arrangement but is usually presented graphically in terms of the thermal effectiveness P and the heat capacity ratio R for each flow arrangement.

Influence of Operating Pressure and Temperature on Selection of Some Types of Heat Exchangers.

Shell and Tube Heat Exchanger

Shell and tube heat exchanger units can be designed for almost any combination of pressure and temperature. In extreme cases, high pressure may impose limitations by fabrication problems associated with material thickness, and by the weight of the finished unit. Differential thermal expansion under steady conditions can induce severe thermal stresses either in the tube bundle or in the shell. Damage due to flow-induced vibration on the shellside is well known. In heat exchanger applications where high heat transfer effectiveness is required, the standard shell and tube design may require a very large amount of heat transfer surface. Depeding o the fluids and operating conditions, other types of heat exchanger design should be investigated.

Compact Heat Exchanger

Compact heat exchangers are constructed from thinner materials, which are manufactured by mechanical bonding, soldering, brazing, welding, and etc. Therefore, they are limited in operating pressures and temperatures.

Gasketed Plate Heat Exchanger and Spiral Exchanger

Gasketed plate heat exchanger and spiral exchanger are limited by pressure and temperature, wherein the limitations are imposed by the capability of the gaskets.

Flow Rate

Flow rate determines the flow area: the higher the flow rate, the higher will be the crossflow area. Higher flow area is required to limit the flow velocity through he conduits and flow passages, and the higher velocity is limited by pressure drop, impingement, erosion, and, in the case of shell and tube exchanger, by shell-side floe-induced vibration. Sometimes a minimum flow velocity is necessary to improve the heat transfer, to eliminate stagnant area, and to minimize fouling.

Flow Arrangement

As defined before, the choice of a particular flow arrangement is dependent upon the required exchanger effectiveness, exchanger construction type, upstream and downstream ducting, package envelope, and other design criteria.

Performance Parameters – Thermal Effectiveness and Pressure Drops

Thermal Effectiveness

For high performance service requiring high thermal effectiveness, use brazed plate-fin exchangers (e.g., cryogenic service) and regenerators (e.g., gas turbine applications), use tube-fin exchangers for slightly less thermal effectiveness in applications, and use shell and tube units for low thermal effectiveness service.

Pressure Drop

Pressure drop is an important parameter in heat exchanger design. Limitations may be imposed either by pumping cost or by process limitations or both. The heat exchanger should be designed in such a way that unproductive pressure drop is avoided to the maximum extent in areas like inlet and outlet bends, nozzles, and manifolds. At the same time, any pressure drop limitation that is imposed must be utilized as nearly as possible for an economic design.

Fouling Tendencies

Fouling is defined as the formation on heat exchanger surfaces of undesirable deposits that impede the heat transfer and increase the resistance to fluid flow, resulting in higher pressure drop. The growth of these deposits causes the thermohydraulic performance of heat exchanger to decline with time. Fouling affects the energy consumption of industrial processes, and it also decides the amount of extra material required to provide extra heat transfer surface to compensate for the effects of fouling. Compact heat exchangers are generally preferred for nonfouling applications. In a shell and tube unit the fluid with more fouling tendencies should be put on the tube side for ease of cleaning. On the shellside with cross baffles, it is sometimes difficult to achieve a good flow distribution if the baffle cut is either too high or too low. Stagnation in any region of low velocity behind the baffles is difficult to avoid if he baffles are cut more than about 20-25%. Plate heat exchangers and spiral plate exchangers are better chosen for fouling services. The flow pattern in plate heat exchanger induces turbulence even at comparable low velocities; in the spiral units, the scrubbing action of fluids on the curved surfaces minimizes fouling.

Type and Phases of Fluids

The phase of the fluids within a unit is an important condition in the selection of the heat exchanger type. Various combinations of fluids phases dealt in heat exchangers are liquid-liquid, liquid-gas, and gas-gas. Liquid phase fluids are generally the simplest to deal with. The high density and favorable values of many transport properties allow high heat- transfer coefficient to be obtained at relatively low-pressure drop.

Maintenance, Inspection, Cleaning, Repair, and Extension Aspects

For instance, consider inspection and manual cleaning; spiral plate exchangers can be made with both sides open at one edge, or with one side open and one closed. They can be made with channels between 5 mm and 25 mm wide, with or without studs. The shell and tube heat exchanger can be with fixed tubesheet or with a removable tube bundle, with small- or large-diameter tubes, or small or wide pitch. Gasketed plate heat exchangers (PHEs) are easy to open, especially when all nozzles are located on the stationary end-plate side. The plate arrangement can be changed for other duties within the frame and nozzle capacity.

Repair of some of the shell and tube exchanger components is possible, but the repair of expansion joint is very difficult. Tubes can be renewed or plugged. Repair of compact heat exchangers of tube-fin type is very difficult except by plugging of the tube. Repair of the plate-fin exchanger is generally very difficult. For these two types of heat exchangers, extension of units for higher thermal duties is generally not possible. All these drawbacks are easily overcome in a PHE. It can be easily repaired, and plates and other parts can be easily replaced. Due to modular construction, PHEs possess the flexibility of enhancing or reducing the heat transfer surface area, modifying the pass arrangement, and addition of more than one duty according to the heat transfer requirements at future date.

Overall Economy

There are two major costs to consider in designing a heat exchanger: the manufacturing cost and the operating costs, including maintenance cost. In general, the less the heat transfer surface area and the less the complexity of he design, the lower is the manufacturing cost. The operating cost is the pumping cost due to pumping devices such as fans, blowers, pumps, and etc. The maintenance costs include costs of spares that require frequent renewal due to corrosion, and costs due to corrosion/fouling prevention and control. Therefore, the heat exchanger design requires a proper balance between thermal sizing and pressure drop.

Fabrication Techniques

Fabrication techniques are likely to be the determining factor o the selection of a heat transfer surface matrix or core. They are the major factors in the initial cost and to a large extent influence the integrity, service life, and ease of maintenance of the finished heat exchanger. For example, shell and tube units are mostly fabricated by welding, plate-fin heat exchanger and automobile aluminum radiators by brazing, copper-brass radiators by soldering, most of the circular tube-fin exchangers by mechanical assembling, etc.

Appendix C.5 Propane and propylene separation

Propane and propylene have similar boiling points (propane: -42.1 °C, propylene, -47.70 °C) and as a result separation of these compounds requires highly complicated units. Distillation is by far the most commonly used separation process in the chemical industry today. The variants that are in use are:

1. Single-Column Process
2. Double-Column Process
3. Heat Pump Process

Single-Column Process: This process requires a large number of trays (150 – 200), resulting in units of about 100 meters. The reflux can be condensed with cooling water (column pressure 16 – 19 bar) or in air coolers (column pressure 21 – 26 bar).

Double-Column Process: For the large throughputs that are common today, the double column process is preferred over the single-column process, since it does requires smaller columns with smaller column diameters, which makes transportation of these units easier. A schematic of the double-column process is given in Figure C.5.1. Only the reflux from the second column is condensed with cooling water. The pressure of the first column is sufficiently high (*ca.* 25 bar) that the overhead vapours (*ca.* 59 °C) can be condensed in the reboiler of the second column and serve as the heat carrier. Heating the first column with warm water is still possible. Both columns provide approximately half the propene product. Since the reboiler for the second column also serves as the condenser of the first column, the first column does not require any cooling water. As a result the cooling water requirements are about half that of the single column process.

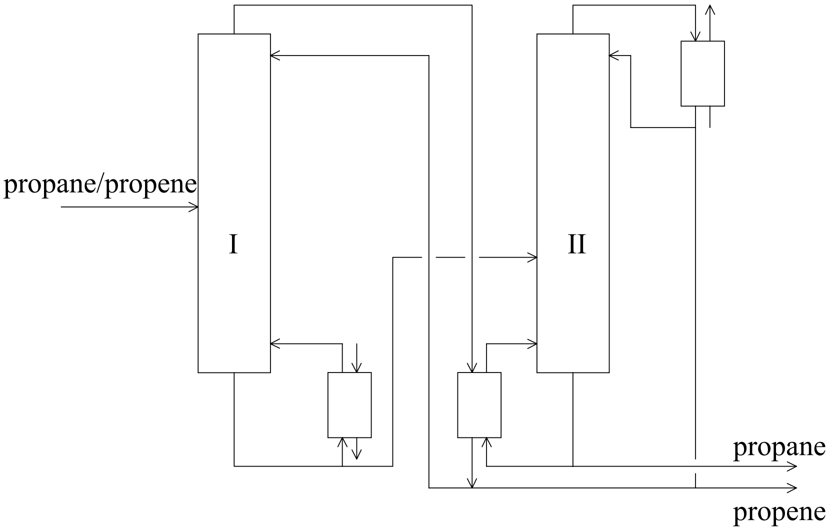

Figure C.5.1. *Schematic of double column process [adapted from Ullman]*

Heat pump process: In the aforementioned processes, the heat for the reboiler is usually available as waste heat from e.g. the steam cracker, and is essentially cost free. If this heat is not available, a heat pump can be used. A schematic of the heat pump process is given in Figure C.5.2. The overhead vapors are heated slightly in the reflux sub cooler, which enables these vapors to be compressed and cooled in the condenser-reboiler.

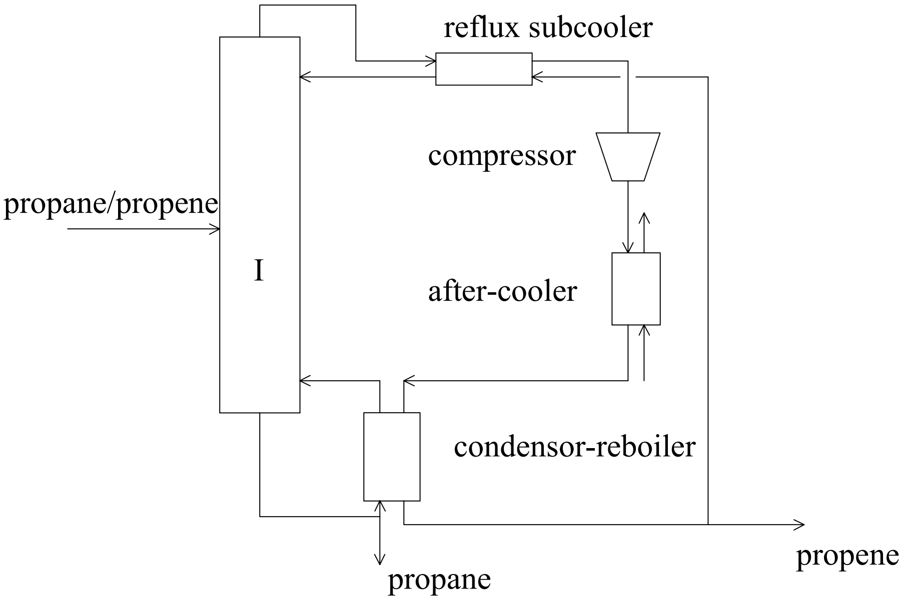

Figure C.5.2 *Schematic of heat pump process [adapted from Ullman]*

Thermodynamic analysis

A simplified scheme for the separation of an equimolar mixture of propylene and propane is given in Figure. C.5.3. The feed is a liquid mixture that is introduced at that point where the liquid has the same composition and temperature.

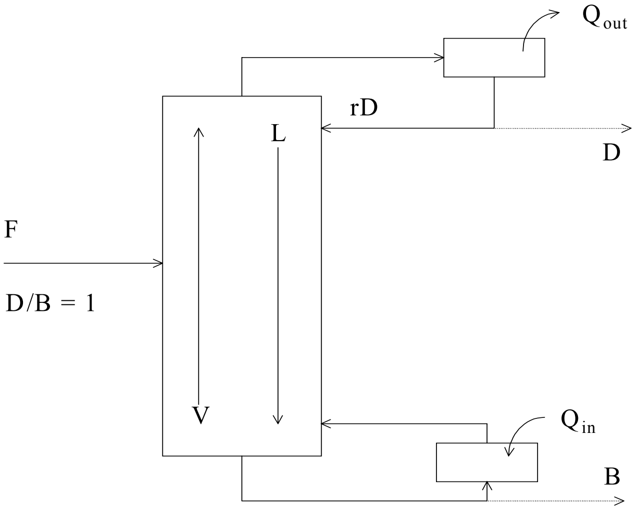

Figure C.5.3. *Distillation at reflux ratio r ($r_{min} < r < \infty$)*

Per mole of feed F, the distillate D amounts to ½ mole and so does the bottom product B. With a reflux ratio r = L/D, in which L is the number of moles of liquid re-introduced at the top of the column, L = rD = ½r moles above the feedpoint and (½r + 1) below the feedpoint. The vapour flow V = ½(r+1) throughout the column. The heat introduced at the bottom of the column is therefore

$$Q_{in} = \tfrac{1}{2}(1+r)\Delta_v H \qquad \text{(C.5.1)}$$

We assume that the heat of vaporisation is roughly the same for both components and that the temperature dependence is negligible over the range of the column. From this, it follows that

$$Q_{in} = Q_{out} \qquad \text{(C.5.2)}$$

in which Q_{out} is the cooling duty of the condenser. Now, it is interesting to note that the overall separation does not require any energy! The number of Joules entering the column equals the number of Joules leaving the column. However, the "quality" of these heat streams, or equivalently, the exergy of these heat streams is not equal, due to the Carnot factor. In an ideal column, that is a column operating under *reversible* conditions, the heat is stripped of its quality and pays for the separation of the liquid mixture into its constituents in the liquid state. The minimum work required to separate the liquid mixture into its constituents is given below (see Figure. C.5.4):

$$W_{sep}^{ideal} = -RT_0 \sum_i x_i \ln x_i \qquad \text{(C.5.3)}$$

where the assumption is made that the mixture behaves in an ideal fashion and is close to the temperature of the surroundings, which, for the propane-propylene mixture is a fair assumption. If we further insert the assumption that the mixture is equimolar, equation (C.4.3) reduces to:

$$W_{sep}^{ideal} = RT_0 \ln 2 \qquad \text{(C.5.4)}$$

which is the minimum amount of exergy that needs to be introduced into the column. And from

$$W_{in}^{\min} = Q_{in}^{\min} T_0 \left(\frac{1}{T_{top}} - \frac{1}{T_{bottom}} \right) \qquad \text{(C.5.5)}$$

which has to equal the minimum amount of work that has to be spent on the separation:

$$W_{in}^{\min} = Q_{sep}^{ideal} \qquad \text{(C.5.6)}$$

This also defines the minimum reflux ratio, *r*, according to

$$Q_{in}^{\min} = \tfrac{1}{2}\left(r^{\min} + 1\right)\Delta_v H \qquad \text{(C.5.7)}$$

Equations (C.4.4), (C.4.5) and (C.4.7) can be combined to yield:

$$r^{\min} = \frac{2 \ln 2}{\ln \alpha_{12}^{ideal}} - 1 \qquad \text{(C.5.8)}$$

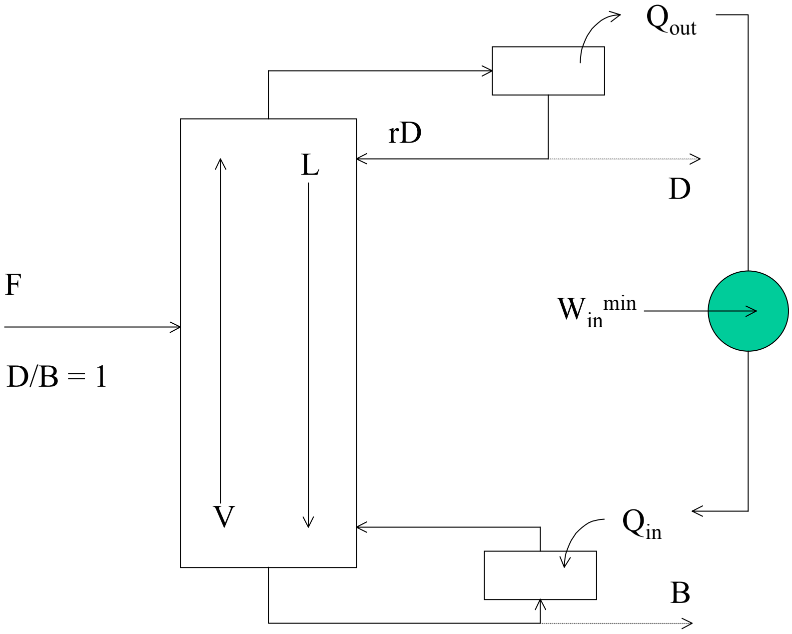

Figure C.5.4. *Minimum work required to upgrade the quality of heat*

We stress that this equation dictates the minimum reflux ratio based purely on thermodynamic arguments. As $\alpha_{12}^{ideal} = 1.11$ in our case, the value of r^{min}=12.28. In general, the mixture will not be equimolar, and if the products are not pure but satisfy a less strict specification, the value of r^{min} will be smaller. Now, a column operated under these conditions will have an efficiency of 100% since, it is using the minimum amount of work necessary to separate the components.

The previous analysis begs the following question. *What will the efficiency be of a "real" propane-propylene distillation column?* To answer this question, we must realize that heat cannot be transferred into the column without a temperature difference. In a "real" column with less stringent product quality constraints, the heat is supplied at 377 K, the bottom temperature of the column is 331 K, the temperature at the top of the column is 320 K and is transferred to the surroundings at 298 K [Seader]. The minimum heat required for separation is, according to equation (C.4.7):

$$Q_{in}^{\min} = \frac{1}{2}\left(r^{\min} + 1\right)\Delta_v H \qquad \text{(C.5.9)}$$

with r^{min} = 9.64 from the data in [Seader]. The separation inside the column does not take place according to thermodynamic ideal processes, and the real heat is larger:

$$Q_{in}^{real} = \frac{1}{2}\left(r^{real} + 1\right)\Delta_v H \qquad \text{(C.5.10)}$$

where r^{real} = 15.9

The heat has to be transferred over a temperature difference of 377 – 331 = 46 K and the resulting lost work can easily be calculated. Then the heat flows from 331 K to 320 K inside the column and is used to perform the separation. Finally, the heat is discarded at the top of the column at 320 K to the surroundings at 298 K. The overall thermodynamic efficiency of the column can be computed as follows:

$$\eta_{overall} = \frac{Q_{in}^{\min}\Delta\left(\frac{1}{T}\right)_{column} T_0}{Q_{in}^{real}\left[\Delta\left(\frac{1}{T}\right)_{bottom} + \Delta\left(\frac{1}{T}\right)_{column} + \Delta\left(\frac{1}{T}\right)_{top}\right]T_0} \qquad \text{(C.5.11)}$$

which yields $\eta_{overall}$ = 0.093:

Closer scrutiny of equation (C.5.11) reveals that the main sources of inefficiency are the temperature driving forces in the condenser and reboiler (the ratio $\frac{Q_{in}^{\min}}{Q_{in}^{real}}$ equals 0.63 and quantifies losses inside the column). The only way of improving the efficiency is to reduce these temperature-driving forces. A noteworthy point is, however, that the heat for the reboiler should be supplied at 377 K, which is often available as waste heat in a chemical plant and integration with other heat sources should therefore be contemplated. For the purpose of this example, however, we will not do so. Another way of improving the single-column process is to use a membrane to split the feed into two different feed streams. See Figure C.4.5.

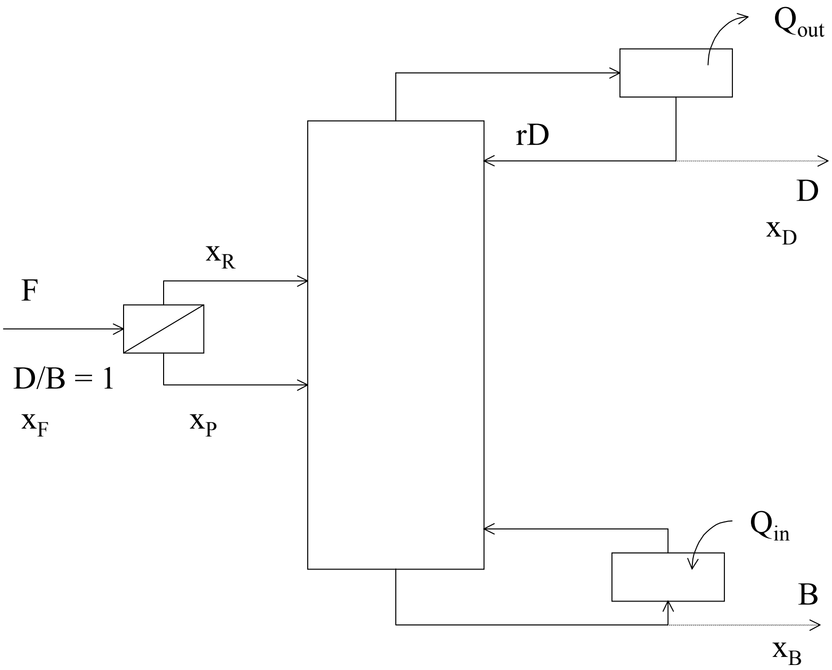

Figure C.4.5 *Hybrid distillation of propane-propylene*

Appendix C.6 Comparison of the Tray and Packed column properties

Table C.6. *Summary of column properties comparison*

	Tray column	Packed column
Diameter	Large	Small
Pressure loss	High (7mbar/stage)	Low (<0.7mbar/stage)
Liquid hold up	Varied over wide range	Very small, thermally unstable substance is less
Gas load	Narrow range	Flexible

Tray column

1. Generally employed in large diameter (larger than 1m)
2. Several down comers necessary
3. Gas load in tray columns must be kept within a relatively narrow range
4. Only valve trays allow greater operational flexibility
5. Liquid load can be varied over a very wide range
6. Allow heat to be added or removed easily

Disadvantages:

1. Relatively high pressure drop (7mbar/equilibrium stage)
2. Decomposition of thermally unstable substances

Packed column

1. Small diameter (smaller than 0.7m), development allows large diameter also
2. Extremely flexible as far as gas load is concerned, but require a minimum liquid load
3. Small pressure loss (0.5mbar/equilibrium stage), more than 1 order magnitude lower than in tray columns
4. Liquid hold up very small, thermally unstable substance is less
5. Countercurrent of gas and liquid, efficiency of mass transfer
6. Ceramic packing. Less corrosion

Disadvantages

1. Not suitable for the treatment of liquids that obtain particulate contaminants or tent to crystallize

Appendix C.7 Recommendations for treatment of light gas

The light gas is consisted of methane, hydrogen and carbon monoxide.
Here are some recommendations for the treatment of light gas.

Membrane Technology

Membrane is a novel technology in separation field. One can select a qualified material for the specified purpose of separation. Hydrogen and carbon monoxide are the composition of synthesis gas and methane is natural gas. Therefore, to separate methane from the stock is the desired way. Unfortunately, after searching the huge amount of literature, we found that due to the fact that the molecular size of methane and carbon monoxide are very close (Diameter of CO: 3.76, Diameter of CH4: 3.80), it is not applicable by separate only by physical sense.
By physical properties membrane separation, only hydrogen can be separated from the stock.

Membrane Reactor

The wide application of membrane reactor is methane steam reforming. In the membrane reactor, two reactions will happen in parallel:

CH_4 steam reforming: $CH_4 + H_2O \rightarrow CO + 3H_2$
Water shifting: $CO + H_2O \rightarrow CO_2 + H_2$

Methane will convert to CO and H_2, which is valuable syngas can be sold to Fischer-Tropsch process. However, this treatment is quite time consuming in calculation of the reactor parameter design and very costly. And this process will bring the by-product CO_2 that requires further treatment for the sustainability point of view.

To sale whole products

Hydrogen and carbon monoxide are Fischer-Tropsch feedstock syngas, and methane is the good fuel gas. Therefore, it is suggested to sale all these three components together to be the feedstock of Fischer-Tropsch process. This is the most economic way for the light gas treatment.

Absorption

In order to separate methane from the stock, one can use the chemical way such as using a right solvent. It is required to cost money for the solvent and to design an extra equipment.

Chromatography

Gas Chromatography (GC) is used for the qualitative and quantitative analysis of complex mixtures of gases, liquids, and sometimes solids. A sample is vaporized and transported by an inert carrier (usually He) gas through a column of sufficient length to provide the separation. There are many different columns with both different

mobile and solid phases used with different carrier gases. The right combination has to be chosen such that the desired separation takes place with reasonable retention times. The different components of the vapor mixture are separated as a result of their different vapor pressures and relative affinities for the bonded liquid phase. As the components of the mixture are separated and elude from the column, they enter a detector, where a signal proportional to the concentration of the component is amplified and displayed. Identification is done by the retention time.
The Chromatography is more commonly used in analysis way and not applicable in the huge amount separation unit. Also the problem is it is very costly.

Conclusion:
In this design, we choose to sale whole products mixture as our decision. The other alternatives can serve as the candidates if there are specified requirements for the treatment, such as purity and H_2/CO ratio.

APPENDIX D

Appendix D. Process Stream Summary

***Table D.1.a** Process stream summary*

STREAM Nr. :		A001	IN	A101		A102		A103		A104	
Name :		Propane Feed In		Propane feed to shell		Gas Product from Shell		Gas product after cooler		Shell gas prod. Compressed	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	7.45	0.1692	9.81	0.2230	7.56	0.1717	7.56	0.1717	7.56	0.1717
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	0.00	0.0000	0.00	0.0000	0.00	0.0001	0.00	0.0001	0.00	0.0001
Propylene	42	0.79	0.0188	1.03	0.0244	3.18	0.0756	3.18	0.0756	3.18	0.0756
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Methane	16	0.00	0.0000	0.00	0.0000	0.00	0.0001	0.00	0.0001	0.00	0.0001
Hydrogen	2	0.00	0.0000	0.00	0.0000	0.10	0.0515	0.10	0.0515	0.10	0.0515
Water	18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Nitrogen	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		8.23	0.1880	10.84	0.2475	10.84	0.2990	10.84	0.2990	10.84	0.2990
Enthalpy	kW	-5975		-7900		-1247		-16528		-12344	
Phase		V		V		V		V		V	
Press.	Bara	1		1		1		1		30	
Temp	oC	540		540		540		25		215	

STREAM Nr. :		A105		A106		A107		A201		A202	
Name :		Comp. gas shell aft cooler		Shell Gas to separation		Liquid shell prod. To dist.		Propane to Tube reactor		Oxygen to tube reactor	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	7.56	0.1717	5.08	0.1155	2.48	0.0563	6.54	0.1487	0.00	0.0000
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	1.79	0.0559
Ethylene	28	0.00	0.0001	0.00	0.0001	0.00	0.0000	0.00	0.0000	0.00	0.0000
Propylene	42	3.18	0.0756	2.22	0.0528	0.96	0.0228	0.68	0.0163	0.00	0.0000
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Methane	16	0.00	0.0001	0.00	0.0001	0.00	0.0000	0.00	0.0000	0.00	0.0000
Hydrogen	2	0.10	0.0515	0.10	0.0504	0.00	0.0011	0.00	0.0000	0.00	0.0000
Water	18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Nitrogen	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.08	0.0029
Total		10.84	0.2990	7.40	0.2189	3.43	0.0801	7.23	0.1650	1.87	0.0589
Enthalpy	kW	-17475		-11090		-6385		-5267		959	
Phase		V/L		V		L		V		V	
Press.	Bara	30		30		30		1		1	
Temp	oC	55		54.7056169		54.7056169		540		540	

Table D.1.a *Process stream summary (Con't)*

STREAM Nr.	:	A203		A204		A205		A206		A207	
Name :		Tube Feed furnace		Feed to tube reactor		Tube gas Prod.		Tube gas prod. After cooler		Process water	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	6.54	0.1487	6.54	0.1487	1.36	0.0309	1.36	0.0309	0.00	0.0000
Oxygen	32	1.79	0.0559	1.79	0.0559	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	0.00	0.0000	0.00	0.0000	2.29	0.0818	2.29	0.0818	0.00	0.0000
Propylene	42	0.68	0.0163	0.68	0.0163	2.47	0.0588	2.47	0.0588	0.00	0.0000
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.62	0.0223	0.62	0.0223	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.00	0.0000	0.18	0.0040	0.18	0.0040	0.00	0.0000
Methane	16	0.00	0.0000	0.00	0.0000	0.58	0.0360	0.58	0.0360	0.00	0.0000
Hydrogen	2	0.00	0.0000	0.00	0.0000	0.05	0.0265	0.05	0.0265	0.00	0.0000
Water	18	0.00	0.0000	0.00	0.0000	1.47	0.0817	0.12	0.0064	1.36	0.0753
Nitrogen	28	0.08	0.0029	0.08	0.0029	0.08	0.0029	0.08	0.0029	0.00	0.0000
Total		9.10	0.2239	9.10	0.2239	9.10	0.3449	7.74	0.2695	1.36	0.0753
Enthalpy	kW	-4308		4957		-4243		-6063		-21804	
Phase		V		V		V		V		L	
Press.	Bara	1		1		1		1		1	
Temp	oC	539.978075		850		850		25		25	

STREAM Nr.	:	A208		A209		A210		401		402	
Name :		Tube gas compressed		Comp.Tube gas after cooler		Process water		Gas prod. To CO_2 rem.Unit		CO2	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	1.36	0.0309	1.36	0.0309	0.00	0.0000	1.36	0.0309	0.00	0.0000
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	2.29	0.0818	2.29	0.0818	0.00	0.0000	2.29	0.0818	0.00	0.0000
Propylene	42	2.47	0.0588	2.47	0.0588	0.00	0.0000	2.47	0.0588	0.00	0.0000
Carbonmon-oxide	28	0.62	0.0223	0.62	0.0223	0.00	0.0000	0.62	0.0223	0.00	0.0000
Carbondi-oxide	44	0.18	0.0040	0.18	0.0040	0.00	0.0000	0.01	0.0003	0.16	0.0037
Methane	16	0.58	0.0360	0.58	0.0360	0.00	0.0000	0.58	0.0360	0.00	0.0000
Hydrogen	2	0.05	0.0265	0.05	0.0265	0.00	0.0000	0.05	0.0265	0.00	0.0000
Water	18	0.12	0.0064	0.01	0.0006	0.10	0.0058	0.01	0.0006	0.00	0.0000
Nitrogen	28	0.08	0.0029	0.08	0.0029	0.00	0.0000	0.08	0.0029	0.00	0.0000
Total		7.74	0.2695	7.64	0.2637	0.10	0.0058	7.48	0.2601	0.16	0.0037
Enthalpy	kW	-1533		-4946		-1665		-3494		-1451	
Phase		V		V		L		V		V	
Press.	Bara	30		30		30		30		30	
Temp	oC	308		30		30		30		30	

***Table D.1.a** Process stream summary (Con’t)*

STREAM Nr. :		301		302		303		304		305	
Name :		Gas Prod. to LG column		Overhead T301		Light gas Product		T301 Reflux		Feed to T302	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	6.44	0.1464	0.00	0.0000	0.00	0.0000	0.00	0.0000	6.44	0.1464
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	2.29	0.0819	7.90	0.2821	0.03	0.0010	7.87	0.2811	2.26	0.0809
Propylene	42	4.69	0.1116	0.00	0.0000	0.00	0.0000	0.00	0.0000	4.69	0.1116
Carbonmon-oxide	28	0.62	0.0223	1.19	0.0426	0.62	0.0223	0.57	0.0203	0.00	0.0000
Carbondi-oxide	44	0.01	0.0003	0.01	0.0002	0.00	0.0000	0.01	0.0002	0.01	0.0003
Methane	16	0.58	0.0361	4.79	0.2994	0.58	0.0361	4.21	0.2633	0.00	0.0000
Hydrogen	2	0.15	0.0769	0.16	0.0797	0.15	0.0769	0.01	0.0027	0.00	0.0000
Water	18	0.01	0.0006	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.01	0.0006
Nitrogen	28	0.08	0.0029	0.14	0.0050	0.08	0.0029	0.06	0.0021	0.00	0.0000
Total		14.88	0.4790	14.19	0.7089	1.47	0.1392	12.72	0.5697	13.42	0.3398
Enthalpy	kW	-14585		-5787		-5786		-5787		-14573	
Phase		V		V		V		V		L	
Press.	Bara	30		15		15		15		15	
Temp	oC	42		-131		-131		-131		10	

STREAM Nr. :		306		307		308		309		310	
Name :		Overhead T302		T302 Reflux		Ethylene Prduct		Bottom product T303		Feed to T303	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	0.01	0.0002	0.01	0.0002	0.00	0.0000	6.44	0.1463	8.91	0.2026
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	2.26	0.0808	0.00	0.0000	2.26	0.0808	0.00	0.0001	0.00	0.0001
Propylene	42	8.53	0.2032	8.52	0.2029	0.01	0.0003	4.67	0.1113	5.63	0.1340
Carbonmon-oxide	28	0.05	0.0018	0.05	0.0018	0.00	0.0000	0.00	0.0000	0.00	0.0000
Carbondi-oxide	44	0.01	0.0003	0.00	0.0000	0.01	0.0003	0.00	0.0000	0.00	0.0000
Methane	16	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Hydrogen	2	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0011
Water	18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.01	0.0006	0.01	0.0006
Nitrogen	28	0.05	0.0019	0.05	0.0019	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		10.93	0.2883	8.63	0.2068	2.29	0.0815	11.12	0.2582	14.56	0.3384
Enthalpy	kW	-14264		11270		2994		-16920		-23305	
Phase		V		L		L		L		L	
Press.	Bara	15		15		15		15		15	
Temp	oC	-39		-39		-39		40		38	

Table D.1.a *Process stream summary (Con't)*

STREAM Nr. :		311		312		313		314		314-W	
Name :		Overhead T303		T303 Reflux		Propylene Prduct		Propane recycle1		Moisture removed	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	0.05	0.0011	0.05	0.0011	0.00	0.0001	8.91	0.2025	0.00	0.0000
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	0.04	0.0015	0.04	0.0014	0.00	0.0001	0.00	0.0000	0.00	0.0000
Propylene	42	105.15	2.5036	100.44	2.3914	4.71	0.1121	0.92	0.0219	0.00	0.0000
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Methane	16	0.00	0.0002	0.00	0.0002	0.00	0.0000	0.00	0.0000	0.00	0.0000
Hydrogen	2	0.05	0.0242	0.05	0.0231	0.00	0.0011	0.00	0.0000	0.00	0.0000
Water	18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.01	0.0006
Nitrogen	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		105.29	2.5306	100.58	2.4172	4.72	0.1133	9.83	0.2244	0.01	0.0006
Enthalpy	kW	0		0		2093		-24058		-168	
Phase		V		L		V		L		L	
Press.	Bara	15		15		15		15		15	
Temp	oC	35		35		35		43		43	

Table D.1.b *Process stream summary of input and output*

STREAM Nr. :		403		404	
Name :		MDEA sol feed		Spent MDEA sol	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s
Propane	44	0.00	0.0000	0.00	0.0000
Oxygen	32	0.00	0.0000	0.00	0.0000
Ethylene	28	0.00	0.0000	0.00	0.0000
Propylene	42	0.00	0.0000	0.00	0.0000
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.16	0.0037
Methane	16	0.00	0.0000	0.00	0.0000
Hydrogen	2	0.00	0.0000	0.00	0.0000
Water	18	15.96	0.8864	15.96	0.8864
Nitrogen	28	0.00	0.0000	0.00	0.0000
MDEA	119.2	15.96	0.1339	15.96	0.1339
Total		31.91	0.8864	32.07	0.8901
Enthalpy	kW	n.a.		n.a.	
Phase		L		L	
Press.	Bara	30		30	
Temp	oC	30		30	

APPENDIX E

Appendix E.1 Aspen Plus simulation results

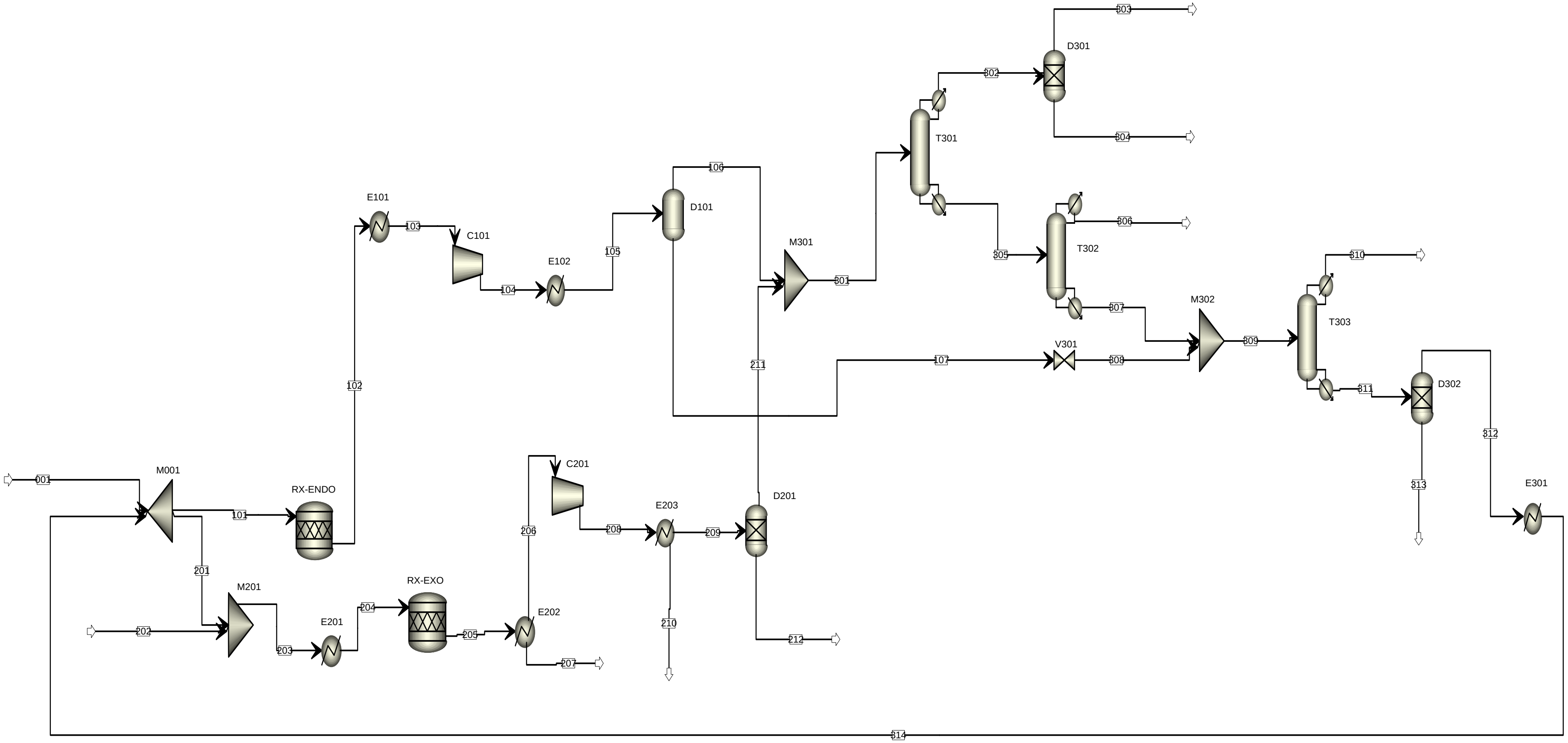

Figure E.1.1 *The flow scheme of Aspen Plus simulation*

Table E.1.1 *The results of Aspen Plus simulation*

Heat and Material Balance Table									
Stream ID		001	101	102	103	104	105	106	107
Temperature	C	540.0	540.0	540.0	25.0	215.1	55.0	54.7	54.7
Pressure	bar	1.000	1.000	1.000	1.000	30.000	30.000	30.000	30.000
Vapor Frac		1.000	1.000	1.000	1.000	1.000	0.730	1.000	0.000
Mole Flow	kmol/hr	675.348	888.943	1073.215	1073.215	1073.215	1073.215	785.355	287.860
Mass Flow	kg/hr	29644.355	39022.384	39022.384	39022.384	39022.384	39022.384	26657.857	12364.527
Volume Flow	cum/hr	45666.790	60110.020	72578.525	26334.839	1379.850	587.287	547.674	28.420
Enthalpy	MMkcal/hr	-5.121	-6.772	-1.069	-14.167	-10.581	-14.979	-9.506	-5.473
Mass Flow	kg/hr								
C_3H_8		26802.447	35329.320	27203.576	27203.576	27203.576	27203.576	18291.666	8911.911
O_2									
N_2									
C_2H_4				10.339	10.339	10.339	10.339	8.799	1.540
C_3H_6		2841.908	3693.064	11431.830	11431.830	11431.830	11431.830	7988.958	3442.872
CH_4				5.912	5.912	5.912	5.912	5.497	0.416
H_2				370.727	370.727	370.727	370.727	362.939	7.788
H_2O									
CO									
CO_2									
MDEA+									
Mole Flow	kmol/hr								
C_3H_8		607.813	801.182	616.910	616.910	616.910	616.910	414.810	202.100
O_2									
N_2									
C_2H_4				0.369	0.369	0.369	0.369	0.314	0.055
C_3H_6		67.535	87.762	271.665	271.665	271.665	271.665	189.849	81.816
CH_4				0.369	0.369	0.369	0.369	0.343	0.026
H_2				183.903	183.903	183.903	183.903	180.040	3.863
H_2O									
CO									
CO_2									
MDEA+									
Mole Frac									
C_3H_8		0.900	0.901	0.575	0.575	0.575	0.575	0.528	0.702
O_2									
N_2									
C_2H_4				343 PPM	343 PPM	343 PPM	343 PPM	399 PPM	191 PPM
C_3H_6		0.100	0.099	0.253	0.253	0.253	0.253	0.242	0.284
CH_4				343 PPM	343 PPM	343 PPM	343 PPM	436 PPM	90 PPM
H_2				0.171	0.171	0.171	0.171	0.229	0.013
H_2O									
CO									
CO_2									
MDEA+									

Table E.1.1 *The results of Aspen Plus simulation (Con’t)*

Heat and Material Balance Table													
Stream ID		201	202	203	204	205	206	207	208	209	210	211	212
Temperature	C	540.0	540.0	540.0	850.0	850.0	25.0	25.0	288.6	30.0	30.0	30.0	30.0
Pressure	bar	1.000	1.000	1.000	1.000	1.000	1.000	1.000	30.000	30.000	30.000	30.000	30.000
Vapor Frac		1.000	1.000	1.000	1.000	1.000	1.000	0.000	1.000	1.000	0.000	1.000	1.000
Mole Flow	kmol/hr	592.629	212.000	804.629	804.629	1238.872	967.996	270.876	967.996	947.059	20.937	933.860	13.198
Mass Flow	kg/hr	26014.923	6741.501	32756.424	32756.424	32755.239	27875.317	4879.922	27875.317	27498.122	377.194	26917.260	580.862
Volume Flow	cum/hr	40073.347	14337.397	54415.038	75172.427	115724.474	23865.336	6.460	1501.437	667.676	0.378	658.313	9.362
Enthalpy	MMkcal/hr	-4.514	0.822	-3.692	4.249	-3.637	-5.197	-18.689	-1.647	-4.239	-1.427	-2.995	-1.244
Mass Flow	kg/hr												
C3H8		23552.880		23552.880	23552.880	4891.314	4891.314	< 0.001	4891.314	4891.314		4891.314	
O2			6444.558	6444.558	6444.558								
N2			296.943	296.943	296.943	296.943	296.943	trace	296.943	296.943		296.943	
C2H4						8246.499	8246.497	0.002	8246.497	8246.497		8246.497	
C3H6		2462.043		2462.043	2462.043	8885.550	8885.549	0.001	8885.549	8885.549		8885.549	
CH4						2072.763	2072.763	< 0.001	2072.763	2072.763		2072.763	
H2						190.757	190.757	trace	190.757	190.757		190.757	
H2O						5294.891	414.984	4879.907	414.984	37.789	377.194	37.789	
CO						2245.138	2245.138	< 0.001	2245.138	2245.138		2245.138	
CO2						631.383	631.372	0.011	631.372	631.372		50.510	580.862
MDEA+													
Mole Flow	kmol/hr												
C3H8		534.121		534.121	534.121	110.923	110.923	trace	110.923	110.923		110.923	
O2			201.400	201.400	201.400								
N2			10.600	10.600	10.600	10.600	10.600	trace	10.600	10.600		10.600	
C2H4						293.953	293.953	< 0.001	293.953	293.953		293.953	
C3H6		58.508		58.508	58.508	211.155	211.155	< 0.001	211.155	211.155		211.155	
CH4						129.202	129.202	< 0.001	129.202	129.202		129.202	
H2						94.627	94.627	trace	94.627	94.627		94.627	
H2O						293.911	23.035	270.876	23.035	2.098	20.937	2.098	
CO						80.154	80.154	trace	80.154	80.154		80.154	
CO2						14.346	14.346	< 0.001	14.346	14.346		1.148	13.198
MDEA+													
Mole Frac													
C3H8		0.901		0.664	0.664	0.090	0.115	10 PPB	0.115	0.117		0.119	
O2			0.950	0.250	0.250								
N2			0.050	0.013	0.013	0.009	0.011	trace	0.011	0.011		0.011	
C2H4						0.237	0.304	318 PPB	0.304	0.310		0.315	
C3H6		0.099		0.073	0.073	0.170	0.218	77 PPB	0.218	0.223		0.226	
CH4						0.104	0.133	66 PPB	0.133	0.136		0.138	
H2						0.076	0.098	7 PPB	0.098	0.100		0.101	
H2O						0.237	0.024	1.000	0.024	0.002	1.000	0.002	
CO						0.065	0.083	4 PPB	0.083	0.085		0.086	
CO2						0.012	0.015	929 PPB	0.015	0.015		0.001	1.000
MDEA+													

Table E.1.1 *The results of Aspen Plus simulation (Con't)*

Heat and Material Balance Table															
Stream ID		301	302	303	304	305	306	307	308	309	310	311	312	313	314
Temperature	C	42.5	-131.4	-131.4	-131.4	9.5	-39.0	39.9	37.4	37.9	34.8	42.7	42.7	42.7	540.0
Pressure	bar	30.000	15.000	15.000	15.000	15.000	15.000	15.000	15.000	15.000	15.000	15.000	15.000	15.000	1.000
Vapor Frac		1.000	1.000	1.000	0.000	0.000	0.000	0.000	0.200	0.060	1.000	0.000	1.000	0.000	1.000
Mole Flow	kmol/hr	1719.215	498.572	494.958	3.614	1220.642	292.954	927.688	287.860	1215.549	407.227	808.322	806.224	2.098	806.224
Mass Flow	kg/hr	53575.117	5275.333	5173.947	101.386	48299.784	8255.705	40044.079	12364.527	52408.606	16977.864	35430.742	35392.953	37.789	35392.953
Volume Flow	cum/hr	1225.285	371.054	369.247	0.173	105.967	19.365	93.977	102.049	214.237	542.593	84.936	1078.658	0.051	54516.577
Enthalpy	MMkcal/hr	-12.501	-4.960	-4.999	0.028	-12.491	2.566	-14.503	-5.473	-19.975	1.794	-20.621	-17.924	-0.144	-6.165
Mass Flow	kg/hr														
C3H8		23182.980	trace		trace	23182.980	7.022	23175.958	8911.911	32087.869	8.116	32079.753	32079.753		32079.753
O2															
N2		296.943	296.943	296.943		trace	trace	trace		trace					
C2H4		8255.295	101.386		101.386	8153.909	8148.609	5.300	1.540	6.841	6.841	trace			
C3H6		16874.507	trace		trace	16874.507	49.528	16824.979	3442.872	20267.851	16954.651	3313.199	3313.199		3313.199
CH4		2078.259	2078.124	2078.124		0.136	0.136	trace	0.416	0.416	0.416	trace			
H2		553.696	553.696	553.696		trace	trace	trace	7.788	7.788	7.788	trace			
H2O		37.789	trace			37.789	trace	37.789		37.789	trace	37.789		37.789	
CO		2245.138	2245.138	2245.138		trace	trace	trace		trace					
CO2		50.510	0.047	0.047		50.462	50.410	0.052		0.052	0.052	trace			
MDEA+															
Mole Flow	kmol/hr														
C3H8		525.733	trace		trace	525.733	0.159	525.573	202.100	727.673	0.184	727.489	727.489		727.489
O2															
N2		10.600	10.600	10.600		trace	trace	trace		trace					
C2H4		294.267	3.614		3.614	290.653	290.464	0.189	0.055	0.244	0.244	trace			
C3H6		401.004	trace		trace	401.004	1.177	399.827	81.816	481.643	402.909	78.735	78.735		78.735
CH4		129.545	129.537	129.537		0.008	0.008	trace	0.026	0.026	0.026	trace			
H2		274.667	274.667	274.667		trace	trace	trace	3.863	3.863	3.863	trace			
H2O		2.098	trace			2.098	trace	2.098		2.098	trace	2.098		2.098	
CO		80.154	80.154	80.154		trace	trace	trace		trace					
CO2		1.148	0.001	0.001		1.147	1.145	0.001		0.001	0.001	trace			
MDEA+															
Mole Frac															
C3H8		0.306	trace		trace	0.431	544 PPM	0.567	0.702	0.599	452 PPM	0.900	0.902		0.902
O2															
N2		0.006	0.021	0.021		trace	trace	trace		trace					
C2H4		0.171	0.007		1.000	0.238	0.992	204 PPM	191 PPM	201 PPM	599 PPM	trace			
C3H6		0.233	trace		3 PPB	0.329	0.004	0.431	0.284	0.396	0.989	0.097	0.098		0.098
CH4		0.075	0.260	0.262		7 PPM	29 PPM	trace	90 PPM	21 PPM	64 PPM	trace			
H2		0.160	0.551	0.555		trace	trace	trace	0.013	0.003	0.009	trace			
H2O		0.001	trace			0.002	trace	0.002		0.002	trace	0.003		1.000	
CO		0.047	0.161	0.162		trace	trace	trace		trace					
CO2		668 PPM	2 PPM	2 PPM		939 PPM	0.004	1 PPM		976 PPB	3 PPM	trace			
MDEA+															

Appendix E.2 Simulation results of distillation columns T301 T302 and T303

- **BLOCK: T301 MODEL: RADFRAC**

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   INLETS     - 301     STAGE  10
   OUTLETS  - 302     STAGE   1
                  -  305     STAGE  20
  PROPERTY OPTION SET:   RK-ASPEN  REDLICH-KWONG-ASPEN EQUATION OF STATE

               ***  MASS AND ENERGY BALANCE  ***
                                      IN            OUT        RELATIVE DIFF.
     TOTAL BALANCE
     MOLE(KMOL/HR )             1719.21        1719.21      -0.132254E-15
     MASS(KG/HR   )             53575.1        53575.1      -0.149389E-14
     ENTHALPY(MMKCAL/H)  -12.5011       -17.4509       0.283641


                 *********************
                 ****  INPUT DATA  ****
                 *********************

  ****   INPUT PARAMETERS   ****

   NUMBER OF STAGES                                         20
   ALGORITHM OPTION                                       STANDARD
   ABSORBER OPTION                                        NO
   INITIALIZATION OPTION                                   STANDARD
   HYDRAULIC PARAMETER CALCULATIONS                 NO
   INSIDE LOOP CONVERGENCE METHOD                  BROYDEN
   DESIGN SPECIFICATION METHOD                        NESTED
   MAXIMUM NO. OF OUTSIDE LOOP ITERATIONS          200
   MAXIMUM NO. OF INSIDE LOOP ITERATIONS            10
   MAXIMUM NUMBER OF FLASH ITERATIONS               50
   FLASH TOLERANCE                                          0.000100000
   OUTSIDE LOOP CONVERGENCE TOLERANCE              0.000100000

  ****   COL-SPECS   ****

   MOLAR VAPOR DIST / TOTAL DIST                          1.00000
   MOLAR BOILUP RATIO                                     0.30000
   BOTTOMS TO FEED RATIO                                  0.71000

  ****    PROFILES    ****

   P-SPEC         STAGE   1  PRES, BAR                    15.0000

                  *******************
                  ****  RESULTS  ****
                  *******************


  ***   COMPONENT SPLIT FRACTIONS   ***

                    OUTLET STREAMS
                    --------------
                               302         305
   COMPONENT:
   C3H8               .25189E-11      1.0000
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 N2                     1.0000          .24305E-09
 C2H4                .12281E-01      .98772
 C3H6                .27710E-10     1.0000
 CH4                   .99993          .65298E-04
 H2                    1.0000          .13345E-13
 H2O                .64603E-20       1.0000
 CO                    1.0000          .72575E-09
 CO2                .93762E-03        .99906


***    SUMMARY OF KEY RESULTS    ***

 TOP STAGE TEMPERATURE         C                             -131.444
 BOTTOM STAGE TEMPERATURE      C                              9.51630
 TOP STAGE LIQUID FLOW         KMOL/HR                        2,046.42
 BOTTOM STAGE LIQUID FLOW      KMOL/HR                        1,220.64
 TOP STAGE VAPOR FLOW          KMOL/HR                         498.572
 BOTTOM STAGE VAPOR FLOW       KMOL/HR                        366.193
 MOLAR REFLUX RATIO                                           4.10455
 MOLAR BOILUP RATIO                                           0.30000
 CONDENSER DUTY (W/O SUBCOOL)  MMKCAL/H                      -6.24869
 REBOILER DUTY                 MMKCAL/H                       1.29891

****   MAXIMUM FINAL RELATIVE ERRORS   ****

 DEW POINT                          0.35192E-08      STAGE=  2
 BUBBLE POINT                       0.36639E-06      STAGE=  1
 COMPONENT MASS BALANCE         0.12473E-06  STAGE=  1 COMP=H2
 ENERGY BALANCE               0.25452E-07  STAGE=  2


****   PROFILES   ****

**NOTE** REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE FLOWS
         FROM THE STAGE EXCLUDING ANY SIDE PRODUCT.  FOR THE FIRST STAGE,
         THE REPORTED VAPOR FLOW IS THE VAPOR DISTILLATE FLOW.  FOR THE
         LAST STAGE, THE REPORTED LIQUID FLOW IS THE LIQUID BOTTOMS FLOW.

