Session 3213

Design of a Propylene Storage Facility

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Senior Design students were to design a propylene storage facility. The problem required solution of mass and energy balances and sizing of heat exhangers, separators, and an adiabatic flash valve. The objective was to minimize the cost per kg. of propylene recovered. Modification and improvement of the initial flow diagram was also encouraged. Solution of this problem was facilitated by the use of CACHE software. Equipment costs were estimated using scaling factors, and cost indexes. Utility costs were also provided to the students.

A degrees-of-freedom analysis reveals that there is one degree of freedom, making this an open ended problem. In order to optimize this process, the production cost vs. the temperature of the cold fluid leaving the first heat exchanger was plotted. By selecting the temperature at the minimum production cost, the problem is completely specified.

Students worked in teams of three to design the propylene facility and submit a formal report. They were given 10 days to complete the assignment. Because of the magnitude of the assignment, delegation of tasks, and parallel completion of tasks was necessary. Weekly private meetings with the instructors and teaching assistant were scheduled with each team to answer questions, and simulate industrial progress meetings with supervisors.

The students reported that the assignment was demanding, but worthwhile. They integrated concepts learned throughout the undergraduate curriculum had experience writing a formal report, meeting a deadline, preparing concise questions and progress reports for the weekly meeting, and working in teams. The problem follows:

Problem Statement: Propylene is stored at 700 psia and 60 by dissolving it in liquid octane. The mole fraction of propylene is 0.515. When recovery of the propylene is desired, it passes through a series of heat exchangers and an adiabatic flash valve to lower the pressure and change the temperature. The final product is 99 mole% propylene vapor at 30 psia and 60°F. The n-octane is returned to storage. A heat exchange network for a propylene storage facility is shown in figure 1 (1). The feed stream, which contains propylene dissolved n-octane, is preheated in the first heat exchanger, H-1, to recover heat by partially condensing a recycled stream consisting of propylene and n-octane. This feed mixture is further heated to its bubble point in a second heat exchanger, H-2, using high pressure steam. The mixture then passes through an adiabatic flash valve and goes to a perfect phase separator, S-2. The vapor leaving the phase separator is parially condensed when it is used as the heat transfer fluid in H- 1. It is further condensed by passing through heat exchanger H-3. The stream then enters a phase separator where the product, 1000 lb. moles/hr (454 kg. mole/hr) of 99 mole% propylene at 30 psia is withdrawn. The liquid stream leaving the phase separator S-2 is cooled to 80°F. Assume the pressure drop across each heat exchanger is 5 psi. Neglect line losses and separators. In order to design the system determine:

1. the degrees of freedom for this system.
2. the cost per pound of propylene recovered. In order to do this you should:

a. determine the temperature, pressure, flowrate, and composition in each of the process streams. Use the CACHE program to calculate the temperature and pressure in streams 5, 6, and 7, and to calculate enthalpy changes across heat exchangers for the propylene, octane mixtures.

b. optimize the size of the heat exchangers to minimize the cost per pound of recovering propylene.

c. determine the size of the phase separators and whether they should be horizontal or vertical.

d. determine the temperature, pressure, flowrate, and composition in each of the process streams. Use the CACHE program to calculate the temperature and pressure in streams 5, 6, and 7, and to calculate enthalpy changes across heat exchangers for the propylene, octane mixtures.

e. Select an appropriate heat transfer fluid for H-3.

Estimate equipment costs using cost indexes and scaling factors. The following figures are also provided:

Chilled Water (40°F)	- $0.093/ton
Steam	- $8.07/ton
yearly operation	-8000 hours
project life	-8 years
salvage value (fs)	- 10%
interest rate (i)	- 10 %
maintenance	- 5% X depreciable capital cost
working capital	- 15% X depreciable capital cost
depreciation	- $(1-f_s)$ X capital costs/ depreciation life
land cost	-0.0167 X capital costs
taxes and insurance	-0.025 X capital costs
administrative costs	-0.15 X maintenance costs
financing costs	- i X (capital cost + working capital+ land cost)

Any improvements to the flow diagram presented are encouraged.

Solution -As mentioned above, a degree of freedom analysis revealed one degree of freedom. The temperature and pressure of stream 1 is given in the problem statement. The pressure in stream 3 is estimated as 695 psia since the pressure drop through the heat exchanger H-1 is assumed to be 5 psi. The pressure in stream 4, which is specified as saturated vapor is thus 690 psia. The CACHE software can be used to determine the temperature in stream 4 because it is saturated vapor at its dewpoint with a known composition and pressure. The distillation (DISTIL) module is accessed and a flash distillation is performed. Temperatures are guessed until the first drop of liquid forms. Thus the temperature of stream 4 is known. The pressure in stream 5 is estimated to be 40 psia because the pressure in the product stream, stream 10 is given as 30 psia and again, pressure drops through the heat exchanger are estimated to be 5 psi, and line losses and pressure drops through separators are assumed to be negligible. The temperature in stream 5 can be determined using the CACHE program. As this is an adiabatic flash valve, temperatures are guessed until the enthalpy in stream 5 equals the enthalpy in stream 4. The phase separator S-2 is considered to be a perfect phase separator. Thus the vapor in stream 7, the liquid in stream 6 and the two-phase mixture in stream 5 are in equilibrium and have the same temperature and pressure. Minimizing the surface area of the heat exchangers minimizes the production costs. Since there is one degree of freedom, the temperature of stream 3 can be specified. Determination of production costs as a function of the temperature in stream 3 reveal that the lower the temperature of stream 3, the lower the production cost. However, the lower limit of the temperature of stream 3 is set by the correction factor, F for heat exchanger H- 1 (2). As the temperature of stream 3 decreases the value of F decreases. The optimum temperature was thus selected as the temperature at which F=0.75, which is the minimum acceptable value reported by Goyal (3). The temperature and pressure in streams 9 and 11 are the same as those of the product, stream 10. The overall composition is the same in streams 7, 8, and 9. Mass balances using a basis of 100 lb. moles/hr. product and calculated mole fractions can be used to determine the flowrates of all the streams. Water was selected as the most economical heat transfer fluid for S-3. Table 1 is a summary