                           ENTHALPY
STAGE TEMPERATURE  PRESSURE          KCAL/MOL          HEAT DUTY
    C          BAR          LIQUID      VAPOR       MMKCAL/H

 1  -131.44       15.000     -6.9313     -9.9478        -6.2486
 2  -67.400       15.000      5.0027     -5.0670
 4  -47.815       15.000      8.3663      6.9148
 5  -47.138       15.000      8.4249      7.3618
 6  -46.810       15.000      8.2803      7.4112
 7  -45.298       15.000      7.2614      7.2839
 8  -36.418       15.000      1.7782      6.4012
 9  -11.228       15.000     -8.6914      1.9513
10  -4.4281       15.000     -7.8347      3.8437
11  -2.9161       15.000     -7.6091      4.6196
19   1.4536       15.000     -8.2415      4.4683
20   9.5163       15.000     -10.233      1.9440         1.2989

STAGE     FLOW RATE              FEED RATE             PRODUCT RATE
        KMOL/HR               KMOL/HR              KMOL/HR
   LIQUID   VAPOR     LIQUID   VAPOR   MIXED     LIQUID    VAPOR
 1  2046.    498.6                               498.5722
 2  2230.    2545.
 4  2371.    2840.
```

```
  5  2374.      2870.
  6  2351.      2872.
  7  2212.      2850.
  8  1826.      2711.
  9  1474.      2325.              1719.2147
 10  1560.       253.8
 11  1578.       339.4
 19  1587.       368.8
 20  1221.       366.2                               1220.6425

   ****  MASS FLOW PROFILES  ****

STAGE     FLOW RATE                FEED RATE             PRODUCT RATE
        KG/HR                  KG/HR              KG/HR
    LIQUID    VAPOR      LIQUID    VAPOR    MIXED      LIQUID    VAPOR
  1 0.4581E+05  5275.                              5275.3340
  2 0.5928E+05 0.5108E+05
  4 0.6601E+05 0.6998E+05
  5 0.6622E+05 0.7128E+05
  6 0.6582E+05 0.7149E+05
  7 0.6336E+05 0.7110E+05
  8 0.5844E+05 0.6864E+05
  9 0.5619E+05 0.6372E+05           .53575+05
 10 0.5907E+05  7888.
 11 0.5974E+05 0.1077E+05
 19 0.6090E+05 0.1209E+05
 20 0.4830E+05 0.1260E+05                          .48300+05

                 ****   MOLE-X-PROFILE     ****
  STAGE     C3H8         N2           C2H4         C3H6         CH4
    1    0.16633E-07   0.36507E-02   0.49355      0.84884E-07   0.46199
    2    0.38278E-06   0.46273E-03   0.87264      0.14763E-05   0.11990
    4    0.36009E-04   0.23949E-03   0.97851      0.85336E-04   0.15891E-01
    5    0.31641E-03   0.23590E-03   0.98072      0.59048E-03   0.12630E-01
    6    0.27491E-02   0.23421E-03   0.97517      0.40426E-02   0.11964E-01
    7    0.22776E-01   0.22735E-03   0.93292      0.26462E-01   0.11521E-01
    8    0.14015       0.20170E-03   0.71333      0.13058       0.10030E-01
    9    0.37938       0.18784E-03   0.31319      0.29310       0.80810E-02
   10    0.36696       0.46244E-04   0.33910      0.28476       0.59140E-02
   11    0.36472       0.91215E-05   0.34584      0.28331       0.34178E-02
   19    0.38237       0.13052E-10   0.31657      0.29847       0.19064E-04
   20    0.43070       0.21106E-11   0.23811      0.32852       0.69300E-05

                 ****   MOLE-X-PROFILE     ****
  STAGE      H2           H2O           CO            CO2
    1    0.48024E-02   0.00000E+00   0.35664E-01   0.34017E-03
    2    0.16473E-02   0.28743E-14   0.45332E-02   0.81127E-03
    4    0.17896E-02   0.12388E-10   0.20636E-02   0.13825E-02
    5    0.17853E-02   0.70223E-09   0.20275E-02   0.16904E-02
    6    0.17782E-02   0.39413E-07   0.20123E-02   0.20502E-02
    7    0.17399E-02   0.21578E-05   0.19526E-02   0.23993E-02
    8    0.16411E-02   0.92359E-04   0.17255E-02   0.22475E-02
    9    0.18904E-02   0.14398E-02   0.15791E-02   0.11462E-02
   10     0.19970E-03   0.13677E-02   0.42719E-03   0.12212E-02
   11     0.17166E-04   0.13535E-02   0.92459E-04   0.12414E-02
   19     0.36363E-13   0.13706E-02   0.27375E-09   0.12028E-02
   20     0.30028E-14   0.17185E-02   0.47656E-10   0.93936E-03


                 ****   MOLE-Y-PROFILE     ****
  STAGE     C3H8         N2           C2H4         C3H6         CH4
```

```
  1   0.26562E-11  0.21261E-01  0.72487E-02  0.22288E-10  0.25981
  2   0.13375E-07  0.71006E-02  0.39828      0.68259E-07  0.42238
  4   0.32873E-05  0.39461E-02  0.79407      0.99057E-05  0.72777E-01
  5   0.29753E-04  0.38918E-02  0.80976      0.70509E-04  0.58272E-01
  6   0.26148E-03  0.38855E-02  0.81175      0.48798E-03  0.55537E-01
  7   0.22681E-02  0.39130E-02  0.80582      0.33353E-02  0.55328E-01
  8   0.18587E-01  0.40962E-02  0.76265      0.21595E-01  0.57192E-01
  9   0.11009      0.47186E-02  0.56189      0.10257      0.63605E-01
 10   0.13253      0.10914E-02  0.67435      0.12272      0.46920E-01
 11   0.13771      0.21258E-03  0.70232      0.12737      0.27161E-01
 19   0.16177      0.30406E-09  0.68600      0.14961      0.15557E-03
 20   0.22127      0.49524E-10  0.57808      0.19830      0.59509E-04

              ****   MOLE-Y-PROFILE    ****
STAGE     H2          H2O          CO           CO2
  1   0.55091      0.00000E+00  0.16077      0.21584E-05
  2   0.11179      0.00000E+00  0.60172E-01  0.27396E-03
  4   0.98170E-01  0.17306E-12  0.30104E-01  0.91599E-03
  5   0.97197E-01  0.10236E-10  0.29638E-01  0.11427E-02
  6   0.97104E-01  0.58033E-09  0.29582E-01  0.13973E-02
  7   0.97854E-01  0.32518E-07  0.29788E-01  0.16919E-02
  8   0.10275      0.17609E-05  0.31164E-01  0.19583E-02
  9   0.11945      0.72549E-04  0.35837E-01  0.17659E-02
 10   0.10984E-01  0.99436E-04  0.91752E-02  0.21413E-02
 11   0.91798E-03  0.10602E-03  0.19637E-02  0.22348E-02
 19   0.18782E-11  0.12703E-03  0.58297E-08  0.23362E-02
 20   0.14757E-12  0.21091E-03  0.10274E-08  0.20808E-02

              ****   K-VALUES          ****
STAGE     C3H8         N2           C2H4         C3H6         CH4
  1   0.15969E-03  5.8237       0.14687E-01  0.26256E-03  0.56238
  2   0.34941E-01  15.345       0.45641      0.46238E-01  3.5228
  4   0.91290E-01  16.477       0.81151      0.11608      4.5797
  5   0.94034E-01  16.498       0.82567      0.11941      4.6140
  6   0.95114E-01  16.589       0.83241      0.12071      4.6422
  7   0.99585E-01  17.211       0.86376      0.12604      4.8022
  8   0.13262      20.308       1.0691       0.16538      5.7022
  9   0.29018      25.120       1.7941       0.34996      7.8710
 10   0.36114      23.602       1.9887       0.43094      7.9337
 11   0.37758      23.305       2.0308       0.44959      7.9469
 19   0.42307      23.296       2.1670       0.50126      8.1603
 20   0.51375      23.464       2.4277       0.60360      8.5872

              ****   K-VALUES          ****
STAGE     H2          H2O          CO           CO2
  1    114.71      0.21845E-05  4.5079       0.63449E-02
  2    67.859      0.34807E-02  13.273       0.33769
  4    54.855      0.13970E-01  14.588       0.66254
  5    54.442      0.14576E-01  14.618       0.67601
  6    54.608      0.14724E-01  14.701       0.68154
  7    56.240      0.15070E-01  15.256       0.70518
  8    62.610      0.19066E-01  18.061       0.87133
  9    63.186      0.50388E-01  22.695       1.5407
 10    55.005      0.72703E-01  21.478       1.7535
 11    53.476      0.78328E-01  21.239       1.8003
 19    51.651      0.92681E-01  21.296       1.9423
 20    49.143      0.12273      21.559       2.2151

              ****   MASS-X-PROFILE    ****
STAGE     C3H8         N2           C2H4         C3H6         CH4
  1   0.32768E-07  0.45689E-02  0.61858      0.15958E-06  0.33112
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  2   0.63496E-06  0.48763E-03  0.92091      0.23369E-05  0.72358E-01
  4   0.57036E-04  0.24099E-03  0.98602      0.12899E-03  0.91573E-02
  5   0.50016E-03  0.23689E-03  0.98628      0.89073E-03  0.72632E-02
  6   0.43300E-02  0.23435E-03  0.97714      0.60762E-02  0.68553E-02
  7   0.35062E-01  0.22234E-03  0.91367      0.38874E-01  0.64525E-02
  8   0.19310      0.17655E-03  0.62526      0.17169      0.50274E-02
  9   0.43899      0.13808E-03  0.23056      0.32365      0.34019E-02
 10    0.42736      0.34212E-04  0.25124      0.31647      0.25057E-02
 11    0.42498      0.67522E-05  0.25637      0.31503      0.14489E-02
 19    0.43933      0.95268E-11  0.23140      0.32725      0.79686E-05
 20    0.47998      0.14942E-11  0.16882      0.34937      0.28097E-05

             ****  MASS-X-PROFILE    ****
STAGE    H2          H2O         CO          CO2
  1   0.43251E-03  0.00000E+00  0.44629E-01  0.66884E-03
  2   0.12492E-03  0.19479E-14  0.47766E-02  0.13431E-02
  4   0.12958E-03  0.80163E-11  0.20762E-02  0.21855E-02
  5   0.12901E-03  0.45350E-09  0.20359E-02  0.26668E-02
  6   0.12803E-03  0.25361E-07  0.20132E-02  0.32229E-02
  7   0.12245E-03  0.13571E-05  0.19093E-02  0.36862E-02
  8   0.10337E-03  0.51987E-04  0.15101E-02  0.30905E-02
  9   0.10000E-03  0.68065E-03  0.11607E-02  0.13237E-02
 10    0.10632E-04  0.65072E-03  0.31602E-03  0.14194E-02
 11    0.91443E-06  0.64432E-03  0.68435E-04  0.14437E-02
 19    0.19100E-14  0.64334E-03  0.19979E-09  0.13792E-02
 20    0.15298E-15  0.78239E-03  0.33735E-10  0.10448E-02

             ****  MASS-Y-PROFILE    ****
STAGE    C3H8        N2          C2H4        C3H6        CH4
  1   0.11070E-10  0.56289E-01  0.19219E-01  0.88638E-10  0.39393
  2   0.29385E-07  0.99102E-02  0.55668      0.14311E-06  0.33761
  4   0.58838E-05  0.44869E-02  0.90421      0.16919E-04  0.47390E-01
  5   0.52815E-04  0.43888E-02  0.91448      0.11944E-03  0.37633E-01
  6   0.46325E-03  0.43730E-02  0.91492      0.82500E-03  0.35796E-01
  7   0.40087E-02  0.43935E-02  0.90606      0.56253E-02  0.35575E-01
  8   0.32367E-01  0.45315E-02  0.84492      0.35887E-01  0.36234E-01
  9   0.17711      0.48224E-02  0.57508      0.15747      0.37227E-01
 10    0.18800      0.98360E-03  0.60860      0.16613      0.24216E-01
 11    0.19136      0.18765E-03  0.62087      0.16890      0.13731E-01
 19    0.21761      0.25983E-09  0.58706      0.19205      0.76131E-04
 20    0.28352      0.40312E-10  0.47122      0.24246      0.27740E-04

             ****  MASS-Y-PROFILE    ****
STAGE    H2          H2O         CO          CO2
  1   0.10496      0.00000E+00  0.42559      0.89774E-05
  2   0.11227E-01  0.00000E+00  0.83972E-01  0.60069E-03
  4   0.80326E-02  0.12655E-12  0.34226E-01  0.16363E-02
  5   0.78875E-02  0.74231E-11  0.33419E-01  0.20244E-02
  6   0.78645E-02  0.42004E-09  0.33290E-01  0.24707E-02
  7   0.79062E-02  0.23479E-07  0.33442E-01  0.29844E-02
  8   0.81800E-02  0.12528E-05  0.34473E-01  0.34036E-02
  9   0.87850E-02  0.47683E-04  0.36622E-01  0.28354E-02
 10    0.71235E-03  0.57629E-04  0.82679E-02  0.30318E-02
 11    0.58314E-04  0.60184E-04  0.17333E-02  0.30993E-02
 19    0.11550E-12  0.69808E-04  0.49812E-08  0.31364E-02
 20    0.86436E-14  0.11040E-03  0.83619E-09  0.26609E-02



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              ***** HYDRAULIC PARAMETERS *****
              *******************************


     *** DEFINITIONS ***

      MARANGONI INDEX = SIGMA - SIGMATO
      FLOW PARAM = (ML/MV)*SQRT(RHOV/RHOL)
      QR = QV*SQRT(RHOV/(RHOL-RHOV))
      F FACTOR = QV*SQRT(RHOV)
        WHERE:
        SIGMA IS THE SURFACE TENSION OF LIQUID FROM THE STAGE
        SIGMATO IS THE SURFACE TENSION OF LIQUID TO THE STAGE
        ML IS THE MASS FLOW OF LIQUID FROM THE STAGE
        MV IS THE MASS FLOW OF VAPOR TO THE STAGE
        RHOL IS THE MASS DENSITY OF LIQUID FROM THE STAGE
        RHOV IS THE MASS DENSITY OF VAPOR TO THE STAGE
        QV IS THE VOLUMETRIC FLOW RATE OF VAPOR TO THE STAGE


              TEMPERATURE
                 C
STAGE       LIQUID FROM       VAPOR TO
  1       -131.44          -67.400
  2       -67.400          -51.158
  4       -47.815          -47.138
  5       -47.138          -46.810
  6       -46.810          -45.298
  7       -45.298          -36.418
  8       -36.418          -11.228
  9       -11.228           23.802
 10        -4.4281          -2.9161
 11        -2.9161          -2.3618
 19         1.4536           9.5163
 20         9.5163           9.5163


          MASS FLOW             VOLUME FLOW        MOLECULAR WEIGHT
          KG/HR                 CUM/HR
STAGE LIQUID FROM  VAPOR TO   LIQUID FROM  VAPOR TO   LIQUID FROM  VAPOR TO
  1  45806.    51081.    91.156     2536.1     22.383     20.071
  2  59283.    64558.    128.31     2845.4     26.584     23.660
  4  66008.    71283.    149.35     3011.4     27.840     24.841
  5  66215.    71491.    150.01     3018.9     27.896     24.890
  6  65823.    71099.    148.85     3025.6     27.997     24.950
  7  63362.    68637.    141.45     3043.0     28.645     25.322
  8  58440.    63715.    124.31     2958.1     32.005     27.410
  9  56187.    61463.    116.63     2891.4     38.109     31.152
 10   59069.    10769.    126.53     421.00     37.865     31.734
 11   59735.    11436.    128.80     443.64     37.844     31.957
 19   60902.    12603.    132.26     469.82     38.380     34.415
 20   48300.    12603.    105.97     469.82     39.569     34.415


          DENSITY               VISCOSITY        SURFACE TENSION
          KG/CUM                CP               DYNE/CM
STAGE LIQUID FROM  VAPOR TO    LIQUID FROM  VAPOR TO       LIQUID FROM
  1  502.50     20.142     0.13447      0.84466E-02   14.281
  2  462.03     22.689     0.97059E-01  0.86153E-02    8.7354
  4  441.96     23.671     0.86187E-01  0.86685E-02    6.7259
  5  441.39     23.681     0.85879E-01  0.86770E-02    6.6565
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  6  442.20       23.499      0.86141E-01  0.87204E-02    6.6716
  7  447.94       22.556      0.88349E-01  0.89623E-02    6.8618
  8  470.10       21.540      0.98510E-01  0.95095E-02    7.7388
  9  481.76       21.257      0.10657      0.10080E-01    8.4143
 10   466.85       25.580      0.98158E-01  0.94588E-02    7.3843
 11   463.79       25.777      0.96570E-01  0.94570E-02    7.1710
 19   460.49       26.824      0.95084E-01  0.95824E-02    6.9703
 20   455.80       26.824      0.93374E-01  0.95824E-02    6.8167


    MARANGONI INDEX   FLOW PARAM         QR         REDUCED F-FACTOR
STAGE    DYNE/CM                    CUM/HR       (GM-L)**.5/MIN
  1                0.17953          518.24       0.18970E+06
  2   -5.5456          0.20349          646.61          0.22589E+06
  4   -.37695          0.21430          716.37          0.24419E+06
  5   -.69390E-01       0.21453          718.81          0.24485E+06
  6   0.15045E-01       0.21342          716.78          0.24445E+06
  7   0.19021          0.20715          700.71          0.24087E+06
  8   0.87696          0.19633          648.21          0.22881E+06
  9   0.67550          0.19203          621.21          0.22218E+06
 10   -1.0300           1.2839          101.36           35488.
 11   -.21331           1.2315          107.62           37540.
 19   -.39655E-01       1.1663           116.85           40555.
 20   -.15356          0.92973          117.48           40555.



               ***************************************
               ***** PACKING SIZING CALCULATIONS *****
               ***************************************


   *******************
   *** SECTION   1 ***
   *******************

   STARTING STAGE NUMBER                                            2
   ENDING STAGE NUMBER                                             19
   CAPACITY CALCULATION METHOD                                  VENDOR
   PRESSURE DROP CALCULATION METHOD                             VENDOR
   LIQUID HOLDUP CALCULATION METHOD                             VENDOR
   PRESSURE PROFILE UPDATED                                        NO

   DESIGN PARAMETERS
   -----------------
   OVERDESIGN FACTOR                                          1.00000
   SYSTEM FOAMING FACTOR                                      1.00000
   FRAC. APP. TO MAXIMUM CAPACITY                             1.00000
   MAXIMUM CAPACITY FACTOR        M/SEC                       MISSING
   DESIGN CAPACITY FACTOR         M/SEC                       MISSING
   PRESSURE DROP FOR THE SECTION   BAR                        MISSING
   PRESSURE DROP PER UNIT HEIGHT   MM-WATER/M                 MISSING

   PACKING SPECIFICATIONS
   ----------------------
   PACKING TYPE                                         PALL-RING
   PACKING MATERIAL                                       METAL
   PACKING SIZE                                           50-MM
   VENDOR                                               RASCHIG
   PACKING SURFACE AREA           SQCM/CC               1.05000
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 PACKING VOID FRACTION                        0.96000
 HETP                   METER                 0.85000
 PACKING HEIGHT              METER            15.3000


               ***** SIZING RESULTS *****

 COLUMN DIAMETER             METER                  2.15301
 MAXIMUM FRACTIONAL CAPACITY                        1.00000
 MAXIMUM CAPACITY FACTOR       M/SEC                0.054844
 PRESSURE DROP FOR THE SECTION  BAR                 0.0052721
 AVERAGE PRESSURE DROP/HEIGHT  MM-WATER/M   3.51378
 MAXIMUM LIQUID HOLDUP/STAGE    CUM                 0.14234


         **** RATING PROFILES AT MAXIMUM COLUMN DIAMETER ****

   HEIGHT
   FROM TOP   FRACTIONAL  PRESSURE   PRESSURE    LIQUID
STAGE  OF SECTION  CAPACITY   DROP      DROP/HEIGHT  HOLDUP     HETP
   METER              BAR       MM-WATER/M   CUM       METER
  2   0.0000E+00  0.8974   0.55373E-03  6.6429     0.1292     0.8500
  3   0.8500      0.9770   0.66984E-03  8.0358     0.1393     0.8500
  4    1.700      0.9967   0.70300E-03  8.4337     0.1419     0.8500
  5    2.550      1.000    0.70908E-03  8.5066     0.1423     0.8500
  6    3.400      0.9959   0.70378E-03  8.4430     0.1416     0.8500
  7    4.250      0.9666   0.66496E-03  7.9773     0.1369     0.8500
  8    5.100      0.8836   0.56537E-03  6.7826     0.1265     0.8500
  9    5.950      0.8380   0.52178E-03  6.2597     0.1223     0.8500
 10    6.800      0.3483   0.15217E-04 0.18255     0.1250     0.8500
 11    7.650      0.3592   0.17070E-04 0.20479     0.1263     0.8500
 12    8.500      0.3630   0.17755E-04 0.21300     0.1268     0.8500
 13    9.350      0.3648   0.18088E-04 0.21700     0.1270     0.8500
 14    10.20      0.3658   0.18266E-04 0.21913     0.1271     0.8500
 15    11.05      0.3663   0.18366E-04 0.22032     0.1272     0.8500
 16    11.90      0.3666   0.18436E-04 0.22117     0.1272     0.8500
 17    12.75      0.3671   0.18538E-04 0.22239     0.1273     0.8500
 18    13.60      0.3686   0.18846E-04 0.22609     0.1275     0.8500
 19    14.45      0.3743   0.20015E-04 0.24011     0.1284     0.8500
```

- **BLOCK: T302   MODEL: RADFRAC**

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--------------------------------
 INLETS   - 305      STAGE  14
 OUTLETS  - 306      STAGE   1
        307     STAGE  24
PROPERTY OPTION SET:   RK-ASPEN  REDLICH-KWONG-ASPEN EQUATION OF STATE

            ***  MASS AND ENERGY BALANCE  ***
                    IN            OUT        RELATIVE DIFF.
 TOTAL BALANCE
   MOLE(KMOL/HR )          1220.64        1220.64       0.000000E+00
   MASS(KG/HR   )       48299.8        48299.8      -0.687679E-12
   ENTHALPY(MMKCAL/H)      -12.4911       -11.9362      -0.444288E-01


              *********************
              ****  INPUT DATA  ****
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                **********************

 ****   INPUT PARAMETERS   ****

  NUMBER OF STAGES                                          24
  ALGORITHM OPTION                                          STANDARD
  ABSORBER OPTION                                            NO
  INITIALIZATION OPTION                                     STANDARD
  HYDRAULIC PARAMETER CALCULATIONS                          NO
  INSIDE LOOP CONVERGENCE METHOD                            BROYDEN
  DESIGN SPECIFICATION METHOD                               NESTED
  MAXIMUM NO. OF OUTSIDE LOOP ITERATIONS                    25
  MAXIMUM NO. OF INSIDE LOOP ITERATIONS                     10
  MAXIMUM NUMBER OF FLASH ITERATIONS                        50
  FLASH TOLERANCE                                           0.000100000
  OUTSIDE LOOP CONVERGENCE TOLERANCE                        0.000100000

 ****   COL-SPECS   ****

  MOLAR VAPOR DIST / TOTAL DIST                    0.0
  MOLAR REFLUX RATIO                           0.36600
  DISTILLATE TO FEED RATIO                       0.24000

 ****    PROFILES   ****

  P-SPEC         STAGE   1  PRES, BAR              15.0000

                *******************
                ****  RESULTS  ****
                *******************


 ***   COMPONENT SPLIT FRACTIONS   ***

                  OUTLET STREAMS
                  --------------
            306         307
  COMPONENT:
  C3H8        .30289E-03   .99970
  N2         1.0000       .50283E-12
  C2H4        .99935       .64987E-03
  C3H6        .29351E-02   .99706
  CH4        1.0000       .61080E-08
  H2         1.0000       .94838E-15
  H2O         .14964E-11   1.0000
  CO         1.0000       .10715E-11
  CO2         .99897       .10339E-02


 ***   SUMMARY OF KEY RESULTS   ***

  TOP STAGE TEMPERATURE          C                        -39.0055
  BOTTOM STAGE TEMPERATURE       C                         39.8810
  TOP STAGE LIQUID FLOW          KMOL/HR                   1,102.66
  BOTTOM STAGE LIQUID FLOW       KMOL/HR                   927.688
  TOP STAGE VAPOR FLOW           KMOL/HR                     0.0
  BOTTOM STAGE VAPOR FLOW        KMOL/HR                   1,234.80
  MOLAR REFLUX RATIO                                       3.76394
  MOLAR BOILUP RATIO                                       1.33105
  CONDENSER DUTY (W/O SUBCOOL)   MMKCAL/H                 -3.38914
  REBOILER DUTY                  MMKCAL/H                  3.94396
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 ****   MANIPULATED VARIABLES   ****

                                   BOUNDS           CALCULATED
                             LOWER        UPPER        VALUE
 MOLAR REFLUX RATIO                 0.10000       15.000       3.7639

 ****    DESIGN SPECIFICATIONS   ****

NO  SPEC-TYPE    QUALIFIERS            UNIT      SPECIFIED    CALCULATED
                                    VALUE        VALUE
 1  MOLE-FRAC    STREAMS: 306                     0.99150       0.99150
           COMPS:   C2H4
           BASE-COMPS:   C3H8
                    O2
                    N2
                    C2H4
                    C3H6
                    CH4
                    H2
                    H2O
                    CO
                    CO2
                    MDEA+

 ****   MAXIMUM FINAL RELATIVE ERRORS   ****

  DEW POINT                    0.63275E-07  STAGE= 14
  BUBBLE POINT                   0.36567E-06  STAGE=  6
  COMPONENT MASS BALANCE         0.19625E-06  STAGE= 14 COMP=CH4
  ENERGY BALANCE                 0.29784E-07  STAGE=  5


 ****    PROFILES   ****

 **NOTE** REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE FLOWS
        FROM THE STAGE EXCLUDING ANY SIDE PRODUCT.  FOR THE FIRST STAGE,
        THE REPORTED VAPOR FLOW IS THE VAPOR DISTILLATE FLOW.  FOR THE
        LAST STAGE, THE REPORTED LIQUID FLOW IS THE LIQUID BOTTOMS FLOW.

                             ENTHALPY
STAGE TEMPERATURE  PRESSURE           KCAL/MOL          HEAT DUTY
    C           BAR          LIQUID      VAPOR       MMKCAL/H

  1  -39.006       15.000       8.7603      11.282        -3.3891
  2  -38.142       15.000       8.3896      11.189
  3  -34.431       15.000       7.0230      10.956
  5  -3.8103       15.000      -0.84896      7.8238
  6   10.247       15.000      -3.4789      5.0304
  7   16.704       15.000      -5.0012      3.1353
 12   21.462       15.000      -9.8613     -0.87630
 13   21.739       15.000      -10.822     -1.5585
 14   22.034       15.000      -11.814     -2.2491
 15   28.582       15.000      -12.804     -4.9781
 22   39.487       15.000      -14.699     -10.441
 23   39.654       15.000      -15.035     -10.802
 24   39.881       15.000      -15.633     -11.392        3.9439

STAGE     FLOW RATE                FEED RATE              PRODUCT RATE
        KMOL/HR                 KMOL/HR                KMOL/HR
```

```
    LIQUID    VAPOR       LIQUID    VAPOR    MIXED      LIQUID    VAPOR
  1  1103.    0.0000E+00                      292.9542
  2  1070.      1396.
  3  970.3      1363.
  5  762.3      1121.
  6  761.5      1055.
  7  768.5      1054.
 12  772.3      1066.
 13  771.5      1065.          .40890-01
 14  2023.      1064.    1220.6016
 15  2066.      1095.
 22  2162.      1233.
 23  2162.      1234.
 24  927.7      1235.                      927.6883

  ****  MASS FLOW PROFILES  ****

STAGE     FLOW RATE                FEED RATE              PRODUCT RATE
          KG/HR                KG/HR               KG/HR
    LIQUID    VAPOR       LIQUID    VAPOR    MIXED      LIQUID    VAPOR
  1 0.3107E+05 0.0000E+00                          8255.7057
  2 0.3050E+05 0.3933E+05
  3 0.2896E+05 0.3876E+05
  5 0.2843E+05 0.3586E+05
  6 0.3002E+05 0.3668E+05
  7 0.3096E+05 0.3828E+05
 12 0.3170E+05 0.3992E+05
 13 0.3172E+05 0.3996E+05              1.4072
 14 0.8327E+05 0.3997E+05  .48298+05
 15 0.8660E+05 0.4322E+05
 22 0.9323E+05 0.5307E+05
 23 0.9330E+05 0.5319E+05
 24 0.4004E+05 0.5326E+05                           .40044+05

                ****   MOLE-X-PROFILE     ****
 STAGE     C3H8         N2          C2H4        C3H6         CH4
    1    0.54356E-03   0.87942E-11   0.99150      0.40176E-02   0.28875E-04
    2    0.38956E-02   0.54437E-12   0.96849      0.23115E-01   0.58232E-05
    3    0.20524E-01   0.13266E-12   0.87562      0.99205E-01   0.20184E-05
    5    0.14590      0.10192E-12   0.36284      0.48914      0.10752E-05
    6    0.20206      0.10578E-12   0.21851      0.57827      0.10117E-05
    7    0.24193      0.10669E-12   0.16250      0.59479      0.97730E-06
   12    0.39946      0.10776E-12   0.13101      0.46888      0.95250E-06
   13    0.43126      0.10807E-12   0.13055      0.43732      0.95305E-06
   14    0.46266      0.10814E-12   0.13006      0.40553      0.95245E-06
   15    0.48660      0.88317E-14   0.79403E-01   0.43245      0.19079E-06
   22    0.53735      0.00000E+00   0.98859E-03   0.46044      0.17246E-11
   23    0.54830      0.00000E+00   0.47474E-03   0.44985      0.32009E-12
   24    0.56654      0.00000E+00   0.20361E-03   0.43099      0.55695E-13

                ****   MOLE-X-PROFILE     ****
 STAGE     H2          H2O          CO          CO2
    1    0.12512E-13   0.10715E-13   0.19857E-09   0.39099E-02
    2    0.00000E+00   0.41303E-12   0.13725E-10   0.44960E-02
    3    0.00000E+00   0.11819E-10   0.34143E-11   0.46474E-02
    5    0.00000E+00   0.23222E-08   0.25364E-11   0.21194E-02
    6    0.00000E+00   0.13594E-07   0.26070E-11   0.11680E-02
    7    0.00000E+00   0.63940E-07   0.26184E-11   0.78351E-03
   12    0.00000E+00   0.76524E-04   0.26358E-11   0.56723E-03
   13    0.00000E+00   0.30226E-03   0.26428E-11   0.56428E-03
   14    0.00000E+00   0.11824E-02   0.26441E-11   0.56151E-03
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  15   0.00000E+00  0.11846E-02  0.23231E-12  0.36177E-03
  22   0.00000E+00  0.12154E-02  0.00000E+00  0.58507E-05
  23   0.00000E+00  0.13754E-02  0.00000E+00  0.28948E-05
  24   0.00000E+00  0.22611E-02  0.00000E+00  0.12779E-05

              ****  MOLE-Y-PROFILE   ****
STAGE   C3H8        N2         C2H4        C3H6        CH4
   1  0.74134E-04  0.13916E-09  0.99576     0.68346E-03  0.14055E-03
   2  0.54356E-03  0.87942E-11  0.99150     0.40176E-02  0.28875E-04
   3  0.31750E-02  0.23178E-11  0.97343     0.19010E-01  0.10778E-04
   5  0.53922E-01  0.23758E-11  0.72701     0.21531     0.84941E-05
   6  0.10554      0.25150E-11  0.53737     0.35446     0.87929E-05
   7  0.14607      0.25195E-11  0.43326     0.41873     0.87525E-05
  12   0.26686      0.24953E-11  0.36799     0.36364     0.86275E-05
  13   0.28975      0.24966E-11  0.36766     0.34103     0.86316E-05
  14   0.31273      0.24968E-11  0.36749     0.31807     0.86356E-05
  15   0.37468      0.19973E-12  0.24005     0.38397     0.17592E-05
  22   0.50371      0.00000E+00  0.32831E-02  0.49261     0.16053E-10
  23   0.51542      0.00000E+00  0.15785E-02  0.48257     0.29788E-11
  24   0.53459      0.00000E+00  0.67843E-03  0.46401     0.51873E-12

              ****  MOLE-Y-PROFILE   ****
STAGE   H2          H2O         CO          CO2
   1  0.57908E-12  0.00000E+00  0.28136E-08  0.33412E-02
   2  0.12512E-13  0.10715E-13  0.19857E-09  0.39099E-02
   3  0.28981E-14  0.32655E-12  0.53460E-10  0.43700E-02
   5  0.33108E-14  0.17449E-09  0.53847E-10  0.37484E-02
   6  0.35179E-14  0.16775E-08  0.56958E-10  0.26165E-02
   7  0.35271E-14  0.98173E-08  0.57048E-10  0.19298E-02
  12   0.34956E-14  0.13892E-04  0.56489E-10  0.14890E-02
  13   0.34975E-14  0.55479E-04  0.56520E-10  0.14865E-02
  14   0.34947E-14  0.21908E-03  0.56527E-10  0.14850E-02
  15   0.00000E+00  0.26864E-03  0.48836E-11  0.10360E-02
  22   0.00000E+00  0.37766E-03  0.00000E+00  0.18741E-04
  23   0.00000E+00  0.42950E-03  0.00000E+00  0.92871E-05
  24   0.00000E+00  0.70988E-03  0.00000E+00  0.41096E-05

              ****  K-VALUES        ****
STAGE   C3H8        N2         C2H4        C3H6        CH4
   1  0.13639      15.824      1.0043      0.17012      4.8675
   2  0.13953      16.155      1.0238      0.17381      4.9586
   3  0.15470      17.471      1.1117      0.19162      5.3402
   5  0.36959      23.309      2.0037      0.44018      7.9002
   6  0.52235      23.777      2.4593      0.61297      8.6914
   7  0.60377      23.615      2.6663      0.70400      8.9558
  12   0.66804      23.155      2.8088      0.77555      9.0577
  13   0.67188      23.102      2.8162      0.77983      9.0568
  14   0.67593      23.089      2.8255      0.78433      9.0667
  15   0.76999      22.615      3.0232      0.88788      9.2205
  22   0.93739      21.421      3.3210      1.0699      9.3080
  23   0.94003      21.394      3.3250      1.0727      9.3063
  24   0.94361      21.391      3.3321      1.0766      9.3136

              ****  K-VALUES        ****
STAGE   H2          H2O         CO          CO2
   1   46.284      0.25693E-01  14.170      0.85453
   2   47.098      0.25942E-01  14.467      0.86964
   3   50.048      0.27628E-01  15.658      0.94032
   5   53.837      0.75138E-01  21.230       1.7686
   6   49.656      0.12340      21.848      2.2401
   7   47.184      0.15354      21.787      2.4630
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  12    44.762      0.18154      21.431      2.6251
  13    44.568      0.18355      21.387      2.6344
  14    44.460      0.18529      21.378      2.6447
  15    41.770      0.22677      21.022      2.8638
  22    37.036      0.31073      20.042      3.2033
  23    36.952      0.31228      20.019      3.2082
  24    36.905      0.31395      20.018      3.2158

               ****   MASS-X-PROFILE    ****
STAGE     C3H8         N2          C2H4         C3H6         CH4
  1   0.85055E-03   0.87419E-11   0.98703      0.59993E-02   0.16438E-04
  2   0.60248E-02   0.53485E-12   0.95292      0.34115E-01   0.32765E-05
  3   0.30321E-01   0.12451E-12   0.82297      0.13986       0.10848E-05
  5   0.17253       0.76569E-13   0.27298      0.55199       0.46257E-06
  6   0.22600       0.75159E-13   0.15548      0.61722       0.41167E-06
  7   0.26479       0.74182E-13   0.11314      0.62121       0.38914E-06
 12    0.42914       0.73544E-13   0.89542E-01   0.48068       0.37227E-06
 13    0.46256       0.73637E-13   0.89085E-01   0.44762       0.37189E-06
 14    0.49565       0.73595E-13   0.88644E-01   0.41459       0.37122E-06
 15    0.51186       0.59019E-14   0.53138E-01   0.43411       0.73015E-07
 22    0.54951       0.00000E+00   0.64316E-03   0.44933       0.64164E-12
 23    0.56038       0.00000E+00   0.30868E-03   0.43874       0.11902E-12
 24    0.57876       0.00000E+00   0.13233E-03   0.42016       0.20700E-13

               ****   MASS-X-PROFILE    ****
STAGE     H2          H2O          CO           CO2
  1   0.89500E-15   0.68498E-14   0.19737E-09   0.61061E-02
  2   0.00000E+00   0.26097E-12   0.13484E-10   0.69398E-02
  3   0.00000E+00   0.71336E-11   0.32040E-11   0.68522E-02
  5   0.00000E+00   0.11219E-08   0.19052E-11   0.25014E-02
  6   0.00000E+00   0.62117E-08   0.18522E-11   0.13038E-02
  7   0.00000E+00   0.28590E-07   0.18204E-11   0.85584E-03
 12    0.00000E+00   0.33586E-04   0.17987E-11   0.60816E-03
 13    0.00000E+00   0.13245E-03   0.18005E-11   0.60405E-03
 14    0.00000E+00   0.51748E-03   0.17993E-11   0.60036E-03
 15    0.00000E+00   0.50910E-03   0.15523E-12   0.37980E-03
 22    0.00000E+00   0.50777E-03   0.00000E+00   0.59713E-05
 23    0.00000E+00   0.57427E-03   0.00000E+00   0.29528E-05
 24    0.00000E+00   0.94370E-03   0.00000E+00   0.13029E-05

               ****   MASS-Y-PROFILE    ****
STAGE     C3H8         N2          C2H4         C3H6         CH4
  1   0.11627E-03   0.13865E-09   0.99355      0.10229E-02   0.80196E-04
  2   0.85055E-03   0.87419E-11   0.98703      0.59993E-02   0.16438E-04
  3   0.49227E-02   0.22829E-11   0.96018      0.28126E-01   0.60799E-05
  5   0.74309E-01   0.20799E-11   0.63738      0.28315       0.42586E-05
  6   0.13389       0.20269E-11   0.43369      0.42911       0.40581E-05
  7   0.17744       0.19443E-11   0.33482      0.48540       0.38680E-05
 12    0.31414       0.18661E-11   0.27560      0.40850       0.36949E-05
 13    0.34064       0.18646E-11   0.27498      0.38260       0.36918E-05
 14    0.36721       0.18625E-11   0.27453      0.35641       0.36891E-05
 15    0.41865       0.14177E-12   0.17064      0.40942       0.71512E-06
 22    0.51606       0.00000E+00   0.21399E-02   0.48162       0.59834E-11
 23    0.52749       0.00000E+00   0.10278E-02   0.47129       0.11091E-11
 24    0.54655       0.00000E+00   0.44127E-03   0.45271       0.19294E-12

               ****   MASS-Y-PROFILE    ****
STAGE     H2          H2O          CO           CO2
  1   0.41519E-13   0.00000E+00   0.28031E-08   0.52299E-02
  2   0.89500E-15   0.68498E-14   0.19737E-09   0.61061E-02
  3   0.20542E-15   0.20684E-12   0.52651E-10   0.67622E-02
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    5   0.20857E-15   0.98237E-10   0.47136E-10   0.51554E-02
    6   0.20401E-15   0.86941E-09   0.45898E-10   0.33127E-02
    7   0.19587E-15   0.48720E-08   0.44019E-10   0.23395E-02
   12   0.18812E-15   0.66813E-05   0.42240E-10   0.17494E-02
   13   0.18797E-15   0.26646E-04   0.42207E-10   0.17442E-02
   14   0.18760E-15   0.10510E-03   0.42162E-10   0.17403E-02
   15   0.00000E+00   0.12263E-03   0.34662E-11   0.11553E-02
   22   0.00000E+00   0.15807E-03   0.00000E+00   0.19163E-04
   23   0.00000E+00   0.17958E-03   0.00000E+00   0.94859E-05
   24   0.00000E+00   0.29650E-03   0.00000E+00   0.41932E-05



               *******************************
               ***** HYDRAULIC PARAMETERS *****
               *******************************


    *** DEFINITIONS ***

     MARANGONI INDEX = SIGMA - SIGMATO
     FLOW PARAM = (ML/MV)*SQRT(RHOV/RHOL)
     QR = QV*SQRT(RHOV/(RHOL-RHOV))
     F FACTOR = QV*SQRT(RHOV)
       WHERE:
       SIGMA IS THE SURFACE TENSION OF LIQUID FROM THE STAGE
       SIGMATO IS THE SURFACE TENSION OF LIQUID TO THE STAGE
       ML IS THE MASS FLOW OF LIQUID FROM THE STAGE
       MV IS THE MASS FLOW OF VAPOR TO THE STAGE
       RHOL IS THE MASS DENSITY OF LIQUID FROM THE STAGE
       RHOV IS THE MASS DENSITY OF VAPOR TO THE STAGE
       QV IS THE VOLUMETRIC FLOW RATE OF VAPOR TO THE STAGE


               TEMPERATURE
                 C
STAGE      LIQUID FROM      VAPOR TO
  1       -39.006         -38.142
  2       -38.142         -34.431
  3       -34.431         -22.714
  5       -3.8103          10.247
  6        10.247          16.704
  7        16.704          19.256
 12         21.462          21.739
 13         21.739          22.033
 14         22.034          28.582
 15         28.582          33.141
 22         39.487          39.654
 23         39.654          39.881
 24         39.881          39.881


        MASS FLOW            VOLUME FLOW        MOLECULAR WEIGHT
         KG/HR             CUM/HR
STAGE LIQUID FROM  VAPOR TO  LIQUID FROM  VAPOR TO  LIQUID FROM  VAPOR TO
  1  39330.    39330.    92.255    1439.6    28.181    28.181
  2  30504.    38759.    70.985    1440.5    28.512    28.441
  3  28961.    37217.    65.600    1428.4    29.849    29.462
  5  28425.    36681.    60.891    1353.8    37.289    34.761
  6  30024.    38280.    65.303    1373.0    39.425    36.301
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  7  30964.     39220.     68.141     1389.1     40.291     36.948
 12   31700.     39956.     70.855     1401.6     41.047     37.509
 13   31719.     39974.     71.024     1401.8     41.113     37.554
 14   83268.     43224.     186.73     1454.6     41.162     39.464
 15   86596.     46552.     197.06     1516.7     41.920     40.905
 22   93233.     53189.     218.43     1647.8     43.121     43.087
 23   93303.     53259.     218.77     1648.9     43.146     43.132
 24   40044.     53259.     93.977     1648.9     43.165     43.132


          DENSITY               VISCOSITY        SURFACE TENSION
          KG/CUM                CP              DYNE/CM
STAGE LIQUID FROM  VAPOR TO    LIQUID FROM  VAPOR TO       LIQUID FROM
  1  426.32      27.320     0.79657E-01  0.87518E-02   5.6008
  2  429.72      26.907     0.80715E-01  0.88425E-02   5.7009
  3  441.48      26.054     0.84768E-01  0.91053E-02   6.0842
  5  466.82      27.094     0.95935E-01  0.95973E-02   7.1105
  6  459.76      27.880     0.93297E-01  0.96431E-02   6.8369
  7  454.41      28.235     0.91329E-01  0.96559E-02   6.6429
 12   447.39      28.507     0.89391E-01  0.96580E-02    6.3749
 13   446.59      28.516     0.89278E-01  0.96582E-02    6.3618
 14   445.94      29.716     0.89339E-01  0.96764E-02    6.3904
 15   439.45      30.694     0.86943E-01  0.96822E-02    6.0332
 22   426.84      32.279     0.82674E-01  0.96808E-02    5.2695
 23   426.49      32.299     0.82622E-01  0.96798E-02    5.2641
 24   426.11      32.299     0.82710E-01  0.96798E-02    5.2980


    MARANGONI INDEX   FLOW PARAM         QR         REDUCED F-FACTOR
STAGE    DYNE/CM                     CUM/HR       (GM-L)**.5/MIN
  1                0.25315         376.70        0.12541E+06
  2  0.10014          0.19693          372.30        0.12454E+06
  3  0.38327          0.18904          357.73        0.12152E+06
  5  0.24154          0.18669          336.06        0.11745E+06
  6  -.27369          0.19314          348.85        0.12083E+06
  7  -.19399          0.19680          357.54        0.12302E+06
 12   -.24455E-01      0.20027          365.64        0.12472E+06
 13   -.13079E-01      0.20050          366.11        0.12476E+06
 14   -.27115          0.49729          388.66        0.13215E+06
 15   -.35717          0.49162          415.61        0.14004E+06
 22   -.14371E-01      0.48203          471.31        0.15603E+06
 23   -.54286E-02      0.48211          472.01        0.15619E+06
 24   0.33986E-01      0.20701          472.23        0.15619E+06




            ****************************************
            ***** PACKING SIZING CALCULATIONS *****
            ****************************************


  *******************
  *** SECTION   1 ***
  *******************


  STARTING STAGE NUMBER                              2
  ENDING STAGE NUMBER                              23
  CAPACITY CALCULATION METHOD                        VENDOR
  PRESSURE DROP CALCULATION METHOD                    VENDOR
  LIQUID HOLDUP CALCULATION METHOD                    VENDOR
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    PRESSURE PROFILE UPDATED                          NO

    DESIGN PARAMETERS
    -----------------
    OVERDESIGN FACTOR                             1.00000
    SYSTEM FOAMING FACTOR                             1.00000
    FRAC. APP. TO MAXIMUM CAPACITY                        1.00000
    MAXIMUM CAPACITY FACTOR        M/SEC                 MISSING
    DESIGN CAPACITY FACTOR        M/SEC                 MISSING
    PRESSURE DROP FOR THE SECTION   BAR                  MISSING
    PRESSURE DROP PER UNIT HEIGHT   MM-WATER/M              MISSING

    PACKING SPECIFICATIONS
    ---------------------
    PACKING TYPE                              PALL-RING
    PACKING MATERIAL                              METAL
    PACKING SIZE                            50-MM
    VENDOR                                RASCHIG
    PACKING SURFACE AREA           SQCM/CC              1.05000
    PACKING VOID FRACTION                           0.96000
    HETP                 METER               0.85000
    PACKING HEIGHT                 METER               18.7000


                 ***** SIZING RESULTS *****

    COLUMN DIAMETER              METER                 1.94576
    MAXIMUM FRACTIONAL CAPACITY                          1.00000
    MAXIMUM CAPACITY FACTOR        M/SEC                0.044094
    PRESSURE DROP FOR THE SECTION   BAR                 0.0081824
    AVERAGE PRESSURE DROP/HEIGHT   MM-WATER/M             4.46187
    MAXIMUM LIQUID HOLDUP/STAGE    CUM                  0.16720


           **** RATING PROFILES AT MAXIMUM COLUMN DIAMETER ****

      HEIGHT
      FROM TOP   FRACTIONAL  PRESSURE    PRESSURE     LIQUID
  STAGE  OF SECTION  CAPACITY    DROP      DROP/HEIGHT  HOLDUP      HETP
      METER               BAR       MM-WATER/M    CUM        METER
   2   0.0000E+00  0.6166   0.21921E-03  2.6298     0.7964E-01  0.8500
   3   0.8500      0.5882   0.20521E-03  2.4618     0.7595E-01  0.8500
   4    1.700      0.5503   0.18697E-03  2.2430     0.7207E-01  0.8500
   5    2.550      0.5506   0.18890E-03  2.2662     0.7317E-01  0.8500
   6    3.400      0.5731   0.20308E-03  2.4362     0.7639E-01  0.8500
   7    4.250      0.5881   0.21265E-03  2.5511     0.7840E-01  0.8500
   8    5.100      0.5950   0.21703E-03  2.6037     0.7932E-01  0.8500
   9    5.950      0.5980   0.21890E-03  2.6260     0.7974E-01  0.8500
  10    6.800      0.5997   0.21980E-03  2.6368     0.7999E-01  0.8500
  11    7.650      0.6008   0.22036E-03  2.6436     0.8017E-01  0.8500
  12    8.500      0.6017   0.22078E-03  2.6487     0.8032E-01  0.8500
  13    9.350      0.6024   0.22107E-03  2.6521     0.8045E-01  0.8500
  14    10.20      0.8500   0.39393E-03  4.7258     0.1499      0.8500
  15    11.05      0.8986   0.46044E-03  5.5238     0.1552      0.8500
  16    11.90      0.9373   0.51993E-03  6.2374     0.1595      0.8500
  17    12.75      0.9634   0.56414E-03  6.7678     0.1626      0.8500
  18    13.60      0.9793   0.59291E-03  7.1129     0.1645      0.8500
  19    14.45      0.9884   0.61012E-03  7.3194     0.1657      0.8500
  20    15.30      0.9936   0.62001E-03  7.4381     0.1663      0.8500
  21    16.15      0.9966   0.62577E-03  7.5071     0.1667      0.8500
  22    17.00      0.9986   0.62939E-03  7.5505     0.1670      0.8500
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23   17.85       1.000     0.63178E-03  7.5792      0.1672      0.8500
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- **BLOCK: T303   MODEL: RADFRAC**

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   INLETS   - 309      STAGE  93
   OUTLETS  - 310      STAGE   1
           311      STAGE 156
  PROPERTY OPTION SET:   RK-ASPEN  REDLICH-KWONG-ASPEN EQUATION OF STATE

                 ***  MASS AND ENERGY BALANCE  ***
                             IN            OUT       RELATIVE DIFF.
   TOTAL BALANCE
     MOLE(KMOL/HR )           1215.55        1215.55       0.000000E+00
     MASS(KG/HR   )          52408.6        52408.6       0.132154E-11
     ENTHALPY(MMKCAL/H)      -19.9753       -18.8269      -0.574924E-01




                    **********************
                    ****  INPUT DATA  ****
                    **********************

  ****   INPUT PARAMETERS   ****

   NUMBER OF STAGES                              156
   ALGORITHM OPTION                             STANDARD
   ABSORBER OPTION                             NO
   INITIALIZATION OPTION                         STANDARD
   HYDRAULIC PARAMETER CALCULATIONS                   NO
   INSIDE LOOP CONVERGENCE METHOD                  BROYDEN
   DESIGN SPECIFICATION METHOD                  NESTED
   MAXIMUM NO. OF OUTSIDE LOOP ITERATIONS            200
   MAXIMUM NO. OF INSIDE LOOP ITERATIONS            10
   MAXIMUM NUMBER OF FLASH ITERATIONS               50
   FLASH TOLERANCE                               0.000100000
   OUTSIDE LOOP CONVERGENCE TOLERANCE                   0.000100000

  ****   COL-SPECS   ****

   MOLAR VAPOR DIST / TOTAL DIST                   1.00000
   MOLAR REFLUX RATIO                         21.1350
   DISTILLATE TO FEED RATIO                      0.40100

  ****    PROFILES   ****

   P-SPEC         STAGE   1  PRES, BAR                15.0000

                     *******************
                     ****  RESULTS  ****
                     *******************


  ***   COMPONENT SPLIT FRACTIONS   ***

                      OUTLET STREAMS
                      --------------
              310         311
   COMPONENT:
   C3H8       .25292E-03   .99975
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 C2H4       1.0000      .76830E-32
 C3H6        .83653      .16347
 CH4        1.0000      .21693E-60
 H2        1.0000      .10998E-98
 H2O        .55068E-51   1.0000
 CO2        1.0000      .64856E-31


***   SUMMARY OF KEY RESULTS   ***

 TOP STAGE TEMPERATURE          C                    34.8061
 BOTTOM STAGE TEMPERATURE       C                    42.7030
 TOP STAGE LIQUID FLOW          KMOL/HR              8,606.74
 BOTTOM STAGE LIQUID FLOW       KMOL/HR              808.322
 TOP STAGE VAPOR FLOW           KMOL/HR              407.227
 BOTTOM STAGE VAPOR FLOW        KMOL/HR              9,016.01
 MOLAR REFLUX RATIO                                  21.1350
 MOLAR BOILUP RATIO                                  11.1540
 CONDENSER DUTY (W/O SUBCOOL)   MMKCAL/H             -27.5735
 REBOILER DUTY                  MMKCAL/H             28.7219


****   MANIPULATED VARIABLES   ****

                                 BOUNDS          CALCULATED
                            LOWER       UPPER       VALUE
DISTILLATE TO FEED RATIO           0.10000      1.0000      0.33501

****   DESIGN SPECIFICATIONS   ****

NO  SPEC-TYPE    QUALIFIERS          UNIT      SPECIFIED    CALCULATED
                                     VALUE        VALUE
 1  MOLE-FRAC    STREAMS: 311                  0.90000      0.90000
            COMPS:   C3H8
            BASE-COMPS:   C3H8
                      O2
                      N2
                      C2H4
                      C3H6
                      CH4
                      H2
                      H2O
                      CO
                      CO2
                      MDEA+

****   MAXIMUM FINAL RELATIVE ERRORS   ****

 DEW POINT                    0.77157E-11  STAGE=  1
 BUBBLE POINT                 0.23596E-09  STAGE=  1
 COMPONENT MASS BALANCE       0.11790E-08  STAGE= 93 COMP=H2
 ENERGY BALANCE               0.14340E-08  STAGE=  1


****   PROFILES   ****

**NOTE** REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE FLOWS
      FROM THE STAGE EXCLUDING ANY SIDE PRODUCT.  FOR THE FIRST STAGE,
      THE REPORTED VAPOR FLOW IS THE VAPOR DISTILLATE FLOW.  FOR THE
      LAST STAGE, THE REPORTED LIQUID FLOW IS THE LIQUID BOTTOMS FLOW.
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                          ENTHALPY
STAGE TEMPERATURE   PRESSURE          KCAL/MOL          HEAT DUTY
     C            BAR          LIQUID      VAPOR       MMKCAL/H

  1   34.806       15.000       1.2431      4.4056       -27.5734
  2   35.336       15.000       1.2559      4.4449
  6   35.358       15.000       1.2453      4.4356
  7   35.359       15.000       1.2419      4.4326
  8   35.360       15.000       1.2382      4.4294
 91   39.418       15.000      -14.176     -10.002
 92   39.499       15.000      -14.468     -10.293
 93   39.597       15.000      -14.736     -10.560
 94   39.610       15.000      -14.751     -10.578
152   42.228       15.000      -24.092     -20.407
153   42.328       15.000      -24.433     -20.781
154   42.430       15.000      -24.776     -21.154
155   42.546       15.000      -25.126     -21.526
156   42.703       15.000      -25.511     -21.906       28.7219

STAGE     FLOW RATE               FEED RATE             PRODUCT RATE
         KMOL/HR                 KMOL/HR             KMOL/HR
     LIQUID    VAPOR      LIQUID    VAPOR    MIXED      LIQUID    VAPOR
  1  8607.     407.2                                407.2269
  2  8646.     9014.
  6  8647.     9054.
  7  8647.     9054.
  8  8647.     9054.
 91  8643.     9051.
 92  8643.     9051.              73.3215
 93  9810.     8977.     1142.2271
 94  9811.     9002.
152  9831.     9021.
153  9831.     9022.
154  9830.     9023.
155  9824.     9022.
156  808.3     9016.                         808.3217

  ****  MASS FLOW PROFILES  ****

STAGE     FLOW RATE               FEED RATE             PRODUCT RATE
         KG/HR                  KG/HR              KG/HR
     LIQUID    VAPOR      LIQUID    VAPOR    MIXED      LIQUID    VAPOR
  1 0.3621E+06 0.1698E+05                           .16978+05
  2 0.3638E+06 0.3791E+06
  6 0.3639E+06 0.3809E+06
  7 0.3639E+06 0.3809E+06
  8 0.3639E+06 0.3809E+06
 91 0.3728E+06 0.3896E+06
 92 0.3729E+06 0.3898E+06         3054.4766
 93 0.4234E+06 0.3869E+06  .49354+05
 94 0.4235E+06 0.3880E+06
152 0.4306E+06 0.3949E+06
153 0.4308E+06 0.3951E+06
154 0.4309E+06 0.3954E+06
155 0.4308E+06 0.3955E+06
156 0.3543E+05 0.3954E+06                      .35431+05

                  ****   MOLE-X-PROFILE     ****
  STAGE     C3H8         C2H4         C3H6          CH4          H2
    1    0.52284E-03   0.18496E-03   0.99905      0.66747E-05   0.23199E-03
    2    0.59535E-03   0.62758E-04   0.99932      0.97326E-06   0.16094E-04
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    6    0.95917E-03   0.12138E-04   0.99902       0.33499E-06   0.10823E-04
    7    0.10724E-02   0.11868E-04   0.99890       0.33493E-06   0.10823E-04
    8    0.11961E-02   0.11789E-04   0.99878       0.33493E-06   0.10824E-04
   91    0.52205       0.11406E-04   0.47789       0.34330E-06   0.11847E-04
   92    0.53178       0.11398E-04   0.46809       0.34341E-06   0.11865E-04
   93    0.54044       0.10050E-04   0.45925       0.22906E-06   0.36656E-05
   94    0.54098       0.32977E-05   0.45872       0.26867E-07   0.10835E-06
  152    0.85689       0.00000E+00   0.14277       0.00000E+00   0.00000E+00
  153    0.86835       0.00000E+00   0.13127       0.00000E+00   0.00000E+00
  154    0.87959       0.00000E+00   0.11987       0.00000E+00   0.00000E+00
  155    0.89037       0.00000E+00   0.10859       0.00000E+00   0.00000E+00
  156    0.90000       0.00000E+00   0.97405E-01   0.00000E+00   0.00000E+00

                 ****   MOLE-X-PROFILE    ****
 STAGE     H2O          CO2
    1    0.00000E+00   0.94573E-06
    2    0.00000E+00   0.33487E-06
    6    0.00000E+00   0.63826E-07
    7    0.00000E+00   0.62100E-07
    8    0.00000E+00   0.61567E-07
   91    0.33213E-04   0.58369E-07
   92    0.10181E-03   0.58310E-07
   93    0.29982E-03   0.51735E-07
   94    0.29983E-03   0.17589E-07
  152    0.33510E-03   0.00000E+00
  153    0.38545E-03   0.00000E+00
  154    0.54324E-03   0.00000E+00
  155    0.10386E-02   0.00000E+00
  156    0.25950E-02   0.00000E+00

                 ****   MOLE-Y-PROFILE    ****
 STAGE     C3H8         C2H4          C3H6          CH4           H2
    1    0.45194E-03   0.59866E-03   0.98940       0.63645E-04   0.94871E-02
    2    0.51963E-03   0.20365E-03   0.99862       0.92485E-05   0.65012E-03
    6    0.83752E-03   0.39393E-04   0.99868       0.31829E-05   0.43702E-03
    7    0.93636E-03   0.38517E-04   0.99858       0.31824E-05   0.43702E-03
    8    0.10445E-02   0.38259E-04   0.99848       0.31823E-05   0.43702E-03
   91    0.48879       0.37837E-04   0.51072       0.31914E-05   0.43816E-03
   92    0.49859       0.37829E-04   0.50090       0.31916E-05   0.43818E-03
   93    0.50754       0.33378E-04   0.49220       0.21285E-05   0.13519E-03
   94    0.50815       0.10953E-04   0.49174       0.24963E-06   0.39947E-05
  152    0.84042       0.00000E+00   0.15947       0.00000E+00   0.00000E+00
  153    0.85303       0.00000E+00   0.14684       0.00000E+00   0.00000E+00
  154    0.86551       0.00000E+00   0.13430       0.00000E+00   0.00000E+00
  155    0.87776       0.00000E+00   0.12188       0.00000E+00   0.00000E+00
  156    0.88951       0.00000E+00   0.10959       0.00000E+00   0.00000E+00

                 ****   MOLE-Y-PROFILE    ****
 STAGE     H2O          CO2
    1    0.00000E+00   0.29111E-05
    2    0.00000E+00   0.10345E-05
    6    0.00000E+00   0.19721E-06
    7    0.00000E+00   0.19188E-06
    8    0.00000E+00   0.19023E-06
   91    0.10319E-04   0.18678E-06
   92    0.31719E-04   0.18673E-06
   93    0.93677E-04   0.16581E-06
   94    0.93719E-04   0.56380E-07
  152    0.11492E-03   0.00000E+00
  153    0.13263E-03   0.00000E+00
  154    0.18750E-03   0.00000E+00
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 155   0.35940E-03   0.00000E+00
 156   0.89904E-03   0.00000E+00

               ****   K-VALUES          ****
STAGE     C3H8          C2H4          C3H6          CH4           H2
   1    0.86440       3.2367        0.99033       9.5352        40.894
   2    0.87282       3.2449        0.99929       9.5026        40.395
   6    0.87317       3.2454        0.99966       9.5017        40.378
   7    0.87318       3.2454        0.99968       9.5016        40.377
   8    0.87319       3.2454        0.99970       9.5016        40.376
  91    0.93629       3.3172        1.0687        9.2963        36.986
  92    0.93758       3.3189        1.0701        9.2936        36.932
  93    0.93912       3.3211        1.0718        9.2922        36.880
  94    0.93933       3.3213        1.0720        9.2915        36.870
 152    0.98077       3.3660        1.1169        9.1677        34.950
 153    0.98236       3.3678        1.1186        9.1640        34.886
 154    0.98400       3.3701        1.1204        9.1620        34.832
 155    0.98584       3.3735        1.1224        9.1654        34.809
 156    0.98835       3.3811        1.1251        9.1858        34.882

               ****   K-VALUES          ****
STAGE     H2O           CO2
   1    0.26080       3.0782
   2    0.26555       3.0893
   6    0.26575       3.0899
   7    0.26576       3.0899
   8    0.26577       3.0899
  91    0.31067       3.2001
  92    0.31154       3.2023
  93    0.31244       3.2051
  94    0.31258       3.2053
 152    0.34295       3.2723
 153    0.34408       3.2748
 154    0.34515       3.2777
 155    0.34605       3.2815
 156    0.34644       3.2882

               ****   MASS-X-PROFILE     ****
STAGE     C3H8          C2H4          C3H6          CH4           H2
   1    0.54803E-03   0.12334E-03   0.99931       0.25453E-05   0.11117E-04
   2    0.62388E-03   0.41839E-04   0.99933       0.37105E-06   0.77098E-06
   6    0.10051E-02   0.80920E-05   0.99899       0.12771E-06   0.51847E-06
   7    0.11237E-02   0.79119E-05   0.99887       0.12768E-06   0.51848E-06
   8    0.12534E-02   0.78588E-05   0.99874       0.12768E-06   0.51849E-06
  91    0.53373       0.74188E-05   0.46624       0.12769E-06   0.55369E-06
  92    0.54345       0.74106E-05   0.45650       0.12768E-06   0.55430E-06
  93    0.55213       0.65323E-05   0.44773       0.85138E-07   0.17120E-06
  94    0.55267       0.21433E-05   0.44721       0.99855E-08   0.50601E-08
 152    0.86269       0.00000E+00   0.13717       0.00000E+00   0.00000E+00
 153    0.87379       0.00000E+00   0.12605       0.00000E+00   0.00000E+00
 154    0.88472       0.00000E+00   0.11505       0.00000E+00   0.00000E+00
 155    0.89537       0.00000E+00   0.10420       0.00000E+00   0.00000E+00
 156    0.90542       0.00000E+00   0.93512E-01   0.00000E+00   0.00000E+00

               ****   MASS-X-PROFILE     ****
STAGE     H2O           CO2
   1    0.00000E+00   0.98935E-06
   2    0.00000E+00   0.35023E-06
   6    0.00000E+00   0.66750E-07
   7    0.00000E+00   0.64945E-07
   8    0.00000E+00   0.64386E-07
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  91   0.13873E-04  0.59558E-07
  92   0.42508E-04  0.59472E-07
  93   0.12514E-03  0.52750E-07
  94   0.12514E-03  0.17934E-07
 152   0.13783E-03  0.00000E+00
 153   0.15846E-03  0.00000E+00
 154   0.22323E-03  0.00000E+00
 155   0.42668E-03  0.00000E+00
 156   0.10666E-02  0.00000E+00

                ****   MASS-Y-PROFILE    ****
STAGE    C3H8        C2H4        C3H6        CH4         H2
   1   0.47801E-03  0.40283E-03  0.99863     0.24491E-04  0.45873E-03
   2   0.54489E-03  0.13586E-03  0.99928     0.35282E-05  0.31165E-04
   6   0.87798E-03  0.26272E-04  0.99907     0.12139E-05  0.20944E-04
   7   0.98160E-03  0.25688E-04  0.99897     0.12137E-05  0.20944E-04
   8   0.10949E-02  0.25516E-04  0.99886     0.12137E-05  0.20944E-04
  91   0.50071      0.24658E-04  0.49924     0.11894E-05  0.20519E-04
  92   0.51051      0.24642E-04  0.48943     0.11889E-05  0.20511E-04
  93   0.51933      0.21728E-04  0.48061     0.79236E-06  0.63237E-05
  94   0.51987      0.71288E-05  0.48008     0.92913E-07  0.18683E-06
 152   0.84665      0.00000E+00  0.15331     0.00000E+00  0.00000E+00
 153   0.85886      0.00000E+00  0.14108     0.00000E+00  0.00000E+00
 154   0.87096      0.00000E+00  0.12897     0.00000E+00  0.00000E+00
 155   0.88287      0.00000E+00  0.11698     0.00000E+00  0.00000E+00
 156   0.89447      0.00000E+00  0.10516     0.00000E+00  0.00000E+00

                ****   MASS-Y-PROFILE    ****
STAGE    H2O         CO2
   1   0.00000E+00  0.30730E-05
   2   0.00000E+00  0.10827E-05
   6   0.00000E+00  0.20634E-06
   7   0.00000E+00  0.20076E-06
   8   0.00000E+00  0.19903E-06
  91   0.43183E-05  0.19096E-06
  92   0.13268E-04  0.19082E-06
  93   0.39160E-04  0.16933E-06
  94   0.39171E-04  0.57567E-07
 152   0.47299E-04  0.00000E+00
 153   0.54554E-04  0.00000E+00
 154   0.77083E-04  0.00000E+00
 155   0.14768E-03  0.00000E+00
 156   0.36934E-03  0.00000E+00



              ********************************
              ***** HYDRAULIC PARAMETERS *****
              ********************************


   *** DEFINITIONS ***

    MARANGONI INDEX = SIGMA - SIGMATO
    FLOW PARAM = (ML/MV)*SQRT(RHOV/RHOL)
    QR = QV*SQRT(RHOV/(RHOL-RHOV))
    F FACTOR = QV*SQRT(RHOV)
      WHERE:
      SIGMA IS THE SURFACE TENSION OF LIQUID FROM THE STAGE
      SIGMATO IS THE SURFACE TENSION OF LIQUID TO THE STAGE
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      ML IS THE MASS FLOW OF LIQUID FROM THE STAGE
      MV IS THE MASS FLOW OF VAPOR TO THE STAGE
      RHOL IS THE MASS DENSITY OF LIQUID FROM THE STAGE
      RHOV IS THE MASS DENSITY OF VAPOR TO THE STAGE
      QV IS THE VOLUMETRIC FLOW RATE OF VAPOR TO THE STAGE


              TEMPERATURE
                C
STAGE       LIQUID FROM       VAPOR TO
  1       34.806          35.336
  2       35.336          35.354
  6       35.358          35.359
  7       35.359          35.360
  8       35.360          35.361
 91        39.418          39.499
 92        39.499          39.583
 93        39.597          39.610
 94        39.610          39.615
152         42.228          42.328
153         42.328          42.430
154         42.430          42.546
155         42.546          42.703
156         42.703          42.703


         MASS FLOW              VOLUME FLOW        MOLECULAR WEIGHT
          KG/HR              CUM/HR
STAGE LIQUID FROM  VAPOR TO   LIQUID FROM  VAPOR TO   LIQUID FROM  VAPOR TO
  1  0.36208E+06 0.37906E+06  817.96      11969.     42.070      42.053
  2  0.36383E+06 0.38081E+06  824.07      12020.     42.080      42.063
  6  0.36389E+06 0.38087E+06  824.30      12021.     42.082      42.064
  7  0.36390E+06 0.38087E+06  824.31      12021.     42.082      42.065
  8  0.36390E+06 0.38088E+06  824.32      12021.     42.082      42.065
 91  0.37280E+06 0.38978E+06  873.20      12077.     43.132      43.067
 92  0.37293E+06 0.38991E+06  874.02      12078.     43.150      43.084
 93  0.42343E+06 0.38800E+06  992.94      12012.     43.163      43.102
 94  0.42348E+06 0.38805E+06  993.12      12013.     43.164      43.104
152  0.43058E+06 0.39515E+06  1030.3      12068.     43.800      43.797
153  0.43081E+06 0.39538E+06  1031.6      12070.     43.822      43.821
154  0.43095E+06 0.39552E+06  1032.6      12071.     43.841      43.841
155  0.43080E+06 0.39537E+06  1032.7      12069.     43.851      43.852
156   35431.     0.39537E+06  84.936     12069.     43.832      43.852


          DENSITY              VISCOSITY        SURFACE TENSION
          KG/CUM                CP              DYNE/CM
STAGE LIQUID FROM  VAPOR TO    LIQUID FROM  VAPOR TO      LIQUID FROM
  1  442.66      31.671      0.85773E-01  0.97396E-02   5.7185
  2  441.50      31.682      0.85343E-01  0.97399E-02   5.6567
  6  441.46      31.683      0.85328E-01  0.97399E-02   5.6544
  7  441.45      31.683      0.85327E-01  0.97399E-02   5.6543
  8  441.45      31.683      0.85327E-01  0.97399E-02   5.6542
 91   426.94      32.274      0.82453E-01  0.96803E-02   5.2049
 92   426.68      32.282      0.82414E-01  0.96795E-02   5.2006
 93   426.45      32.302      0.82397E-01  0.96801E-02   5.2029
 94   426.41      32.303      0.82388E-01  0.96800E-02   5.2014
152   417.91      32.744      0.80592E-01  0.96340E-02   4.9314
153   417.62      32.758      0.80538E-01  0.96326E-02   4.9245
154   417.35      32.766      0.80508E-01  0.96317E-02   4.9242
155   417.14      32.758      0.80558E-01  0.96326E-02   4.9443
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156   417.14       32.758      0.80855E-01  0.96326E-02    5.0282


     MARANGONI INDEX   FLOW PARAM         QR         REDUCED F-FACTOR
STAGE    DYNE/CM                    CUM/HR       (GM-L)**.5/MIN
   1                0.25550          3322.4         0.11226E+07
   2   -.61847E-01      0.25594          3342.0         0.11276E+07
   6   -.99414E-04      0.25596          3342.7         0.11278E+07
   7   -.10102E-03      0.25596          3342.7         0.11278E+07
   8   -.10822E-03      0.25596          3342.7         0.11278E+07
  91   -.75253E-02      0.26297          3453.7         0.11435E+07
  92   -.43230E-02      0.26308          3455.5         0.11437E+07
  93   -.31442E-01      0.30036          3438.7         0.11378E+07
  94   -.14190E-02      0.30037          3439.2         0.11379E+07
 152   -.89730E-02      0.30501          3518.6         0.11509E+07
 153   -.68501E-02      0.30517          3521.3         0.11513E+07
 154   -.30202E-03      0.30530          3523.4         0.11516E+07
 155   0.20072E-01      0.30534          3523.4         0.11513E+07
 156   0.83856E-01      0.25113E-01      3523.4         0.11513E+07




             ****************************************
             ***** PACKING SIZING CALCULATIONS *****
             ****************************************


   *******************
   *** SECTION   1 ***
   *******************

   STARTING STAGE NUMBER                                 2
   ENDING STAGE NUMBER                                 155
   CAPACITY CALCULATION METHOD                           VENDOR
   PRESSURE DROP CALCULATION METHOD                      VENDOR
   LIQUID HOLDUP CALCULATION METHOD                      VENDOR
   PRESSURE PROFILE UPDATED                              NO

   DESIGN PARAMETERS
   -----------------
   OVERDESIGN FACTOR                                1.00000
   SYSTEM FOAMING FACTOR                            1.00000
   FRAC. APP. TO MAXIMUM CAPACITY                   1.00000
   MAXIMUM CAPACITY FACTOR        M/SEC             MISSING
   DESIGN CAPACITY FACTOR         M/SEC             MISSING
   PRESSURE DROP FOR THE SECTION   BAR              MISSING
   PRESSURE DROP PER UNIT HEIGHT   MM-WATER/M       MISSING

   PACKING SPECIFICATIONS
   ----------------------
   PACKING TYPE                               PALL-RING
   PACKING MATERIAL                           METAL
   PACKING SIZE                               80-MM
   VENDOR                                     RASCHIG
   PACKING SURFACE AREA          SQCM/CC      0.78000
   PACKING VOID FRACTION                      0.96000
   HETP                   METER               0.85000
   PACKING HEIGHT                METER        130.900
```

```
                ***** SIZING RESULTS *****

  COLUMN DIAMETER              METER                4.58676
  MAXIMUM FRACTIONAL CAPACITY                          1.00000
  MAXIMUM CAPACITY FACTOR        M/SEC                0.059233
  PRESSURE DROP FOR THE SECTION   BAR                  0.11602
  AVERAGE PRESSURE DROP/HEIGHT    MM-WATER/M              9.03798
  MAXIMUM LIQUID HOLDUP/STAGE     CUM                   0.71584


         **** RATING PROFILES AT MAXIMUM COLUMN DIAMETER ****

    HEIGHT
    FROM TOP    FRACTIONAL  PRESSURE    PRESSURE     LIQUID
STAGE  OF SECTION  CAPACITY    DROP       DROP/HEIGHT  HOLDUP      HETP
    METER                BAR        MM-WATER/M    CUM        METER
  2   0.0000E+00  0.9252    0.68429E-03  8.2092     0.6172    0.8500
  3   0.8500     0.9254    0.68460E-03  8.2129     0.6173    0.8500
  4    1.700     0.9254    0.68463E-03  8.2132     0.6173    0.8500
  5    2.550     0.9254    0.68464E-03  8.2134     0.6173    0.8500
  6    3.400     0.9254    0.68465E-03  8.2135     0.6173    0.8500
  7    4.250     0.9254    0.68466E-03  8.2137     0.6173    0.8500
  8    5.100     0.9254    0.68467E-03  8.2138     0.6173    0.8500
  9    5.950     0.9254    0.68469E-03  8.2140     0.6173    0.8500
 10    6.800     0.9254    0.68470E-03  8.2141     0.6174    0.8500
 11    7.650     0.9254    0.68472E-03  8.2143     0.6174    0.8500
 12    8.500     0.9254    0.68473E-03  8.2145     0.6174    0.8500
 13    9.350     0.9254    0.68475E-03  8.2147     0.6174    0.8500
 14    10.20     0.9255    0.68477E-03  8.2149     0.6174    0.8500
 15    11.05     0.9255    0.68479E-03  8.2152     0.6174    0.8500
 16    11.90     0.9255    0.68481E-03  8.2155     0.6174    0.8500
 17    12.75     0.9255    0.68484E-03  8.2158     0.6174    0.8500
 18    13.60     0.9255    0.68487E-03  8.2161     0.6174    0.8500
 19    14.45     0.9255    0.68490E-03  8.2165     0.6175    0.8500
 20    15.30     0.9255    0.68493E-03  8.2168     0.6175    0.8500
 21    16.15     0.9256    0.68496E-03  8.2173     0.6175    0.8500
 22    17.00     0.9256    0.68500E-03  8.2178     0.6175    0.8500
 23    17.85     0.9256    0.68505E-03  8.2183     0.6175    0.8500
 24    18.70     0.9256    0.68509E-03  8.2188     0.6175    0.8500
 25    19.55     0.9257    0.68514E-03  8.2194     0.6176    0.8500
 26    20.40     0.9257    0.68520E-03  8.2201     0.6176    0.8500
 27    21.25     0.9257    0.68526E-03  8.2208     0.6176    0.8500
 28    22.10     0.9258    0.68533E-03  8.2216     0.6177    0.8500
 29    22.95     0.9258    0.68540E-03  8.2225     0.6177    0.8500
 30    23.80     0.9259    0.68548E-03  8.2235     0.6177    0.8500
 31    24.65     0.9259    0.68557E-03  8.2245     0.6178    0.8500
 32    25.50     0.9260    0.68566E-03  8.2256     0.6178    0.8500
 33    26.35     0.9260    0.68576E-03  8.2269     0.6179    0.8500
 34    27.20     0.9261    0.68588E-03  8.2282     0.6179    0.8500
 35    28.05     0.9262    0.68600E-03  8.2297     0.6180    0.8500
 36    28.90     0.9263    0.68613E-03  8.2313     0.6181    0.8500
 37    29.75     0.9264    0.68628E-03  8.2331     0.6181    0.8500
 38    30.60     0.9264    0.68644E-03  8.2350     0.6182    0.8500
 39    31.45     0.9265    0.68661E-03  8.2370     0.6183    0.8500
 40    32.30     0.9267    0.68680E-03  8.2393     0.6184    0.8500
 41    33.15     0.9268    0.68700E-03  8.2417     0.6185    0.8500
 42    34.00     0.9269    0.68722E-03  8.2444     0.6186    0.8500
 43    34.85     0.9271    0.68746E-03  8.2473     0.6187    0.8500
 44    35.70     0.9272    0.68773E-03  8.2504     0.6188    0.8500
 45    36.55     0.9274    0.68801E-03  8.2538     0.6190    0.8500
 46    37.40     0.9276    0.68832E-03  8.2575     0.6191    0.8500
```

47	38.25	0.9278	0.68865E-03	8.2615	0.6193	0.8500
48	39.10	0.9280	0.68901E-03	8.2658	0.6194	0.8500
49	39.95	0.9282	0.68940E-03	8.2705	0.6196	0.8500
50	40.80	0.9284	0.68982E-03	8.2756	0.6198	0.8500
51	41.65	0.9287	0.69028E-03	8.2810	0.6200	0.8500
52	42.50	0.9290	0.69077E-03	8.2869	0.6203	0.8500
53	43.35	0.9293	0.69129E-03	8.2932	0.6205	0.8500
54	44.20	0.9296	0.69186E-03	8.3000	0.6208	0.8500
55	45.05	0.9300	0.69247E-03	8.3073	0.6211	0.8500
56	45.90	0.9304	0.69312E-03	8.3151	0.6214	0.8500
57	46.75	0.9308	0.69382E-03	8.3235	0.6217	0.8500
58	47.60	0.9312	0.69456E-03	8.3324	0.6221	0.8500
59	48.45	0.9317	0.69535E-03	8.3419	0.6224	0.8500
60	49.30	0.9322	0.69619E-03	8.3520	0.6228	0.8500
61	50.15	0.9327	0.69709E-03	8.3627	0.6233	0.8500
62	51.00	0.9332	0.69804E-03	8.3741	0.6237	0.8500
63	51.85	0.9338	0.69903E-03	8.3861	0.6242	0.8500
64	52.70	0.9344	0.70009E-03	8.3987	0.6247	0.8500
65	53.55	0.9351	0.70119E-03	8.4120	0.6252	0.8500
66	54.40	0.9357	0.70235E-03	8.4259	0.6257	0.8500
67	55.25	0.9364	0.70356E-03	8.4404	0.6263	0.8500
68	56.10	0.9372	0.70482E-03	8.4555	0.6269	0.8500
69	56.95	0.9379	0.70613E-03	8.4712	0.6275	0.8500
70	57.80	0.9387	0.70749E-03	8.4875	0.6281	0.8500
71	58.65	0.9395	0.70888E-03	8.5042	0.6287	0.8500
72	59.50	0.9403	0.71031E-03	8.5214	0.6294	0.8500
73	60.35	0.9411	0.71178E-03	8.5390	0.6300	0.8500
74	61.20	0.9419	0.71328E-03	8.5569	0.6307	0.8500
75	62.05	0.9428	0.71480E-03	8.5752	0.6314	0.8500
76	62.90	0.9437	0.71633E-03	8.5936	0.6321	0.8500
77	63.75	0.9445	0.71788E-03	8.6122	0.6328	0.8500
78	64.60	0.9454	0.71944E-03	8.6309	0.6335	0.8500
79	65.45	0.9462	0.72100E-03	8.6496	0.6341	0.8500
80	66.30	0.9471	0.72255E-03	8.6682	0.6348	0.8500
81	67.15	0.9479	0.72409E-03	8.6867	0.6355	0.8500
82	68.00	0.9488	0.72562E-03	8.7050	0.6362	0.8500
83	68.85	0.9496	0.72712E-03	8.7230	0.6368	0.8500
84	69.70	0.9504	0.72859E-03	8.7407	0.6375	0.8500
85	70.55	0.9512	0.73003E-03	8.7579	0.6381	0.8500
86	71.40	0.9520	0.73144E-03	8.7748	0.6387	0.8500
87	72.25	0.9527	0.73280E-03	8.7911	0.6393	0.8500
88	73.10	0.9534	0.73412E-03	8.8069	0.6398	0.8500
89	73.95	0.9541	0.73538E-03	8.8221	0.6404	0.8500
90	74.80	0.9547	0.73658E-03	8.8365	0.6409	0.8500
91	75.65	0.9553	0.73768E-03	8.8497	0.6414	0.8500
92	76.50	0.9558	0.73857E-03	8.8604	0.6418	0.8500
93	77.35	0.9767	0.81731E-03	9.8050	0.6968	0.8500
94	78.20	0.9768	0.81764E-03	9.8089	0.6969	0.8500
95	79.05	0.9768	0.81773E-03	9.8101	0.6969	0.8500
96	79.90	0.9769	0.81783E-03	9.8112	0.6970	0.8500
97	80.75	0.9769	0.81793E-03	9.8124	0.6970	0.8500
98	81.60	0.9770	0.81803E-03	9.8137	0.6971	0.8500
99	82.45	0.9770	0.81815E-03	9.8151	0.6971	0.8500
100	83.30	0.9771	0.81827E-03	9.8165	0.6971	0.8500
101	84.15	0.9772	0.81841E-03	9.8182	0.6972	0.8500
102	85.00	0.9772	0.81855E-03	9.8199	0.6972	0.8500
103	85.85	0.9773	0.81871E-03	9.8218	0.6973	0.8500
104	86.70	0.9774	0.81888E-03	9.8238	0.6974	0.8500
105	87.55	0.9774	0.81906E-03	9.8260	0.6974	0.8500
106	88.40	0.9775	0.81926E-03	9.8284	0.6975	0.8500
107	89.25	0.9776	0.81947E-03	9.8309	0.6976	0.8500

108	90.10	0.9777	0.81970E-03	9.8337	0.6977	0.8500
109	90.95	0.9778	0.81995E-03	9.8366	0.6978	0.8500
110	91.80	0.9780	0.82021E-03	9.8398	0.6978	0.8500
111	92.65	0.9781	0.82050E-03	9.8432	0.6980	0.8500
112	93.50	0.9782	0.82081E-03	9.8469	0.6981	0.8500
113	94.35	0.9784	0.82114E-03	9.8509	0.6982	0.8500
114	95.20	0.9785	0.82149E-03	9.8552	0.6983	0.8500
115	96.05	0.9787	0.82187E-03	9.8597	0.6985	0.8500
116	96.90	0.9789	0.82228E-03	9.8647	0.6986	0.8500
117	97.75	0.9791	0.82273E-03	9.8700	0.6988	0.8500
118	98.60	0.9793	0.82320E-03	9.8756	0.6989	0.8500
119	99.45	0.9795	0.82370E-03	9.8817	0.6991	0.8500
120	100.3	0.9798	0.82425E-03	9.8882	0.6993	0.8500
121	101.1	0.9800	0.82483E-03	9.8952	0.6995	0.8500
122	102.0	0.9803	0.82545E-03	9.9026	0.6997	0.8500
123	102.8	0.9806	0.82611E-03	9.9105	0.7000	0.8500
124	103.7	0.9809	0.82681E-03	9.9190	0.7002	0.8500
125	104.5	0.9813	0.82756E-03	9.9280	0.7005	0.8500
126	105.4	0.9816	0.82836E-03	9.9376	0.7008	0.8500
127	106.2	0.9820	0.82921E-03	9.9478	0.7011	0.8500
128	107.1	0.9824	0.83012E-03	9.9586	0.7014	0.8500
129	107.9	0.9828	0.83107E-03	9.9701	0.7018	0.8500
130	108.8	0.9833	0.83208E-03	9.9822	0.7021	0.8500
131	109.6	0.9838	0.83315E-03	9.9950	0.7025	0.8500
132	110.5	0.9843	0.83428E-03	10.009	0.7029	0.8500
133	111.3	0.9848	0.83546E-03	10.023	0.7033	0.8500
134	112.2	0.9853	0.83671E-03	10.038	0.7038	0.8500
135	113.0	0.9859	0.83802E-03	10.053	0.7042	0.8500
136	113.9	0.9865	0.83939E-03	10.070	0.7047	0.8500
137	114.7	0.9871	0.84082E-03	10.087	0.7052	0.8500
138	115.6	0.9878	0.84230E-03	10.105	0.7057	0.8500
139	116.4	0.9885	0.84385E-03	10.123	0.7063	0.8500
140	117.3	0.9892	0.84546E-03	10.143	0.7068	0.8500
141	118.1	0.9899	0.84712E-03	10.163	0.7074	0.8500
142	119.0	0.9906	0.84883E-03	10.183	0.7080	0.8500
143	119.8	0.9914	0.85059E-03	10.204	0.7086	0.8500
144	120.7	0.9922	0.85240E-03	10.226	0.7092	0.8500
145	121.5	0.9929	0.85426E-03	10.248	0.7099	0.8500
146	122.4	0.9938	0.85614E-03	10.271	0.7105	0.8500
147	123.2	0.9946	0.85807E-03	10.294	0.7112	0.8500
148	124.1	0.9954	0.86002E-03	10.317	0.7118	0.8500
149	124.9	0.9962	0.86199E-03	10.341	0.7125	0.8500
150	125.8	0.9971	0.86397E-03	10.365	0.7132	0.8500
151	126.6	0.9979	0.86594E-03	10.388	0.7138	0.8500
152	127.5	0.9987	0.86788E-03	10.412	0.7145	0.8500
153	128.3	0.9995	0.86965E-03	10.433	0.7151	0.8500
154	129.2	1.000	0.87093E-03	10.448	0.7156	0.8500
155	130.0	0.9999	0.87068E-03	10.445	0.7158	0.8500

Appendix E.3 Reactor Design

For this project, we propose to use Shell and Tube reactor. The endothermic reaction, dehydrogenation of propane, takes place in the shell side and exothermic reaction, oxidative dehydrogenation of propane, takes place in the tube side. In this chapter, it shows the idea how to estimate the volume of reactor, volume and type of catalyst used and the material of reactor.

E.3.1. Tube side:

There are many reactions take place in tube side. The rate expressions of reactions are showed

1. Propane oxidation to propylene

$$3C_3H_{8(g)} + 9O_2 \rightarrow C_3H_{6(g)} + 3CO + 3CO_2 + 9H_2O_{(g)} \qquad \text{(E.3.1.1)}$$

$$(-r_1) = k_{O_2}P_{O_2}\left[1+\frac{k_{O_2}P_{O_2}}{2\cdot k_{C_3H_8}P_{C_3H_8}} - \sqrt{\left(\frac{k_{O_2}P_{O_2}}{2\cdot k_{C_3H_8}P_{C_3H_8}}\right)^2 + \frac{k_{O_2}P_{O_2}}{k_{C_3H_8}P_{C_3H_8}}}\right] \qquad \text{(E.3.1.2)}$$

2. Propylene oxidation to CO

$$2C_3H_{6(g)} + 7.5O_2 \rightarrow 3CO_{(g)} + 3CO_{2(g)} + 6H_2O_{(g)} \qquad \text{(E.3.1.3)}$$

$$(-r_2) = k_2 P^{m}_{C_3H_6}\ P^{n}_{O_2} \qquad \text{(E.3.1.4)}$$

3. CO oxidation into CO_2

$$CO_{(g)} + 0.5O_{2_{(g)}} \rightarrow CO_{2(g)} \qquad \text{(E.3.1.5)}$$

$$(-r_3) = k_3 P^{m_5}_{CO}\ P^{n_5}_{O_2} \qquad \text{(E.3.1.6)}$$

4. Cracking of propane

$$C_3H_{8(g)} \rightarrow CH_{4(g)} + C_2H_{4(g)} \qquad \text{(E.3.1.7)}$$

$$(-r_4) = \frac{k_4\left(P_{C_3H_8} - \dfrac{P_{C_3H_6}P_{H_2}}{K_1}\right)}{1+P_{C_3H_6}K_2} \qquad \text{(E.3.1.8)}$$

E.3.2. Shell side:

There are many reactions take place in shell side. The rate expressions of reactions are showed

1. Cracking of propane

$$C_3H_{8(g)} \rightarrow C_2H_{4(g)} + CH_{4(g)} \qquad \text{(E.3.2.1)}$$

(A) (B) (C)

$$(-r_5) = \frac{k_4^S\left(P^S_{C_3H_8} - \dfrac{P^S_{C_3H_6}P^S_{H_2}}{K_1}\right)}{1 + P^S_{C_3H_6}K_2} \qquad \text{(E.3.2.2)}$$

2. Propane dehydrogenation

$$C_3H_{8(g)} \rightarrow C_3H_{6(g)} + H_{2(g)} \qquad \text{(E.3.2.3)}$$

(A) (D) (E)

$$(-r_6) = k_6 P^{O,S}_{C_3H_8} \qquad \text{(E.3.2.4)}$$

We consider that the Tube side of this reactor is plug flow reactor system both sides, and then make the mass balance and energy balance at the steady state condition.

E.3.3. Tube side:

Mass Balance for component:

For Propane

In = Out + Conversion + hold up

$$F_T C_{C_3H_8}\Big|_{V=Vo} = F_T C_{C_3H_8}\Big|_{V=Vo+dV} + (-r_{C_3H_8})dV_T + V_T\frac{dC_{C_3H_8}}{dt}$$

$$\boxed{-\frac{dC_{C_3H_8}}{dV_T} = \left(\frac{(-r_1)+(-r_4)}{F_T}\right) \qquad \text{(E.3.3.1)}}$$

For oxygen

$$\boxed{-\frac{dC_{O_2}}{dV_T} = \left(\frac{3(-r_1)+3.75(-r_2)+0.5(-r_3)}{F_T}\right) \qquad \text{(E.3.3.2)}}$$

For propylene

$$\boxed{-\frac{dC_{C_3H_6}}{dV_T} = \left(\frac{-1/3(-r_1)+(-r_2)}{F_T}\right) \qquad \text{(E.3.3.3)}}$$

For carbon monoxide

$$\boxed{-\frac{dC_{CO}}{dV_T} = \left(\frac{-(-r_1)-1.5(-r_2)+(-r_3)}{F_T}\right) \qquad \text{(E.3.3.4)}}$$

For carbon dioxide

$$-\frac{dC_{CO_2}}{dV_T} = \left(\frac{-(-r_1)-1.5(-r_2)-(-r_3)}{F_T}\right) \qquad (E.3.3.5)$$

For water

$$-\frac{dC_{H_2O}}{dV_T} = \left(\frac{-3(-r_1)-3(-r_2)}{F_T}\right) \qquad (E.3.3.6)$$

For methane

$$-\frac{dC_{CH_4}}{dV_T} = \left(\frac{-(-r_4)}{F_T}\right) \qquad (E.3.3.7)$$

For ethylene

$$-\frac{dC_{C_2H_4}}{dV_T} = \left(\frac{-(-r_4)}{F_T}\right) \qquad (E.3.3.8)$$

Energy balance:

In = Out + heat transfer + heat generated + heat accumulate

$$\begin{aligned}\rho_{T,mix}F_TC_{P,mix}T_{Ti} &= \rho_{T,mix}F_TC_{P,mix}T_T + U\cdot dA\cdot(T_T - T_S) + \\ &+\left[\Delta H_1(-r_1)+\Delta H_2(-r_2)+\Delta H_3(-r_3)+\Delta H_4(-r_4)\right]dV_T\end{aligned}$$

$$\begin{aligned}\rho_{T,mix}F_TC_{P,mix}T_{Ti} &= \rho_{T,mix}F_TC_{P,mix}T_T + \frac{4U}{D_{ia.,Tube}}(T_T - T_S)dV_T + \\ &+\left[\Delta H_1(-r_1)+\Delta H_2(-r_2)+\Delta H_3(-r_3)+\Delta H_4(-r_4)\right]dV_T\end{aligned}$$

$$-\frac{dT_T}{dV_T} = \left(\frac{\frac{4U}{D_{ia.,Tube}}(T_T - T_S)+\left[\Delta H_1(-r_1)+\Delta H_2(-r_2)+\Delta H_3(-r_3)+\Delta H_4(-r_4)\right]}{\rho_{T,mix}F_TC_{P,mix}}\right) \qquad (E.3.3.9)$$

E.3.4. Shell side:

For the Shell side, we also consider like plug flow reactor system, which there is temperature and concentration profile. The volume of shell reactor is temperature and conversion dependence. So, taking mass balance and energy balance at the steady state condition is to model this system as following.

Mass Balance for component:

In = Out + Conversion + hold up

$$F_S C_{C_3H_8}\Big|_{V=V_o} = F_S C_{C_3H_8}\Big|_{V=V_o+dV} + (-r_{C_3H_8})dV_S + V_S \frac{dC_{C_3H_8}}{dt}$$

For propane

$$\boxed{-\frac{dC_{C_3H_8}}{dV_S} = \left(\frac{(-r_5)+(-r_6)}{F_S}\right) \qquad (E.3.4.1)}$$

For ethylene

$$\boxed{-\frac{dC_{C_2H_4}}{dV_S} = \frac{-(-r_5)}{F_S} \qquad (E.3.4.2)}$$

For methane

$$\boxed{-\frac{dC_{CH_4}}{dV_S} = \frac{-(-r_5)}{F_S} \qquad (E.3.4.3)}$$

For propylene

$$\boxed{-\frac{dC_{C_3H_6}}{dV_S} = \frac{-(-r_6)}{F_S} \qquad (E.3.4.4)}$$

For hydrogen

$$\boxed{-\frac{dC_{H_2}}{dV_S} = \frac{-(-r_6)}{F_S} \qquad (E.3.4.5)}$$

Energy balance:

In = Out + heat transfer + heat generated + heat acc.

$$\begin{aligned}\rho_S F_S C_{P,S} T_{Si} &= \rho_S F_S C_{P,S} T_S - U dA (T_T - T_S) + \\ &\quad + \left[\Delta H_5 (-r_5) + \Delta H_6 (-r_6)\right] dV_S\end{aligned}$$

$$\begin{aligned}\rho_S F_S C_{P,S} T_{Si} &= \rho_S F_S C_{P,S} T_S - \frac{4U}{D_{ia,Tube}} (T_T - T_S) dV_T + \\ &\quad + \left[\Delta H_5 (-r_5) + \Delta H_6 (-r_6)\right] dV_s\end{aligned}$$

$$\boxed{\frac{-dT_S}{dV_S} = \left(\frac{\dfrac{-4U}{D_{ia,Tube}}(T_T - T_S)\cdot\dfrac{dV_T}{dV_S} + \left[\Delta H_5(-r_5) + \Delta H_6(-r_6)\right]}{\rho_S F_S C_{P,S}}\right) \quad (E.3.4.6)}$$

From the mathematic model above, they are the series of ordinary differential equations that consume time to solve precisely. In this conceptual design, the reactor will be simplified by considering the main reaction, and the reactor both sides take place in gas phase heterogeneously with fixed bed catalyst, that heat capacity of gas phase is usually less than the liquid phase. Then, the heat exchange between shell and tube could be by heat conduction of each side in order to retain its temperature. Therefore the reactions of both sides are assumed as isothermal reactions. In addition to this, catalyst deactivation and coke formation will not be considered during design as well. The following section shows the simplified calculation of reactor both sides.

E.3.5. Propane oxidative dehydrogenation in tube side

For the reactions in tube side, oxidation dehydrogenation of propane will be considered in the main reaction, the kinetic rate expression and catalyst data show below

$$(-r_{C_3H_8}) = k_1 P_{C_3H_8}^{0.6} \qquad \text{(E.3.5.1)}$$

$$k_1 = 0.004\ \text{mol}_{C_3H_8} / \text{g-cat} \cdot \text{min (atm)}^{0.6}$$

Catalyst data:

Type:	V_2O_5
Specific surface:	$95 m^2$/g-cat
Void of bed:	0.5
Particle density:	$1500 kg/m^3$

Effective diffusivity of propane to V_2O_5 is less, and then assumes the catalyst effectiveness is ~1.

We can rewrite the rate reaction constant, which is catalyst mass basis into the volume of fluid basis.

$$k_1 = 0.004\ \frac{\text{mol}_{C_3H_8}}{\text{g-cat} \cdot \text{min (atm)}^{0.6}} \cdot \frac{1}{95}\frac{\text{g-cat}}{m^2} \cdot \frac{1}{60}\frac{\text{min}}{\text{sec}} \qquad \text{(E.3.5.2)}$$

$$= 7.02 \times 10^{-7}\ \frac{\text{mol}_{C_3H_8}}{(\text{atm})^{0.6} \cdot m^2 \cdot \text{sec}}$$

$$CS = 95\ \frac{m^2}{\text{g-cat}} \cdot 1{,}500\ \frac{\text{kg-cat}}{m^3_{\text{cat.}}} \cdot (1-0.5)\ \frac{m^3_{\text{cat.}}}{m^3_{\text{reactor.}}} \cdot 1{,}000 \frac{\text{g-cat}}{\text{kg-cat}} \qquad \text{(E.3.5.3)}$$

$$= 7.125 \times 10^{7}\ \frac{m^2}{m^3_{\text{reactor.}}}$$

$$k_1' = 7.02 \times 10^{-7}\ \frac{\text{mol}_{C_3H_8}}{(\text{atm})^{0.6} \cdot m^2 \cdot \text{sec}} \times CS \qquad \text{(E.3.5.4)}$$

$$= 50\ \frac{\text{mol}_{C_3H_8}}{(\text{atm})^{0.6} \cdot m^3_{\text{reactor}} \cdot \text{sec}}$$

From the differential mass balance in terms of the void volume, the general expression shows as

$$F_{A0} dX_A = (-r_A'')dV_P$$

$$(-r_A'')dV_P = (-r_A) \cdot void\ fraction. \cdot dV_{Reactor}$$

so,

$$F_{A0} dX_A = (-r_A) \cdot void\ fraction \cdot dV_{Reactor}$$

$$\mathrm{F}_{C_3H_{8,0}} dX_{C_3H_8} = k_1^{'} \cdot P_{C_3H_8}^{0.6} \cdot void\ fraction \cdot dV_{\mathrm{Re}actor}$$

Assume the gas in reaction perform like ideal gas, hence the term $P_{C_3H_8}^{0.6}$ can be changed in term of concentration $P_{C_3H_8}^{0.6} = \left(C_{C_3H_8} RT\right)^{0.6}$

Then,

$$\mathrm{F}_{C_3H_{8,0}} dX_{C_3H_8} = k_1^{'} \cdot \left(C_{C_3H_8} RT\right)^{0.6} \cdot void\ fraction \cdot dV_{\mathrm{Re}actor}$$

$$= k_1^{'} \cdot \left(C_{C_3H_{8,0}} (1 - X_{C_3H_8}) RT\right)^{0.6} \cdot void\ fraction \cdot dV_{\mathrm{Re}actor}$$

$$\int_{X=0}^{X=0.8} \frac{1}{\left(1 - X_{C_3H_8}\right)^{0.6}} dX_{C_3H_8} = \int_{\mathrm{V}=0}^{\mathrm{V}} \frac{k_1^{'}}{\mathrm{F}_{C_3H_{8,0}}} \cdot \left(C_{C_3H_{8,0}} RT\right)^{0.6} \cdot void\ fraction \cdot dV_{\mathrm{Re}actor}$$

In order to design as the base case, from ASPEN result, we can define

$F_{C_3H_8,0} =$ 534.121 kmol/hr

$$C_{C_3H_8,0} = \left(\frac{0.664/1.01}{R \cdot T}\right) \frac{\mathrm{kmol}}{\mathrm{m}^3}$$

$$= \left(\frac{0.657}{R \cdot T}\right) \frac{\mathrm{kmol}}{\mathrm{m}^3}$$

therefore,

$$\int_{X=0}^{X=0.8} \frac{1}{\left(1 - X_{C_3H_8}\right)^{0.6}} dX_{C_3H_8} = \int_{\mathrm{V}=0}^{\mathrm{V_{Reactor}}} \frac{50 \cdot 3600}{534.121 \cdot 1000} \cdot \left(0.657\right)^{0.6} \cdot (1 - 0.5) \cdot dV_{\mathrm{Re}actor}$$

$$1.1867 = 0.131 V_{\mathrm{Re}actor}$$

$$V_{\mathrm{Re}actor} = 9.06\ \mathrm{m}^3$$

The void fraction is 0.5, then we can estimate the volume of fixed catalyst to be used that it is $\frac{V_{\mathrm{Re}actor} - V_{catalyst}}{V_{\mathrm{Re}actor}} = 0.5$, so, $V_{catalyst} = 4.53\ \mathrm{m}^3$ and amount of catalyst is 6.8 tons

E.3.6. Propane dehydrogenation in shell side

Vapor-phase dehydrogenation of propane in shell side assumed as a tubular flow reactor considered with two reactions:

$$C_3H_{8(g)} \rightarrow C_2H_{4(g)} + CH_{4(g)} \qquad \text{(E.3.6.1)}$$

$$C_3H_{8(g)} \rightarrow C_3H_{6(g)} + H_{2(g)} \qquad \text{(E.3.6.2)}$$

The rate equations are:

$$(-\mathrm{r}_{1,\mathrm{C_3H_8}}) = \mathrm{k_1 P_{C_3H_8}} \qquad \text{(E.3.6.3)}$$

$$(-\mathrm{r}_{2,\mathrm{C_3H_8}}) = \frac{\mathrm{k_2 (P_{C_3H_8}} - \frac{\mathrm{P_{C_3H_6} P_{H_2}}}{\mathrm{K_1}})}{1 + \mathrm{P_{C_3H_6} K_2}} \qquad \text{(E.3.6.4)}$$

$$k_i = k_{0i} \exp\left(\frac{-Ea_i}{R}\left(\frac{1}{T}-\frac{1}{T_0}\right)\right) \quad \text{(E.3.6.5)}$$

$$K_i = K_{0i} \exp\left(\frac{-\Delta H}{R}\left(\frac{1}{T}-\frac{1}{T_0^{'}}\right)\right) \quad \text{(E.3.6.6)}$$

where $P_{C_3H_8}$ = partial pressure of propane, bar

$P_{C_3H_6}$ = partial pressure of propylene, bar

P_{H_2} = partial pressure of hydrogen, bar

K_1 , K_2 = equilibrium constants for the two reactions in term of partial pressures

k_1, k_2 = rate reaction constants for the two reactions, sec

Ea = activated energy , kJ/mol

ΔH = enthalpy, kJ/mol

T_0 , $T_0^{'}$ = temperature reference, K at experiment 600°C

R = gas constant, J/mol K

The data from the literature [Applied catalyst A :General 248(2003) 105-106, J.Gascon] shows in Table E.3.1 below.
Catalyst specification:

Shape : Pellet
Diameter : 5.0mm
High : 4.9 mm
Specific surface : 214m^2 /g-cat
Void fraction : 0.5
Catalyst density : 1,500kg/m^3

Table E.3.1 *Kinetic parameter*

Parameter	Value(units)
k_{S01}	0.0516(mmol/(g s)) : k_1 = 17.62 molC_3H_8/m^3 s at T = 540°C
K_{02}	3450 (mmol/l) : K_{02} = 0.00000068 m^3/mol at T = 540°C
Ea_1	35.5(kJ/mol)
ΔH	-595(kJ/mol) at 600°C
Ea_2	308 (kJ/mol)
k_{S02}	10^E-5 (mmol/(g s) : k_2 = 0.0033 s^{-1} at T = 540°C

K_1 is estimated at 540 C, from the relation of Gibb's free energy.
K_1=0.089

Then, we can simplify the reaction rates as

$$(-r_1) = 17.62 \cdot P_{C_3H_8} \quad \text{(E.3.6.7)}$$

and

$$(-r_2) = \frac{0.0033(P_{C_3H_8} - \frac{P_{C_3H_6}P_{H_2}}{0.089})}{1+P_{C_3H_6}6.8\cdot 10^{-7}} \qquad (E.3.6.8)$$

The design equation for the plug flow is

$$\frac{V}{F} = \int_{x_A=0}^{x_A=x_{AF}} \frac{dX_A}{(-r_A)}$$

If the mass balances are based on propane, then the two equations are

$$\frac{V}{F} = \int \frac{dX_1}{(-r_1)} \qquad (E.3.6.9)$$

$$\frac{V}{F} = \int \frac{dX_2}{(-r_2)} \qquad (E.3.6.10)$$

where the conversion X_1 is the gram moles of propane disappearing by the reaction 1 per gram mole of the feed, and the conversion X_2 is the gram moles of propane disappearing by the reaction 2 per gram mole of the feed.

Based on 1.0 mole of feed of entering propane, the moles of each component at conversions X_1 and X_2 are:

Component	**mole at conversions X1 ans X2**
C_3H_8	1-X1-X2
C_3H_6	X2
H_2	X2
CH_4	X1
C_2H_4	X1
Total	1+X1+X2

Since the total mole equals to 1+X1+X2, the mole fractions of each components are also given by these quantities. If the components are assumed to behave as ideal gases, then the partial pressures are:

Component	**Partial pressure p_i**
C_3H_8	1-X1-X2 / (1+X1+X2)
C_3H_6	X2/ (1+X1+X2)
H_2	X2/ (1+X1+X2)
CH_4	X1 / (1+X1+X2)
C_2H_4	X1 / (1+X1+X2)

Substituting the partial pressure of the components of the components in eq. E.3.6.11 and eq. E.3.6.12 gives

$$(-r_1) = 17.62\left(\frac{1-X_1-X_2}{1+X_1+X_2}\right) \qquad (E.3.6.11)$$

$$(-r_2) = \frac{0.0033\left(\left(\frac{1-X_1-X_2}{1+X_1+X_2}\right) - \frac{X_2{}^2}{0.089(1+X_1+X_2)^2}\right)}{\left(1+\frac{X_2}{(1+X_1+X_2)}6.8\cdot 10^{-7}\right)} \qquad (E.3.6.12)$$

Substituting eq. E.3.6.11 and eq. E.3.6.12 in the eq. E.3.6.13 and eq. E.3.6.14, and the composition of the components are computed for various values of V/F.

$$\frac{dX_1}{d(V/F)} = r_1(X_1,X_2) = 17.62 \left(\frac{1-X_1-X_2}{1+X_1+X_2}\right) \qquad \text{(E.3.6.13)}$$

$$\frac{dX_2}{d(V/F)} = r_2(X_1,X_2) = \frac{0.0033\left(\left(\frac{1-X_1-X_2}{1+X_1+X_2}\right) - \frac{X_2^{\,2}}{0.089\left(1+X_1+X_2\right)^2}\right)}{\left(1+\frac{X_2}{\left(1+X_1+X_2\right)}6.8\cdot 10^{-7}\right)} \qquad \text{(E.3.6.14)}$$

Calculating by function ODE45, MATLAB, resulting in Table E.3.2 and Figure E.3.1 shows the plot of the rates of each reaction as a function of V/F. In both instance, the rates decrease toward zero as V/F increases.

Table E.3.2 *Conversion versus V/F for the dehydrogenation of propane*

V/F	X1	X2	XT	rate1	rate2
0.0000	0.0000	0.0000	0.0000	1.76E+01	0.0033
0.0018	0.0307	5.8E-06	0.0307	1.66E+01	0.0031
0.0036	0.0597	1.12E-05	0.0597	1.56E+01	0.0028
0.0054	0.087	1.63E-05	0.0870	1.48E+01	0.0025
0.0072	0.113	2.12E-05	0.1130	1.40E+01	0.0022
0.0126	0.1835	3.44E-05	0.1835	1.22E+01	0.0014
0.0180	**0.245**	**4.59E-05**	**0.2450**	**1.07E+01**	**0.0006**
0.0234	0.2994	5.61E-05	0.2995	9.50E+00	-0.0002

The reactor volume required to process from Aspen Plus Simulation, 889kmol/hr of propane is estimated from Table E.3.2. For the total conversion of 23 %,

V/F = 0.018
V = 0.018F
= 0.018*889*1000/3600
= 4.45m^3

Catalyst particle void in the reactor is 0.5, so the volume of catalyst in shell side is equal to 2.225 m^3 and catalyst used 3.3 tons.

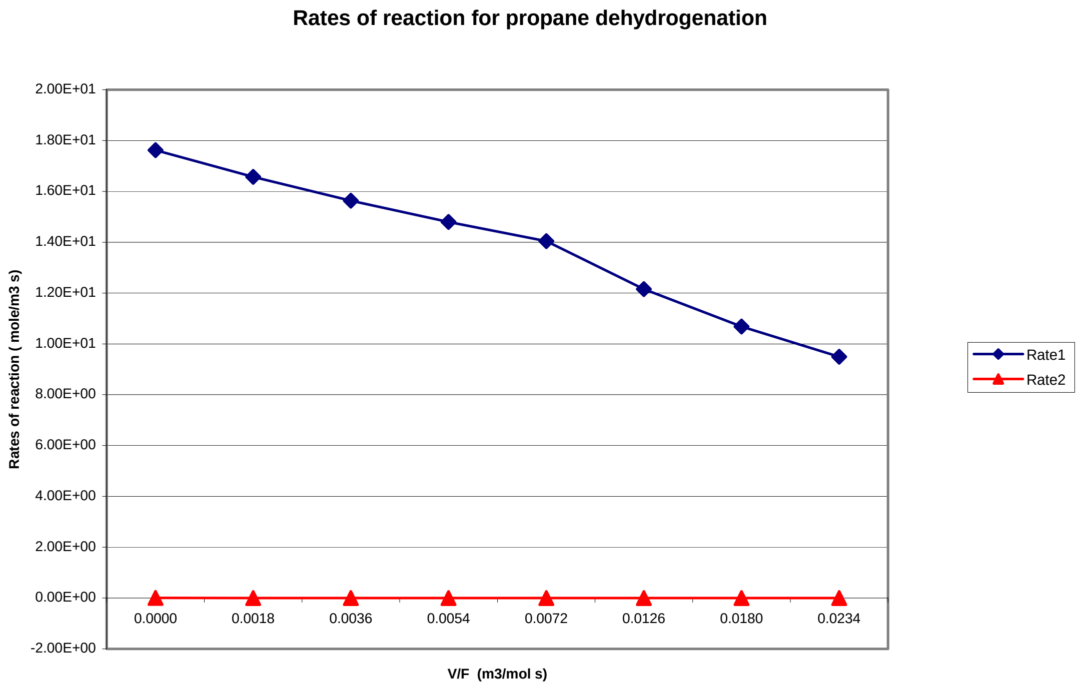


Figure E.3.1 *Rates of reaction for propane dehydrogenation*

Since, in this design stage, we use the simplified model, the heat transfer coefficient of shell and tube is assumed to a general figure approximately. The heat transfer coefficient used in further calculation like cost evaluation is assumed as 500 W/m^2 ^{o}C. The heat exchange between shell and tube is roughly equal to the heat reaction of exothermic, which will be removed, as mention is Appendix A.2. The amount of heat is $20\mathrm{kcal/mol}_{C_3H_8}$ ($84\ \mathrm{kJ/mol}_{C_3H_8}$) and feed propane through the exothermic reaction is 804.63 kmol/hr based on the required product 200 kton annual. Therefore the removed heat is 18.7 x 10^3 kW. Therefore the total area of the tube can be estimated.

$$Q = UA\Delta T_{LM}$$

$$A = \frac{Q}{U\Delta T_{LM}} = \frac{18.7 \cdot 10^3 kW}{500W/m^{2\circ}C \cdot (850-540)} = 121m^2$$

Then, the ratio of area and volume results in the diameter of the tube side 0.3 m.
The number and length of tube can be calculated listed in the Table E.3.3. We choose 20 tubes with the 6.41-meter long.

Table E.3.3*. Number and length of tube reacto*

Dia. Tube (0.3 m)	Numbe of tube	Height of Tube (m)	Area (m2)	Volume (m3)
0.3	10	12.82	121	9.06
0.3	**20**	**6.41**	**121**	**9.06**
0.3	30	4.27	121	9.06
0.3	40	3.20	121	9.06
0.3	50	2.56	121	9.06
0.3	60	2.14	121	9.06
0.3	70	1.83	121	9.06
0.3	80	1.60	121	9.06
0.3	90	1.42	121	9.06
0.3	100	1.28	121	9.06

0.3	110	1.17	121	9.06
0.3	120	1.07	121	9.06

And for the shell side, the total volume can estimate the diameter of reactor.

$Total\ Volume = 4.45 + 9.06 = 13.51 m^3$

$$Dia.of \mathrm{Reactor} = \sqrt{\frac{4 \cdot \mathrm{Volume}_{\mathrm{total}}}{\pi \cdot height}} = 1.63 \text{ m.}$$

Appendix E.4 Shell and Tube Heat Exchanger Design

The design of shell and tube heat exchanger involves the determination of the heat transfer coefficient and pressure drop on both the tube side and the shell side. A large number of methods are available for determining the shellside performance. Before the design procedure, some guidelines for shellside design and points to be raised while specifying a heat exchanger are listed, followed by preliminary sizing of a shell and tube heat exchanger.

Guideline for Shell-Side Design

Recommend guidelines for shellside design include:

1. Accept TEMA fabrication clearances and tolerances, and enforce these standards during fabrication.
2. For segmental baffles employ 20%battle cuts.
3. Employ no-tubes-in-window design to eliminate the damage from flow-induced vibration.
4. Evaluate the heat transfer in the clean condition and pressure drop in the maximum fouled condition.
5. Employ sealing devices to minimize bypassing between the bundle and shell for pull through floating heat exchanger, and through pass partition lanes.
6. Ration of baffle spacing to shell diameter may be restricted between 0.2 and 1.0. Baffle spacing much greater than the shell diameter must be carefully evaluated.
7. Avoid shell longitudinal baffles that are not welded to the shell; all other sealing methods are inadequate.

Specifying the Right Heat Exchanger

When specifying an exchanger for design, various factors to be considered or questions that should be raised are listed by Gutterman:

1. Type of heat transfer, i.e., boiling, condensing, or single-phase heat transfer.
2. Since the heat exchanger has two pressure chambers, which chamber should receive the cold fluid?
3. More viscous fluid shall be routed on the shell side to obtain better heat transfer.
4. It is customary to assign the higher pressure to the tube side to minimize shell thickness.
5. Consider various potential and possible upset conditions in assigning the design pressure and/or design temperature.
6. Pass arrangements on the shell side and tube side to obtain maximum heat transfer?
7. Have considered the tube size and thickness?
8. What is the acceptable pressure drop on the tube side and shell side? Is the sum of the pumping cost and the initial equipment cost minimized?
9. Have considered the maximum allowable pressure drop to obtain the maximum heat transfer/
10. Are the tubeside and shellside velocities are high enough for good heat transfer and to minimize fouling but well below the limits that can cause erosion-corrosion on the tubeside, and impingement attack and flow-induced vibration on the shell side?
11. Have considered the nozzle sizes and adequate shell escape area? Are the nozzle orientations consistent with tube layout pattern?
12. Is the baffle arrangement designed to promote good flow distribution on the shell side and hence good heat transfer, and to minimize fouling and flow-induced vibration?
13. Does the design provide for efficient expulsion of noncondensables that may degrade the performance?
14. Is the service corrosive or dirty? If so, have specified corrosion-resistant materials and reasonable fouling factors?
15. Does the design minimize fouling?

Design Considerations for a Shell and Tube Heat Exchanger

The basic criterion that a given or designed heat exchanger should satisfy is that it should perform the given heat duty within the allowable pressure drop. The design is also to satisfy additional criteria such as:

1. Withstand operating conditions, startup, shutdown, and upset conditions that influence the thermal and mechanical design.
2. Maintenance and servicing.
3. Multiple shell arrangement.
4. Cost.
5. Size limitations.

In terms of five factors just mentioned, multishell arrangement needs some comments on it. Consider the advantages of a multishell arrangement to allow one unit to be taken out of service for maintenance without severely upsetting the rest of the plant. For part load operation, multiple shells result into an economical operation. Shipping and handling may dictate restrictions on the overall size or weight of the unit, resulting in multiple shells for an application.

Thermal Design Procedure

The overall design procedure of a shell and tube heat exchanger is quite lengthy, and hence it is necessary to break down this procedure into distinct steps:

1. Approximate sizing of shell and tube heat exchanger.
2. Evaluation of geometric parameters also known as auxiliary calculations.
3. Correction factors for heat transfer and pressure drop.
4. Shell-side heat transfer coefficient and pressure drop.
5. Tube-side heat transfer coefficient and pressure drop.
6. Evaluation of the design, i.e., comparison of the results with the design specification.

The approximate design involves arriving at a tentative set of heat exchanger parameters, and if the design is accepted after rating then this becomes the final design. Various stages of approximate design include the following:

1. Compute overall heat transfer coefficient.
2. Compute heat transfer rate required.
3. Compute the heat transfer area required.
4. Design the geometry.

Basic logic structure for process heat exchanger design is shown as Figure E.4.1:

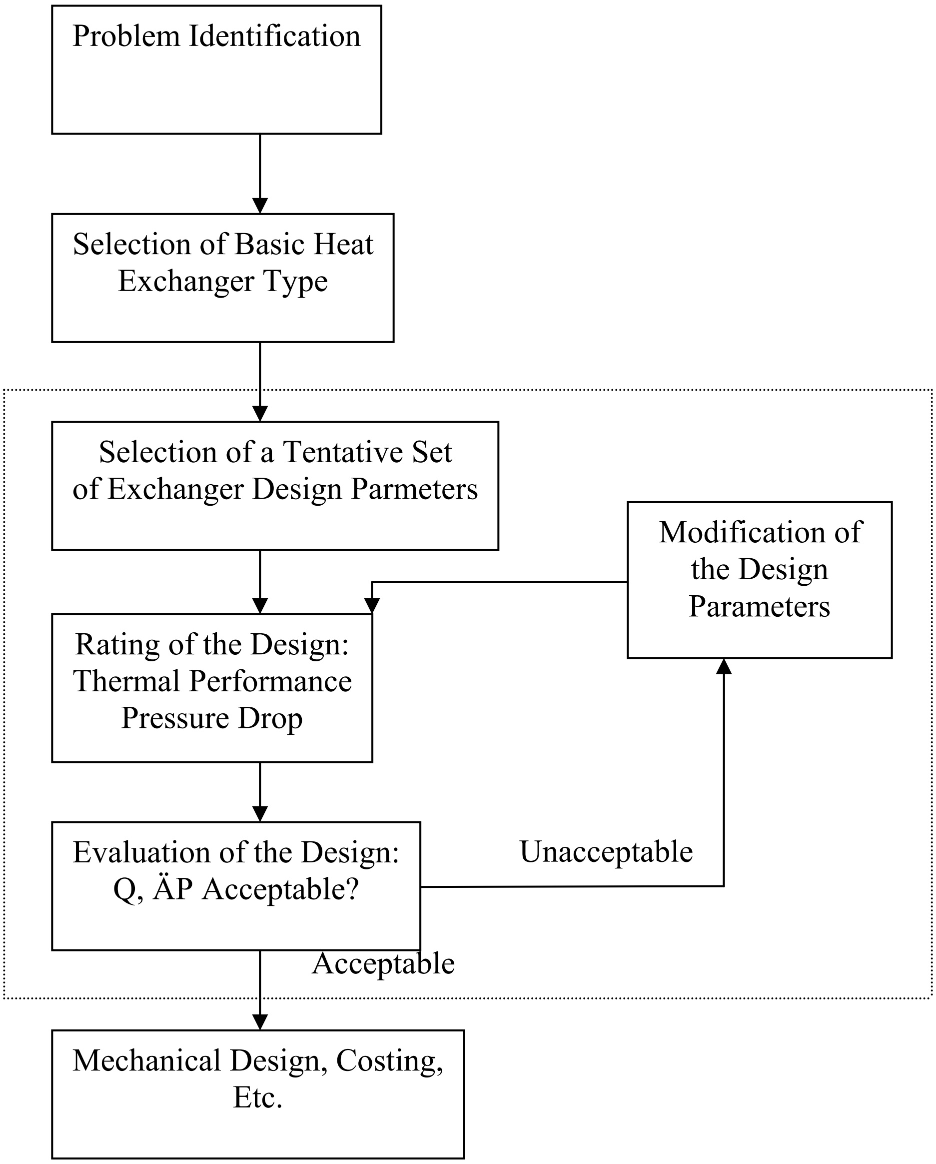


Figure E.4.1 *Basic logic structure for process heat exchanger design*

Example of Design

Here present the design of one shell and tube heat exchanger in dehydrogenation process. For the physical properties and conditions of the components in shell and tube is shown as Table E.4.1.

Table E.4.1. *Physical properties and conditions of components*

Shell- side Stream No. 201	Inlet	Mean	Outlet
Phase	Gas	Gas	Gas
Temperature ^{o}C	40	59	78
Specific heat kJ/kg ^{o}C	2.01	2.05	2.09
Thermal conductivity W/m ^{o}C	0.135	0.134	0.133
Density kg/m^3	840	820	800
Viscosity mNsm^{-2}	4.3	3.2	2.4
Tube-side Stream No. 202	Inlet	Mean	Outlet
Phase	Gas	Gas	Gas

Temperature ^{o}C	200	145	90
Specific heat kJ/kg ^{o}C	2.72	2.47	2.26
Thermal conductivity W/m ^{o}C	0.130	0.132	0.135
Density kg/m^3	690	730	770
Viscosity mNsm^{-2}	0.22	0.43	0.80

Estimation of Heat Load

The heat load is calculated in the general case from

$$\dot{m}_h c_{ph}(T_{hi} - T_{ho}) = \dot{m}_c c_{pc}(T_{co} - H_{ci}) = q \qquad \text{(E.4.1)}$$

the heat load also can be got from Aspen Plus:

q = 1509.4 kW

Estimation of Log Mean Temperature Difference

Determine the logarithmic mean temperature difference for countercurrent flow using the temperature as defined earlier:

$$q_T = UA_T \frac{\Delta T_2 - \Delta T_1}{\ln\left(\dfrac{\Delta T_2}{\Delta T_1}\right)} = UA_T \overline{\Delta T_L} \qquad \text{(E.4.2)}$$

where

$\Delta T_1 = T_{ho} - T_{ci}$

$\Delta T_2 = T_{hi} - T_{co}$

q_T = total heat transfer rate

A_T = total heat transfer area

$$\overline{\Delta T_L} = \frac{\Delta T_2 - \Delta T_1}{\ln\left(\dfrac{\Delta T_2}{\Delta T_1}\right)} = \frac{(200-78)-(90-40)}{\ln\left(\dfrac{200-78}{90-40}\right)} = 80.7°\text{C}$$

LMTD Correction Factor

Values of *F* can be found from the thermal relation charts. However, for estimation purpose, a reasonable estimate may often be obtained without restoring to the charts.

1. For a single tube pass, purely countercurrent heat exchanger, *F*=1.0
2. For a single shell with any even number of tube side passes, *F* should be between 0.8 and 1.0

The correction factor is a function of the shell and tube fluid temperatures, and the number of tube and shells passes. It is normally correlated as a function of two dimensionless temperature ratios:

$$R = \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} = \frac{200-90}{78-40} = 2.9$$

$$S = \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} = \frac{78-40}{200-40} = 2.4$$

R is equal to the shell-side fluid flow rate times the fluid mean specific heat; divided by the tube-side fluid flow rate times the tube-side fluid specific heat.

S is a measure of the temperature efficiency of the exchanger.

For a 1 shell 2 tube pass exchanger, the correction factor is given by:

$$F = \frac{\sqrt{\left(R^2+1\right)}\ln\left[\dfrac{1-S}{1-RS}\right]}{\left(R-1\right)\ln\left[\dfrac{2-S\left(R+1-\sqrt{R^2+1}\right)}{2-S\left(R+1+\sqrt{R^2+1}\right)}\right]} = 0.876$$

Method to Determine Number of Shells

Quickly check the limits:

$2T_{ho} \geq T_{ci} + T_{co}$ hot fluid on the shell side

$2T_{co} \leq T_{hi} + T_{ho}$ cold fluid on the shell side

If the limits are approached, it is necessary to use multiple 1-2*N* shells in series. There is a rapid graphical technique for estimating a sufficient number of 1-2*N* shells in series. The terminal temperatures of the two streams are plotted on the ordinates of an arithmetic graph paper sheet as shown as Figure E.4.2:

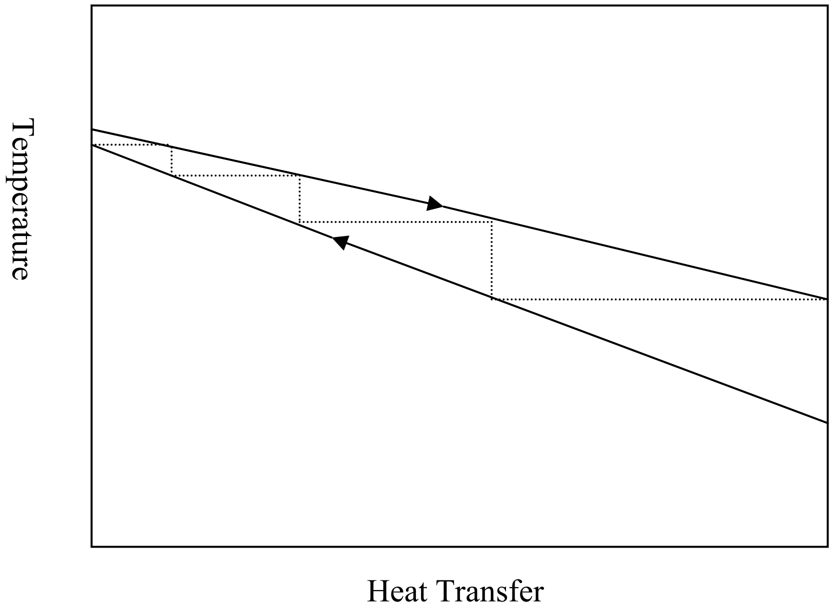


***Figure E.4.2** Heat transfer efficiency v.s. temperature*

The distance between the ordinates is arbitrary. Starting with the cold fluid outlet temperature, a horizontal line is laid off until it intercepts the hot fluid line. From that point a vertical line is draw to the cold line at or below the cold fluid inlet temperature. The number of horizontal lines (including the one that intersects the right-hand ordinate) is equal to the number of shells in series that is clearly sufficient to perform the duty. Following this procedure will usually result in a number of shells having an overall *F* close to 0.8.

Estimation of U

First, need to guess a number of *U*. The assumption can be chosen from Table E.4.2. [Coulson & Richardson's Volume 6]. Here start with 300 W/m^{2o}C

***Table E.4.2** Typical overall coefficients*

Shell and tube exchangers

Hot fluid	Cold fluid	U (W/m^2 °C)
Heat exchangers		
Water	Water	800–1500
Organic solvents	Organic solvents	100–300
Light oils	Light oils	100–400
Heavy oils	Heavy oils	50–300
Gases	Gases	10–50
Coolers		
Organic solvents	Water	250–750
Light oils	Water	350–900
Heavy oils	Water	60–300
Gases	Water	20–300
Organic solvents	Brine	150–500
Water	Brine	600–1200
Gases	Brine	15–250
Heaters		
Steam	Water	1500–4000
Steam	Organic solvents	500–1000
Steam	Light oils	300–900
Steam	Heavy oils	60–450
Steam	Gases	30–300
Dowtherm	Heavy oils	50–300
Dowtherm	Gases	20–200
Flue gases	Steam	30–100
Flue	Hydrocarbon vapours	30–100
Condensers		
Aqueous vapours	Water	1000–1500
Organic vapours	Water	700–1000
Organics (some non-condensables)	Water	500–700
Vacuum condensers	Water	200–500
Vaporisers		
Steam	Aqueous solutions	1000–1500
Steam	Light organics	900–1200
Steam	Heavy organics	600–900

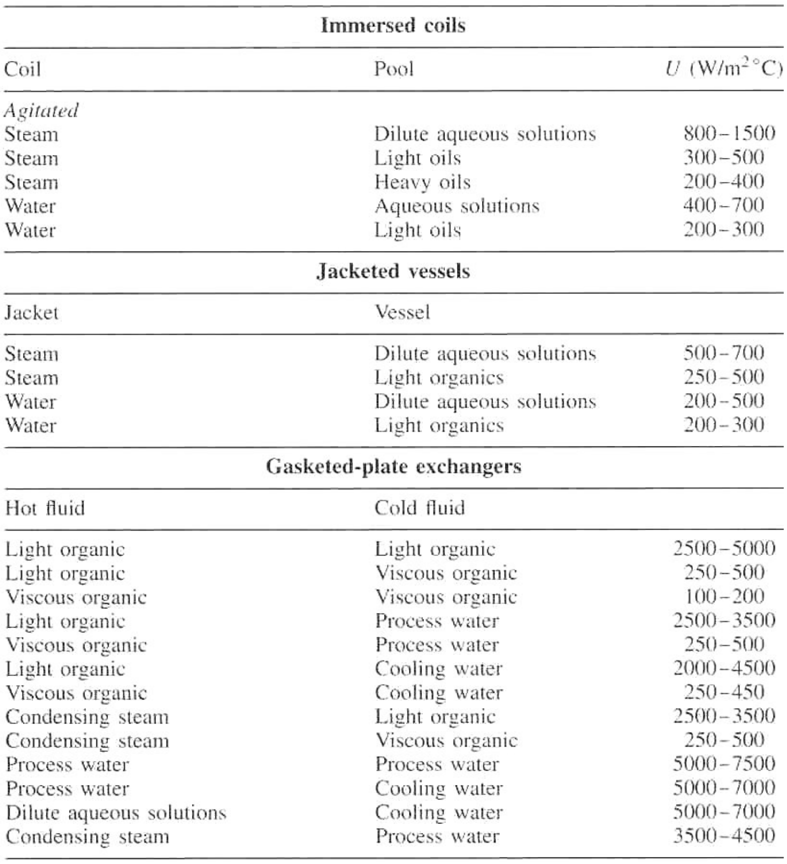
Air-cooled exchangers

Process fluid	
Water	300–450
Light organics	300–700
Heavy organics	50–150
Gases, 5–10 bar	50–100
10–30 bar	100–300
Condensing hydrocarbons	300–600

Immersed coils

Coil	Pool	
Natural circulation		
Steam	Dilute aqueous solutions	500–1000
Steam	Light oils	200–300
Steam	Heavy oils	70–150
Water	Aqueous solutions	200–500
Water	Light oils	100–150

(continued overleaf)

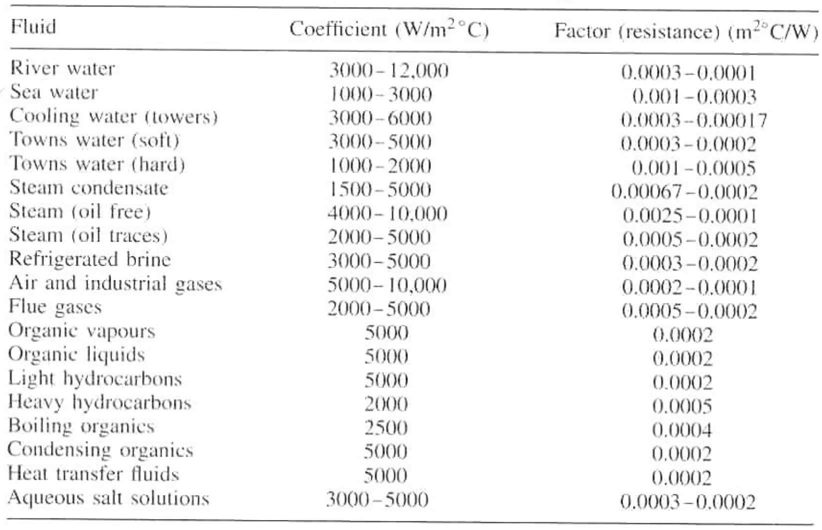


	Immersed coils	
Coil	Pool	U (W/m^2°C)
Agitated		
Steam	Dilute aqueous solutions	800–1500
Steam	Light oils	300–500
Steam	Heavy oils	200–400
Water	Aqueous solutions	400–700
Water	Light oils	200–300
	Jacketed vessels	
Jacket	Vessel	
Steam	Dilute aqueous solutions	500–700
Steam	Light organics	250–500
Water	Dilute aqueous solutions	200–500
Water	Light organics	200–300
	Gasketed-plate exchangers	
Hot fluid	Cold fluid	
Light organic	Light organic	2500–5000
Light organic	Viscous organic	250–500
Viscous organic	Viscous organic	100–200
Light organic	Process water	2500–3500
Viscous organic	Process water	250–500
Light organic	Cooling water	2000–4500
Viscous organic	Cooling water	250–450
Condensing steam	Light organic	2500–3500
Condensing steam	Viscous organic	250–500
Process water	Process water	5000–7500
Process water	Cooling water	5000–7000
Dilute aqueous solutions	Cooling water	5000–7000
Condensing steam	Process water	3500–4500

Fouling Resistance

Fouling resistance values may be chosen from TEMA Table E.4.3 [Coulson & Richardson's Volume 6].

***Table E.4.3** Fouling factors (coefficients), typical values*



Fluid	Coefficient (W/m^2°C)	Factor (resistance) (m^2°C/W)
River water	3000–12,000	0.0003–0.0001
Sea water	1000–3000	0.001–0.0003
Cooling water (towers)	3000–6000	0.0003–0.00017
Towns water (soft)	3000–5000	0.0003–0.0002
Towns water (hard)	1000–2000	0.001–0.0005
Steam condensate	1500–5000	0.00067–0.0002
Steam (oil free)	4000–10,000	0.0025–0.0001
Steam (oil traces)	2000–5000	0.0005–0.0002
Refrigerated brine	3000–5000	0.0003–0.0002
Air and industrial gases	5000–10,000	0.0002–0.0001
Flue gases	2000–5000	0.0005–0.0002
Organic vapours	5000	0.0002
Organic liquids	5000	0.0002
Light hydrocarbons	5000	0.0002
Heavy hydrocarbons	2000	0.0005
Boiling organics	2500	0.0004
Condensing organics	5000	0.0002
Heat transfer fluids	5000	0.0002
Aqueous salt solutions	3000–5000	0.0003–0.0002

Calculation of A_o

Once q, U, LMTD, and F are known, the total outside heat transfer area A_o can be calculated:

$$A_o = \frac{q}{UF\left(LMTD\right)} \qquad \text{(E.4.3)}$$

$$A_o = \frac{1509.4\times10^3}{300\times0.876\times80.7} = 71.15\ \text{m}^2$$

Determination of Shell Size and Tube Length from Heat Transfer Area, A_o

The objective is to find the right number of tubes of diameter, D_o, and the shell diameter, D_s, to accommodate the number of tubes, N_t, with given tube length, L.

$$A_o = \pi D_o N_t L \qquad \text{(E.4.4)}$$

The problem now arises of how to interpret the value of value A_o in terms of tube length and shell diameter, when both values are not known.

For estimation purpose, the tube number N_t is given by

$$N_t = 0.785\left(\frac{CTP}{CL}\right)\frac{D_s^2}{(PR)^2 D_o^2} \qquad \text{(E.4.5)}$$

where CL is the tube layout constant given by

$C_l = 0.87$ for $\theta_{\text{tp}} = 30^\circ$ and 60°

$C_l = 1.0$ for $\theta_{\text{tp}} = 45^\circ$ and 90°

CTP is the tube count calculation constant that accounts for the incomplete coverage of the shell diameter by tubes, due to necessary clearances between the shell and the outer tube circle and tube omissions due to tube pass lanes for multitude pass design.

one-tube pass: $CTP = 0.93$

two-tube pass: $CTP = 0.9$

three-tube pass: $CTP = 0.85$

PR is tube pitch ratio ($=P_t/D_o$). Usually, PR is 1.25.

Substituting equation (26) into (25), the result is given by

$$A_o = 0.785\pi\left(\frac{CTP}{CL}\right)\frac{D_s^2 L}{(PR)^2 D_o} \qquad \text{(E.4.6)}$$

From equation (27), an expression for the shell diameter in terms of main constructional diameters can be obtained as:

$$D_s = 0.637\sqrt{\frac{CL}{CTP}}\left[\frac{A_o (PR)^2 D_o}{L}\right]^{1/2} \qquad \text{(E.4.7)}$$

Heat transfer surface A_o can be obtained by various combinations of parameter L and D_o for any given tube layout pattern. An initial assumption of these values is necessary.

Here use 19.05 mm outside diameter, 14.83 mm inside diameter, and 5 m long tubes on a square pitch arrangement.

Area of one tube:

$$A_1 = \pi \times 0.01905 \times 5 = 0.2992 \text{ m}^2$$

Number of tubes = 71.15/0.2992 = 237, say 240

So, for 2 passes, tubes per pass, N_t = 120

$$D_s = 0.637\sqrt{\frac{CL}{CTP}}\left[\frac{A_o (PR)^2 D_o}{L}\right]^{1/2} = 0.637 \times \sqrt{\frac{1}{0.9}} \times \left[\frac{71.15 \times (1.25)^2 \times 0.01905}{5}\right] = 0.437 \text{ m}$$

Tube-Side Heat Transfer Coefficient

A general equation that can be used for exchanger design is:

$$Nu = C\,\text{Re}^{0.8}\,\text{Pr}^{0.33}\left(\frac{\mu}{\mu_w}\right)^{0.14} \qquad \text{(E.4.8)}$$

where

Nu = Nusselt number $= \dfrac{h_i d_e}{k_f}$

Re = Reynolds number $= \dfrac{\rho u_t d_e}{\mu}$

Pr = Prandtl number $= \dfrac{C_p \mu}{k_f}$

$C = 0.021$ for gases
$C = 0.023$ for non-viscous liquids
$C = 0.027$ for viscous liquids

and

h_i = inside cofficient, W/m°C,
d_e = equivalent diameter, m = D_i
u_t = fluid velocity, m/s,
k_f = fluid thermal conductivity, W/m°C,
G_t = mass velocity, mass flow per unit area, kg/m^{2}s,
μ = fluid viscosity at the bulk fluid temperature, Ns/m^2,
μ_w = fluid viscosity at the wall,
C_p = fluid specific heat, heat capacity, J/kg°C.

Calculation of u_t:

tube cross-sectional area $= \dfrac{\pi}{4}(0.01483)^2 = 0.0001727 \text{ m}^2$

so area per pass $= 120 \times 0.0001727 = 0.02073 \text{ m}^2$

volumetric flow rate $= \dfrac{19.44}{820} = 0.0237 \ {}^{\text{m}^3}\!/_{\text{s}}$

tube-side velocity, $u_t = \dfrac{0.0237}{0.02073} = 1.14 \ {}^{\text{m}}\!/_{\text{s}}$

It is often convenient to correlate heat transfer data in terms of a heat transfer j factor. The heat transfer factor is defined as:

$$j_h = St\,\mathrm{Pr}^{0.67}\left(\frac{\mu}{\mu_w}\right)^{-0.14} \qquad \text{(E.4.9)}$$

St = Stanton number $= 0.0225\exp\left[-0.0225(\ln \mathrm{Pr})^2\right]\mathrm{Re}^{-0.205}\,\mathrm{Pr}^{-0.505}$

eq. E.4.8 can be rearranged as:

$$\frac{h_i D_i}{k_f} = j_h\,\mathrm{Re}\,\mathrm{Pr}^{0.33}\left(\frac{\mu}{\mu_w}\right)^{0.14} \qquad \text{(E.4.10)}$$

After having Re and L/D_i, from Figure E.4.3 [Coulson & Richardson's volume 6], it can get the j_h value. By eq. E.4.10, h_i can be calculated.

Calculation of h_i:

$$\mathrm{Re} = \frac{820 \times 1.15 \times 0.01483}{3.2 \times 10^{-3}} = 4388$$

$$\mathrm{Pr} = \frac{2.996 \times 10^{3} \times 3.2 \times 10^{-3}}{0.134} = 71.57$$

$$\frac{L}{D_i} = \frac{5}{0.01483} = 337$$

From Figure E.4.3,

$$j_h = 3.3 \times 10^{-3}$$

$$N_u = 3.3 \times 10^{-3} \left(4388\right)\left(71.57\right)^{0.33} = 59.27$$

$$h_i = 59.27 \times \left(\frac{0.134}{0.01483}\right) = 536 \ \mathrm{W}/\mathrm{m}^{2}\,{}^{\mathrm{o}}\mathrm{C}$$

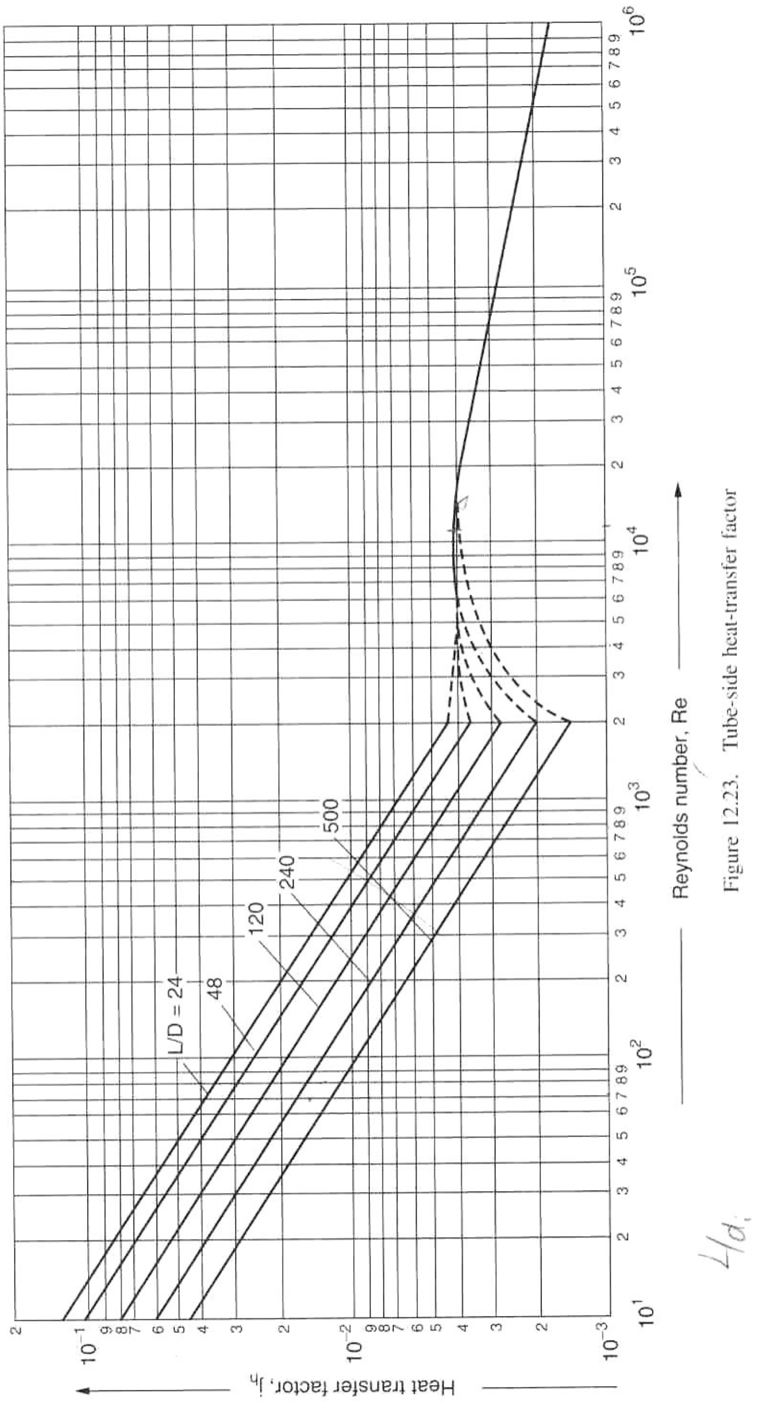


Figure 12.23. Tube-side heat-transfer factor

FigureE.4.3 *Tube side heat transfer factor*

Shell-Side Heat Transfer Coefficient

Calculate the area for cross-flow A_s for the hypothetical row of tubes at the shell equator, given by:

$$A_s = \frac{\left(P_t - D_o\right) D_s l_B}{P_t} \qquad \text{(E.4.11)}$$

where

P_t = tube pitch

D_o = tube outside diameter

D_s = shell inside diameter

l_B = baffle spacing

The term $(P_t-D_o)/P_t$ is the ratio of the clearance between tubes and the total distance between tube centers. The baffle spacing used arrange from 0.2 to 1.0 shell diameter. A close baffle spacing will give higher heat transfer coefficient but at the expense of higher pressure drop. The optimum spacing will usually be between 0.3 to 0.5 times the shell diameter. The baffle cut is the height of the segment removed to form the baffle, expressed as a percentage of the baffle disc diameter. Generally, a baffle cut of 20 to 25% will be the optimum, giving good heat transfer rate, without excessive pressure drop.

$$A_s = \frac{(1.25\times 0.01905 - 0.01905)}{1.25\times 0.01905}\times 0.437\times \frac{0.437}{5} = 0.009548 \text{ m}^2$$

Calculate the shell-side mass velocity G_s and the linear velocity u_s:

$$G_s = \frac{W_s}{A_s}$$

$$u_s = \frac{G_s}{\rho}$$

where

W_s = fluid flow rate on the shell-side, kg/s

ρ = shell-side fluid density, kg/m^3

Calculate the shell-side equivalent diameter (hydraulic diameter). For a square pitch arrangement:

$$d_e = \frac{1.27}{D_o}\left(P_t^2 - 0.785D_o^2\right) \qquad \text{(E.4.12)}$$

For an equilateral triangular pitch arrangement:

$$d_e = \frac{1.10}{D_o}\left(P_t^2 - 0.917D_o^2\right) \qquad \text{(E.4.13)}$$

Calculate the shell-side Reynolds number, given by:

$$\text{Re} = \frac{G_s d_e}{\mu} = \frac{u_s d_e \rho}{\mu} \qquad \text{(E.4.14)}$$

For the calculated Reynolds number, read the value of j_h from Figure E.4.4 [Coulson & Richardson's Volume 6] for the selected baffle cut and tube arrangement, and calculate the shell-side heat transfer coefficient h_o from:

$$\frac{h_o D_i}{k_f} = j_h \,\text{Re}\,\text{Pr}^{0.33}\left(\frac{\mu}{\mu_w}\right)^{0.14} \qquad \text{(E.4.15)}$$

Calculation of h_o:

$$d_e = \frac{1.27}{D_o}\left(P_t^2 - 0.785D_o^2\right) = \frac{1.27}{0.01905}\left(\left(1.25\times0.01905\right)^2 - 0.785\left(0.01905\right)_o^2\right) = 0.01881\text{ m}$$

$$\text{volumetric flow rate} = \frac{5.556}{730} = 0.00761\ {}^{\text{m}^3}\!/_{\text{s}}$$

$$\text{shell-side velocity} = \frac{0.0761}{0.009548} = 0.797\ {}^{\text{m}}\!/_{\text{s}}$$ From

$$\text{Re} = \frac{730\times0.797\times0.01881}{0.43\times10^{-3}} = 25453$$

$$\text{Pr} = \frac{3.515\times10^3\times0.43\times10^{-3}}{0.132} = 11.45$$

Figure E.4.4,

$$j_h = 6\times10^{-3}$$

$$h_o = \left(\frac{0.132}{0.0188}\right)\times6\times10^{-3}\times25453\times\left(11.45\right)^{0.33} = 2396\ {}^{\text{W}}\!/_{\text{m}^2\,{}^\circ\text{C}}$$

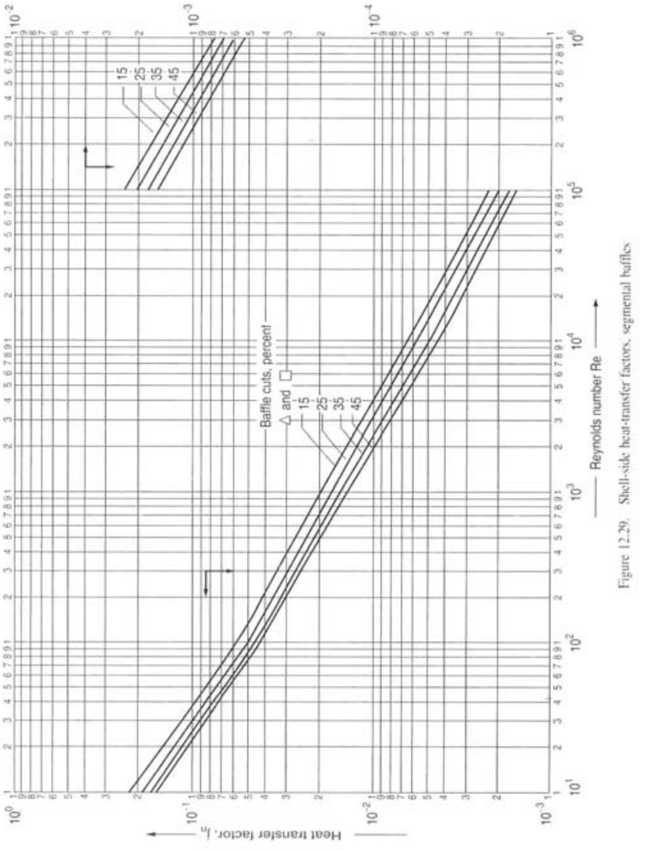


Figure E.4.4 *Shell side heat transfer factors, segmental baffles*

Check the Estimated U_o

The greatest uncertainty in preliminary calculations is estimating the overall heat transfer coefficient. U_o can be calculated from individual values of heat transfer coefficient on shell side and tube side, wall resistance, and fouling resistance:

$$U_o = \frac{1}{\frac{1}{h_{di}}\left(\frac{D_o}{D_i}\right) + \frac{1}{h_i}\left(\frac{D_o}{D_i}\right) + \frac{x_w}{k_m}\left(\frac{D_o}{\overline{D_L}}\right) + \frac{1}{h_o} + \frac{1}{h_{do}}} \qquad \text{(E.4.16)}$$

can be modified as:

$$U_o = \frac{1}{\frac{1}{h_{di}}\left(\frac{D_o}{D_i}\right) + \frac{1}{h_i}\left(\frac{D_o}{D_i}\right) + \frac{D_o}{2k_m}\ln\left(\frac{D_o}{D_i}\right) + \frac{1}{h_o} + \frac{1}{h_{do}}}$$

$$= \frac{1}{\frac{1}{0.0002}\left(\frac{0.01905}{0.01483}\right) + \frac{1}{536}\left(\frac{0.01905}{0.01483}\right) + \frac{0.01905}{2\times 45}\ln\left(\frac{0.01905}{0.01483}\right) + \frac{1}{2396} + \frac{1}{0.0002}}$$

$$= 300.6921 \ \mathrm{W}/\mathrm{m^2\,{}^oC}$$

Check the calculation:

If $0 < \frac{U_{o,calc} - U_{o,ass}}{U_{o,ass}} < 30\%$, the initial assumption of U_o is acceptable.

$$\frac{U_{o,calc} - U_{o,ass}}{U_{o,ass}} \times 100\% = \frac{300.6921 - 300}{300} = 0.23\%$$

Therefore, the initial assumption of U_o is acceptable.

Appendix E.5 T302 Column sizing report

Table E.5.1 *T302 column sizing report*

Item	**Sizing Data**		**Comments**
Parameter			
M_G	10.9	kg/s	Gas phase flow rate
Rho_G	21.7	kg/m3	Gas phase density
M_L	22.05	kg/s	Liquid phase flow rate
Rho_L	426	kg/m3	Liquid phase density
Visc_L	1.00E-04	Pa s	Liquid phase dynamic viscoscity
Capacity Factor Correlation			
F_{LG}	0.455	[-]	$(M_L /M_G)(\text{Rho}_G/\text{Rho}_L)^{0.5}$
Design Pressure drop	4.1	mbar/m	distillation range 4-8 mbar/m
C_{rp}	0.65	[-]	Read from the Figure 8.2.3.1
F_{rp}	66	1/m	use dia.50 mm pall ring in order to saving cost
$F_{G,oper}$	1.1874	[-]	$[\text{Crp*(RhoL-RhoG)/(13*Frp(viscL/RhoL)}^{0.1}]^{0.5}$
Dimension of Column			
Dia	1.586	m	$[\text{4*MG/(pi*FGOper*Rho}^{0.5})]$
HETP	0.85	m	This valid for Rasching and pall ring in distillation with moderate value of surface tension . low viscoscity and pressure drop below 6 mbar/m
No. of ideal stages	24	[-]	
Height of packing	20.4	m	
Height of top	1	m	
Height of bottom	1	m	
Height of skirt	1.5	m	
Height of column	23.9	m	

Appendix E.6 Calculation of CO_2 removal equipment

T401 *CO_2 absorber*
From CO_2 solubility data are given as blow[2]:

Table E.6.1 *Solubility of CO2 in 50 mass % MDEA*

CO_2 loading (mol CO_2/mol MDEA*)	(per cent w/w solution**)	P_{CO2} (kPa)		
		25°C	50°C	75°C
0.2547	0.0470	8.27		
0.2988	0.0551	10.34		
0.4923	0.0909	19.72		
0.0150	0.0028		0.78	
0.0442	0.0082		2.47	
0.0740	0.0137		4.87	
0.1315	0.0243		11.67	
0.1916	0.0354		17.36	
0.2420	0.0447		24.46	
0.3190	0.0589		38.75	
0.3854	0.0711		53.04	
0.4529	0.0836		70.92	
0.4884	0.0901		76.19	
0.0162	0.0030			3.62
0.0334	0.0062			7.92
0.0420	0.0078			9.72
0.0537	0.0099			13.72
0.0770	0.0142			21.31
0.1010	0.0186			31.11
0.1330	0.0245			45.39
0.1656	0.0306			61.88
0.1946	0.0359			78.87

* Operation at atmospheric pressure.
** The solution is 50 mass% MDEA which is dissolved in water.

Use data from Table E.6.1, Figure E.6.1 is plotted.

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[2] http://www.cape.canterbury.ac.nz/Apcche_Proceedings/ APCChE/Data/802rev.pdf

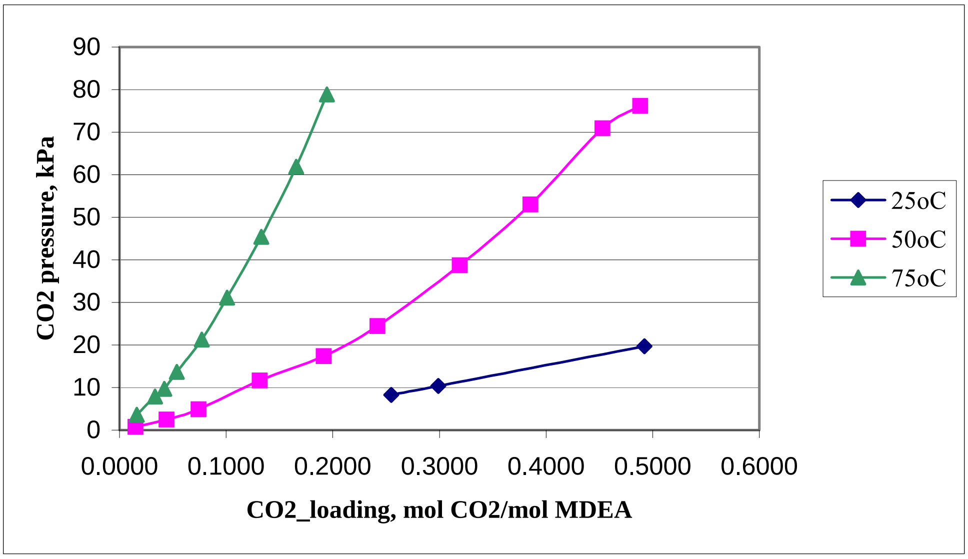


Figure E.6.1 *Solubility of CO2 in 50 mass % MDEA solution*

Due to the volatility and related high concentration of MDEA and followed low temperature separation process, 25°C absorption temperature is designed.

Partial pressure of CO_2 in the feed = 1.5%*101.325 = 1.52[kPa]

Partial pressure of CO_2 in the exit gas at 99 percent recovery =1.52*0.01= 0.0152 [kPa]

Assume the equilibrium line at 25°C in Figure E.6.1 is straight line so Figure E.6.2 [volume 6 p597] can be used to estimate the number of stages needed.

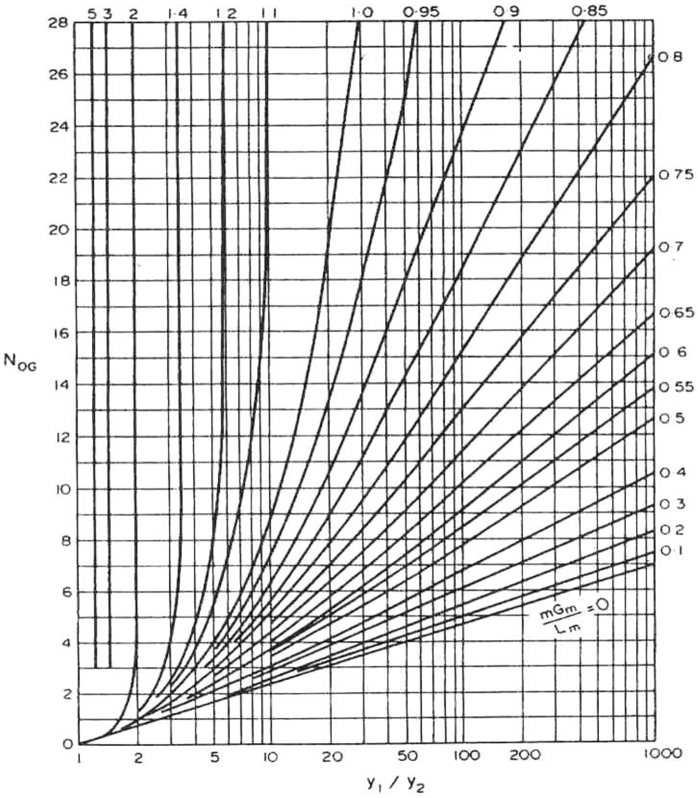


Figure E.6.2 *Number of transfer units* N_{OG} *as a function of y1/y2 with* mG_m / L_m *as parameter*

From the data above: partial pressure at 0.2988 [mol CO2/mol MDEA] = 10.34 [kPa].

Mol. fraction in vapor $=\dfrac{10.34}{101.325}=0.1020$

Mol. fraction in liquid $=\dfrac{0.2988}{0.2988+1+1*119.2/18}=0.0377$

The slope of operation line $m=\dfrac{0.1020}{0.0377}=2.71$

Using Figure E.6.2 the number of stages required at different MDEA solution flow rates will be determined and the "optimum" rate chosen:

$$\frac{y_1}{y_2}=\frac{p_1}{p_2}=\frac{14.54}{0.1454}=100$$

where y_1 and y_2 = the mol fractions of the solute in the gas at the bottom and top of the column, respectively

$m\dfrac{G_m}{L_m}$	0.1	0.2	0.3	0.4	0.5	0.55	0.6	0.65	0.7	0.75	0.8	0.85	0.9
N_{OG}	5.0	5.4	6.1	6.7	7.8	8.4	9.2	10.2	11.3	13.0	15.1	18.3	23.6

where G_m = molar gas flow-rate per unit cross-sectional area
L_m = molar liquid flow-rate per unit cross-sectional area
N_{OG} = the number of overall gas-phase transfer units

Below 0.55 there is only a small decrease in the number of stages required with increasing liquid rate; above 0.7 the number of stages increases rapidly with decreasing liquid rate. It can be seen that the "optimum" will be between mG_m/L_m=0.55 to 0.7, as would be expected.

Check the liquid outlet composition at 0.55 and 0.7:

Material balance $L_m x_1 = G_m(y_1 - y_2)$

So $x_1=\dfrac{G_m}{L_m}(1.5\%*0.99)=\dfrac{m}{2.71}\dfrac{G_m}{L_m}(0.0148)$

at $m\dfrac{G_m}{L_m}=0.55,\ x_1=0.003$ mol fraction

at $m\dfrac{G_m}{L_m}=0.7,\ x_1=0.0038$ mol fraction

Use 0.7, as the higher concentration will favor the stripper design and operation, without significantly increasing the number of stages needed in the absorber.

$N_{OG}=11.3$, =12 (say)

Column diameter

From the data of mass balance,

Gas flow-rate (stream 209) = 7.64kg/s, = 0.263kmol/s

Carbon dioxide mole fraction in gas inlet is 0.015.

Liquid flow-rate = 2.71/0.7*0.263 = 1.02kmol/s, = 31.91kg/s

Select 50mm metal Pall rings.

From Table E.6.2 [volume 6], F_p = 66m^{-1}

Gas density ρ_V = 27498.122/667.676 = 41.18kg/m^3

***Table E.6.2** Design data for various packings*

	Size		Bulk density (kg/m^3)	Surface area a (m^2/m^3)	Packing factor F_pm^{-1}
	in.	mm			
Raschig rings ceramic	0.50	13	881	368	2100
	1.0	25	673	190	525
	1.5	38	689	128	310
	2.0	51	651	95	210
	3.0	76	561	69	120
Metal (density for carbon steel)	0.5	13	1201	417	980
	1.0	25	625	207	375
	1.5	38	785	141	270
	2.0	51	593	102	190
	3.0	76	400	72	105
Pall rings metal (density for carbon steel)	0.625	16	593	341	230
	1.0	25	481	210	160
	1.25	32	385	128	92
	2.0	51	353	102	66
	3.5	76	273	66	52
Plastics (density for polypropylene)	0.625	16	112	341	320
	1.0	25	88	207	170
	1.5	38	76	128	130
	2.0	51	68	102	82
	3.5	89	64	85	52
Intalox saddles ceramic	0.5	13	737	480	660
	1.0	25	673	253	300
	1.5	38	625	194	170
	2.0	51	609	108	130
	3.0	76	577		72

Assume: Liquid density ρ_L = 1000kg/m^3

Liquid viscosity μ_L = 10^{-3}Ns/m^2

$$F_{LV} = \frac{L_W^*}{V_W^*}\sqrt{\frac{\rho_V}{\rho_L}} = \frac{69.97}{7.64}\sqrt{\frac{41.18}{1000}} = 1.86$$

where L_W^* = liquid mass flow-rate per unit column cross-sectional area, kg/m^{2}s

V_W^* = gas mass flow-rate per unit column cross-section area, kg/m^{2}s

F_{LV} = flow factor

Design for a pressure drop of 20mm H_2O/m packing

From Figure E.6.3 [volume 6]

$$K_4 = 0.22$$

At flooding, $K_4 = 0.42$

Here K_4 is the function [volume 6]:

$$K_4 = \frac{13.1(V_W^*)^2 F_P\left(\mu_L/\rho_L\right)^{0.1}}{\rho_V(\rho_L - \rho_V)}$$

where F_P = packing factor, characteristic of the size and type of packing, m^{-1}

μ_L = liquid viscosity, Ns/m^2

ρ_L, ρ_V = liquid and vapor densities, kg/m^3

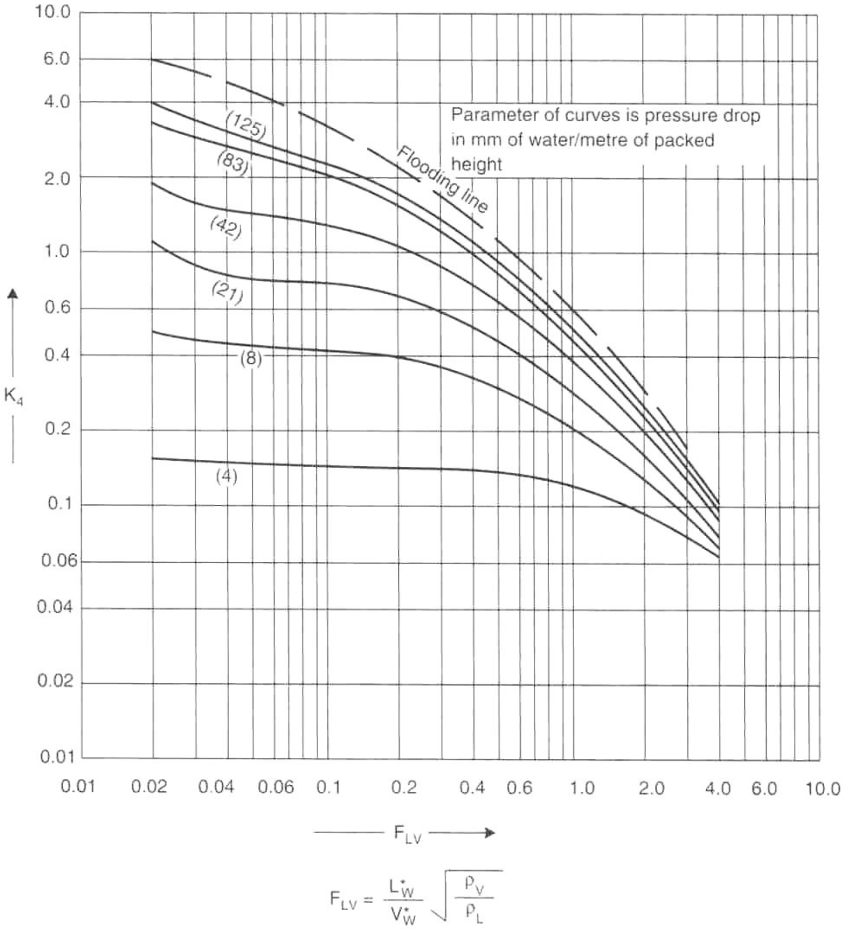


Figure E.6.3*Generalized pressure drop correlation, adapt from a figure by Norton Co. with permission*

Percentage flooding = $\sqrt{\frac{0.22}{0.42}} = 0.72$, satisfactory.

From equation above

$$V_W^* = \left[\frac{K_4\rho_V(\rho_L-\rho_V)}{13.1F_P\left(\mu_L/\rho_L\right)^{0.1}}\right]^{0.5}$$

$$= \left[\frac{0.22*41.18(1000-41.18)}{13.1*66\left(10^{-3}/1000\right)^{0.1}}\right]^{0.5}$$

$$= 6.32\text{kg/m}^2\text{s}$$

Column area required = 7.64/6.32 = 1.21m^2

Diameter = $\sqrt{\frac{4}{\pi}*1.21} = 1.24\text{m}$

Round off to 1.30m

Column area = $\frac{\pi}{4}*1.30^2 = 1.33\text{m}^2$

Column height

Onda’s method

The equation for the effective area is:

$$\frac{a_w}{a} = 1 - \exp\left[-1.45\left(\frac{\sigma_C}{\sigma_L}\right)^{0.75}\left(\frac{L_W^*}{a\mu_L}\right)^{0.1}\left(\frac{L_W^{*\,2}a}{\rho_L{}^2 g}\right)^{-0.05}\left(\frac{L_W^{*\,2}}{\rho_L\sigma_L a}\right)^{0.2}\right]$$

and for the mass coefficients:

$$k_L\left(\frac{\rho_L}{\mu_L g}\right)^{1/3} = 0.0051\left(\frac{L_W^*}{a_w\mu_L}\right)^{2/3}\left(\frac{\mu_L}{\rho_L D_L}\right)^{-1/2}(ad_P)^{0.4}$$

$$\frac{k_G}{a}\frac{RT}{D_v} = K_5\left(\frac{V_W^*}{a\mu_V}\right)^{0.7}\left(\frac{\mu_V}{\rho_V D_v}\right)^{1/3}(ad_P)^{-2.0}$$

where K_5 = 5.23 for packing sizes above 15mm, and 2.00 for sized below 15mm
a_w = effective interfacial area of packing per unit volume, m^2/m^3
a = actual area of packing per unit volume m^2/m^3
d_P = packing size, m
σ_C = critical surface tension for the particular packing material, N/m [data see volume 6]
σ_L = liquid surface tension, N/m
k_G = gas film mass transfer coefficient, $kmol/m^2s$ atm or $kmol/m^2s$ bar
k_L = Liquid film mass transfer coefficient, $kmol/m^2s\ (kmol/m^3)$=m/s
R = gas constant, bar $m^3/kmol$ K
T = operation temperature, K
g = gravity, $9.81m/s^2$
D_v, D_l= Mass diffusivity in vapor or liquid phase, m^2/s

Gas and liquid diffusivities can be calculated using given equation as below [volume 6, chapter 8, p291]

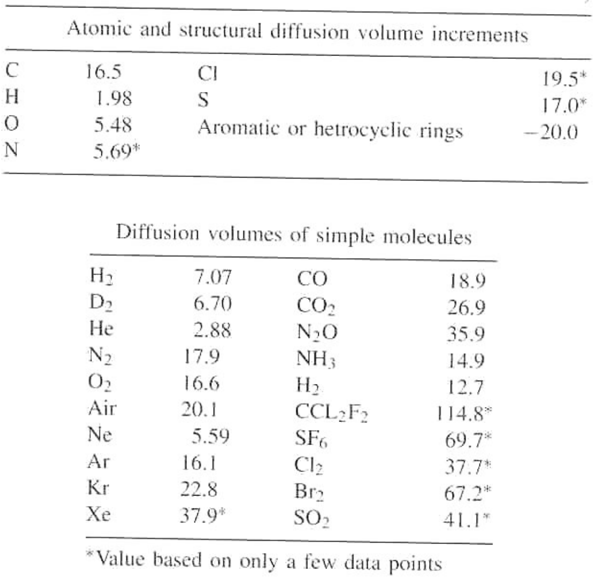
$$D_v = \frac{1.013e\text{-}7T^{1.75}(1/M_a + 1/M_b)^{1/2}}{P\left[\left(\sum_a v_i\right)^{1/3} + \left(\sum_b v_i\right)^{1/3}\right]^2}$$

where M_a,M_b = molecular weights of components a and b
$\sum_a v_i$, $\sum_b v_i$ = the summation of the special diffusion volume coefficients for components a and b, given in Table E.6.3 [volume 6]

$$D_L = \frac{1.173e\text{-}13(\phi M)^{0.5}T}{\mu V_M{}^{0.6}}$$

where ϕ = an association factor for the solvent
2.6 for water
1.9 for methanol
1.5 for ethanol
1.0 for unassociated solvents
M = Molecular weight of solvent
μ = viscosity of solvent, mNs/m^2
V_m = molar volume of the solute at its boiling point, $m^3/kmol$. This can be estimated from the group contributions given in Table E.6.4 [volume 6]

Table E.6.3 *Special atomic diffusion volumes*

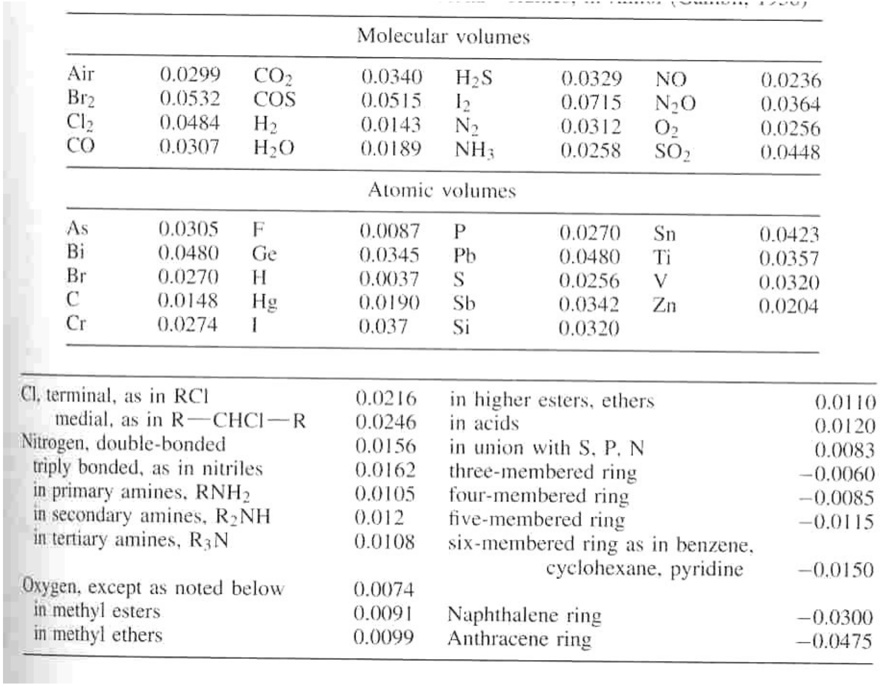


Atomic and structural diffusion volume increments			
C	16.5	Cl	19.5*
H	1.98	S	17.0*
O	5.48	Aromatic or hetrocyclic rings	−20.0
N	5.69*		

Diffusion volumes of simple molecules			
H_2	7.07	CO	18.9
D_2	6.70	CO_2	26.9
He	2.88	N_2O	35.9
N_2	17.9	NH_3	14.9
O_2	16.6	H_2	12.7
Air	20.1	CCL_2F_2	114.8*
Ne	5.59	SF_6	69.7*
Ar	16.1	Cl_2	37.7*
Kr	22.8	Br_2	67.2*
Xe	37.9*	SO_2	41.1*

*Value based on only a few data points

Table E.6.4 *structural contributions to molar volumes, m^3/mol*



Molecular volumes							
Air	0.0299	CO_2	0.0340	H_2S	0.0329	NO	0.0236
Br_2	0.0532	COS	0.0515	I_2	0.0715	N_2O	0.0364
Cl_2	0.0484	H_2	0.0143	N_2	0.0312	O_2	0.0256
CO	0.0307	H_2O	0.0189	NH_3	0.0258	SO_2	0.0448

Atomic volumes							
As	0.0305	F	0.0087	P	0.0270	Sn	0.0423
Bi	0.0480	Ge	0.0345	Pb	0.0480	Ti	0.0357
Br	0.0270	H	0.0037	S	0.0256	V	0.0320
C	0.0148	Hg	0.0190	Sb	0.0342	Zn	0.0204
Cr	0.0274	I	0.037	Si	0.0320		

Cl, terminal, as in RCl	0.0216	in higher esters, ethers	0.0110
medial, as in R—CHCl—R	0.0246	in acids	0.0120
Nitrogen, double-bonded	0.0156	in union with S, P, N	0.0083
triply bonded, as in nitriles	0.0162	three-membered ring	−0.0060
in primary amines, RNH_2	0.0105	four-membered ring	−0.0085
in secondary amines, R_2NH	0.012	five-membered ring	−0.0115
in tertiary amines, R_3N	0.0108	six-membered ring as in benzene, cyclohexane, pyridine	−0.0150
Oxygen, except as noted below	0.0074		
in methyl esters	0.0091	Naphthalene ring	−0.0300
in methyl ethers	0.0099	Anthracene ring	−0.0475

$$D_v = \frac{1.013e\text{-}7 * 303.13^{1.75}(1/44 + 1/29)^{1/2}}{30\left[(26.9)^{1/3} + (20.1)^{1/3}\right]^2}$$

$$= 5.45e\text{-}7\ \mathrm{m^2/s}$$

$$D_L = \frac{1.173e\text{-}13 * (1.0 * 31.28)^{0.5} * 303.13}{1.002 * 0.0340^{0.6}}$$

$$= 1.51e\text{-}9\mathrm{m^2/s}$$

The film transfer unit heights are given by:

$$H_G = \frac{G_m}{k_G a_w P}$$

$$H_L = \frac{L_m}{k_L a_w C_t}$$

where P = column operating pressure, bar

$Ct =$ total concentration, kmol/m^3

$H_L, H_G =$ overall height of a transfer unit and the individual film transfer units

Assume the flow of gas and liquid is essentially constant throughout the column, the height of packing required, Z, is given by:

$$Z = \frac{G_m}{K_G a C_t} \int_{y2}^{y1} \frac{dy}{y - y_e}$$

$$Z = H_{OG} N_{OG}$$

$$H_{OG} = H_G + m \frac{G_m}{L_m} H_L$$

where $Z =$ height of packing required, m

$y_e =$ concentration in the gas that would be in equilibrium with the liquid concentration at any point

$H_{OG} =$ height of an overall gas-phase transfer unit

$$L_W^* = 31.91/1.33 = 23.99\text{kg/m}^2\text{s}$$

$$R = 0.08314\text{bar m}^3\text{/kmol K}$$

Assume σ_L equals water $= 70\text{e-}3\text{N/m}$

$$g = 9.81\text{m/s}^2$$

$$d_P = 50\text{e-}3\text{m}$$

From Table E.6.2, for 50mm Pall rings

$$a = 102\text{m}^2\text{/m}^3$$

For metal materials,

$$\sigma_C = 75\text{e-}3\text{N/m}$$

$$\frac{a_w}{a} = 1 - \exp\left[-1.45\left(\frac{75e\text{-}3}{70e\text{-}3}\right)^{0.75}\left(\frac{23.99}{102*1e\text{-}3}\right)^{0.1}\left(\frac{23.99^2*102}{1000^2*9.81}\right)^{-0.05}\left(\frac{23.99^2}{1000*70e\text{-}3*102}\right)^{0.2}\right]$$

$$= 0.87$$

$$a_w = 0.87*102 = 88.74\text{m}^2\text{/m}^3$$

$$k_L\left(\frac{1000}{10^{-3}*9.81}\right)^{1/3} = 0.0051\left(\frac{52.61}{88.74*10^{-3}}\right)^{2/3}\left(\frac{10^{-3}}{1000*1.51e\text{-}9}\right)^{-1/2}(102*50e\text{-}3)^{0.4}$$

$$k_L = 5.88\text{e-}4\text{m/s}$$

V_W^* on actual column diameter = 7.64/1.33 = 5.74kg/m^{2}s

$$\frac{k_G}{102}\frac{0.08314*303.15}{5.45e\text{-}7} = 5.23\left(\frac{5.74}{102*0.018e\text{-}3}\right)^{0.7}\left(\frac{0.018e\text{-}3}{41.18*5.45e\text{-}7}\right)^{1/3}(102*50e\text{-}3)^{-2.0}$$

$$k_G = 1.15e\text{-}4\text{kmol/sm}^2\text{bar}$$

$$G_m = \frac{5.74}{27498.122/947.059} = 0.198\text{kmol/m}^2\text{s}$$

$$L_m = \frac{52.61}{0.1312*119.2 + (1 - 0.1312)*18} = 1.68\text{kmol/m}^2\text{s}$$

$$H_G = \frac{0.198}{1.15e\text{-}4*88.74*30} = 0.65\text{m}$$

$$H_L = \frac{1.68}{5.46\text{e-}4*88.74*(1000/31.28)} = 1.08\text{m}$$

$$H_{OG} = 0.65 + 0.7*1.08 = 1.4\text{m}$$

$$Z = 1.4*12 = 16.8\text{m}$$

Round up packed bed height to 17m.

Then the column volume is deduced: $V = \pi \frac{1.3^2}{4}$*17=22.6m^3

After the calculation:
Column diameter: 1.30m
Column height: 17m
Column volume: 22.6m^3

Appendix E.7 Gas-liquid separators calculation

1. *D-101 Gas sep. Drum*

From stream tables: at 30bar, 55°C, liquid density 435.1kg/m^3, and vapor density 48.7kg/m^3.

$u_t = 0.07\left[\left(435.1-48.7\right)/48.7\right]^{1/2} = 0.1972$ m/s

Vapor volumetric flow-rate is 547.3m^3/hr.

Take $h_v = D_v/2$ and $L_v/D_v = 4$

Cross-sectional area for vapor flow $= \dfrac{\pi D_v^2}{4} * 0.5 = 0.393 D_v^2$

Vapor velocity, $u_v = 547.3/3600/(0.393 D_v^2) = 0.387 D_v^{-2}$

Vapor residence time required for the droplets to settle to liquid surface

$= h_v / u_t = 0.5 D_v / 0.1972 = 2.54 D_v$

Actual residence time = vessel length/vapor velocity

$= L_v / u_v = 4 D_v / 0.387 D_v^{-2} = 10.34 D_v^3$

For satisfactory separation required residence time = actual.

So, $2.54 D_v = 10.34 D_v^3$

$D_v = 0.50$ m , say 0.6m.

Liquid hold-up time,

Liquid volumetric flow-rate is 28.4m^3/hr.

liquid cross-sectional area $= \pi * (0.6)^2 / 4 * 0.5 = 0.141 \text{ m}^2$

Length, $L_v = 4 * 0.6 = 2.4$ m

Hold-up volume $= 0.141 * 2.4 = 0.339 \text{ m}^3$

Hold-up time = liquid volume/liquid flow-rate

$= 0.339/(28.4/3600) = 43.0 \text{ s} = 0.72$ minutes

This is unsatisfactory, 10 minutes minimum required.

Accordingly, need to increase the liquid volume. This is best way to increase the vessel diameter. Keep liquid height at half the vessel diameter and try to find a suitable diameter that can reach enough residence time. After several trying, new *D$_v$* is given.

New $D_v = 1.5$ m

Liquid residence time checking:

Liquid cross-sectional area $= \pi * (1.5)^2 / 4 * 0.5 = 0.88 \text{ m}^2$

Length, $L_v = 4 * 1.5 = 6$ m

New liquid volume $= 0.88 * 6 = 5.28 \text{ m}^3$

New residence time $= 5.28/(28.4/3600) = 669.3 \text{ s} = 11.2$ minutes

This is satisfactory.

The vessel volume is $\pi \dfrac{D_v^2}{4} L_v = \pi * \dfrac{(1.5)^2}{4} * 6 = 10.6$ m

Result:

Vessel diameter: 1.5m.

Vessel length: 6m.

Vessel volume: 10.6m^3.

2. *D-301 T301 Reflux Drum*

From data of Aspan simulation, at 15bar, -131.44°C, liquid density 502.50kg/m^3, vapor density 14.22kg/m^3.

$u_t = 0.07\left[(502.5 - 14.22)/14.22\right]^{1/2} = 0.4102$ m/s

Vapor volumetric flow-rate is 5275kg/hr =261.9m^3/hr

Take $h_v = D_v/2$ and $L_v/D_v = 3$

Cross-sectional area for vapor flow $= \dfrac{\pi D_v^2}{4} * 0.5 = 0.393 D_v^2$

Vapor velocity, $u_v = 261.9/3600/(0.393 D_v^2) = 0.185 D_v^{-2}$

Vapor residence time required for the droplets to settle to liquid surface

$= h_v/u_t = 0.5 D_v/0.4102 = 1.22 D_v$

Actual residence time = vessel length/vapor velocity

$= L_v/u_v = 3D_v/0.185 D_v^{-2} = 16.2 D_v^3$

For satisfactory separation required residence time = actual.

So, $1.22 D_v = 16.2 D_v^3$

$D_v = 0.42$ m, say 0.45m.

Liquid hold-up time,

Liquid volumetric flow-rate is 45810kg/hr =91.2m^3/hr.

liquid cross-sectional area $= \pi * (0.45)^2/4 * 0.5 = 0.0795$ m^2

Length, $L_v = 3 * 0.45 = 1.35$ m

Hold-up volume $= 0.0795 * 1.5 = 0.1073$ m^3

Hold-up time = liquid volume/liquid flow-rate

$= 0.1073/(91.2/3600) = 4.23$ s $<< 10$ minutes

As same as D101 design, increase the vessel diameter to reach higher residence time. After several trying, new D_v is given.

New $D_v = 2.4$ m

Recalculate:

Liquid hold-up time,

Liquid volumetric flow-rate is 91.2m^3/hr.

liquid cross-sectional area $= \pi * (2.4)^2/4 * 0.5 = 2.26$ m^2

Length, $L_v = 3 * 2.4 = 7.2$ m,

Hold-up volume $= 2.26 * 7.2 = 16.272$ m^3

Hold-up time = liquid volume/liquid flow-rate

$= 16.272/(91.2/3600) = 642.3$ s $= 10.7$ minutes

This is satisfactory.

The vessel volume is $\pi \dfrac{D_v^2}{4} L_v = \pi * \dfrac{(2.4)^2}{4} * 7.2 = 32.6$ m^3

Result:

Vessel diameter: 2.4m.
Vessel length: 7.2m.
Vessel volume: 32.6m^3.

Appendix E.8 Equipment Summary & Specification Sheets

REACTORS, COLUMNS & VESSELS – SUMMARY

EQUIPMENT NR. : **NAME :**	**RX001** **Shell Side Reactor**	**RX002** **Tube Side Reactor**	**D101** **Gas sep. Drum**	**D102** **Gas Buffer Feed Drum**	**D201** **Gas Comp. Suc.Drum**
	Vertical	Multi tube Vertical	Horizontal	Vertical	Vertical
Pressure [bara] :	1	1	30	30	1
Temp. [^oC] :	540	850	55	42.5	25
Volume [m^3] :	4.45	0.45 *(1)*	10.6	42.4	51.5
Diameter [m] :	1.63	0.30	1.5	3	3.2
L or H [m] :	6.41	6.41	6	6	6.4
Internals					
- Tray Type :	-	-			
- Tray Number :	-	-			
- Fixed Packing					
Type :	Fixed bed	Fixed bed			
Shape :	-	-			
- Catalyst					
Type :	Pt on Zeolite	V_2O_5			
Shape :	Sphere	Sphere			
-					
-					
-					
<u>Number</u>					
- Series :	-	-			
- Parallel :	2 *(2)*	2 *(2)*			
Materials of Construction :	SS304	SS304	CS	CS	CS
Other :					

Remarks:

(1) There are 20 tubes in one reactor, which is the total volume of tube side 9.06 m^3.

(2) The number of Reactor is 2, one for operation, the other one for de coke.

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

REACTORS, COLUMNS & VESSELS – SUMMARY

EQUIPMENT NR. : **NAME :**	**D301** **T301 Reflux Drum**	**D302** **T302 Reflux Drum**	**D401** **MDEA Sol Drum**	**T301** **Light Gas Column**	**T302** **Ethylene Column**
	Horizontal	Horizontal	Vertical	Vertical column	Vertical Column
Pressure [bara] :	15	15	30	15	15
Temp. [^{o}C] :	-131.44	-39	30	-131/10	-39/39
Volume [m^3] :	32.6	3.45	3.45	-	-
Diameter [m] :	2.4	1.3	1.3	2.15	1.95
L or H [m] :	7.2	2.6	2.6	15.30	18.70
Internals					
- Tray Type :				-	
- Tray Number :				-	
- Fixed Packing					
Type :				Packed column	Packed column
Shape :				Pall ring	Pall ring
- Catalyst					
Type :				-	-
Shape :				-	-
-					
-					
-					
<u>Number</u>					
- Series :				-	-
- Parallel :				-	-
Materials of Construction :	CS	CS	CS	CS	CS
Other :					

Remarks:

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

REACTORS, COLUMNS & VESSELS – SUMMARY

EQUIPMENT NR. : **NAME :**	**T303** **Propylene Column**	**T401** **CO_2 Absorber**	**T402** **CO_2 Regenerator**		
	Vertical Column	Vertical Column	Vertical Column		
Pressure [bara] :	15	30	1		
Temp. [^oC] :	34/43	30	101-104		
Volume [m^3] :	-	22.6	62.9		
Diameter [m] :	4.5	1.3	3.65		
L or H [m] :	131.00	17	4.25		
Internals					
- Tray Type :	-	-	-		
- Tray Number :	-	-	-		
- Fixed Packing					
Type :	Packed column	Packed column	Packed column		
Shape :	Pall ring	Pall ring	Pall ring		
- Catalyst					
Type :	-	-	-		
Shape :	-	-	-		
-					
-					
-					
<u>Number</u>					
- Series :	-				
- Parallel :	-				
Materials of Construction :	CS	CS/Coating	CS		
Other :					

Remarks:

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

HEAT EXCHANGERS & FURNACES – SUMMARY

EQUIPMENT NR. : **NAME :**	**E001** **Propane feed heater 1**	**E002** **Propane feed heater 2**	**E003** **Propane feed heater 3**	**E004** **Propane feed heater 4**	**E005** **Propane feed heater 5**
	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Electric heater (1)	Shell and tubes heat exchanger	Electric heater (1)
Substance					
- Tubes :	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4		C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4/CO/CO_2/H_2O	
- Shell :	*n*-C_3	*n*-C_3		*n*-C_3	
Duty [kW] :	7890	6632	359	7536	1645
Heat Exchange area [m²] :	24966	15785		7569	
Number					
- Series :	1	1		1	
- Parallel :	-			-	
Pressure [bara]					
- Tubes :	1	1		1	
- Shell :	1	1		1	
Temperature In / Out [°C]					
- Tubes :	308.0 / 32.029	540.0 / 308.0		850.0 / 550.0	
- Shell :	25.0 / 298.0	298.0 / 528.0		540.0 / 795.0	
Special Materials of Construction :	Tubes : CS Shell : CS	Tubes : CS Shell : CS		Tubes : CS Shell : CS	
Other :					

Remarks:

Heat exchangers are calculated with heat integration

(1) Electric heater is used for start-up. During normally operation, it could be used for controllability.

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

HEAT EXCHANGERS & FURNACES – SUMMARY

EQUIPMENT NR. : **NAME :**	**E006** **Propane feed heater 6**	**E101** **Shell product cooler**	**E102** **Compressed gas product cooler**	**E103** **Compressed gas product cooler2**	**E201** **Tube product cooler**
	Electric heater (1)	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger
Substance **- Tubes :** **- Shell :**		C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4 Cooling W.	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4 C_3 recycle	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4 Cooling W.	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4/CO/CO_2/H_2O C_3 recycle.
Duty [kW] :	1760	200	4800.7	294.3	4916
Heat Exchange area [m^2] :		167	4153	36	4106
Number **- Series :** **- Parallel :**		1 -	1 -	1 -	1 -
Pressure [bara] **- Tubes :** **- Shell :**		1 4.0	1 15	1 4	1.0 1.0
Temperature In / Out [^{o}C] **- Tubes :** **- Shell :**		32.029 / 20.0 25.0 / 26.0	227.0 / 65.0 43.0 / 217.0	65.0 / 50.0 20.0 / 27.2	550.0 / 308.0 298.0 / 476.0
Special Materials of Construction :		Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS
Other :					

Remarks:

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

HEAT EXCHANGERS & FURNACES – SUMMARY

EQUIPMENT NR. : **NAME :**	**E202** **Tube product Cooler2**	**E203** **Tube product Cooler3**	**E204** **Compressed tube gas cooler**	**E301A** **T301** **Light Gas Column Reboiler**	**E301B** **T301** **Light Gas Column Condenser**
	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger
Substance **- Tubes :** **- Shell :**	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4/CO/CO_2/H_2O C_3 recycle	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4/CO/CO_2/H_2O Cooling W.	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4/CO/ CO_2/H_2O Cooling W.	C_3-/$C_{3=}$/$C_{2=}$ Hot Water.	Light gas Expanded H_2
Duty [kW] :	2242.3	2961	4967	1515	7290
Heat Exchange area [m^2] :	12659	609	363	85	1060
Number **- Series :** **- Parallel :**	1	1 -	1 -	1 -	1 -
Pressure [bara] **- Tubes :** **- Shell :**	30 1	1.45 4	30.0 4.0	15.0 4.0	15.0 5.0
Temperature In / Out [^{o}C] **- Tubes :** **- Shell :**	308.0 / 219.0 217.0 / 298.0	219.0 / 25.0 20.0 / 121.0	288.6 / 30.0 20.0 / 100.0	9.52 / 9.52 65.0 / 50.0	-131.0 / -131.0 -250.0/ -200.0
Special Materials of Construction :	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS
Other :					

Remarks:

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

HEAT EXCHANGERS & FURNACES – SUMMARY

EQUIPMENT NR. : **NAME :**	**E302A** **T302** **C2 Column Reboiler**	**E302B** **T302** **C2 Column Condenser**	**E303A** **T303** **C3 Column Reboiler**	**E303B** **T303** **C3 Column Condenser**	**E303C** **Heat Compressor after cooler**
	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger
Substance					
- Tubes :	C_3= /C_{3-}/C_2=	C_3= /C_{3-}/C_2=	C_3-	C_3=	C_3=
-					
- Shell :	Hot water	$C_{2=}$ expanded	C_3=	C_3-	Cooling W.
Duty [kW] :	4601	3954	61	604	543
Heat Exchange area [m^2] :	722	526	1073	503	145
Number					
- Series :	1	1	1	1	1
- Parallel :	-	-	-	-	-
Pressure [bara]					
- Tubes :	15.0	15.0	15.0	15.0	15.0
- Shell :	4.0	4.0	15.0	15.0	4.0
Temperature In / Out [^{o}C]					
- Tubes :	39.65 / 39.65	-39.0 / -39.0	42.5 / 42.7	35 / 38.0	45.9 / 43.2
- Shell :	65.0 / 50.0	-75.0 / -70.0	43.2 / 42.5	48.8 / 35.0	20.0 / 25.0
Special Materials of Construction :	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS
Other :					

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : Date :	**CPD3297** 16 December 2003

HEAT EXCHANGERS & FURNACES – SUMMARY

EQUIPMENT NR. : **NAME :**	**E401** **MDEA Cooler**	**E402** **CO_2 Stripper Reboiler**	**AE101** **Shell product cooler**	**AE102** **Compressed gas product cooler**	**AE202** **Tube product Cooler**
	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger	Shell and tubes heat exchanger
Substance **- Tubes :** **- Shell :**	MDEA CW	Spent MDEA MS	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4 BFW.	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4 BFW.	C_3-/$C_{3=}$/$C_{2=}$/ $H_2/CH_4/CO/CO_2/H_2O$ BFW
Duty [kW] :	8,132	34,377	14722	4140	20727
Heat Exchange area [m^2] :	464	1,034	621	172	756
Number **- Series :** **- Parallel :**	- -	1 -	1 -	1 -	1
Pressure [bara] **- Tubes :** **- Shell :**	30 4	30 4	1 4.0	30 4	1 4
Temperature In / Out [^oC] **- Tubes :** **- Shell :**	101.97/30 20/70	101 / 104 150 / 105	540.0 / 25.0 20.0 / 250.0	227.0 / 55.0 20.0 / 100.0	850.0 / 25.0 20.0 / 500.0
Special Materials of Construction :	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS
Other :					

Remarks:

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

HEAT EXCHANGERS & FURNACES – SUMMARY

EQUIPMENT NR. : **NAME :**	**AE203** **Tube product Cooler**				
	Shell and tubes heat exchanger				
Substance **- Tubes :** **- Shell :**	C_3-/$C_{3=}$/$C_{2=}$/ H_2/CH_4/CO/CO_2/H_2O BFW				
Duty [kW] :	4970				
Heat Exchange area [m^2] :	538				
Number **- Series :** **- Parallel :**	1 -				
Pressure [bara] **- Tubes :** **- Shell :**	30 4				
Temperature In / Out [°C] **- Tubes :** **- Shell :**	308.0 / 30.0 20.0 / 250.0				
Special Materials of Construction :	Tubes : CS Shell : CS				
Other :					

Remarks:

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

PUMPS, BLOWERS & COMPRESSORS – SUMMARY

EQUIPMENT NR. : **NAME :**	**C101** **Shell Product Compressor**	**C201** **Tube Product Compressor**	**C303** **Propylene heat Compressor**	**P101A/B** **(AP202)** **Process water pump1:**	**P102A/B** **Process water pump2**
Type : **Number :**	Centrifugal*(1)* 1	Centrifugal*(1)* 1	Centrifugal 1	Centrifugal 2	Centrifugal 2
Medium transferred :	C_3= / *n*-C_3/ C_2=/Light gas	C_3= / *n*-C_3/ C_2=/Light gas	C_3=	water	water
Capacity **[kg/s] :** **Inlet [m^3/s] :**	- 7.32	- 6.63	- 3.37	0.365	0.365
Density Inlet [kg/m^3]	1.48	1.17	31.24	1,000	1,000
Pressure [bara] **Suct. / Disch. :**	1.0/ 30.0	1.0/30.0	15.0/17.0	1.0 / 3.0	1.0 / 3.0
Temperature **In / Out [°C] :**	25.0 / 215.0	25.0/308.2	35.0/45.9	25 / 25	25 / 25
Power [kW] **- Theor. :** **- Actual :**	3156 4170	3161 4516	647.3 899.0	0.172 0.245	0.172 0.245
Number **- Theor. :** **- Actual :**	1	1	1	2*(2)*	2*(2)*
Special Materials of Construction :	MS casing	MS casing	MS casing	MS casing	MS casing
Other :	Double mechanical seals	Double mechanical seals	Double mechanical seals	Double mechanical seals	Double mechanical seals

Remarks:

(1) 2 stage compressor.

(2) One installed spare included

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

PUMPS, BLOWERS & COMPRESSORS – SUMMARY

EQUIPMENT NR. : **NAME :**	**P103A/B** **Process water pump3**	**P301A/B** **T301Bottom pump**	**P301C/D** **T301Reflux pump**	**P302A/B** **T302Bottom pump**	**P302C/D** **T302Reflux pump**
Type :	Centrifugal	Centrifugal	Centrifugal	Centrifugal	Centrifugal
Number :	2	2	2	2	2
Medium transferred :	water	C_3= / n-C_3 C_2=	C_3= / n-C_3 C_2=	C_3= / n-C_3	n-C_3 / C_3=
Capacity [kg/s] :	0.365	13.42	12.72	11.12	8.63
Inlet [m^3/s] :		-	-		
Density Inlet [kg/m^3]	1,000	456	502	426	426
Pressure [bara] Suct. / Disch. :	1.0 / 3.0	15.0 / 16.5	15.0/16.5	15.0 / 23.0	15.0 / 17.0
Temperature In / Out [°C] :	25 / 25	9.5	-131	39.9 / 41.2	-39
Power [kW]					
- Theor. :	0.172	4.38	5.06	21	4
- Actual :	0.245	6.25	7.31	30	6
Number					
- Theor. :					
- Actual :	2*(1)*	2*(1)*	2*(1)*	2*(1)*	2*(1)*
Special Materials of Construction :	MS casing	MS casing	MS casing	MS casing	MS casing
Other :	Double mechanical seals	Double mechanical seals	Double mechanical seals	Double mechanical seals	Double mechanical seals

Remarks:

(1) One installed spare included.

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

PUMPS, BLOWERS & COMPRESSORS – SUMMARY

EQUIPMENT NR. : **NAME :**	**P303A/B** **T303Bottom pump:**	**P402A/B** **MDEA recycle pump1**	**P403A/B** **MDEA recycle pump2**		
Type : **Number :**	Centrifugal 2	Centrifugal 2	Centrifugal 2		
Medium transferred :	*n*-C_3	MDEA sol^n	MDEA sol^n		
Capacity **[kg/s] :** **Inlet [m^3/s] :**	9.84	32	32		
Density Inlet [kg/m^3]	417	990	990		
Pressure [bara] **Suct. / Disch. :**	15.0 /17.0	1.0/2.5	1.0/30.0		
Temperature **In / Out [^{o}C] :**	43	101	30		
Power [kW] **- Theor. :** **- Actual :**	4.7 7	5.2 7.38	92 131		
Number **- Theor. :** **- Actual :**	*2(1)*	*2(1)*	*2(1)*		
Special Materials of Construction :	MS casing	MS casing	MS casing		
Other :	Double mechanical seals	Double mechanical seals	Double mechanical seals		

Remarks:

(1) One installed spare included.

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

Specification Sheets

SHELL & TUBE REACTOR– SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**RX001 & RX002** *(1)*	**In Series**	**:**	1
NAME	**:**	**Shell & Tube Reactor**	**In Parallel**	**:**	none

General Data

Service	**:**	**- Reactor**		
Type	**:**	**- Shell and Tube**		
Position	**:**	**- ~~Horizontal~~** **- Vertical**		
Capacity	**[m^3]**	**:** 13.51		**(Calc.)**
Heat Exchange Area	**[m^2]**	**:** 121		**(Calc.)**
Overall Heat Transfer Coefficient	**[W/m^2·^{o}C]**	**:** 500		**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^{o}C]**	**:** -		
Passes Tube Side		**:** 1		
Passes Shell Side		**:** 1		
Correction Factor LMTD (min. 0.75)		**:** -		
Corrected LMTD	**[^{o}C]**	**:** -		

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	$C_{3=}/C_{2=}/CH_4/H_2/C_{3-}$	$C_{3=}/C_{2=}/CH_4/H_2/C_{3-}/CO/CO_2/H_2O$
Mass Stream	**[kg/s]**	**:**	10.84	9.10
Mass Stream to				
- Evaporize	**[kg/s]**	**:**	-	-
- Condense	**[kg/s]**	**:**	-	-
Average Specific Heat	**[kJ/kg·^{o}C]**	**:**	3.499	3.512
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^{o}C]**	**:**	540	850
Temperature OUT	**[^{o}C]**	**:**	-	-
Pressure	**[bara]**	**:**	1	1
Material		**:**	SS304	SS304

Remarks:

(1) Novel reactor type

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

VESSEL – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**D101**	**In Series**	:	1
NAME	:	**Gas sep. Drum**	**In Parallel**	:	none

General Data

Service		:	- ~~**Buffer**~~ / ~~**Storage**~~ / **Separation** / ~~**Reaction**~~
Type		:	**- Drum**
Position		:	**- Horizontal** - ~~**Vertical**~~
Internals		:	**- Demister** / ~~**Plate**~~ / ~~**Coil**~~ / _________
Heating/Cooling medium		:	**- none** / ~~**Open**~~ / ~~**Closed**~~ / ~~**External Hxgr**~~ /________
- Type		:	n.a.
- Quantity	**[kg/s]**	:	n.a.
- Press./Temp.'s	**[bara/^{o}C]**	:	n.a.
Vessel Diameter (ID)	**[m]**	:	1.5
Vessel Height	**[m]**	:	6
Vessel Tot. Volume	**[m^3]**	:	10.6
Vessel Material		:	Carbon steel
Other		:	

Process Conditions

Stream Data		**Feed**		**Top**		**Bottom**	
Temperature	**[^{o}C]**	55		54.7		54.7	
Pressure	**[bara]**	30		30		30	
Density	**[kg/m^3]**	66.4		48.7		435.1	
Mass Flow	**[kg/s]**	10.8		7.4		3.4	
Composition		mol%	wt%	mol%	wt%	mol%	wt%
C_3H_8		0.575	0.55	0.528	0.69	0.702	0.72
C_2H_4		343ppm	0.21	399ppm	0.00	191ppm	0.00
C_3H_6		0.253	0.23	0.242	0.30	0.284	0.28
CH_4		343ppm	0.00	436ppm	0.00	90ppm	0.00
H_2		0.171	0.01	0.229	0.01	0.013	0.00

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B.Wang Y.Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

VESSEL – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**D102**	**In Series**	**:**	1
NAME	**:**	**Gas Buffer Feed Drum**	**In Parallel**	**:**	none

General Data

Service		**:**	**- Buffer / ~~Storage~~ / ~~Separation~~ / ~~Reaction~~**
Type		**:**	**- Drum**
Position		**:**	**- ~~Horizontal~~** **- Vertical**
Internals		**:**	**- ~~Demister~~ / ~~Plate~~ / ~~Coil~~ / _________**
Heating/Cooling medium		**:**	**- none / ~~Open~~ / ~~Closed~~ / ~~External Hxgr~~ /________**
- Type		**:**	n.a.
- Quantity	**[kg/s]**	**:**	n.a.
- Press./Temp.'s	**[bara/^{o}C]**	**:**	n.a.
Vessel Diameter (ID)	**[m]**	**:**	3
Vessel Height	**[m]**	**:**	6
Vessel Tot. Volume	**[m^3]**	**:**	42.4
Vessel Material		**:**	Carbon steel
Other		**:**	

Process Conditions

Stream Data		**Feed 1**	**Feed 2**	**Out**
Temperature	**[^{o}C]**	54.7	30	42.5
Pressure	**[bara]**	30	30	30
Density	**[kg/m^3]**	48.7	40.9	43.7
Mass Flow	**[kg/s]**	7.4	7.5	14.9

Composition	mol%	wt%	mol%	wt%	mol%	wt%
C_3H_8	0.528	0.68	0.019	0.18	0.306	0.43
N_2	0.00	0.00	0.011	0.01	0.006	0.01
C_2H_4	399ppm	0.00	0.315	0.31	0.171	0.15
C_3H_6	0.242	0.30	0.226	0.33	0.233	0.31
CH_4	436ppm	0.00	0.138	0.08	0.075	0.04
H_2	0.229	0.01	0.101	0.01	0.160	0.01
H_2O	0.00	0.01	0.002	0.00	0.001	0.00
CO	0.00	0.00	0.086	0.08	0.047	0.04
CO_2	0.00	0.00	0.001	0.00	668ppm	0.00

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B.Wang Y.Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

VESSEL – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**D201**	**In Series** :	1
NAME	:	**Gas Comp.Suc.Drum**	**In Parallel** :	none

General Data

Service : **- Buffer / ~~Storage~~ / ~~Separation~~ / ~~Reaction~~**

Type : **- Drum**

Position : **- ~~Horizontal~~**
- Vertical

Internals : **- ~~Demister~~ / ~~Plate~~ / ~~Coil~~ / _________**
Heating/Cooling medium : **- none / ~~Open~~ / ~~Closed~~ / ~~External Hxgr~~ /________**
- Type : n.a.
- Quantity **[kg/s]** : n.a.
- Press./Temp.'s **[bara/°C]** : n.a.
Vessel Diameter (ID) **[m]** : 3.2
Vessel Height **[m]** : 6.4
Vessel Tot. Volume **[m^3]** : 51.5
Vessel Material : Carbon steel
Other :

Process Conditions

Stream Data		**Feed**		**Outlet**	
Temperature	**[°C]**	25		25	
Pressure	**[bara]**	1		1	
Density	**[kg/m^3]**	1.2		1.2	
Mass Flow	**[kg/s]**	7.7		7.7	
Composition		mol%	wt%	mol%	wt%
C_3H_8		0.115	0.18	0.115	0.18
N_2		0.011	0.01	0.011	0.01
C_2H_4		0.304	0.30	0.304	0.30
C_3H_6		0.218	0.32	0.218	0.32
CH_4		0.133	0.07	0.133	0.07
H_2		0.098	0.01	0.098	0.01
H_2O		0.024	0.01	0.024	0.01
CO		0.083	0.08	0.083	0.08
CO_2		0.015	0.02	0.015	0.02

Remarks:

Designers : Montree l. O. Muraza W.K. Lin B.Wang Y.Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

VESSEL – SPECIFICATION SHEET

EQUIPMENT NUMBER : **D301** **In Series** : 1
NAME : **T301 Reflux Drum** **In Parallel** : none

General Data

Service : **- ~~Buffer~~ / ~~Storage~~ / Separation / ~~Reaction~~**

Type : **Drum**

Position : **- Horizontal**
- ~~Vertical~~

Internals : **- Demister / ~~Plate~~ / ~~Coil~~ / _________**
Heating/Cooling medium : **- none / ~~Open~~ / ~~Closed~~ / ~~External Hxgr~~ /________**
- Type : n.a.
- Quantity **[kg/s]** : n.a.
- Press./Temp.'s **[bara/°C]** : n.a.
Vessel Diameter (ID) **[m]** : 2.4
Vessel Height **[m]** : 7.2
Vessel Tot. Volume **[m^3]** : 32.6
Vessel Material : Carbon steel
Other :

Process Conditions

Stream Data		**Feed**	**Outlet**	**Reflux**
Temperature	**[°C]**	-131.44	-131.44	-131.44
Pressure	**[bara]**	15	15	15
Density	**[kg/m^3]**	20.14	14.2	502.50
Mass Flow	**[kg/s]**	14.2	1.5	12.7

Composition	mol%	wt%	mol%	wt%	mol%	wt%
C_3H_8	0.000	0.000	trace	0.00	0.000	0.000
N_2	0.008	0.010	0.021	0.06	0.004	0.005
C_2H_4	0.418	0.555	0.007	0.02	0.494	0.619
C_3H_6	0.000	0.000	trace	0.00	0.000	0.000
CH_4	0.395	0.337	0.260	0.39	0.462	0.331
H_2	0.115	0.011	0.551	0.10	0.005	0.000
H_2O	0.000	0.000	trace	0.00	0.000	0.000
CO	0.064	0.085	0.161	0.43	0.036	0.045
CO_2	0.000	0.001	2ppm	0.00	0.000	0.001

Remarks:

Designers : Montree l. O. Muraza W.K. Lin
B.Wang Y.Zou

Project ID-Number : **CPD3297**
Date : 16 December 2003

VESSEL – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**D302**	**In Series**	:	1
NAME	:	**T302 Reflux Drum**	**In Parallel**	:	none

General Data

Service		:	**- ~~Buffer~~ / ~~Storage~~ / Separation / ~~Reaction~~**
Type		:	**- Drum**
Position		:	**- Horizontal** **- ~~Vertical~~**
Internals		:	**- ~~Demister~~ / ~~Plate~~ / ~~Coil~~ / _________**
Heating/Cooling medium		:	**- none / ~~Open~~ / ~~Closed~~ / ~~External Hxgr~~ /________**
- Type		:	n.a.
- Quantity	**[kg/s]**	:	n.a.
- Press./Temp.'s	**[bara/°C]**	:	n.a.
Vessel Diameter (ID)	**[m]**	:	1.3
Vessel Height	**[m]**	:	2.6
Vessel Tot. Volume	**[m^3]**	:	3.45
Vessel Material		:	Carbon steel
Other		:	

Process Conditions

Stream Data		**Feed**	**Outlet**	**Reflux**
Temperature	**[°C]**	-39	-39	-39
Pressure	**[bara]**	15	15	15
Density	**[kg/m^3]**	426.3	426.3	426.3
Mass Flow	**[kg/s]**	11.0	2.3	8.7

Composition	mol%	wt%	mol%	wt%	mol%	wt%
C_3H_8	544ppm	0.579	544ppm	0.579	544ppm	0.579
N_2	trace	0.000	trace	0.000	trace	0.000
C_2H_4	0.992	0.000	0.992	0.000	0.992	0.000
C_3H_6	0.004	0.420	0.004	0.420	0.004	0.420
CH_4	29ppm	0.000	29ppm	0.000	29ppm	0.000
H_2	trace	0.000	trace	0.000	trace	0.000
H_2O	trace	0.001	trace	0.001	trace	0.001
CO	trace	0.000	trace	0.000	trace	0.000
CO_2	0.004	0.000	0.004	0.000	0.004	0.000

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B.Wang Y.Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

VESSEL – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**D401**	**In Series**	:	1
NAME	:	**MDEA Sol Drum**	**In Parallel**	:	none

General Data

Service : **- Buffer / ~~Storage~~ / ~~Separation~~ / ~~Reaction~~**

Type : **-**

Position : **- ~~Horizontal~~**
- Vertical

Internals		:	**- ~~Demister~~ / ~~Plate~~ / ~~Coil~~ / _________**
Heating/Cooling medium		:	**- none / ~~Open~~ / ~~Closed~~ / ~~External Hxgr~~ /________**
- Type		:	n.a.
- Quantity	**[kg/s]**	:	n.a.
- Press./Temp.'s	**[bara/°C]**	:	n.a.
Vessel Diameter (ID)	**[m]**	:	1.3
Vessel Height	**[m]**	:	2.6
Vessel Tot. Volume	**[m^3]**	:	3.45
Vessel Material		:	Carbon steel
Other		:	

Process Conditions

Stream Data		**Feed**		**Bottom**	
Temperature	**[°C]**	30		30	
Pressure	**[bara]**	30		30	
Density	**[kg/m^3]**	656.45		656.45	
Mass Flow	**[kg/s]**	70.0		70.0	
Composition		wt%	wt%	mol%	wt%
CO_2		10ppm	0.00	10ppm	0.00
H_2O		0.870	0.503	0.870	0.503
MDEA		0.130	0.497	0.130	0.497

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B.Wang Y.Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

DISTILLATION COLUMN – SPECIFICATION SHEET

EQUIPMENT NUMBER : T-301
NAME : Light Gas/Main products separation

General Data

Service : **- distillation / ~~extraction~~ / ~~absorption~~ / ———**
Column Type : **- packed / ~~tray~~ / ~~spray~~ / ———**
Packed Type : **- ~~cap —~~ / pall-ring / ~~valve~~ / ———**
Tray Number *(1)*
- **Theoretical** : 20
- **Actual** :
- **Feed (actual)** : 10

Tray Distance (HETP) [m] : 0.850 **Tray Material** :
Column Diameter [m] : 2.15 **Column Material** : **CS** *(2)*
Column Height [m] : 15.3
Heating : **- ~~none~~ / ~~open steam~~ / reboiler / ———** *(3)*

Process Conditions

Stream Details	**Feed**	**Top**	**Bottom**	**Reflux / ~~Absorbent~~**	**~~Extractant~~ / ~~side stream~~**
Temp. [°C]	: 42.5	: -131.4	: 9.5	: -131.4	
Pressure [bara]	: 30	: 15	: 15	: 15	
Density [kg/m^3]	: 43.72	: 14.22	: 455.8	: 455.8	
Mass Flow [kg/s]	: 14.88	: 1.47	: 13.41	: 12.72	

Composition	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Propane	30.6	43.27	Trace	Trace	43.1	48.0				
Nitrogen	0.6	0.55	2.1	5.63	Trace	Trace				
Ethylene	17.1	15.41	0.7	1.92	23.8	16.88				
Propylene	23.3	31.5	Trace	Trace	32.9	34.94				
Methane	7.5	3.88	26.0	39.39	Trace	Trace				
Hydrogen	16.0	1.03	55.1	10.50	Trace	Trace				
Water	0.1	0.07	Trace	Trace	0.2	0.08				
Carbon monoxide	4.7	4.19	16.1	42.56	Trace	Trace				
Carbon dioxide	0.07	0.09	Trace	Trace	0.09	0.10				

Column Internals

Trays *(5)*
Number of ~~caps~~ / sieve holes / ——— : …
Active Tray Area [m^2] : …
Weir Length [mm] : …
Diameter of chute pipe / hole / ——— [mm] : …

Packing
Type : **pall-ring**
Material : **metal**
Volume [m^3] :
Length [m] :
Width [m] :
Height [m] : **15.3**

Remarks:
(1) Tray numbering from top to bottom.
(2) CS = Carbon Steel.

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

DISTILLATION COLUMN – SPECIFICATION SHEET

EQUIPMENT NUMBER : T-302
NAME : Ethylene/Propylene separation

General Data

Service : - distillation / ~~extraction~~ / ~~absorption~~ / ———
Column Type : - packed / ~~tray~~ / ~~spray~~ / ———
Packed Type : - ~~cap~~ / pall-ring / ~~valve~~ / ———
Tray Number *(1)*
- **Theoretical :** 24
- **Actual :**
- **Feed (actual) :** 14

Tray Distance (HETP) [m] : 0.85 **Tray Material :**
Column Diameter [m] : 1.95 **Column Material : CS** *(2)*
Column Height [m] : 18.7
Heating : - ~~none~~ / ~~open steam~~ / reboiler / ———

Process Conditions

Stream Details	Feed	Top	**Bottom**	**Reflux / ~~Absorbent~~**	**~~Extractant~~ / ~~side stream~~**
Temp. [°C]	: 9.5	: -39	: 39.9	: 3.764	
Pressure [bara]	: 15	: 15	: 15	: 15	
Density [kg/m^3]	: 455.8	: 426.3	: 426.1	: 426.1	
Mass Flow [kg/s]	: 13.42	: 2.29	: 11.12	: 21.77	

Composition	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Ethylene	23.8	16.8	99.2	98.7	0.002	0.01	99.2	98.7		
Propane	32.9	34.9	0.4	0.6	43.1	42.0-	0.4	0.6		
———										
———										
———										
———										

Column Internals *(4)*

Trays
Number of
~~caps~~ / sieve holes / ——— : …
Active Tray Area [m^2] : …
Weir Length [mm] : …
Diameter of
chute pipe / hole / ——— [mm] : …

Packing
Type : Pall ring
Material : metal
Volume [m^3] :
Length [m] :
Width [m] :
Height [m] : 18.7

Remarks:
(1)Tray numbering from top to bottom.
(2)CS = Carbon Steel.

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : Date :	**CPD3297** 16 December 2003

DISTILLATION COLUMN – SPECIFICATION SHEET

EQUIPMENT NUMBER : T-303
NAME : Propylene/propane separation

General Data

Service : **- distillation / ~~extraction~~ / ~~absorption~~ / ———**
Column Type : **- packed / ~~tray~~ / ~~spray~~ / ———**
Packed Type : **- ~~cap~~ / pall-ring / ~~valve~~ / ———**
Tray Number *(1)*
- Theoretical :
- Actual : 156
- Feed (actual) : 93
Tray Distance (HETP) [m] : 0.85 **Tray Material** :
Column Diameter [m] : 4.59 **Column Material** : **CS** *(2)*
Column Height [m] : 23.0
Heating : **- ~~none~~ / ~~open steam~~ / reboiler / ———**

Process Conditions

Stream Details	Feed		Top		**Bottom**		**Reflux / ~~Absorbent~~**		**~~Extractant~~ / ~~side stream~~**	
Temp. [^oC]	: 37.9		: 34.8		: 42.7		: 35			
Pressure [bara]	: 15		: 15		: 15		: 15			
Density [kg/m^3]	: 244.63		: 31.29		: 417.144		: 417.144			
Mass Flow [kg/s]	: 14.56		: 4.72		: 9.84		: 100.54			
Composition	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Propylene	39.6	38.7	98.9	99.8	9.7	9.4	98.9	99.8		
Propane	59.9	61.2	0.05	0.05	90.0	90.5	0.05	0.05		
———										
———										
———										
———										

Column Internals *(4)*

Trays **Number of**
~~caps~~ / sieve holes / ——— : …
Active Tray Area [m^2] : …
Weir Length [mm] : …
Diameter of
chute pipe / hole / ——— [mm] : …

Packing **Type** :
Material :
Volume [m^3] :
Length [m] :
Width [m] :
Height [m] :130.9

Remarks:
(1)Tray numbering from top to bottom.
(2)CS = Carbon Steel.

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

ABSORBER COLUMN – SPECIFICATION SHEET

EQUIPMENT NUMBER : T-401
NAME : Carbon dioxide Absorber

General Data

Service : **- ~~distillation~~ / ~~extraction~~ / ~~absorption~~ / ———**
Column Type : **- packed / ~~tray~~ / ~~spray~~ / ———**
Packed Type : **- ~~cap~~ / pall-ring / ~~valve~~ / ———**
Tray Number
- Theoretical :
- Actual : 20
- Feed (actual) : Stream 209: from bottom; Stream 401: from top
Tray Distance (HETP) [m] : 0.850 **Tray Material** :
Column Diameter [m] : 1.3 **Column Material : CS**
Column Height [m] : 17.00
Heating : **- ~~none~~ / ~~open steam~~ / reboiler /**

Process Conditions

Stream Details	**Feed Stream 209**		**Feed Stream 403**		**Outlet Stream 401**		**Outlet Stream 402**		**~~Reflux / Absorbent~~**	
Temp. [^oC]	30		30		30		30			
Pressure [bara]	30		30		30		30			
Density [kg/m^3]	41.18		1002		40.89		1002			
Mass Flow [kg/s]	7.64		31.91		7.48		32.07			
Composition	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Propane	0.117	0.178	0.000	0.000	0.119	0.182	0.000	0.000		
Ethylene	0.310	0.300	0.000	0.000	0.315	0.307	0.000	0.000		
Propylene	0.223	0.323	0.000	0.000	0.226	0.331	0.000	0.000		
Carbon monoxide	0.084	0.081	0.000	0.000	0.086	0.084	0.000	0.000		
Carbon dioxide	0.015	0.024	0.000	0.000	0.000	0.000	0.145	0.193		
Methane	0.136	0.076	0.000	0.000	0.139	0.077	0.000	0.000		
Hydrogen	0.100	0.007	0.000	0.000	0.102	0.007	0.000	0.000		
Water	0.002	0.001	0.869	0.500	0.002	0.001	0.742	0.403		
Nitrogen	0.011	0.010	0.000	0.000	0.011	0.011	0.000	0.000		
MDEA	0.000	0.000	0.131	0.500	0.000	0.000	0.112	0.403		

Column Internals *(4)*

~~Trays~~ Number of
~~caps~~ / sieve holes / ——— : …
Active Tray Area [m^2] : …
Weir Length [mm] : …
Diameter of
chute pipe / hole / ——— [mm]

Packing Type : Pall ring
Material :metal
Volume [m^3] :
Length [m] :
Width [m] :
Height [m] : 17

Remark:

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

DISTILLATION COLUMN – SPECIFICATION SHEET

EQUIPMENT NUMBER : T-402
NAME : CO_2 –MDEA Sol stripper

General Data

Service : - Stripper/ ~~extraction~~ / ~~absorption~~ / ———
Column Type : - packed / ~~tray~~ / ~~spray~~ / ———
Packed Type : - ~~cap~~ / pall-ring / ~~valve~~ / ———
Tray Number
- **Theoretical** : -
- **Actual** : 7
- **Feed (actual)** : 1

Tray Distance (HETP) [m] : 0.85 **Tray Material :**
Column Diameter [m] : 3.65 **Column Material : CS**
Column Height [m] : 4.25
Heating : - ~~none~~ / ~~open steam~~ / reboiler / ———

Process Conditions

Stream Details	Feed	Top	**Bottom**	**Reflux / ~~Absorbent~~**	**~~Extractant~~ / ~~side stream~~**
Temp. [^{o}C]	: 30	:101	: 105	:	
Pressure [bara]	: 30	: 1.0	:1.0	:	
Density [kg/m^3]	: 1002	: 0.58	:1002	:	
Mass Flow [kg/s]	: 32.07	: 0.16	31.9	:	

Composition	**mol%**	**wt%**	**Mol%**	**wt%**	**Mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Water	32.4	14.9	trace	trace	87	50	-	-		
MDEA	4.9	14.9	trace	trace	13	50	-	-		
CO_2	62.7	70.2	99.99	99.99	trace	trace	-	-		
———										
———										
———										

Column Internals

Trays Number of	**Packing** Type : pall ring
~~caps~~ / sieve holes / ——— : …	**Material : Plastic**
Active Tray Area [m^2] : …	**Volume [m^3] :**
Weir Length [mm] : …	**Length [m] :**
Diameter of	**Width [m] :**
chute pipe / hole / ——— [mm] : …	**Height [m] : 4.25**

Remarks:

.

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-001**	**In Series**	**:**	1
NAME	**:**	**Feed Heater1**	**In Parallel**	**:**	none

General Data

Service	**:**	- **Heat Exchanger** - ~~**Cooler**~~ - ~~**Condenser (water cooled)**~~	- ~~**Vaporizer**~~ - ~~**Reboiler**~~
Type	**:**	- ~~**Fixed Tube Sheets**~~ - ~~**Floating Head**~~ - ~~**Hair Pin**~~ - **Double Tube**	- ~~**Plate Heat Exchanger**~~ - ~~**Finned Tubes**~~ - ~~**Thermosyphon**~~
Position	**:**	- **Horizontal** - ~~**Vertical**~~	

Capacity	**[kW]**	**:**	7890	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	24966	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	**:**	50	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	**:**	8	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^oC]**	**:**	6.32	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	No. 001	No. 102
Mass Stream	**[kg/s]**	**:**	8.23	10.84
Mass Stream to				
- ~~**Evaporate**~~	~~**[kg/s]**~~	**:**		
- ~~**Condense**~~	~~**[kg/s]**~~	**:**		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	**:**	3.51	2.64
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^oC]**	**:**	25.0	308.0
Temperature OUT	**[^oC]**	**:**	298.0	32.029
Pressure	**[bara]**	**:**	1	1
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**E-002**	**In Series** :	1
NAME :	**Feed Heater2**	**In Parallel** :	none

General Data

Service :
- **Heat Exchanger** - **~~Vaporizer~~**
- **~~Cooler~~** - **~~Reboiler~~**
- **~~Condenser (water cooled)~~**

Type :
- **~~Fixed Tube Sheets~~** - **~~Plate Heat Exchanger~~**
- **~~Floating Head~~** - **~~Finned Tubes~~**
- **~~Hair Pin~~** - **~~Thermosyphon~~**
- **Double Tube**

Position :
- **Horizontal**
- **~~Vertical~~**

Capacity	**[kW]**	:	6632	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	15785	**(Calc.)**
Overall Heat Transfer Coefficient	**[W/m^2·^{o}C]**	:	50	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^{o}C]**	:	11	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	0.75	
Cor**rected LMTD**	**[^{o}C]**	:	8.25	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	No. 002	No. 101
Mass Stream	**[kg/s]**	:	8.23	10.84
Mass Stream to				
- **~~Evaporate~~**	**~~[kg/s]~~**	:	-	-
- **~~Condense~~**	**~~[kg/s]~~**	:	-	-
Average Specific Heat	**[kJ/kg·^{o}C]**	:	3.50	2.64
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^{o}C]**	:	298.0	540.0
Temperature OUT	**[^{o}C]**	:	528.0	308.0
Pressure	**[bara]**	:	1.0	1.0
Material		:	CS	CS

Remarks:

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**E-003**		**In Series**	: 1
NAME	:	**Feed Heater3**	**(1)**	**In Parallel**	: none

General Data

Service :
- **Electric heater** - ~~**Vaporizer**~~
- ~~**Cooler**~~ - ~~**Reboiler**~~
- ~~**Condenser (water cooled)**~~

Type :
- ~~**Fixed Tube Sheets**~~ - ~~**Plate Heat Exchanger**~~
- ~~**Floating Head**~~ - ~~**Finned Tubes**~~
- ~~**Hair Pin**~~ - ~~**Thermosyphon**~~
- ~~**Double Tube**~~

Position :
- ~~**Horizontal**~~
- ~~**Vertical**~~

Capacity	**[kW]**	:	359	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	-	**(Calc.)**
Overall Heat Transfer Coefficient	**[W/m^2·^{o}C]**	:	-	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^{o}C]**	:	-	
Passes Tube Side		:		
Passes Shell Side		:	-	
Correction Factor LMTD (min. 0.75)		:	-	
Corrected LMTD	**[^{o}C]**	:	-	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	No. 003	electric
Mass Stream	**[kg/s]**	:	8.23	-
Mass Stream to				
- ~~**Evaporate**~~	~~**[kg/s]**~~	:	-	-
- ~~**Condense**~~	~~**[kg/s]**~~	:	-	-
Average Specific Heat	**[kJ/kg·^{o}C]**	:	3.5	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^{o}C]**	:	527.5	-
Temperature OUT	**[^{o}C]**	:	540.0	-
Pressure	**[bara]**	:	1	1
Material		:	CS	-

Remarks:

(1) Use during start up and used for temperature controllability during normal operation, target temperature is 540 C

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : Date :	**CPD3297** 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-004**	**In Series**	**:**	1
NAME	**:**	**Feed Heater4**	**In Parallel**	**:**	none

General Data

Service	**:**	**- Heat Exchanger** **- ~~Cooler~~** **- ~~Condenser (water cooled)~~**	**- ~~Vaporizer~~** **- ~~Reboiler~~**
Type	**:**	**- ~~Fixed Tube Sheets~~** **- ~~Floating Head~~** **- ~~Hair Pin~~** **- ~~Double Tube~~**	**- ~~Plate Heat Exchanger~~** **- ~~Finned Tubes~~** **- ~~Thermosyphon~~**
Position	**:**	**- Horizontal** **- ~~Vertical~~**	

Capacity	**[kW]**	**:**	7536	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	7569	**(Calc.)**
Overall Heat Transfer Coefficient	**[W/m^2·^{o}C]**	**:**	50	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^{o}C]**	**:**	27	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^{o}C]**	**:**	20.25	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	No. 007+008	No. 201
Mass Stream	**[kg/s]**	**:**	9.10	9.10
Mass Stream to				
- ~~Evaporate~~	**~~[kg/s]~~**	**:**		
- ~~Condense~~	**~~[kg/s]~~**	**:**		-
Average Specific Heat	**[kJ/kg·^{o}C]**	**:**	3.5	3.4
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^{o}C]**	**:**	540.0	850.0
Temperature OUT	**[^{o}C]**	**:**	795.0	550.0
Pressure	**[bara]**	**:**	1	1
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**E-005**	**In Series**	:	1
NAME	:	**Feed Heater5 (1)**	**In Parallel**	:	none

General Data

Service :
- **Electric heater**
- ~~**Vaporizer**~~
- ~~**Cooler**~~
- ~~**Reboiler**~~
- ~~**Condenser (water cooled)**~~

Type :
- ~~**Fixed Tube Sheets**~~
- ~~**Plate Heat Exchanger**~~
- ~~**Floating Head**~~
- ~~**Finned Tubes**~~
- ~~**Hair Pin**~~
- ~~**Thermosyphon**~~
- ~~**Double Tube**~~

Position :
- ~~**Horizontal**~~
- ~~**Vertical**~~

Capacity	**[kW]**	:	1645	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	-	**(Calc.)**
Overall Heat Transfer Coefficient	**[W/m^2·^{o}C]**	:	-	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^{o}C]**	:	-	
Passes Tube Side		:	-	
Passes Shell Side		:	-	
Correction Factor LMTD (min. 0.75)		:	-	
Corrected LMTD	**[^{o}C]**	:	-	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	No. 009	electric
Mass Stream	**[kg/s]**	:	9.10	-
Mass Stream to				
- ~~**Evaporate**~~	~~**[kg/s]**~~	:		
- ~~**Condense**~~	~~**[kg/s]**~~	:		-
Average Specific Heat	**[kJ/kg·^{o}C]**	:	3.26	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^{o}C]**	:	540.0	-
Temperature OUT	**[^{o}C]**	:	850.0	-
Pressure	**[bara]**	:	1	-
Material		:	CS	-

Remarks:

(1) Use during start up and used for temperature controllability during normal operation, target temperature is 850 C

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
	Date	:	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-006**	**In Series**	**:**	1
NAME	**:**	**Feed Heater6 (1)**	**In Parallel**	**:**	none

General Data

Service : - **Electric heater** - **~~Vaporizer~~**
- **~~Cooler~~** - **~~Reboiler~~**
- **~~Condenser (water cooled)~~**

Type : - **~~Fixed Tube Sheets~~** - **~~Plate Heat Exchanger~~**
- **~~Floating Head~~** - **~~Finned Tubes~~**
- **~~Hair Pin~~** - **~~Thermosyphon~~**
- **~~Double Tube~~**

Position : - **~~Horizontal~~**
- **~~Vertical~~**

Capacity	**[kW]**	**:**	1760	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	-	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	**:**	-	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	**:**	-	
Passes Tube Side		**:**	-	
Passes Shell Side		**:**	-	
Correction Factor LMTD (min. 0.75)		**:**	-	
Corrected LMTD	**[^oC]**	**:**	-	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	No. 317	eletric
Mass Stream	**[kg/s]**	**:**	9.83	-
Mass Stream to				
- **~~Evaporate~~**	**~~[kg/s]~~**	**:**		
- **~~Condense~~**	**~~[kg/s]~~**	**:**		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	**:**	2.8	-
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^oC]**	**:**	476.0	-
Temperature OUT	**[^oC]**	**:**	540.0	-
Pressure	**[bara]**	**:**	1	-
Material		**:**	CS	-

Remarks:

(1) Use during start up and used for temperature controllability during normal operation, target temperature is 540 C

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-101**	**In Series**	**:**	1
NAME	**:**	**Shell Product cooler**	**In Parallel**	**:**	none

General Data

Service		**:**	- ~~Heat Exchanger~~ - **Cooler** - ~~Condenser~~	- ~~Vaporizer~~ - ~~Reboiler~~
Type		**:**	- ~~Fixed Tube Sheets~~ - ~~Floating Head~~ - ~~Hair Pin~~ - **Double Tube**	- ~~Plate Heat Exchanger~~ - ~~Finned Tubes~~ - ~~Thermosyphon~~
Position		**:**	- **Horizontal** - ~~Vertical~~	
Capacity	**[kW]**	**:**	200	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	167	**(Calc.)**
Overall Heat Transfer Coefficient	**[W/m^2·^{o}C]**	**:**	300	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^{o}C]**	**:**	5	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^{o}C]**	**:**	3.75	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Cooling water	No.103
Mass Stream	**[kg/s]**	**:**	10.0	10.84
Mass Stream to				
- ~~Evaporate~~	~~[kg/s]~~	**:**		
- ~~Condense~~	~~[kg/s]~~	**:**		-
Average Specific Heat	**[kJ/kg·^{o}C]**	**:**	4.2	3.52
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^{o}C]**	**:**	20.0	32.0
Temperature OUT	**[^{o}C]**	**:**	26.0	25
Pressure	**[bara]**	**:**	4	1
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number Date	: :	**CPD3297** 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**E-102**	**In Series**	:	1
NAME	:	**Compressed gas product cooler**	**In Parallel**	:	none

General Data

Service	:	- ~~**Heat Exchanger**~~ - **Cooler** - ~~**Condenser**~~	- ~~**Vaporizer**~~ - ~~**Reboiler**~~
Type	:	- ~~**Fixed Tube Sheets**~~ - ~~**Floating Head**~~ - ~~**Hair Pin**~~ - **Double Tube**	- ~~**Plate Heat Exchanger**~~ - ~~**Finned Tubes**~~ - ~~**Thermosyphon**~~
Position	:	- **Horizontal** - ~~**Vertical**~~	

Capacity	**[kW]**	:	4800.7	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	4513	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	:	100	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	:	15	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	0.75	
Corrected LMTD	**[^oC]**	:	11.25	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	No. 314	No. 105
Mass Stream	**[kg/s]**	:	9.84	10.84
Mass Stream to				
- ~~**Evaporate**~~	~~**[kg/s]**~~	:		
- ~~**Condense**~~	~~**[kg/s]**~~	:		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	:	3.2	2.69
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^oC]**	:	43.0	227.0
Temperature OUT	**[^oC]**	:	217.0	65.0
Pressure	**[bara]**	:	15	1
Material		:	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number Date	: :	**CPD3297** 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	: **E-103**	**In Series**	: 1
NAME	: **Compressed gas product cooler2**	**In Parallel**	: none

General Data

Service	:	- ~~**Heat Exchanger**~~ - **Cooler** - ~~**Condenser**~~	- ~~**Vaporizer**~~ - ~~**Reboiler**~~
Type	:	- ~~**Fixed Tube Sheets**~~ - ~~**Floating Head**~~ - ~~**Hair Pin**~~ - **Double Tube**	- ~~**Plate Heat Exchanger**~~ - ~~**Finned Tubes**~~ - ~~**Thermosyphon**~~
Position	:	- **Horizontal** - ~~**Vertical**~~	

Capacity	**[kW]**	:	294.3	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	36	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	:	300	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	:	36	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	0.75	
Corrected LMTD	**[^oC]**	:	27	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Cooling water	No. 106
Mass Stream	**[kg/s]**	:	9.83	10.84
Mass Stream to				
- ~~**Evaporate**~~	~~**[kg/s]**~~	:		
- ~~**Condense**~~	~~**[kg/s]**~~	:		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	:	4.2	2.69
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^oC]**	:	20.0	65.0
Temperature OUT	**[^oC]**	:	27.2	55.0
Pressure	**[bara]**	:	4	1
Material		:	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number :	**CPD3297**
		Date :	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-201**	**In Series**	**:**	1
NAME	**:**	**Tube product cooler**	**In Parallel**	**:**	none

General Data

Service	**:**	- ~~**Heat Exchanger**~~	- ~~**Vaporizer**~~
		- **Cooler**	- ~~**Reboiler**~~
		- ~~**Condenser**~~	
Type	**:**	- ~~**Fixed Tube Sheets**~~	- ~~**Plate Heat Exchanger**~~
		- ~~**Floating Head**~~	- ~~**Finned Tubes**~~
		- ~~**Hair Pin**~~	- ~~**Thermosyphon**~~
		- **Double Tube**	
Position	**:**	- **Horizontal**	
		- ~~**Vertical**~~	

Capacity	**[kW]**	**:**	4916	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	4106	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	**:**	50	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	**:**	32	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^oC]**	**:**	24	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	No.316	No. 202
Mass Stream	**[kg/s]**	**:**	9.84	9.10
Mass Stream to				
- ~~**Evaporate**~~	~~**[kg/s]**~~	**:**		
- ~~**Condense**~~	~~**[kg/s]**~~	**:**		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	**:**	2.80	3.40
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^oC]**	**:**	298.0	550.0
Temperature OUT	**[^oC]**	**:**	476.0	308.0
Pressure	**[bara]**	**:**	1	1
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number Date	: :	**CPD3297** 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**E-202**	**In Series**	:	1
NAME	:	**Tube product cooler2**	**In Parallel**	:	none

General Data

Service :
- ~~Heat Exchanger~~
- Cooler
- ~~Condenser~~
- ~~Vaporizer~~
- ~~Reboiler~~

Type :
- ~~Fixed Tube Sheets~~
- ~~Floating Head~~
- ~~Hair Pin~~
- Double Tube
- ~~Plate Heat Exchanger~~
- ~~Finned Tubes~~
- ~~Thermosyphon~~

Position :
- Horizontal
- ~~Vertical~~

Capacity	**[kW]**	:	2242.3	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	12659	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	:	50	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	:	5	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	0.75	
Corrected LMTD	**[^oC]**	:	3.75	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	No.317	No. 203
Mass Stream	**[kg/s]**	:	9.84	9.10
Mass Stream to				
- ~~**Evaporate**~~	~~**[kg/s]**~~	:		
- ~~**Condense**~~	~~**[kg/s]**~~	:		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	:	3.5	3.42
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^oC]**	:	217.0	308.0
Temperature OUT	**[^oC]**	:	298.0	219.0
Pressure	**[bara]**	:	1	30
Material		:	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER : **E-203** **In Series** : 1
NAME : **Tube product cooler3** **In Parallel** : none

General Data

Service	:	- ~~**Heat Exchanger**~~ - **Cooler** - ~~**Condenser**~~	- ~~**Vaporizer**~~ - ~~**Reboiler**~~
Type	:	- ~~**Fixed Tube Sheets**~~ - ~~**Floating Head**~~ - ~~**Hair Pin**~~ - **Double Tube**	- ~~**Plate Heat Exchanger**~~ - ~~**Finned Tubes**~~ - ~~**Thermosyphon**~~
Position	:	- **Horizontal** - ~~**Vertical**~~	

Capacity	**[kW]**	:	4259.7	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	609	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	:	300	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	:	31	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	0.75	
Corrected LMTD	**[^oC]**	:	23.25	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Cooling water	No. 204
Mass Stream	**[kg/s]**	:	11.65	9.10
Mass Stream to				
- Evaporate	**[kg/s]**	:	4.26	
~~**- Condense**~~	~~**[kg/s]**~~	:		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	:	4.2	2.82
Heat of Evap. / Condensation	**[kJ/kg]**	:	2200	-
Temperature IN	**[^oC]**	:	20.0	219.0
Temperature OUT	**[^oC]**	:	121.0	25.0
Pressure	**[bara]**	:	4	1.45
Material		:	CS	CS

Remarks:

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-204**	**In Series**	**:**	1
NAME	**:**	**Compressed tube gas cooler**	**In Parallel**	**:**	none

General Data

Service	:	**- ~~Heat Exchanger~~** **- Cooler** **- ~~Condenser~~**	**- ~~Vaporizer~~** **- ~~Reboiler~~**
Type	:	**- ~~Fixed Tube Sheets~~** **- ~~Floating Head~~** **- ~~Hair Pin~~** **- Double Tube**	**- ~~Plate Heat Exchanger~~** **- ~~Finned Tubes~~** **- ~~Thermosyphon~~**
Position	:	**- Horizontal** **- ~~Vertical~~**	

Capacity	**[kW]**	**:**	4967	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	363	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	**:**	300	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	**:**	61	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^oC]**	**:**	45.75	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Cooling water	No. 207
Mass Stream	**[kg/s]**	**:**	14.7	7.74
Mass Stream to				
- ~~Evaporate~~	**~~[kg/s]~~**	**:**		
- ~~Condense~~	**~~[kg/s]~~**	**:**		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	**:**	4.2	2.57
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^oC]**	**:**	20.0	30.0
Temperature OUT	**[^oC]**	**:**	100.0	288.6
Pressure	**[bara]**	**:**	4	30
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number :	**CPD3297**
		Date :	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-301A**	**In Series**	**:**	1
NAME	**:**	**Light Gas Column Reboiler**	**In Parallel**	**:**	none

General Data

Service	**:**	**- ~~Heat Exchanger~~**	**- ~~Vaporizer~~**
		- ~~Cooler~~	**- Reboiler**
		- ~~Condenser~~	
Type	**:**	**- ~~Fixed Tube Sheets~~**	**- ~~Plate Heat Exchanger~~**
		- ~~Floating Head~~	**- ~~Finned Tubes~~**
		- ~~Hair Pin~~	**- Thermosyphon**
		- ~~Double Tube~~	**-**
Position	**:**	**- ~~Horizontal~~**	
		- Vertical	

Capacity	**[kW]**	**:**	1515	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	85	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2\cdot{}^oC$]**	**:**	500	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	**:**	48	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^oC]**	**:**	36	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Hot water	Bottoms from T301
Mass Stream	**[kg/s]**	**:**	24.05	16.92
Mass Stream to				
- Evaporate	**[kg/s]**	**:**		3.5
- Condense	**[kg/s]**	**:**		-
Average Specific Heat	**[$kJ/kg\cdot{}^oC$]**	**:**	4.2	
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	432.86
Temperature IN	**[^oC]**	**:**	65.0	9.52
Temperature OUT	**[^oC]**	**:**	50.0	9.52
Pressure	**[bara]**	**:**	4	15
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin	Project ID-Number	:	**CPD3297**
	B. Wang Y. Zou	Date	:	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**E-301B**	**In Series**	**:** 1
NAME :	**Light Gas Column Condenser**	**In Parallel**	**:** none

General Data

Service :
- ~~**Heat Exchanger**~~ - ~~**Vaporizer**~~
- ~~**Cooler**~~ - ~~**Reboiler**~~
- **Condenser (water cooled)**

Type :
- ~~**Fixed Tube Sheets**~~ - ~~**Plate Heat Exchanger**~~
- ~~**Floating Head**~~ - ~~**Finned Tubes**~~
- ~~**Hair Pin**~~ - ~~**Thermosyphon**~~
- **Double Tube**

Position :
- **Horizontal**
- ~~**Vertical**~~

Capacity	**[kW]**	**:**	7290	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	1060	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	**:**	100	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	**:**	92	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^oC]**	**:**	69	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	H2 expanded	No. 302
Mass Stream	**[kg/s]**	**:**	11.3	14.19
Mass Stream to				
- ~~**Evaporate**~~	~~**[kg/s]**~~	**:**		
- **Condense**	**[kg/s]**	**:**		12.72
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	**:**	12.9	-
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	573.11
Temperature IN	**[^oC]**	**:**	-250.0	-131.0
Temperature OUT	**[^oC]**	**:**	-200.0	-131.0
Pressure	**[bara]**	**:**	5	15
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : Date :	**CPD3297** 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**E-302A**	**In Series :**	1
NAME :	**C2 Column Reboiler**	**In Parallel :**	none

General Data

Service	:	- ~~**Heat Exchanger**~~ - ~~**Cooler**~~ - ~~**Condenser**~~	- ~~**Vaporizer**~~ - **Reboiler**
Type	:	- ~~**Fixed Tube Sheets**~~ - ~~**Floating Head**~~ - ~~**Hair Pin**~~ - ~~**Double Tube**~~	- ~~**Plate Heat Exchanger**~~ - ~~**Finned Tubes**~~ - **Thermosyphon** -
Position	:	- ~~**Horizontal**~~ - **Vertical**	

Capacity	**[kW]**	:	4601	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	733	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2\cdot{}^oC$]**	:	500	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	:	17	
Passes Tube Side		:	1	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	0.75	
Corrected LMTD	**[^oC]**	:	12.75	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Hot water	Bottoms from T302
Mass Stream	**[kg/s]**	:	73.04	25.92
Mass Stream to				
- Evaporate	**[kg/s]**	:		14.77
~~**- Condense**~~	~~**[kg/s]**~~	:		-
Average Specific Heat	**[$kJ/kg\cdot{}^oC$]**	:	4.2	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	177.51
Temperature IN	**[^oC]**	:	65.0	39.65
Temperature OUT	**[^oC]**	:	50.0	39.65
Pressure	**[bara]**	:	4	15
Material		:	CS	CS

Remarks:

Designers : Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : **CPD3297** Date : 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER : **E-302B** — **In Series** : 1
NAME : **C2 Column Condenser** — **In Parallel** : none

General Data

Service : - ~~Heat Exchanger~~ - ~~Vaporizer~~
- ~~Cooler~~ - ~~Reboiler~~
- Condenser

Type : - ~~Fixed Tube Sheets~~ - ~~Plate Heat Exchanger~~
- ~~Floating Head~~ - ~~Finned Tubes~~
- ~~Hair Pin~~ - ~~Thermosyphon~~
- Double Tube

Position : - Horizontal
- ~~Vertical~~

Capacity	**[kW]**	:	3954	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	526	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	:	300	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	:	33	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	0.75	
Corrected LMTD	**[^oC]**	:	24.75	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	C2= expanded	No. 306
Mass Stream	**[kg/s]**	:	8.23	10.93
Mass Stream to				
- ~~Evaporate~~	~~[kg/s]~~	:		
- **Condense**	**[kg/s]**	:		10.93
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	:	96.09	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	364.76
Temperature IN	**[^oC]**	:	-75.0	-39.0
Temperature OUT	**[^oC]**	:	-70.0	-39.0
Pressure	**[bara]**	:	4	15
Material		:	CS	CS

Remarks:

Designers : Montree l. O. Muraza W.K. Lin
B. Wang Y. Zou

Project ID-Number : **CPD3297**
Date : 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-303A**	**In Series**	**:** 1
NAME	**:**	**C3 Column Reboiler**	**In Parallel**	**:** none

General Data

Service	**:**	- ~~**Heat Exchanger**~~ - ~~**Cooler**~~ - ~~**Condenser**~~	- ~~**Vaporizer**~~ - **Reboiler**
Type	**:**	- ~~**Fixed Tube Sheets**~~ - ~~**Floating Head**~~ - ~~**Hair Pin**~~ - ~~**Double Tube**~~	- ~~**Plate Heat Exchanger**~~ - ~~**Finned Tubes**~~ - **Thermosyphon** -
Position	**:**	- ~~**Horizontal**~~ - **Vertical**	

Capacity	**[kW]**	**:**	61	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	1073	**(Calc.)**
Overall Heat Transfer Coefficient	**[W/m^2·^{o}C]**	**:**	400	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^{o}C]**	**:**	0.19	
Passes Tube Side		**:**	1	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^{o}C]**	**:**	0.14	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	No.313	Bottom from T303
Mass Stream	**[kg/s]**	**:**	18.1	119.7
Mass Stream to				
- Evaporate	**[kg/s]**	**:**		-
~~**- Condense**~~	~~**[kg/s]**~~	**:**		-
Average Specific Heat	**[kJ/kg·^{o}C]**	**:**	4.8	3.24
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^{o}C]**	**:**	43.2	42.5
Temperature OUT	**[^{o}C]**	**:**	42.5	42.7
Pressure	**[bara]**	**:**	15	15
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : Date :	**CPD3297** 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-303B**	**In Series**	**:**	1
NAME	**:**	**C3 Column Condenser**	**In Parallel**	**:**	none

General Data

Service	**:**	**- ~~Heat Exchanger~~** **- ~~Cooler~~** **- Condenser (water cooled)**	**- ~~Vaporizer~~** **- ~~Reboiler~~**
Type	**:**	**- ~~Fixed Tube Sheets~~** **- ~~Floating Head~~** **- ~~Hair Pin~~** **- Double Tube**	**- ~~Plate Heat Exchanger~~** **- ~~Finned Tubes~~** **- ~~Thermosyphon~~**
Position	**:**	**- Horizontal** **- ~~Vertical~~**	

Capacity	**[kW]**	**:**	604	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	503	**(Calc.)**
Overall Heat Transfer Coefficient	**[W/m^2·^{o}C]**	**:**	400	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^{o}C]**	**:**	4	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^{o}C]**	**:**	3	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Propylene reflux	No. 311
Mass Stream	**[kg/s]**	**:**	22.8	105.2
Mass Stream to				
- ~~Evaporate~~	**~~[kg/s]~~**	**:**		
- Condense	**[kg/s]**	**:**	22.8	-
Average Specific Heat	**[kJ/kg·^{o}C]**	**:**	19.1	19.1
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	319.8	-
Temperature IN	**[^{o}C]**	**:**	48.8	35.0
Temperature OUT	**[^{o}C]**	**:**	35.0	38.0
Pressure	**[bara]**	**:**	15	15
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number :	**CPD3297**
		Date :	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-303C**	**In Series**	**:** 1
NAME	**:**	**Heat Compressor after cooler**	**In Parallel**	**:** none

General Data

Service	**:**	- ~~**Heat Exchanger**~~ - **Cooler** - ~~**Condenser**~~	- ~~**Vaporizer**~~ - ~~**Reboiler**~~
Type	**:**	- ~~**Fixed Tube Sheets**~~ - ~~**Floating Head**~~ - ~~**Hair Pin**~~ - **Double Tube**	- ~~**Plate Heat Exchanger**~~ - ~~**Finned Tubes**~~ - ~~**Thermosyphon**~~
Position	**:**	- **Horizontal** - ~~**Vertical**~~	

Capacity	**[kW]**	**:**	543	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	145	**(Calc.)**
Overall Heat Transfer Coefficient	**[W/m^2·^{o}C]**	**:**	1000	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^{o}C]**	**:**	5	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^{o}C]**	**:**	3.75	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Cooling water	From E-303B
Mass Stream	**[kg/s]**	**:**	25.9	105.29
Mass Stream to				
- Evaporate	**[kg/s]**	:	-	-
- Condense	**[kg/s]**	**:**	-	-
Average Specific Heat	**[kJ/kg·^{o}C]**	**:**	4.2	1.91
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^{o}C]**	**:**	20.0	45.9
Temperature OUT	**[^{o}C]**	**:**	25.0	43.2
Pressure	**[bara]**	**:**	1	17
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number :	**CPD3297**
		Date :	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**E-401**	**In Series**	:	1
NAME	:	**MDEA Cooler**	**In Parallel**	:	none

General Data

Service	:	- ~~Heat Exchanger~~ - Cooler - ~~Condenser~~	- ~~Vaporizer~~ - ~~Reboiler~~
Type	:	- ~~Fixed Tube Sheets~~ - ~~Floating Head~~ - ~~Hair Pin~~ - ~~Double Tube~~	- ~~Plate Heat Exchanger~~ - ~~Finned Tubes~~ - Thermosyphon -
Position	:	- ~~Horizontal~~ - Vertical	

Capacity	**[kW]**	:	8,132	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	464	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	:	1,000	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	:	23	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	0.75	
Corrected LMTD	**[^oC]**	:	17.25	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Cooling Water	Bottoms from T-402
Mass Stream	**[kg/s]**	:	38.7	31.91
Mass Stream to				
- Evaporize	**[kg/s]**	:	-	-
- ~~Condense~~	**~~[kg/s]~~**	:	-	-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	:	4.2	3.54
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^oC]**	:	20	101.97
Temperature OUT	**[^oC]**	:	70	30.0
Pressure	**[bara]**	:	4	30
Material		:	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number Date	: :	**CPD3297** 16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-402**	**In Series**	**:**	1
NAME	**:**	**Stripper Reboiler**	**In Parallel**	**:**	none

General Data

Service	**:**	**- ~~Heat Exchanger~~** **- ~~Cooler~~** **- ~~Condenser~~**	**- ~~Vaporizer~~** **- Reboiler**
Type	**:**	**- ~~Fixed Tube Sheets~~** **- ~~Floating Head~~** **- ~~Hair Pin~~** **- ~~Double Tube~~**	**- ~~Plate Heat Exchanger~~** **- ~~Finned Tubes~~** **- ~~Thermosyphon~~** **-**
Position	**:**	**- Horizontal** **- ~~Vertical~~**	

Capacity	**[kW]**	**:**	15756	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	233	**(Calc.)**
Overall Heat Transfer Coefficient	**[W/m^2·^{o}C]**	**:**	500	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^{o}C]**	**:**	180	
Passes Tube Side		**:**	1	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^{o}C]**	**:**	135	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Hot water	Bottoms from T-402
Mass Stream	**[kg/s]**	**:**	181.9	32.1
Mass Stream to				
- Evaporize	**[kg/s]**	**:**		18.36
- ~~Condense~~	**~~[kg/s]~~**	**:**		-
Average Specific Heat	**[kJ/kg·^{o}C]**	**:**	4.6	-
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	858.17
Temperature IN	**[^{o}C]**	**:**	150	102.0
Temperature OUT	**[^{o}C]**	**:**	105	104.6
Pressure	**[bara]**	**:**	4	1
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**AE-101**	**In Series**	:	1
NAME	:	**Shell Product cooler**	**In Parallel**	:	none

General Data

Service :
- ~~Heat Exchanger~~
- Cooler
- ~~Condenser~~
- ~~Vaporizer~~
- ~~Reboiler~~

Type :
- ~~Fixed Tube Sheets~~
- ~~Floating Head~~
- ~~Hair Pin~~
- Double Tube
- ~~Plate Heat Exchanger~~
- ~~Finned Tubes~~
- ~~Thermosyphon~~

Position :
- Horizontal
- ~~Vertical~~

Capacity	**[kW]**	:	14722	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	621	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	:	450	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	:	70	
Passes Tube Side		:	2	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	0.75	
Corrected LMTD	**[^oC]**	:	52.5	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	BFW	No.A102
Mass Stream	**[kg/s]**	:	15.24	10.84
Mass Stream to				
- ~~**Evaporate**~~	~~**[kg/s]**~~	:		
- ~~**Condense**~~	~~**[kg/s]**~~	:		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	:	4.2	2.63
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^oC]**	:	20.0	540.0
Temperature OUT	**[^oC]**	:	250.0	25
Pressure	**[bara]**	:	4	1
Material		:	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**AE-102**	**In Series**	**:**	1
NAME	**:**	**Compressed gas product cooler**	**In Parallel**	**:**	none

General Data

Service	:	- ~~Heat Exchanger~~ - Cooler - ~~Condenser~~	- ~~Vaporizer~~ - ~~Reboiler~~
Type	:	- ~~Fixed Tube Sheets~~ - ~~Floating Head~~ - ~~Hair Pin~~ - Double Tube	- ~~Plate Heat Exchanger~~ - ~~Finned Tubes~~ - ~~Thermosyphon~~
Position	:	- Horizontal - ~~Vertical~~	

Capacity	**[kW]**	**:**	4140	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	172	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	**:**	450	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	**:**	71	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^oC]**	**:**	53.25	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Cooling water	No. A104
Mass Stream	**[kg/s]**	**:**	12.32	10.84
Mass Stream to				
- ~~**Evaporate**~~	~~**[kg/s]**~~	**:**		
- ~~**Condense**~~	~~**[kg/s]**~~	**:**		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	**:**	4.2	2.22
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^oC]**	**:**	20.0	227.0
Temperature OUT	**[^oC]**	**:**	100.0	55.0
Pressure	**[bara]**	**:**	4	30
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number :	**CPD3297**
		Date :	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**AE-202**	**In Series**	**:**	1
NAME	**:**	**Tube product cooler**	**In Parallel**	**:**	none

General Data

Service	**:**	**- ~~Heat Exchanger~~**	**- ~~Vaporizer~~**
		- Cooler	**- ~~Reboiler~~**
		- ~~Condenser~~	
Type	**:**	**- ~~Fixed Tube Sheets~~**	**- ~~Plate Heat Exchanger~~**
		- ~~Floating Head~~	**- ~~Finned Tubes~~**
		- ~~Hair Pin~~	**- ~~Thermosyphon~~**
		- Double Tube	
Position	**:**	**- Horizontal**	
		- ~~Vertical~~	

Capacity	**[kW]**	**:**	20727	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	756	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	**:**	450	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	**:**	81	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^oC]**	**:**	60.75	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	BFW	No. A205
Mass Stream	**[kg/s]**	**:**	10.28	9.10
Mass Stream to				
- ~~Evaporate~~	**~~[kg/s]~~**	**:**		
- ~~Condense~~	**~~[kg/s]~~**	**:**		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	**:**	4.2	2.77
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^oC]**	**:**	20.0	850.0
Temperature OUT	**[^oC]**	**:**	500.0	25.0
Pressure	**[bara]**	**:**	4	1
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:** **AE-203**		**In Series**	**:** 1
NAME	**:** **Tube product cooler**		**In Parallel**	**:** none

General Data

Service :
- ~~Heat Exchanger~~
- Cooler
- ~~Condenser~~
- ~~Vaporizer~~
- ~~Reboiler~~

Type :
- ~~Fixed Tube Sheets~~
- ~~Floating Head~~
- ~~Hair Pin~~
- **Double Tube**
- ~~Plate Heat Exchanger~~
- ~~Finned Tubes~~
- ~~Thermosyphon~~

Position :
- **Horizontal**
- ~~Vertical~~

Capacity	**[kW]**	**:**	4970	**(Calc.)**
Heat Exchange Area	**[m^2]**	**:**	538	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	**:**	450	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	**:**	27	
Passes Tube Side		**:**	2	
Passes Shell Side		**:**	1	
Correction Factor LMTD (min. 0.75)		**:**	0.75	
Corrected LMTD	**[^oC]**	**:**	20.25	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	BFW	No. A208
Mass Stream	**[kg/s]**	**:**	5.14	7.74
Mass Stream to				
- Evaporate	**[kg/s]**	**:**	4.26	
- ~~Condense~~	**~~[kg/s]~~**	**:**		-
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	**:**	4.2	2.3
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	2200	-
Temperature IN	**[^oC]**	**:**	20.0	308.0
Temperature OUT	**[^oC]**	**:**	250.0	30.0
Pressure	**[bara]**	**:**	4	30
Material		**:**	CS	CS

Remarks:

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number : Date :	**CPD3297** 16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**P101 A/B (AP202)**	**Operating**	: 1
NAME	:	**T301 Bottom Pump**	**Installed Spare**	: 1

Service : Process water pump
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid			:	Propylene / Propane/Ethylene		
Temperature	**(T)**	**[^{o}C]**	:	25.0		
Density	**(ρ)**	**[kg/m^3]**	:	1,000		
Viscosity	**(η)**	**[N·s/m^2]**	:	0.0006		
Vapour Pressure	**(p_v)**	**[bara]**	:	-	**at Temperature [^{o}C]**	: 50.0

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	0.365*10^{-3}
Suction Pressure	**(p_s)**	**[bara]**	:	1.0
Discharge Pressure	**(p_d)**	**[bara]**	:	3.0
Theoretical Power		**[kW]**	:	0.172
Pump Efficiency		**[-]**	:	0.7
Power at Shaft		**[kW]**	:	0.245

Construction Details *(1)*

RPM		1700	**Nominal diameter**		
Drive		Electrical	**Suction Nozzle**	**[…]**	
Type electrical motor			**Discharge Nozzle**	**[…]**	
Tension	**[V]**	380	**Cooled Bearings**		**Yes / No**
Rotational direction		**Clock /**	**Cooled Stuffing Box**		**Yes / No**
		~~Counter Cl.~~	**Smothering Gland**		**Yes / No**
Foundation Plate		**Combined /**	**If yes**		
		~~two parts~~	**- Seal Liquid**		**Yes / No**
Flexible Coupling		Yes	**- Splash Rings**		**Yes / No**
Pressure Gauge Suction		No	**- Packing Type**		
Pressure Gauge Discharge		Yes	**- Mechanical Seal**		**Yes / No**
Min. Overpressure above			**- N.P.S.H.**	**[m]**	
p_v/p_m	**[bar]**	0.1	**$\{= p_m \cdot \rho \cdot g\}$**		

Construction Materials *(2)*

Pump House		MS	**Wear Rings**		
Pump Rotor		HT Steel	**Shaft Box**		
Shaft		HT Steel			
Special provisions		none			
Operating Pressure	**[bara]**	3.0	**Test Pressure**	**[bara]**	

Remarks:

(1) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(2) MS = Mild Steel; HT Steel = High Tensile Steel

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number :	**CPD3297**
		Date :	16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**P102 A/B**	**Operating**	:	1
NAME	:	**T301 Bottom Pump**	**Installed Spare**	:	1

Service : Process water pump
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid			:	Propylene / Propane/Ethylene			
Temperature	**(T)**	**[^{o}C]**	:	25.0			
Density	**(ρ)**	**[kg/m^3]**	:	1,000			
Viscosity	**(η)**	**[N·s/m^2]**	:	0.0006			
Vapour Pressure	**(p_v)**	**[bara]**	:	-	**at Temperature [^{o}C]**	:	50.0

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	$0.365*10^{-3}$
Suction Pressure	**(p_s)**	**[bara]**	:	1.0
Discharge Pressure	**(p_d)**	**[bara]**	:	3.0
Theoretical Power		**[kW]**	:	0.172
Pump Efficiency		**[-]**	:	0.7
Power at Shaft		**[kW]**	:	0.245

Construction Details *(1)*

RPM		: 1700	**Nominal diameter**			
Drive		: Electrical	**Suction Nozzle**	**[…]**	:	
Type electrical motor		:	**Discharge Nozzle**	**[…]**	:	
Tension	**[V]**	: 380	**Cooled Bearings**		:	**Yes / No**
Rotational direction		: **Clock / ~~Counter Cl.~~**	**Cooled Stuffing Box**		:	**Yes / No**
			Smothering Gland		:	**Yes / No**
Foundation Plate		: **Combined / ~~two parts~~**	**If yes**			
			- Seal Liquid		:	**Yes / No**
Flexible Coupling		: Yes	**- Splash Rings**		:	**Yes / No**
Pressure Gauge Suction		: No	**- Packing Type**		:	
Pressure Gauge Discharge		: Yes	**- Mechanical Seal**		:	**Yes / No**
Min. Overpressure above p_v/p_m	**[bar]**	: 0.1	**- N.P.S.H.** $\{ = p_m \cdot \rho \cdot g \}$	**[m]**	:	

Construction Materials *(2)*

Pump House		: MS	**Wear Rings**	:
Pump Rotor		: HT Steel	**Shaft Box**	:
Shaft		: HT Steel		
Special provisions		: none		
Operating Pressure	**[bara]**	: 3.0	**Test Pressure [bara]**	:

Remarks:

(3) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(4) MS = Mild Steel; HT Steel = High Tensile Steel

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number :	**CPD3297**
		Date :	16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER : **P103 A/B** **Operating :** 1
NAME : **T301 Bottom Pump** **Installed Spare :** 1

Service : Process water pump
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid			Propylene / Propane/Ethylene
Temperature	**(T)**	**[°C]**	25.0
Density	**(ρ)**	**[kg/m^3]**	1,000
Viscosity	**(η)**	**[N·s/m^2]**	0.0006
Vapour Pressure	**(p_v)**	**[bara]**	-

at Temperature [°C] : 50.0

Power

Capacity	**(Φ_v)**	**[m^3/s]**	$0.365*10^{-3}$
Suction Pressure	**(p_s)**	**[bara]**	1.0
Discharge Pressure	**(p_d)**	**[bara]**	3.0
Theoretical Power		**[kW]**	0.172
Pump Efficiency		**[-]**	0.7
Power at Shaft		**[kW]**	0.245

Construction Details *(1)*

RPM : 1700
Drive : Electrical
Type electrical motor :
Tension [V] : 380
Rotational direction : **Clock / ~~Counter Cl.~~**
Foundation Plate : **Combined / ~~two parts~~**
Flexible Coupling : Yes
Pressure Gauge Suction : No
Pressure Gauge Discharge : Yes
Min. Overpressure above p_v/p_m [bar] : 0.1

Nominal diameter
Suction Nozzle […] :
Discharge Nozzle […] :
Cooled Bearings : **Yes / No**
Cooled Stuffing Box : **Yes / No**
Smothering Gland : **Yes / No**
If yes
- Seal Liquid : **Yes / No**
- Splash Rings : **Yes / No**
- Packing Type :
- Mechanical Seal : **Yes / No**
- N.P.S.H. [m] :
$\{ = p_m \cdot \rho g \}$

Construction Materials *(2)*

Pump House : MS
Pump Rotor : HT Steel
Shaft : HT Steel
Special provisions : none
Operating Pressure [bara] : 3.0

Wear Rings :
Shaft Box :
Test Pressure [bara] :

Remarks:

(5) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.
(6) MS = Mild Steel; HT Steel = High Tensile Steel

Designers : Montree l. O. Muraza W.K. Lin
B. Wang Y. Zou

Project ID-Number : **CPD3297**
Date : 16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER : **P301 A/B**
NAME : **T301 Bottom Pump**
Operating : 1
Installed Spare : 1

Service : Bottom pump
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid : Propylene / Propane/Ethylene
Temperature **(T)** **[^{o}C]** : 9.5
Density **(ρ)** **[kg/m^3]** : 456
Viscosity **(η)** **[N·s/m^2]** : 0.0001
Vapour Pressure **(p_v)** **[bara]** : 20.8 **at Temperature [^{o}C]** : 50.0

Power

Capacity **(Φ_v)** **[m^3/s]** : 0.029
Suction Pressure **(p_s)** **[bara]** : 15.0
Discharge Pressure **(p_d)** **[bara]** : 16.5

Theoretical Power **[kW]** : 4.38

Pump Efficiency **[-]** : 0.7
Power at Shaft **[kW]** : 6.25

Construction Details (1)

RPM : 3000
Drive : Electrical
Type electrical motor :
Tension **[V]** : 380
Rotational direction : **Clock / ~~Counter Cl.~~**
Foundation Plate : **Combined / ~~two parts~~**
Flexible Coupling : Yes
Pressure Gauge Suction : No
Pressure Gauge Discharge : Yes
Min. Overpressure above p_v/p_m **[bar]** : 0.1

Nominal diameter
Suction Nozzle **[…]** :
Discharge Nozzle **[…]** :
Cooled Bearings : **Yes / No**
Cooled Stuffing Box : **Yes / No**
Smothering Gland : **Yes / No**
If yes
- Seal Liquid : **Yes / No**
- Splash Rings : **Yes / No**
- Packing Type :
- Mechanical Seal : **Yes / No**
- N.P.S.H. **[m]** :
$\{ = p_m \cdot \rho g \}$

Construction Materials (2)

Pump House : MS
Pump Rotor : HT Steel
Shaft : HT Steel
Special provisions : none

Operating Pressure **[bara]** : 16.5

Wear Rings :
Shaft Box :

Test Pressure **[bara]** :

Remarks:

(7) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.
(8) MS = Mild Steel; HT Steel = High Tensile Steel

Designers : Montree l. O. Muraza W.K. Lin
B. Wang Y. Zou

Project ID-Number : **CPD3297**
Date : 16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**P301 C/D**	**Operating :**	1
NAME :	**T301 Reflux Pump**	**Installed Spare :**	1

Service : Reflux pump
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid			:	Propylene / Propane/Ethylene
Temperature	**(T)**	**[^{o}C]**	:	-131
Density	**(ρ)**	**[kg/m^3]**	:	502
Viscosity	**(η)**	**[N·s/m^2]**	:	0.0001
Vapour Pressure	**(p_v)**	**[bara]**	:	20.8 **at Temperature [^{o}C]** : 50.0

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	0.025
Suction Pressure	**(p_s)**	**[bara]**	:	15.0
Discharge Pressure	**(p_d)**	**[bara]**	:	16.5
Theoretical Power		**[kW]**	:	5.06
Pump Efficiency		**[-]**	:	0.7
Power at Shaft		**[kW]**	:	7.31

Construction Details *(1)*

RPM :	3000	**Nominal diameter**	
Drive :	Electrical	**Suction Nozzle [...]** :	
Type electrical motor :		**Discharge Nozzle [...]** :	
Tension [V] :	380	**Cooled Bearings** :	**Yes / No**
Rotational direction :	**Clock /**	**Cooled Stuffing Box** :	**Yes / No**
	~~Counter Cl.~~	**Smothering Gland** :	**Yes / No**
Foundation Plate :	**Combined /**	**If yes**	
	~~two parts~~	**- Seal Liquid** :	**Yes / No**
Flexible Coupling :	Yes	**- Splash Rings** :	**Yes / No**
Pressure Gauge Suction :	No	**- Packing Type** :	
Pressure Gauge Discharge :	Yes	**- Mechanical Seal** :	**Yes / No**
Min. Overpressure above		**- N.P.S.H. [m]** :	
p_v/p_m [bar] :	0.1	**{ = $p_m \cdot \rho \cdot g$ }**	

Construction Materials *(2)*

Pump House :	MS	**Wear Rings** :	
Pump Rotor :	HT Steel	**Shaft Box** :	
Shaft :	HT Steel		
Special provisions :	none		
Operating Pressure [bara] :	16.5	**Test Pressure [bara]** :	

Remarks:

(9) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(10) MS = Mild Steel; HT Steel = High Tensile Steel

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number :	**CPD3297**
		Date :	16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**P302 A/B**	**Operating**	:	1
NAME	:	**T302 Bottom Pump**	**Installed Spare**	:	1

Service : **Bottom** pump
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid			:	Propylene / Propane/Ethylene			
Temperature	**(T)**	**[^{o}C]**	:	39.9/41.2			
Density	**(ρ)**	**[kg/m^3]**	:	426			
Viscosity	**(η)**	**[N·s/m^2]**	:	0.0001			
Vapour Pressure	**(p_v)**	**[bara]**	:	20.8	**at Temperature [^{o}C]**	:	50.0

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	0.026
Suction Pressure	**(p_s)**	**[bara]**	:	15.0
Discharge Pressure	**(p_d)**	**[bara]**	:	23.0
Theoretical Power		**[kW]**	:	21
Pump Efficiency		**[-]**	:	0.7
Power at Shaft		**[kW]**	:	30

Construction Details *(1)*

RPM		:	3000	**Nominal diameter**			
Drive		:	Electrical	**Suction Nozzle**	**[…]**	:	
Type electrical motor		:		**Discharge Nozzle**	**[…]**	:	
Tension	**[V]**	:	380	**Cooled Bearings**		:	**Yes / No**
Rotational direction		:	**Clock /**	**Cooled Stuffing Box**		:	**Yes / No**
			~~Counter Cl.~~	**Smothering Gland**		:	**Yes / No**
Foundation Plate		:	**Combined /**	**If yes**			
		~~two parts~~		**- Seal Liquid**		:	**Yes / No**
Flexible Coupling		:	Yes	**- Splash Rings**		:	**Yes / No**
Pressure Gauge Suction		:	No	**- Packing Type**		:	
Pressure Gauge Discharge		:	Yes	**- Mechanical Seal**		:	**Yes / No**
Min. Overpressure above				**- N.P.S.H.**	**[m]**	:	
p_v/p_m	**[bar]**	:	0.1	**{ = $p_m \cdot \rho \cdot g$ }**			

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		:	
Pump Rotor		:	HT Steel	**Shaft Box**		:	
Shaft		:	HT Steel				
Special provisions		:	none				
Operating Pressure	**[bara]**	:	23	**Test Pressure**	**[bara]**	:	

Remarks:

(11) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(12) MS = Mild Steel; HT Steel = High Tensile Steel

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**P302 C/D**	**Operating**	:	1
NAME	:	**T302 Reflux Pump**	**Installed Spare**	:	1

Service	:	Reflux pump
Type	:	Centrifugal
Number	:	2

Operating Conditions & Physical Data

Pumped liquid			:	Propylene / Propane/Ethylene			
Temperature	**(T)**	**[^{o}C]**	:	-39			
Density	**(ρ)**	**[kg/m^3]**	:	426			
Viscosity	**(η)**	**[N·s/m^2]**	:	0.0001			
Vapour Pressure	**(p_v)**	**[bara]**	:	20.8	**at Temperature [^{o}C]**	:	50.0

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	0.021
Suction Pressure	**(p_s)**	**[bara]**	:	15.0
Discharge Pressure	**(p_d)**	**[bara]**	:	17.0
Theoretical Power		**[kW]**	:	4
Pump Efficiency		**[-]**	:	0.7
Power at Shaft		**[kW]**	:	6

Construction Details *(1)*

RPM		:	3000	**Nominal diameter**				
Drive		:	Electrical	**Suction Nozzle**	**[…]**	:		
Type electrical motor		:		**Discharge Nozzle**	**[…]**	:		
Tension	**[V]**	:	380	**Cooled Bearings**		:	**Yes / No**	
Rotational direction		:	**Clock /** ~~**Counter Cl.**~~	**Cooled Stuffing Box**		:	**Yes / No**	
				Smothering Gland		:	**Yes / No**	
Foundation Plate		:	**Combined /** ~~**two parts**~~	**If yes**				
				- Seal Liquid		:	**Yes / No**	
Flexible Coupling		:	Yes	**- Splash Rings**		:	**Yes / No**	
Pressure Gauge Suction		:	No	**- Packing Type**		:		
Pressure Gauge Discharge		:	Yes	**- Mechanical Seal**		:	**Yes / No**	
Min. Overpressure above p_v/p_m	**[bar]**	:	0.1	**- N.P.S.H.** $\{= p_m \cdot \rho \cdot g\}$	**[m]**	:		

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		:	
Pump Rotor		:	HT Steel	**Shaft Box**		:	
Shaft		:	HT Steel				
Special provisions		:	none				
Operating Pressure	**[bara]**	:	17.0	**Test Pressure**	**[bara]**	:	

Remarks:

(13) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(14) MS = Mild Steel; HT Steel = High Tensile Steel

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**P303 A/B**	**Operating**	:	1
NAME	:	**T303 Bottom Pump**	**Installed Spare**	:	1

Service	:	Reflux pump
Type	:	Centrifugal
Number	:	2

Operating Conditions & Physical Data

Pumped liquid			:	Propylene / Propane/Ethylene			
Temperature	**(T)**	**[^oC]**	:	43			
Density	**(ρ)**	**[kg/m^3]**	:	417			
Viscosity	**(η)**	**[$N{\cdot}s/m^2$]**	:	0.0001			
Vapour Pressure	**(p_v)**	**[bara]**	:	20.8	**at Temperature [^oC]**	:	50.0

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	0.024
Suction Pressure	**(p_s)**	**[bara]**	:	15.0
Discharge Pressure	**(p_d)**	**[bara]**	:	17.0
Theoretical Power		**[kW]**	:	4.7
Pump Efficiency		**[-]**	:	0.7
Power at Shaft		**[kW]**	:	7.0

Construction Details *(1)*

RPM		:	3000	**Nominal diameter**			
Drive		:	Electrical	**Suction Nozzle**	**[…]**	:	
Type electrical motor		:		**Discharge Nozzle**	**[…]**	:	
Tension	**[V]**	:	380	**Cooled Bearings**		:	**Yes / No**
Rotational direction		:	**Clock / ~~Counter Cl.~~**	**Cooled Stuffing Box**		:	**Yes / No**
				Smothering Gland		:	**Yes / No**
Foundation Plate		:	**Combined / ~~two parts~~**	**If yes**			
				- Seal Liquid		:	**Yes / No**
Flexible Coupling		:	Yes	**- Splash Rings**		:	**Yes / No**
Pressure Gauge Suction		:	No	**- Packing Type**		:	
Pressure Gauge Discharge		:	Yes	**- Mechanical Seal**		:	**Yes / No**
Min. Overpressure above p_v/p_m	**[bar]**	:	0.1	**- N.P.S.H.** $\{=p_m{\cdot}\rho{\cdot}g\}$	**[m]**	:	

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		:
Pump Rotor		:	HT Steel	**Shaft Box**		:
Shaft		:	HT Steel			
Special provisions		:	none			
Operating Pressure	**[bara]**	:	17.0	**Test Pressure**	**[bara]**	:

Remarks:

(15) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(16) MS = Mild Steel; HT Steel = High Tensile Steel

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**P401 A/B**	**Operating**	:	1
NAME	:	**T401 Bottom Pump**	**Installed Spare**	:	1

Service : Reflux pump
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid			:	Propylene / Propane/Ethylene			
Temperature	**(T)**	**[^{o}C]**	:	30			
Density	**(ρ)**	**[kg/m^3]**	:	990			
Viscosity	**(η)**	**[N·s/m^2]**	:	0.0006			
Vapour Pressure	**(p_v)**	**[bara]**	:	-	**at Temperature [^{o}C]**	:	50.0

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	0.02
Suction Pressure	**(p_s)**	**[bara]**	:	30.0
Discharge Pressure	**(p_d)**	**[bara]**	:	31.5
Theoretical Power		**[kW]**	:	3.0
Pump Efficiency		**[-]**	:	0.7
Power at Shaft		**[kW]**	:	4.5

Construction Details *(1)*

RPM		:	3000	**Nominal diameter**			
Drive		:	Electrical	**Suction Nozzle**	**[…]**	:	
Type electrical motor		:		**Discharge Nozzle**	**[…]**	:	
Tension	**[V]**	:	380	**Cooled Bearings**		:	**Yes / No**
Rotational direction		:	**Clock /** **~~Counter Cl.~~**	**Cooled Stuffing Box**		:	**Yes / No**
				Smothering Gland		:	**Yes / No**
Foundation Plate		:	**Combined /** **~~two parts~~**	**If yes**			
				- Seal Liquid		:	**Yes / No**
Flexible Coupling		:	Yes	**- Splash Rings**		:	**Yes / No**
Pressure Gauge Suction		:	No	**- Packing Type**		:	
Pressure Gauge Discharge		:	Yes	**- Mechanical Seal**		:	**Yes / No**
Min. Overpressure above				**- N.P.S.H.**	**[m]**	:	
p_v/p_m	**[bar]**	:	0.1	**$\{ = p_m \cdot \rho \cdot g \}$**			

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		:	
Pump Rotor		:	HT Steel	**Shaft Box**		:	
Shaft		:	HT Steel				
Special provisions		:	none				
Operating Pressure	**[bara]**	:	31.5	**Test Pressure**	**[bara]**	:	

Remarks:

(17) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(18) MS = Mild Steel; HT Steel = High Tensile Steel

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	:	**CPD3297**
		Date	:	16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**P402 A/B**	**Operating**	: 1
NAME	:	**MDEA Recycle Pump 1**	**Installed Spare**	: 1

Service : Reflux pump
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid			:	Propylene / Propane/Ethylene		
Temperature	**(T)**	**[^{o}C]**	:	30		
Density	**(ρ)**	**[kg/m^3]**	:	990		
Viscosity	**(η)**	**[N·s/m^2]**	:	0.0006		
Vapour Pressure	**(p_v)**	**[bara]**	:	-	**at Temperature [^{o}C]**	: 50.0

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	0.020
Suction Pressure	**(p_s)**	**[bara]**	:	30.0
Discharge Pressure	**(p_d)**	**[bara]**	:	31.5
Theoretical Power		**[kW]**	:	3.0
Pump Efficiency		**[-]**	:	0.7
Power at Shaft		**[kW]**	:	4.5

Construction Details *(1)*

RPM		:	3000	**Nominal diameter**			
Drive		:	Electrical	**Suction Nozzle**	**[…]**	:	
Type electrical motor		:		**Discharge Nozzle**	**[…]**	:	
Tension	**[V]**	:	380	**Cooled Bearings**		:	**Yes / No**
Rotational direction		:	**Clock / ~~Counter Cl.~~**	**Cooled Stuffing Box**		:	**Yes / No**
				Smothering Gland		:	**Yes / No**
Foundation Plate		:	**Combined / ~~two parts~~**	**If yes**			
				- Seal Liquid		:	**Yes / No**
Flexible Coupling		:	Yes	**- Splash Rings**		:	**Yes / No**
Pressure Gauge Suction		:	No	**- Packing Type**		:	
Pressure Gauge Discharge		:	Yes	**- Mechanical Seal**		:	**Yes / No**
Min. Overpressure above				**- N.P.S.H.**	**[m]**	:	
p_v/p_m	**[bar]**	:	0.1	**{ = $p_m \cdot \rho \cdot g$ }**			

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		:
Pump Rotor		:	HT Steel	**Shaft Box**		:
Shaft		:	HT Steel			
Special provisions		:	none			
Operating Pressure	**[bara]**	:	31.5	**Test Pressure**	**[bara]**	:

Remarks:

(19) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(20) MS = Mild Steel; HT Steel = High Tensile Steel

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number :	**CPD3297**
		Date :	16 December 2003

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**P403 A/B**	**Operating**	: 1
NAME	:	**MDEA Recycle Pump2**	**Installed Spare**	: 1

Service : Reflux pump
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid			:	Propylene / Propane/Ethylene		
Temperature	**(T)**	**[°C]**	:	30		
Density	**(ρ)**	**[kg/m^3]**	:	990		
Viscosity	**(η)**	**[N·s/m^2]**	:	0.0006		
Vapour Pressure	**(p_v)**	**[bara]**	:	-	**at Temperature [°C]**	: 50.0

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	0.020
Suction Pressure	**(p_s)**	**[bara]**	:	30.0
Discharge Pressure	**(p_d)**	**[bara]**	:	31.5
Theoretical Power		**[kW]**	:	3.0
Pump Efficiency		**[-]**	:	0.7
Power at Shaft		**[kW]**	:	4.5

Construction Details *(1)*

RPM		: 3000	**Nominal diameter**		
Drive		: Electrical	**Suction Nozzle**	**[…]**	:
Type electrical motor		:	**Discharge Nozzle**	**[…]**	:
Tension	**[V]**	: 380	**Cooled Bearings**		: **Yes / No**
Rotational direction		: **Clock /**	**Cooled Stuffing Box**		: **Yes / No**
		~~Counter Cl.~~	**Smothering Gland**		: **Yes / No**
Foundation Plate		: **Combined /**	**If yes**		
		~~two parts~~	**- Seal Liquid**		: **Yes / No**
Flexible Coupling		: Yes	**- Splash Rings**		: **Yes / No**
Pressure Gauge Suction		: No	**- Packing Type**		:
Pressure Gauge Discharge		: Yes	**- Mechanical Seal**		: **Yes / No**
Min. Overpressure above			**- N.P.S.H.**	**[m]**	:
p_v/p_m	**[bar]**	: 0.1	**$\{= p_m \cdot \rho \cdot g\}$**		

Construction Materials *(2)*

Pump House		: MS	**Wear Rings**		:
Pump Rotor		: HT Steel	**Shaft Box**		:
Shaft		: HT Steel			
Special provisions		: none			
Operating Pressure	**[bara]**	: 31.5	**Test Pressure**	**[bara]**	:

Remarks:

(21) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(22) MS = Mild Steel; HT Steel = High Tensile Steel

Designers :	Montree l. O. Muraza W.K. Lin B. Wang Y. Zou	Project ID-Number	: **CPD3297**
		Date	: 16 December 2003

APPENDIX F

Appendix F Process safety

F.1. Hazard and Operability Studies (HAZOP)

1. The explanation of guidewords using in HAZOP

Table F.1 *Standard guidewords and their generic meanings*

Guide word	**Meaning**
No (not, none)	None of the design intent is achieved
More (more of, higher)	Quantitative increase in a parameter
Less (less of , lower)	Quantitative decrease in a parameter
As well as (more than)	An additional activity occurs
Part of	Only some of the design intention is achieved
Reverse	Logical opposite of the design intention occurs
Other than (other)	Complete substitution – another activity takes place
Other useful guidewords include:	
Where else	Applicable for flows, transfers, sources and destinations
Before/after	The step (or some part of it) is effected out of sequence
Early/late sequence	The timing is different from the intention
Faster/slower	The step is done/not done with the right timing

2. The procedure of HAZOP is given in HAZOP guide/volume 6

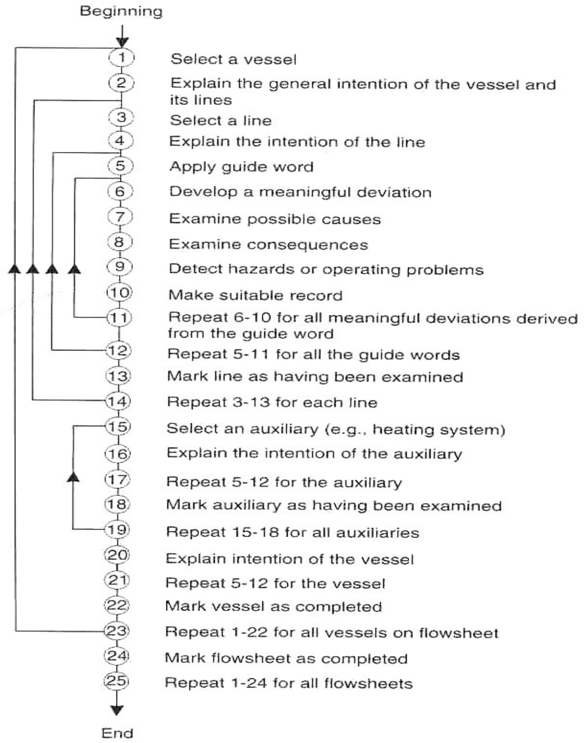


Figure F.1 *the procedure of HAZOP*

F.2. Dow Fire and Explosion Index (F & EI)

1. The procedure for calculating unit hazard factor F & EI shown here is referred from Figure F.2. [Guide]

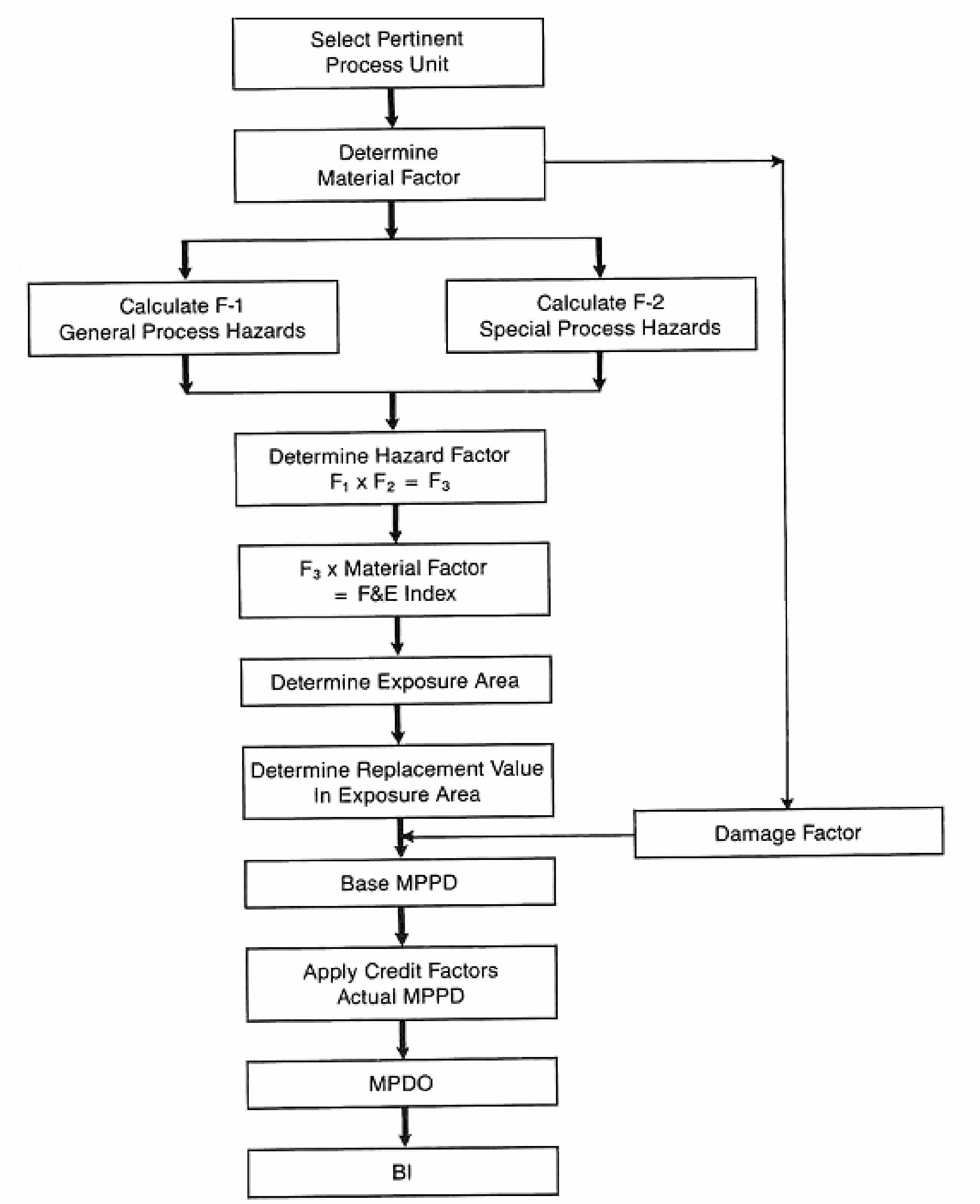
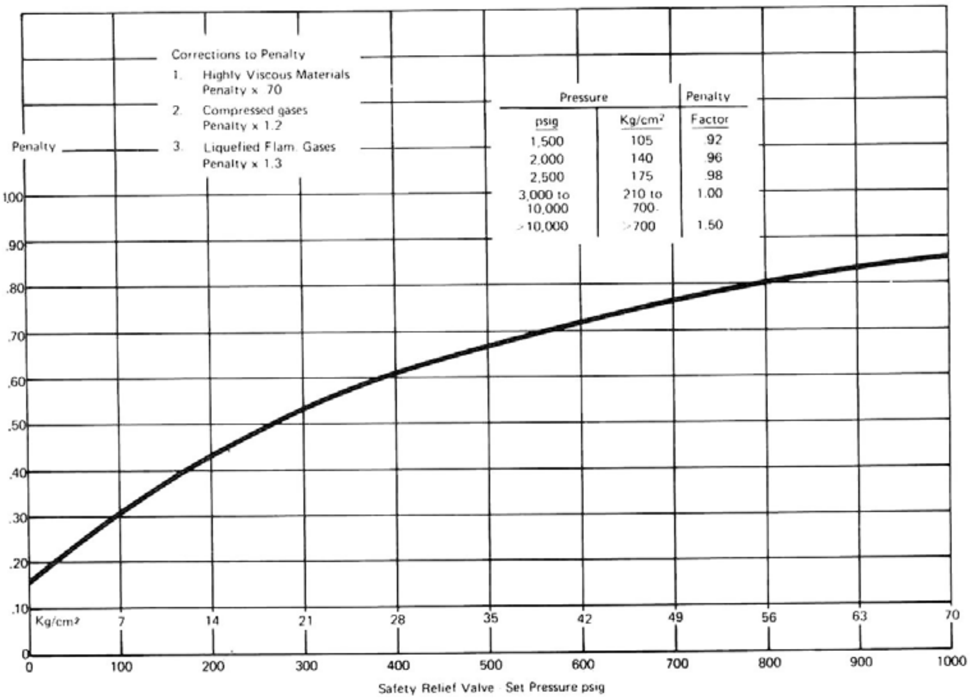


Figure F.2 *The procedure for calculating unit hazard factor F & EI*

2. To determine general process hazards and special process hazards, some figures from Guide are used in calculation.

For *relief pressure* item, Figure A10.3 determines the relation of pressure penalty and set pressure.

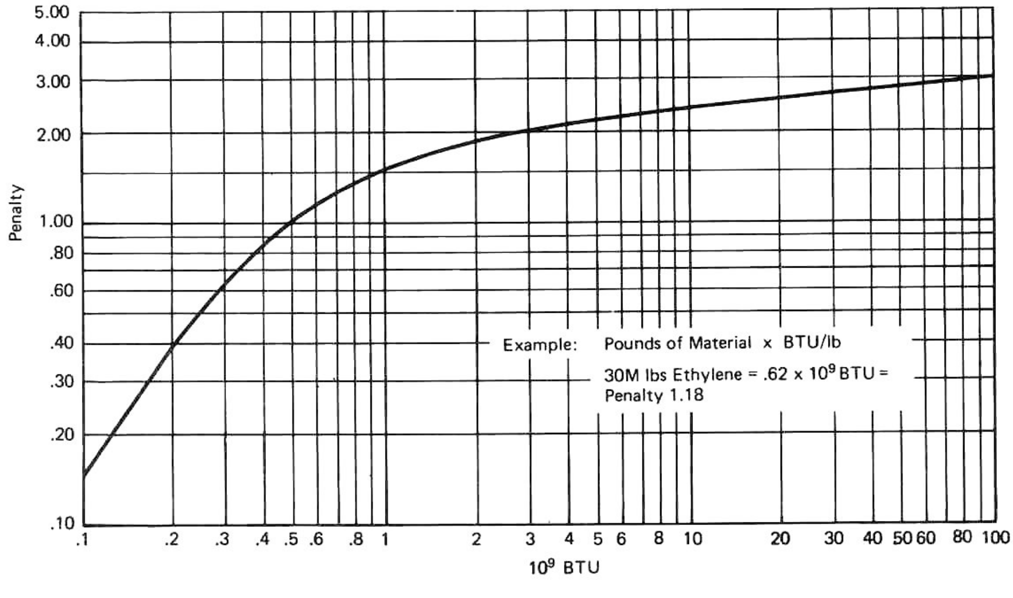


Corrections to Penalty
1. Highly Viscous Materials Penalty x .70
2. Compressed gases Penalty x 1.2
3. Liquefied Flam. Gases Penalty x 1.3

Pressure		Penalty
psig	Kg/cm²	Factor
1,500	105	.92
2,000	140	.96
2,500	175	.98
3,000 to 10,000	210 to 700	1.00
>10,000	>700	1.50

Figure F.3 *Pressure penalty for flammable and combustible liquids*

For *quantity of flammable and unstable materials* item, flammability vs. penalty is given in Figure F.4



Example: Pounds of Material x BTU/lb
30M lbs Ethylene = .62 x 10^9 BTU = Penalty 1.18

Figure F.4 *Liquids or Gases flammability in process*

3. In order to check if shell and tube reactor is the Pertinent Process Unit, other three distillation columns' F & EI are also calculated. The results list as below:

Table F.2 *Dow Fire and Explosion Index Form of light ends distillation column*

PREPARED BY CPD 3297	APPROVED BY		DATE 03-Dec-03
SITE Light ends distillation column	LOCATION Grey area		
MATERIALS IN PROCESS UNIT	Propane, nitrogen, ethylene, propylene, methane, hydrogen, water, carbon monoxide		
STATE OF OPERATION _X_DESIGN _NORMAL OP __SHUTDOWN	BASIC MATERIAL(S) FOR MATERIAL FACTOR	Ethylene	

MATERIAL FACTOR (Table I or Appendices A or B) Note requirements when unit temp over 60^{o}C		24
1. GENERAL PROCESS HAZARDS	**Penalty Factor Range**	**Penalty Factor Used**[1]
BASE FACTOR……………………………………………………………………………………….	1.00	1.00
A. EXOTHERMIC CHEMICAL REACTIONS	0.30 to 1.25	0.00
B. ENDOTHERMIC PROCESSES	0.20 to 0.40	0.00
C. MATERIAL HANDLING AND TRANSFER	0.25 to 1.05	0.50
D. ENCLOSED OR INDOOR PROCESS UNITS	0.25 to 0.90	0.00
E. ACCESS	0.20 to 0.35	0.00
F. DRAINAGE AND SPILL CONTROL		0.00
GENERAL PROCESS HAZARDS FACTOR (F_1)……………………………………………………		1.50
2. SPECIAL PROCESS HAZARDS		
BASE FACTOR……………………………………………………………………………………….	1.00	1.00
A. TOXIC MATERIAL(S)	0.20 to 0.80	0.40
B. SUB ATMOSPHERIC PRESSURE (>500 mm Hg)	0.50	0.00
C. OPERATION IN OR NEAR FLAMMABLE RANGE _ Inerted _X_ Not Inerted		
1. TANK FARMS STORAGE FLAMMABLE LIQUIDS	0.50	
2. PROCESS UPSET OR PURGE FAILURE	0.30	
3. ALWAYS IN FLAMMABLE RANGE	0.80	0.80
D. DUST EXPLOSION (See Table 3)	0.25 to 2.00	0.00
E. PRESSURE (See Fig 2) Operating Pressure 217.6 psig		0.45
F. LOW TEMPERATURE	0.20 to 0.30	0.00
G. QUALITY OF FLAMMABLE / UNSTABLE MATERIAL : Quantity 19686 lb H_C = 21.5*103BTU/lb		
1. LIQUIDS OR GASES IN PROCESS (See Fig 3)		0.84
2. LIQUIDS OR GASES IN STORAGE (See Fig 4)		
3. COMBUSTABLE SOLIDS IN STORAGE, DUST IN PROCESS (See Fig 5)		
H. CORROSION AND EROSION	0.10 to 0.75	0.00
I. LEAKAGE - JOINTS AND PACKING	0.10 to 1.50	0.00
J. USE OF FIRED EQUIPMENT (See Fig 6)		0.00
K. HOT OIL EXCHANGE SYSTEMS (See Table 5)	0.15 to 1.15	0.00
L. ROTATING EQUIPMENT	0.50	0.00
SPECIAL PROCESS HAZARDS (F_2)………………………………………………………………………………………		3.49
PROCESS UNITS FACTOR HAZARDS (F_1 x F_2) = F_3…………………………………………………………………………		5.24
FIRE AND EXPLOSION INDEX (F_3 x MF = F&EI)…………………………………………………………………………		125.64

Table F.3 *Dow Fire and Explosion Index Form of ethylene distillation column*

PREPARED BY CPD 3297	APPROVED BY	DATE 03-Dec-03
SITE Ethylene distillation column	LOCATION Grey area	
MATERIALS IN PROCESS UNIT Propane, ethylene, propylene, water		
STATE OF OPERATION _X_DESIGN _NORMAL OP __SHUTDOWN	BASIC MATERIAL(S) FOR MATERIAL FACTOR	Ethylene
MATERIAL FACTOR (Table I or Appendices A or B) Note requirements when unit temp over 60°C		24
1. GENERAL PROCESS HAZARDS	**Penalty Factor Range**	**Penalty Factor Used**[1]
BASE FACTOR	1.00	1.00
A. EXOTHERMIC CHEMICAL REACTIONS	0.30 to 1.25	0.00
B. ENDOTHERMIC PROCESSES	0.20 to 0.40	0.00
C. MATERIAL HANDLING AND TRANSFER	0.25 to 1.05	0.50
D. ENCLOSED OR INDOOR PROCESS UNITS	0.25 to 0.90	0.00
E. ACCESS	0.20 to 0.35	0.00
F. DRAINAGE AND SPILL CONTROL		0.00
GENERAL PROCESS HAZARDS FACTOR (F_1)		1.50
2. SPECIAL PROCESS HAZARDS		
BASE FACTOR	1.00	1.00
A. TOXIC MATERIAL(S)	0.20 to 0.80	0.40
B. SUB ATMOSPHERIC PRESSURE (>500 mm Hg)	0.50	0.00
C. OPERATION IN OR NEAR FLAMMABLE RANGE _Inerted _X_ Not Inerted		
1. TANK FARMS STORAGE FLAMMABLE LIQUIDS	0.50	
2. PROCESS UPSET OR PURGE FAILURE	0.30	
3. ALWAYS IN FLAMMABLE RANGE	0.80	0.80
D. DUST EXPLOSION (See Table 3)	0.25 to 2.00	0.00
E. PRESSURE (See Fig 2) Operating Pressure 217.6 psig		0.45
F. LOW TEMPERATURE	0.20 to 0.30	0.00
G. QUALITY OF FLAMMABLE / UNSTABLE MATERIAL : Quantity 17747 lb / kg H_C = 21.5*103 BTU/lb		
1. LIQUIDS OR GASES IN PROCESS (See Fig 3)		0.80
2. LIQUIDS OR GASES IN STORAGE (See Fig 4)		
3. COMBUSTABLE SOLIDS IN STORAGE, DUST IN PROCESS (See Fig 5)		
H. CORROSION AND EROSION	0.10 to 0.75	0.00
I. LEAKAGE - JOINTS AND PACKING	0.10 to 1.50	0.00
J. USE OF FIRED EQUIPMENT (See Fig 6)		0.00
K. HOT OIL EXCHANGE SYSTEMS (See Table 5)	0.15 to 1.15	0.00
L. ROTATING EQUIPMENT	0.50	0.00
SPECIAL PROCESS HAZARDS (F_2)		3.45
PROCESS UNITS FACTOR HAZARDS (F_1 x F_2) = F_3		5.18
FIRE AND EXPLOSION INDEX (F_3 x MF = F&EI)		124.20

Table F.4 *Dow Fire and Explosion Index Form of propylene distillation column*

PREPARED BY	CPD 3297	APPROVED BY	DATE 03-Dec-03
SITE	Propylene distillation column	LOCATION Grey area	
MATERIALS IN PROCESS UNIT	Propane, propylene, water		
STATE OF OPERATION _X_DESIGN __NORMAL OP __SHUTDOWN		BASIC MATERIAL(S) FOR MATERIAL FACTOR	Proylene
MATERIAL FACTOR (Table I or Appendices A or B) Note requirements when unit temp over 60°C			21
1. GENERAL PROCESS HAZARDS		**Penalty Factor Range**	**Penalty Factor Used[1]**
BASE FACTOR…		1.00	1.00
A. EXOTHERMIC CHEMICAL REACTIONS		0.30 to 1.25	0.00
B. ENDOTHERMIC PROCESSES		0.20 to 0.40	0.00
C. MATERIAL HANDLING AND TRANSFER		0.25 to 1.05	0.50
D. ENCLOSED OR INDOOR PROCESS UNITS		0.25 to 0.90	0.00
E. ACCESS		0.20 to 0.35	0.00
F. DRAINAGE AND SPILL CONTROL			0.00
GENERAL PROCESS HAZARDS FACTOR (F1)…			1.50
2. SPECIAL PROCESS HAZARDS			
BASE FACTOR…		1.00	1.00
A. TOXIC MATERIAL(S)		0.20 to 0.80	0.40
B. SUB ATMOSPHERIC PRESSURE (>500 mm Hg)		0.50	0.00
C. OPERATION IN OR NEAR FLAMMABLE RANGE	__Inerted_X_NotInerted		
1. TANK FARMS STORAGE FLAMMABLE LIQUIDS		0.50	
2. PROCESS UPSET OR PURGE FAILURE		0.30	
3. ALWAYS IN FLAMMABLE RANGE		0.80	0.80
D. DUST EXPLOSION (See Table 3)		0.25 to 2.00	0.00
E. PRESSURE (See Fig 2)	Operating Pressure 217.6psig		0.45
F. LOW TEMPERATURE		0.20 to 0.30	0.00
G. QUALITY OF FLAMMABLE / UNSTABLE MATERIAL :	Quantity 19257 lb / kg Hc = 21.5*103 BTU/lb		
1. LIQUIDS OR GASES IN PROCESS (See Fig 3)			0.84
2. LIQUIDS OR GASES IN STORAGE (See Fig 4)			
3. COMBUSTABLE SOLIDS IN STORAGE, DUST IN PROCESS (See Fig 5)			
H. CORROSION AND EROSION		0.10 to 0.75	0.00
I. LEAKAGE - JOINTS AND PACKING		0.10 to 1.50	0.00
J. USE OF FIRED EQUIPMENT (See Fig 6)			0.00
K. HOT OIL EXCHANGE SYSTEMS (See Table 5)		0.15 to 1.15	0.00
L. ROTATING EQUIPMENT		0.50	0.00
SPECIAL PROCESS HAZARDS (F2)…			3.49
PROCESS UNITS FACTOR HAZARDS (F1 x F2) = F3…			5.24
FIRE AND EXPLOSION INDEX (F3 x MF = F&EI)…			109.94

APPENDIX G

Appendix G Economics

Purchased Equipment Costs (PCE) with Lang method, can be found in the Table G.1.

Table G.1 *Purchased Equipment Costs*

Purchased Equipment Costs (PCE) in December 2003			
Type	Name	Equipment	Cost (US$)
Reactor	RX001	Shell & Tube	113,643
	RX002	Shell & Tube	113,643
Total			227,286
Drum	D101	Vessel	12,354
	D102	Vessel	23,488
	D201	Vessel	26,382
	D301	Vessel	24,217
	D302	Vessel	6,298
	D401	Vessel	5,212
Total			97,951
Column	T301	Packed Column	115,198
	T302	Packed Column	128,018
	T303	Packed Column	3,669,718
	T401	Packed Column	38,049
	T402	Packed Column	38,049
Total			3,989,032
Heat exchanger	E202	Shell & Tube	117,196
	E203	Shell & Tube	74,385
	E101	Shell & Tube	96,306
	E102	Shell & Tube	26,667
	E301A	Shell & Tube	24,795
	E301B	Shell & Tube	164,194
	E302A	Shell & Tube	113,555
	E302B	Shell & Tube	82,650
	E303A	Shell & Tube	166,315
	E303B	Shell & Tube	77,971
	E303C	Shell & Tube	22,456
	E402	Shell & Tube	61,987
	E401	Shell & Tube	148,149
Total			1,028,474
Furnace	E201	Process, Cylindrical	1,118,956
Compressor	C101	Reciprocating	935,706
	C201	Reciprocating	998,811
	C303	Reciprocating	209,010
Total			2,143,526
Total Purchased cost (US $)			8,605,226

From the *Table 6.1* [Coulson& Richardson, Volumn6], we can calculate Direct Capital Cost and Fixed Capital Cost in the Table G.2.

Table G.2 *Capital costs estimation*

Item	Process type Fluids
1. Major equipment as total purchased cost	
f1 : Equipment erection	0.40
f2 : Piping	0.70
f3 : Instrumentation	0.20
f4 : Electrical	0.10
f5 : Buildings,process	0.15
f6 : Utilities	0.20
f7 : Storages	0.15
f8 : Site development	0.05
f9 : Ancillary buildings	0.15
$\sum$ (f1+f2+f3+…+f9)	2.10
2. Total physical plant cost (PPC)	
PPC = PCE(1+ f1 + f2 + …+ f9)	
PPC = Direct Cost	26,676,201
3. Indirect cost	
f10 : Design and engineering	0.20
f11 : Contractor's fee	0.05
f12 : Contingency	0.05
$\sum$ (f10+f11+f12)	0.30
Indirect Capital Cost	8,002,860
Fixed Capital = PPC (1 + f10 + f11 +f12)	
Fixed Capital = Direct + Indirect cost	34,679,062

Costs	2003
	US$
1. Direct Capital Cost	26,676,201
2. Indirect Capital Cost	8,002,860
3. Fixed Capital Cost	34,679,062

Raw material and utilities costs, which is used in the process, determined in Table G.3

Table G.3 *Raw material and utilities costs*

Stream hrs/annum = 8040

Raw Materials	***Str.No.***	***kg/s***	***m3/a***	***ton/hrs***	***ton/a***	***t/t Alkenes***	***Price US$/unit***	***Unit***	***Cost US$/a @2003***	***Cost Million US$/a @2003***
Propane	001	8.23	-	29.64	238,341	1.1748	160	Ton	38,194,083	38.194
Oxygen	008	1.87	-	6.74	54,194	0.2671	143	Ton	7,771,432	7.771
Catalyst1 (V2O5/CeO2/SA5205)					6.8	0.00003	106,549	Ton	724,535	0.725
Catalyst2 (Pt on MFI zeolite)					3.3	0.00002	181,939	Ton	600,399	0.600
Total catalyst cost								Ton	1,324,934	1.325
Total Raw Material cost (IN)									47,290,450	47.290

Product	***Str.No.***	***kg/s***	***m3/a***	***ton/hrs***	***ton/a***	***t/t Alkenes***	***Price US$/ton***	***Unit***	***Income US$/a @2003***	***Income Million US$/a @2003***
Ethylene	308	2.29	-	8.26	66,376	0.3272	518	ton	34,382,700	34.383
Propylene	313	4.72		16.98	136,502	0.6728	408	ton	55,692,827	55.693
Light gas for syngas plant	303	1.47		5.28	42,414	0.2091	64	ton	2,718,717	2.719
CO2 for EOR	402	0.16		0.58	4,670	0.0230	6.5	ton	30,356	0.030
Water for EOR	208-5	1.46		5.26	42,267	0.2083	0.01	ton	423	0.000
Total Income (OUT)									92,825,022	92.825

Utilities	***Load***	***Cost US$/unit***	***Unit***	***Cost US$/a @2003***	***Cost Million US$ @2003***
Water (t/a)	11,188,806	0.01	Ton	111,888	0.112
Electric (kWh/a)	342,596,748	0.04	kWh	14,469,791	14.470
Total Utilities cost				14,581,679	14.582

Economic Criteria

a) Net Cash Flow (NCF) can be calculated from Gross Income and Production Costs.

$$\text{Net Cash Flow}_{\text{annual}} = \sum (\text{Gross Income}_{\text{annual}} - \text{Production Costs}_{\text{annual}})$$

$$\text{Gross Income}_{\text{annual}} = \Sigma \left(\text{Products}_{\text{annual}} \text{ x } \frac{\text{Price}}{\text{Unit}} \right)$$

$$= 92{,}825{,}022 \text{ US\$/a}$$

$$= 92.83 \text{ US\$ million}$$

$$\text{Production Cost}_{\text{annual}} = 79.10 \text{ US\$ million}$$

$$\text{Net Cash Flow}_{\text{annual}} = 92.83 - 79.10 \text{ US\$ million}$$

$$= 13.72 \text{ US\$ million}$$

b) Rate of Return (ROR) and Pay Out (Back) Time (POT or PBP) from the total Investment and NCF along ref(1) approach.

$$\text{ROR} = \frac{\text{Accu. Cash Flow}}{\text{(Project life*Tot.Investment)}}$$

$$\text{Accu.Cash Flow} = \text{(Net Cash Flow*Plant life) - Investment Cost +}$$

$$\text{+ Salvage Value}$$

$$= 13.72 \frac{\text{US\$ million}}{\text{year}} \text{*15 year) - 37.45 US\$ million +}$$

$$\text{+ (8\%Fixed Capital Cost) US\$ million}$$

$$= ((13.72*15) - 37.45 +(0.1*34.68))$$

$$= 171.12 \text{ US\$ million}$$

Project life time = Construction time + Plant life time + Salvage

= 2 + 15 +1

= 18 years

So,

$$\text{ROR} = \frac{171.12 \text{ US\$ million}}{18 \text{ years * } 37.45 \text{ US\$ million}}$$

$$= 25.38 \%$$

and,

POT or PBP = 6 years

In calculating cash flow (NCF), the project is usually considered as an isolated system, and taxes on profits and the effect of depreciation of the investment are not considered, since tax rates are not constant and depend on government policy as well as the rates of depreciation. Depreciation rates also depend on the accounting practices of the particular company. Therefore during evaluating projects, the effect of government policy must be taken into account at some stages particularly when considering projects in different country.

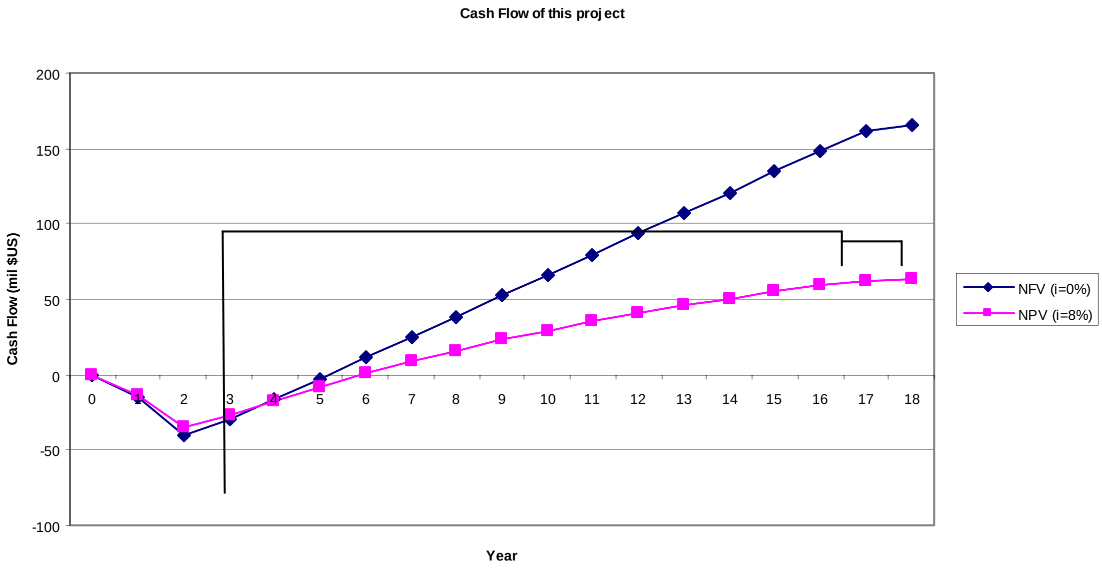


Figure G.1 *Cash flow of this project*

APPENDIX H

Appendix H.1 PFS for process with heat integration

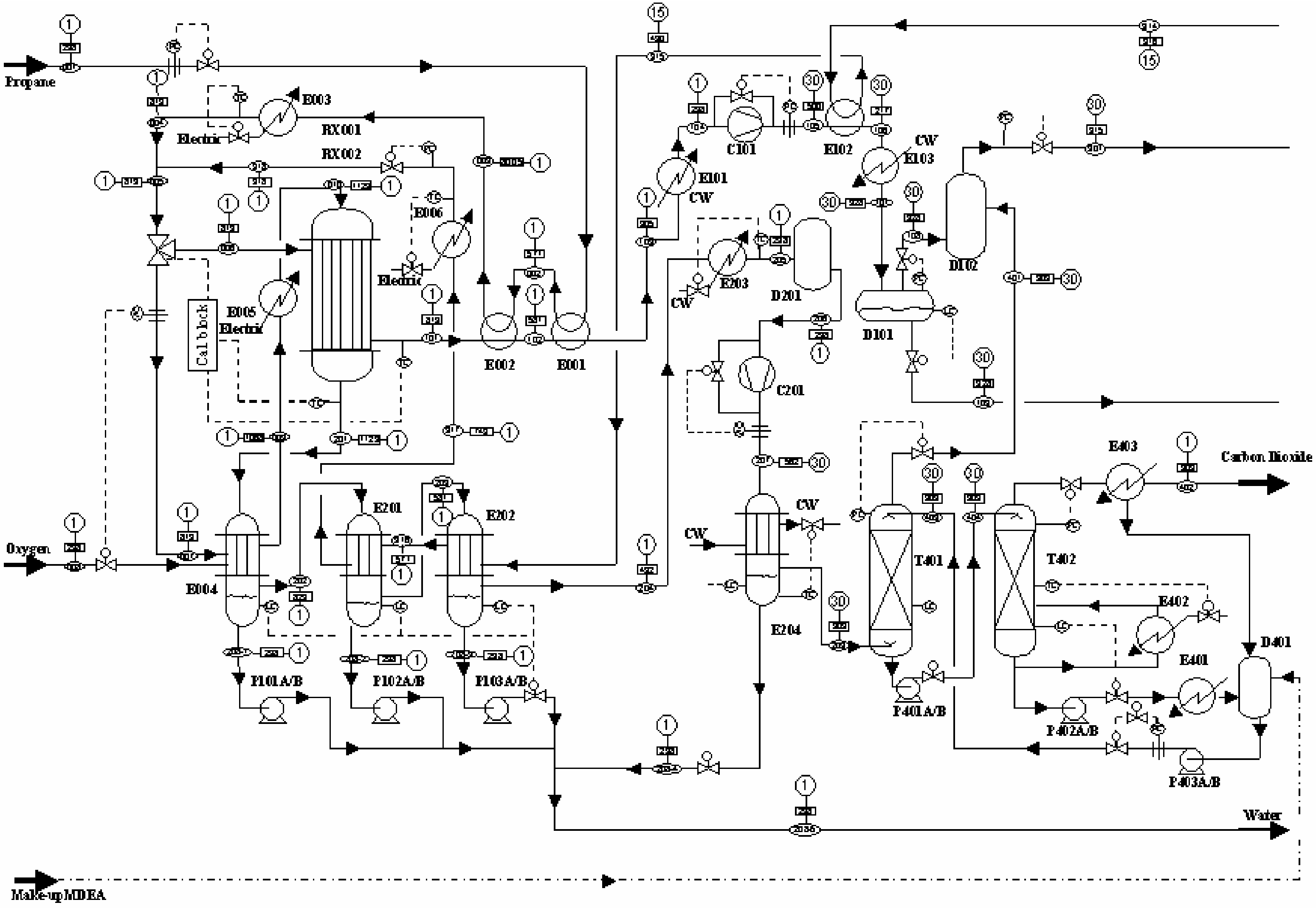


Figure H.1.1 *Process flow scheme Part 1.*

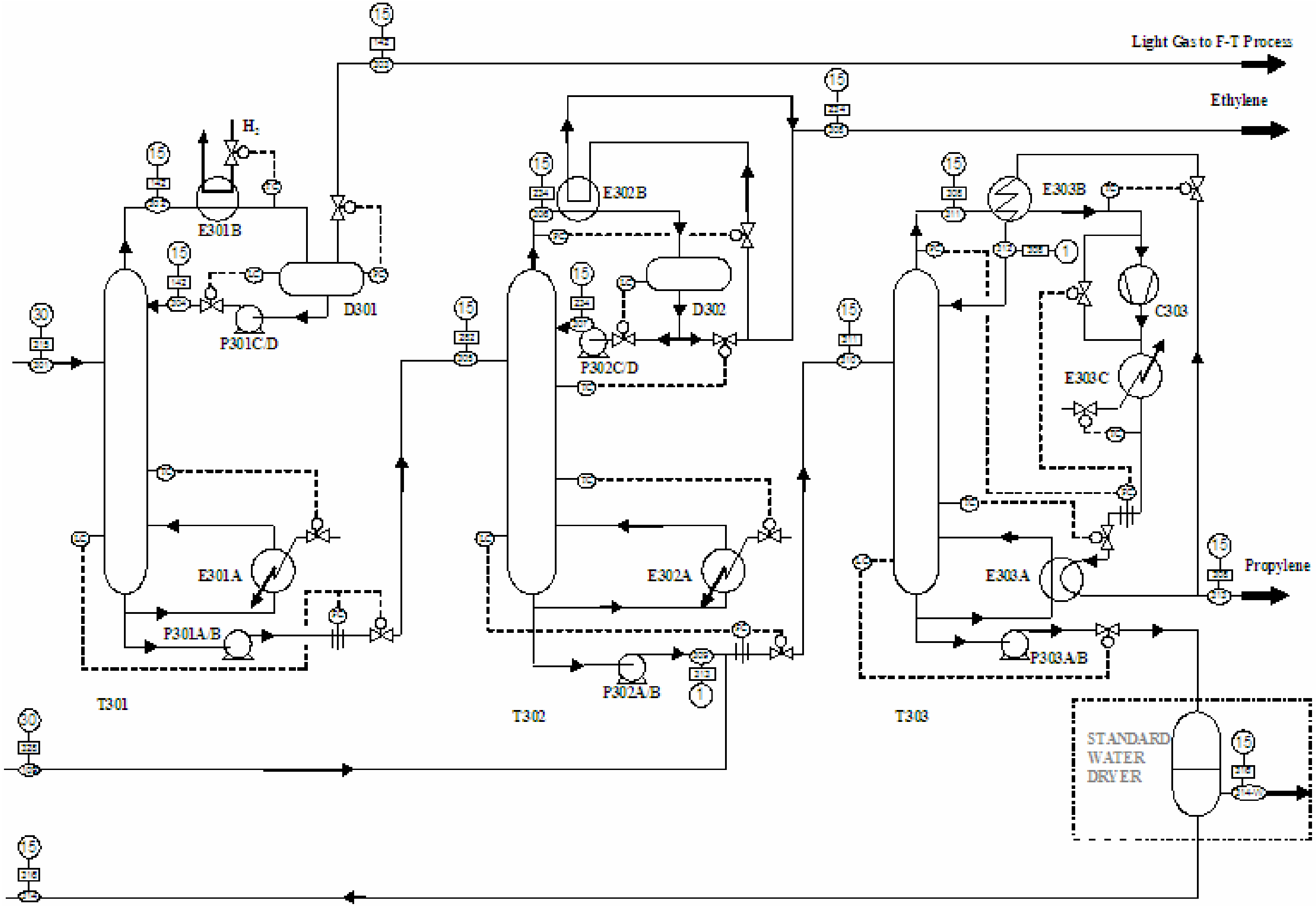


Figure H.1.2 *Process flow scheme Part 2*

Appendix H.2 Summary of utilities for the process with Heat integration

Table H.2. *Summary of utilities for process with Heat integration*

SUMMARY OF UTILITIES															
EQUIPMENT		UTILITIES													REMARKS
Nr.	Name	Heating					Cooling				Power				
		Load	Consumption (t/h)				Load	Consumption (t/h)			Actual	Consumption (t/h, kWh/h)			
			Steam			Hot		Cooling	Air	Refrig.	Load	Steam (t/h)		Electr.	
		kW	LP	MP	HP	Water	kW	Water			kW	HP	MP	kWh/h	
E003	Propane feed heater 3	359									359				[1] At E-301B, the H2 expanded is used as refrigerant [2] At 302B, the Ethylene is used as refrigerant [3]Energy requirement at E303A & E303B is fulfilled by heat pump
E005	Propane feed heater 5	1,645									1,645				
E006	Oxygen feed heater 1	1,760									1,760				
E001	Feed Heater1	7,890													
E002	Feed Heater2	6,632													
E004	Feed Heater4	7,536													
E101	Shell product cooler						200	36							
E102	Compressed gas product cooler						4,801								
E103	Compressed gas product cooler 2						294	35							
E201	Tube product cooler						4,916								
E202	Tube product cooler2						2,242								
E203	Tube product cooler 3						2,961	42							
E204	Compressed tube gas cooler						4,967	53							

Table H.2. *Summary of utilities for process with Heat integration (con't)*

SUMMARY OF UTILITIES															
EQUIPMENT		UTILITIES													REMARKS
Nr.	Name	Heating					Cooling				Power				
		Load kW	Consumption (t/h)				Load kW	Consumption (t/h)			Actual Load kW	Consumption (t/h, kWh/h)			
			Steam			Hot Water		Cooling Water	Air	Refrig.		Steam (t/h)		Electr. kWh/h	
			LP	MP	HP							HP	MP		
E301A	Light Gas Column Reboiler	1,515				87									
E301B	Light Gas Column Condenser						7,290			[1]					
E302A	C2 Column Reboiler	4,601				263									
E302B	C2 Column Condenser						3,954			[2]					
E303A	C3 Column Reboiler	61												[3]	
E303B	C3 Column Condenser						604							[3]	
E303C	Heat Compresor after cooler						543	93							
E402	T402 Reboiler	34,377				655									
E401	MDEA cooler						8,132	139							

Table H.2. *Summary of utilities for process with Heat integration (con't)*

SUMMARY OF UTILITIES															
EQUIPMENT		UTILITIES													REMARKS
Nr.	Name	Heating					Cooling				Power				
		Load	Consumption (t/h)				Load	Consumption (t/h)			Actual	Consumption (t/h, kWh/h)			
			Steam			Hot		Cooling	Air	Refrig.	Load	Steam (t/h)		Electr.	
		kW	LP	MP	HP	Water	kW	Water			kW	HP	MP	kWh/h	
C101	Shell Product Compressor										4,171				
C201	Tube Product Compressor										4,516				
C303	Propylene heat Compressor										670				
P301 A/B	T301Bottom pump										6.25				
P301C/D	T301Reflux pump										7.31				
P101A/B	Process water pump 1										0.25				
P102A/B	Process water pump 2										0.25				
P103A/B	Process water pump 3										0.25				
P401A/B	T401Bottom pump										4.50				
P402A/B	MDEA recycle pump										4.50				
P302A/B	T302Bottom pump										30.00				
P302C/D	T302Reflux pump										6.00				
P303A/B	T303Bottom pump										7.00				
P401A/B	T401Bottom pump										4.50				
P402A/B	MDEA recycle pump	66,376									4.50				
TOTAL	132,753	0	0	0	1,004	40,905	399	0	[2]	13,196	0	0	[1]		

Appendix H.3 Heat and Mass balance for the process with Heat integration

Table H.3.1a *Mass and heat balance for the process with heat integration*

STREAM Nr. :		001	IN	002		003		004		005=004+318	
Name :		Propane Feed In		Preheated Propane1		Preheated Propane2		Preheated Propane3		Total Propane Feed	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	7.45	0.1692	7.45	0.1692	7.45	0.1692	7.45	0.1692	16.36	0.3717
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Propylene	42	0.79	0.0188	0.79	0.0188	0.79	0.0188	0.79	0.0188	1.71	0.0407
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Methane	16	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Hydrogen	2	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Water	18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Nitrogen	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		8.23	0.1880	8.23	0.1880	8.23	0.1880	8.23	0.1880	18.07	0.4124
Enthalpy	kW	-17379		-12289		-6334		-5975		-13167	
Phase		V		V		V		V		V	
Press.	Bara	1		1		1		1		1	
Temp	oC	25		298		527.5		540		540	

Table H.3.1a *Mass and heat balance for the process with heat integration (Con't)*

STREAM Nr. :		006		007		008		009		010	
Name :		Propane to Shell Rx		Propane splitted to Tube		Oxygen feed		Material to Tube		Feed to Tube	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	9.81	0.2230	6.54	0.1487	0.00	0.0000	6.54	0.1487	6.54	0.1487
Oxygen	32	0.00	0.0000	0.00	0.0000	1.79	0.0559	1.79	0.0559	1.79	0.0559
Ethylene	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Propylene	42	1.03	0.0244	0.68	0.0163	0.00	0.0000	0.68	0.0163	0.68	0.0163
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Methane	16	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Hydrogen	2	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Water	18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Nitrogen	28	0.00	0.0000	0.00	0.0000	0.08	0.0029	0.08	0.0029	0.08	0.0029
Total		10.84	0.2475	7.23	0.1650	1.87	0.0589	9.10	0.2239	9.10	0.2239
Enthalpy	kW	-7900		-5267		959		-4308		4957	
Phase		V		V		V		V		V	
Press.	Bara	1		1		1		1		1	
Temp	oC	540		540		25		795		850	

Table H.3.1a *Mass and heat balance for the process with heat integration (Con’t)*

STREAM Nr. :		101		102		103		104		105	
Name :		Shell gas product1		Shell gas product2		Shell gas product3		Shell gas product4		Shell gas compressed	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	7.56	0.1717	7.56	0.1717	7.56	0.1717	7.56	0.1717	7.56	0.1717
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	0.003	0.0001	0.003	0.0001	0.003	0.0001	0.003	0.0001	0.003	0.0001
Propylene	42	3.18	0.0756	3.18	0.0756	3.18	0.0756	3.18	0.0756	3.18	0.0756
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Methane	16	0.002	0.0001	0.002	0.0001	0.002	0.0001	0.002	0.0001	0.002	0.0001
Hydrogen	2	0.10	0.0515	0.10	0.0515	0.10	0.0515	0.10	0.0515	0.10	0.0515
Water	18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Nitrogen	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		10.84	0.2990	10.84	0.2990	10.84	0.2990	10.84	0.2990	10.84	0.2990
Enthalpy	kW	-1247		-9325		-16392		-16528		-12344	
Phase		V		V		V		V		V	
Press.	Bara	1		1		1		1		30	
Temp	oC	540		308		32		25		227	

Table H.3.1a *Mass and heat balance for the process with heat integration (Con't)*

STREAM Nr. :		106		107		108		109		201	
Name :		Shell gas compressed2		Shell gas comp.after cooler		Shell gas to sep.Unit		Shell liq to sep. Unit		Tube gas product1	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	7.56	0.1717	7.56	0.1717	5.08	0.1155	2.48	0.0563	1.36	0.0309
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	0.003	0.0001	0.003	0.0001	0.002	0.0001	0.000	0.0000	2.291	0.0818
Propylene	42	3.18	0.0756	3.18	0.0722	2.22	0.0528	0.96	0.0228	2.47	0.0588
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.62	0.0223
Carbondi-oxide	44	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.18	0.0040
Methane	16	0.002	0.0001	0.002	0.0000	0.002	0.0001	0.000	0.0000	0.576	0.0360
Hydrogen	2	0.10	0.0515	0.10	0.0023	0.10	0.0504	0.00	0.0011	0.05	0.0265
Water	18	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	1.47	0.0817
Nitrogen	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.08	0.0029
Total		10.84	0.2990	10.84	0.2464	7.40	0.2189	3.43	0.0801	9.10	0.3449
Enthalpy	kW	-16373		-17475		-11090		-6385		-4243	
Phase		V		L/V		V		L		V	
Press.	Bara	30		30		30		30		1	
Temp	oC	65		55		55		55		850	

***Table H.3.1a** Mass and heat balance for the process with heat integration (Con't)*

STREAM Nr. :		202		203		204		205		206	
Name :		Tube gas product2		Tube gas product3		Tube gas product4		Tube gas product5		Tube gas compressed	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	1.36	0.0309	1.36	0.0309	1.36	0.0309	1.36	0.0309	1.36	0.0309
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	2.291	0.0818	2.291	0.0818	0.082	0.0029	2.291	0.0818	2.291	0.0521
Propylene	42	2.47	0.0588	2.47	0.0588	2.29	0.0545	2.47	0.0588	2.47	0.0561
Carbonmon-oxide	28	0.62	0.0223	0.62	0.0223	2.47	0.0882	0.62	0.0223	0.62	0.0142
Carbondi-oxide	44	0.18	0.0040	0.18	0.0040	0.58	0.0131	0.18	0.0040	0.18	0.0040
Methane	16	0.576	0.0360	0.576	0.0360	0.053	0.0033	0.576	0.0360	0.576	0.0131
Hydrogen	2	0.05	0.0265	0.05	0.0265	0.12	0.0576	0.05	0.0265	0.05	0.0012
Water	18	1.02	0.0566	0.57	0.0315	0.62	0.0346	0.12	0.0064	0.12	0.0026
Nitrogen	28	0.08	0.0029	0.08	0.0029	0.18	0.0063	0.08	0.0029	0.08	0.0019
Total		8.65	0.3198	8.19	0.2946	7.74	0.2915	7.74	0.2695	7.74	0.1760
Enthalpy	kW	-12982		-18961		-3102		-6063		-6063	
Phase		V		V		V		V		V	
Press.	Bara	1		1		1		1		1	
Temp	oC	550		308		219		25		25	

Table H.3.1a *Mass and heat balance for the process with heat integration (Con't)*

STREAM Nr.	:	207		208(208-5)	=208-1 to -4	209		301		302	
Name :		Tube gas compressed2		Water discharged		Tube gas to CO_2 removal		Propane to Tube		Overhead T301	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	1.36	0.0309	0.00	0.0000	1.36	0.0309	6.44	0.1464	0.00	0.0000
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	0.082	0.0029	0.00	0.0000	2.29	0.0818	2.29	0.0819	7.90	0.2821
Propylene	42	2.29	0.0545	0.00	0.0000	2.47	0.0588	4.69	0.1116	0.00	0.0000
Carbonmon-oxide	28	2.47	0.0882	0.00	0.0000	0.62	0.0223	0.62	0.0223	1.19	0.0426
Carbondi-oxide	44	0.58	0.0131	0.00	0.0000	0.18	0.0040	0.01	0.0003	0.01	0.0002
Methane	16	0.053	0.0033	0.00	0.0000	0.58	0.0360	0.58	0.0361	4.79	0.2994
Hydrogen	2	0.12	0.0576	0.00	0.0000	0.05	0.0265	0.15	0.0769	0.16	0.0797
Water	18	0.62	0.0346	1.46	0.0332	0.01	0.0006	0.01	0.0006	0.00	0.0000
Nitrogen	28	0.18	0.0063	0.00	0.0000	0.08	0.0029	0.08	0.0029	0.14	0.0050
Total		7.74	0.2915	1.46	0.0332	7.64	0.2637	14.88	0.4790	14.19	0.7089
Enthalpy	kW	-1921		-20359		-4946		-14585		-5787	
Phase		V		L		V		V		V	
Press.	Bara	30		3		30		30		15	
Temp	oC	289		25		30		42		-131	

Table H.3.1a *Mass and heat balance for the process with heat integration (Con't)*

STREAM Nr. :		303		304		305		306		307	
Name :		Light gas Product		Reflux T301		Feed to T302		Overhead T302		Reflux T302	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	0.00	0.0000	0.00	0.0000	6.44	0.1464	0.01	0.0002	0.01	0.0002
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	0.03	0.0010	7.87	0.2811	2.26	0.0809	2.26	0.0808	0.00	0.0000
Propylene	42	0.00	0.0000	0.00	0.0000	4.69	0.1116	8.53	0.2032	8.52	0.2029
Carbonmon-oxide	28	0.62	0.0223	0.57	0.0203	0.00	0.0000	0.05	0.0018	0.05	0.0018
Carbondi-oxide	44	0.00	0.0000	0.01	0.0002	0.01	0.0003	0.01	0.0003	0.00	0.0000
Methane	16	0.58	0.0361	4.21	0.2633	0.00	0.0000	0.00	0.0000	0.00	0.0000
Hydrogen	2	0.15	0.0769	0.01	0.0027	0.00	0.0000	0.00	0.0000	0.00	0.0000
Water	18	0.00	0.0000	0.00	0.0000	0.01	0.0006	0.00	0.0000	0.00	0.0000
Nitrogen	28	0.08	0.0029	0.06	0.0021	0.00	0.0000	0.05	0.0019	0.05	0.0019
Total		1.47	0.1392	12.72	0.5697	13.42	0.3398	10.93	0.2883	8.63	0.2068
Enthalpy	kW	-5786		-5787		-14573		-14264		11270	
Phase		V		V		L		V		L	
Press.	Bara	15		15		15		15		15	
Temp	oC	-131		-131		10		-39		-39	

Table H.3.1a *Mass and heat balance for the process with heat integration (Con't)*

STREAM Nr. :		308		309		310		311		312	
Name :		Ethylene Prduct		Bottom product T303		Feed to T303		Overhead T303		Reflux T302	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	0.00	0.0000	6.44	0.1463	8.91	0.2026	0.05	0.0011	0.05	0.0010927
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0
Ethylene	28	2.26	0.0808	0.00	0.0001	0.00	0.0001	0.04	0.0015	0.04	0.0014474
Propylene	42	0.01	0.0003	4.67	0.1113	5.63	0.1340	105.15	2.5036	100.44	2.391449
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0
Carbondi-oxide	44	0.01	0.0003	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	7.027E-06
Methane	16	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0002	0.00	0.000154
Hydrogen	2	0.00	0.0000	0.00	0.0000	0.00	0.0011	0.05	0.0242	0.05	0.0230689
Water	18	0.00	0.0000	0.01	0.0006	0.01	0.0006	0.00	0.0000	0.00	6.849E-39
Nitrogen	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0
Total		2.29	0.0815	11.12	0.2582	14.56	0.3384	105.29	2.5306	100.58	2.4172
Enthalpy	kW	2994		-16920		-23305		0		0	
Phase		L		L		L		V		L	
Press.	Bara	15		15		15		15		15	
Temp	oC	-39		40		38		35		35	

Table H.3.1a *Mass and heat balance for the process with heat integration (Con't)*

STREAM Nr.	:	313		314		314-W		315		316	
Name :		Propylene Prduct		Propane recycle1		Moisture removed		Propane recycle2		Propane recycle3	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	0.00	0.0001	8.91	0.2025	0.00	0.0000	8.91	0.2025	8.91	0.2025
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	0.00	0.0001	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Propylene	42	4.71	0.1121	0.92	0.0219	0.00	0.0000	0.92	0.0219	0.92	0.0219
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Methane	16	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Hydrogen	2	0.00	0.0011	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Water	18	0.00	0.0000	0.00	0.0000	0.01	0.0006	0.00	0.0000	0.00	0.0000
Nitrogen	28	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		4.72	0.1133	9.83	0.2244	0.01	0.0006	9.83	0.2244	9.83	0.2244
Enthalpy	kW	2093		-24058		-168		-17002		-14849	
Phase		V		L		L		V		V	
Press.	Bara	15		15		15		15		15	
Temp	oC	35		43		43		217		298	

***Table H.3.1a** Mass and heat balance for the process with heat integration (Con’t)*

STREAM Nr.	:	317		318		401		402	
Name :		Propane recycle4		Propane recycle5		CO2 less gas product		CO2 By product	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	8.91	0.2025	8.91	0.2025	1.36	0.0309	0.00	0.0000
Oxygen	32	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Ethylene	28	0.00	0.0000	0.00	0.0000	2.29	0.0818	0.00	0.0000
Propylene	42	0.92	0.0219	0.92	0.0219	2.47	0.0588	0.00	0.0000
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000	0.62	0.0223	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.00	0.0000	0.01	0.0003	0.16	0.0037
Methane	16	0.00	0.0000	0.00	0.0000	0.58	0.0360	0.00	0.0000
Hydrogen	2	0.00	0.0000	0.00	0.0000	0.05	0.0265	0.00	0.0000
Water	18	0.00	0.0000	0.00	0.0000	0.01	0.0006	0.00	0.0000
Nitrogen	28	0.00	0.0000	0.00	0.0000	0.08	0.0029	0.00	0.0000
Total		9.83	0.2244	9.83	0.2244	7.48	0.2601	0.16	0.0037
Enthalpy	kW	-9407		-7193		-3494		-1451	
Phase		V		V		V		V	
Press.	Bara	15		1		30		1	
Temp	oC	476		540		30		101	

Table H.3.1a *Mass and heat balance for the process with heat integration (Con’t)*

STREAM Nr. :		403		404	
Name :		MDEA sol feed		Spent MDEA sol	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s
Propane	44	0.00	0.0000	0.00	0.0000
Oxygen	32	0.00	0.0000	0.00	0.0000
Ethylene	28	0.00	0.0000	0.00	0.0000
Propylene	42	0.00	0.0000	0.00	0.0000
Carbonmon-oxide	28	0.00	0.0000	0.00	0.0000
Carbondi-oxide	44	0.00	0.0000	0.16	0.0037
Methane	16	0.00	0.0000	0.00	0.0000
Hydrogen	2	0.00	0.0000	0.00	0.0000
Water	18	15.96	0.8864	15.96	0.8864
Nitrogen	28	0.00	0.0000	0.00	0.0000
MDEA	119.2	15.96	0.1339	15.96	0.1339
Total		31.91	0.8864	32.07	0.8901
Press.	Bara	30		30	
Temp	oC	30		30	

Table H.3.1bOverall *mass and heat balance for the process with heat integration (Con’t)*

Overall Component Mass Balance & Stream Heat balance							
STREAM Nr. :		001+008		303+308+313+402+208+314W			OUT-IN
Name :		Total Plant	IN	Total Plant	OUT	Total Plant	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Propane	44	7.45	0.1692	0.00	0.0001	-7.44	-0.1691
Oxygen	32	1.79	0.0559	0.00	0.0000	-1.79	-0.0559
Ethylene	28	0.00	0.0000	2.29	0.0819	2.29	0.0819
Propylene	42	0.79	0.0188	4.72	0.1125	3.93	0.0937
Carbonmon-oxide	28	0.00	0.0000	0.62	0.0223	0.62	0.0223
Carbondi-oxide	44	0.00	0.0000	0.18	0.0040	0.18	0.0040
Methane	16	0.00	0.0000	0.58	0.0361	0.58	0.0361
Hydrogen	2	0.00	0.0000	0.16	0.0780	0.16	0.0780
Water	18	0.00	0.0000	1.47	0.0817	1.47	0.0817
Nitrogen	28	0.08	0.0029	0.08	0.0029	0.00	0.0000
Total		10.11	0.2439	10.11	0.4165	0.00	

Appendix H.4 Economy

Profitability analysis is required to evaluate the economic aspect of a project design. Some crucial measures such as Purchased Equipment (PCE), Total Investment cost, production cost, profit margin, return on investment (ROI), payback period (PBP). The cash flow, net present value (NPV) and the investor's rate of return (IRR) (also know as the discounted cash-flow rate of return (DCFRR) to count the time value of money.

Margin is defined as
follows:

$$\text{Margin}=\text{Total Value (Products, Wastes } OUT\text{) - Total Value (Feedstock's, Process Chemicals, } IN\text{)}$$

In order to prevent confusion, in this conceptual stage, profitability analysis would not include some local regulation such as, local taxes, depreciation, subsidy, grant etc. Thereby, they will be taken on board in further stage.

In this chapter, it presents the economic indexes such as investment cost, margin, and economic criteria in order to view the feasibility of proposed design.

Investment

Lang method is used in order to estimate the investment of the designed process. Investment is considered in term of direct capital cost, indirect capital cost, fixed capital cost, license cost, and working capital cost.
Some commodity chemicals can be obtained from the Chemical Marketing Reporter and Chemical week magazine.
Direct Capital Cost: The summation of the material costs required to build the complete Process that are incurred in the construction of a plan, in addition to the cost of equipment are

1. Equipment erection, including foundations and minor structural work.
2. Piping, including insulation and painting
3. Electrical, power and lighting.
4. Instruments, local and control room
5. Process building and structures
6. Ancillary buildings, offices, laboratory buildings, workshops.
7. Storages, raw materials and finished products.
8. Utilities, provision of plant for steam, water, air, firefighting services
9. Site and site preparation

Indirect Capital Cost: The costs that are first, for Design and engineering, which cover the cost of design and the cost of engineering, purchasing, procurement and construction supervision. Secondly, for Contractor's fees and for Contingency allowance.

Fixed Capital Cost is the summation of Direction Capital Cost and Indirection Capital Cost. Purchased Equipment Costs (PSE) with Lang method and capital cost can be found in the Table H.4.1. The summary of the capital investment cost presents in Table H.4.2.

Purchased Equipment Costs (PCE) with Lang method, can be found in the Table H.4.1.

Table H.4.1 *Purchased Equipment Costs*

Purchased Equipment Costs (PCE) in December 2003			
Type	Name	Equipment	Cost (US$)
Reactor	RX001	Shell & Tube	113,643
	RX002	Shell & Tube	113,643
Total			227,286
Drum	D101	Vessel	12,354
	D102	Vessel	23,488
	D201	Vessel	26,382
	D301	Vessel	24,217
	D302	Vessel	6,298
	D401	Vessel	5,212
Total			97,951
Column	T301	Packed Column	115,198
	T302	Packed Column	128,018
	T303	Packed Column	3,669,718
	T401	Packed Column	38,049
	T402	Packed Column	38,049
Total			3,989,032
Heat exchanger	E-001	Shell & Tube	3,868,958
	E-002	Shell & Tube	2,446,242
	E-004	Shell & Tube	1,173,004
	E-101	Shell & Tube	35,126
	E-102	Shell & Tube	643,563
	E-103	Shell & Tube	30,994
	E-201	Shell & Tube	636,338
	E-202	Shell & Tube	1,961,731
	E-203	Shell & Tube	65,577
	E-204	Shell & Tube	61,987
	E-301A	Shell & Tube	24,795
	E-301B	Shell & Tube	164,194
	E-302A	Shell & Tube	113,555
	E-302B	Shell & Tube	82,650
	E-303A	Shell & Tube	166,315
	E-303B	Shell & Tube	77,971
	E-303C	Shell & Tube	22,456
	E-402	Shell & Tube	61,987
	E-401	Shell & Tube	148,149
Total			11,785,592
Furnace	E-003	Process, Cylindrical	34,504
	E-005	Process, Cylindrical	111,404
	E-006	Process, Cylindrical	117,354
Total			263,262
Compressor	C101	Reciprocating	935,706
	C201	Reciprocating	998,811
	C303	Reciprocating	209,010
Total			2,143,526

From the table 6.1 (Reference: Coulson& Richardson, Volumn6), we can calculate Direct Capital Cost and Fixed Capital Cost in the Table H.4.2.

Table H.4.2 *Capital costs estimation*

Item	Process type Fluids
1. Major equipment as total purchased cost	
f1 : Equipment erection	0.40
f2 : Piping	0.70
f3 : Instrumentation	0.20
f4 : Electrical	0.10
f5 : Buildings,process	0.15
f6 : Utilities	0.20
f7 : Storages	0.15
f8 : Site development	0.05
f9 : Ancillary buildings	0.15
∑ (f1+f2+f3+…+f9)	2.10
2. Total physical plant cost (PPC)	
PPC = PCE(1+ f1 + f2 + …+ f9)	
PPC = Direct Cost	57,370,614
3. Indirect cost	
f10 : Design and engineering	0.20
f11 : Contractor's fee	0.05
f12 : Contingency	0.05
∑ (f10+f11+f12)	0.30
Indirect Capital Cost	17,211,184
Fixed Capital = PPC (1 + f10 + f11 +f12)	
Fixed Capital = Direct + Indirect cost	74,581,798

Costs	2003
	US$
1. Direct Capital Cost	57,370,614
2. Indirect Capital Cost	17,211,184
3. Fixed Capital Cost	74,581,798

To estimate the total investment cost of this process, the total investment is fixed capital cost and working capital. Therefore the total capital cost is summarized in Table H.4.3.

Table H.4.3 *Total Capital Cost*

Total Investment Costs	Year' 2003 US$ million
1. Fixed Capital Cost	74.58
2.Working Capital Cost	5.97
Total Investment Costs	80.55

Note: Working Capital Cost means the additional investments for start up until income starts such as initial catalyst charge, raw material & intermediates, finished product inventories.

The raw material and utilities are presented in Table H.4.4. Meanwhile, Production cost is shown in Table H.4.5.

Table H.4.4 *The raw material and utilities cost*

Stream hrs/annum = 8040

Raw Materials	Str.No.	kg/s	m3/a	ton/hrs	ton/a	t/t Alkenes	Price US$/unit	Unit	Cost US$/a @2003	Cost Million US$/a @2003
Propane	1	8.23	-	29.64	238,341	1.1748	160	Ton	38,194,083	38.194
Oxygen	8	1.87	-	6.74	54,194	0.2671	143	Ton	7,771,432	7.771
Catalyst1 (V2O5/CeO2/SA5205					6.8	0.00003	106,549	Ton	724,535	0.725
Catalyst2 (Pt on MFI zeolite)					3.3	0.00002	181,939	Ton	600,399	0.6
Total catalyst cost								Ton	1,324,934	1.325
Total Raw Material cost (IN)									47,290,450	47.29

Product	Str.No.	kg/s	m3/a	ton/hrs	ton/a	t/t Alkenes	Price US$/ton	Unit	Income US$/a @2003	Income Million US$/a @2003
Ethylene	308	2.29	-	8.26	66,376	0.3272	518	ton	34,382,700	34.383
Propylene	313	4.72		16.98	136,502	0.6728	408	ton	55,692,827	55.693
Light gas for syngas plant	303	1.47		5.28	42,414	0.2091	64	ton	2,718,717	2.719
CO2 for EOR	402	0.16		0.58	4,670	0.023	6.5	ton	30,356	0.03
Water for EOR	208-5	1.46		5.26	42,267	0.2083	0.01	ton	423	0
Total Income (OUT)									92,825,022	92.825

Table H.4.4 *The raw material and utilities cost (Con't)*

Utilities	***Load***	***Cost US$/unit***	***Unit***	***Cost US$/a @2003***	***Cost Million US$ @2003***
Water (t/a)	11,281,221	0.010	Ton	112,812	0.113
Steam (t/a)		12.000	Ton		0.000
Electric (kWh/a)	106,097,231	0.04	kWh	4,481,084	4.481
Fuel (t/a)	0				
Total Utilities cost				4,593,896	**4.594**

Table H.4.5 *Production cost*

Production Cost	***Cost US$/a @2003***	***Cost Million US$/a @2003***	***% of Total Production Cost***
Variable cost			
1. Raw Materials	***47,290,450***	47.290	61.37
2. Miscellaneous materials	372,909	0.373	0.48
3. Utilities	***4,593,896***	4.594	5.96
4. Shipping and Packaging		0.000	0.00
Sub Variable cost A	*52,257,255*	52.257	67.82
Fixed cost			
5. Maintenance	3,729,090	3.729	4.84
6. Operating Labor	1,500,000	1.500	1.95
7. Laboratory costs	300,000	0.300	0.39
8. Supervision	300,000	0.300	0.39
9. Plant Overheads	750,000	0.750	0.97
10. Capital charges	7,458,180	7.458	9.68
11 Insurance	745,818	0.746	0.97
12. Local taxes	1,491,636	1.492	1.94
13. Royalties	745,818	0.746	0.97
Sub Fixed cost B	*17,020,542*	17.021	22.09
Direct production costs A+B	69,277,796	69.278	89.91
14. Sales Expense	3,463,890	3.464	4.50
15. General overheads	3,463,890	3.464	4.50
16. Research and Development	851,027	0.851	1.10
Sub-total C	*7,778,807*	7.779	10.09
Annual production cost = A+B+C	***77,056,603***	77.057	100.00
Annual production rate (ton/annum)	202,878		
Production cost (Pound/kg) =	***380***		

Economic criteria

In order to determine the economic, Process Cash Flow, Rate of Return (ROR) and Pay Out (Back) Time (POT) of the investment are summarized in Table E.4.6.

Table E.4.6 Economic Criteria

Economic criteria	
Cash Flow M US$	15.77
Rate of Return (ROR) [%]	11.17
Pay Out Time yrs	10
DCFROR %	15.31

Economic Criteria

c) Net Cash Flow (NCF) from Gross Income and Production Costs.

$$\text{Net Cash Flow}_{\text{annual}} = \sum (\text{Gross Income}_{\text{annual}} - \text{Production Costs}_{\text{annual}})$$

$$\text{Gross Income}_{\text{annual}} = \Sigma \left(\text{Products}_{\text{annual}} \text{ x } \frac{\text{Price}}{\text{Unit}} \right)$$

$$= 92{,}825{,}022 \text{ US\$/a}$$

$$= 93.83 \text{ US\$ million}$$

$$\text{Production Cost }_{\text{annual}} = 77.057 \text{ US\$ million}$$

$$\text{Net Cash Flow}_{\text{annual}} = 93.83 - 77.05 \text{ US\$ million}$$

$$= 15.77 \text{ US\$ million}$$

d) Rate of Return (ROR) and Pay Out (Back) Time (POT or PBP) from the total Investment and NCF along approach.

$$\text{ROR} = \frac{\text{Accu. Cash Flow}}{(\text{Project life*Tot.Investment})}$$

$$\text{Accu.Cash Flow} = (\text{Net Cash Flow*Plant life}) - \text{Total Investment Cost} + $$

$$+ \text{Salvage Value}$$

$$= 15.77 \frac{\text{US\$ million}}{\text{year}} \text{*15 year}) - 80.55 \text{ US\$ million} +$$

$$+ (8\%\text{Fixed Capital Cost) US\$ million}$$

$$= ((15.77*15) - 80.55 + (0.1* 74.58))$$

$$= 161.94 \text{ US\$ million}$$

$$\text{Project life time} = \text{Construction time} + \text{Plant life time} + \text{Salvage}$$
$$= 2 + 15 + 1$$
$$= 18 \text{ years}$$

So,

$$\text{ROR} = \frac{161.94 \text{ US\$ million}}{18 \text{ years} * 80.55 \text{ US\$ million}}$$
$$= 11.17\ \%$$

and,

$$\text{POT} = 10 \text{ years}$$

In calculating cash flow (NCF), The project is usually considered as an isolated system, and taxes on profits and the effect of depreciation of the investment are not considered, since tax rates are not constant and depend on government policy as well as the rates of depreciation. Depreciation rates also depend on the accounting practices of the particular company. Therefore during evaluating projects, the effect of government policy must be taken into account at some stages particularly when considering projects in different country.

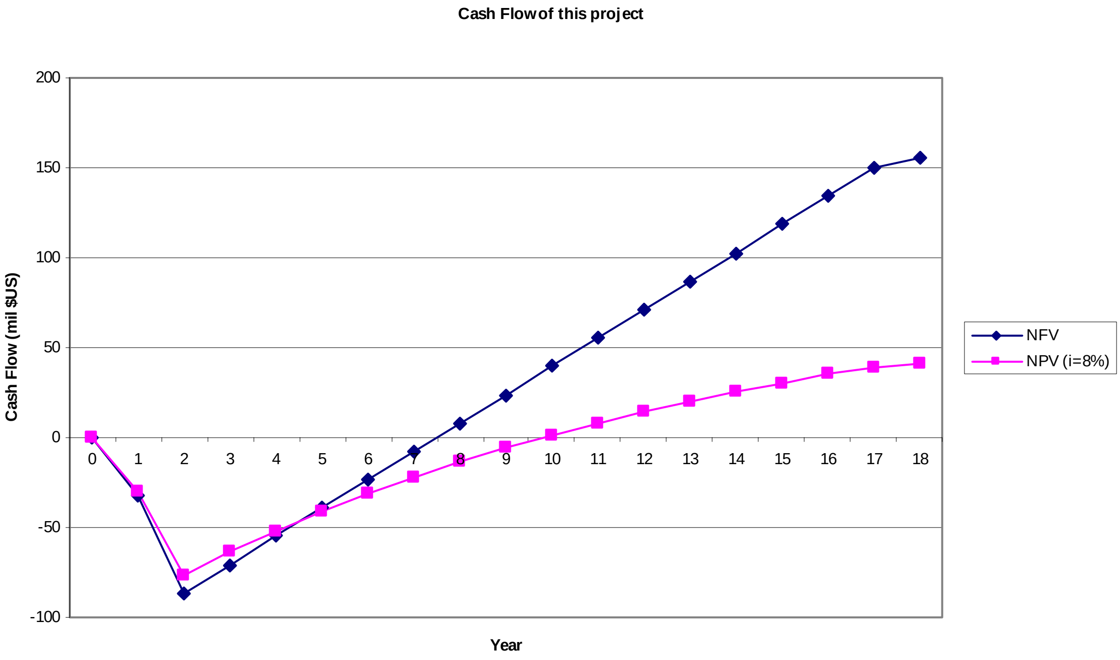


Figure H.4.1. The discounted cash flow (i=0% and i=8%)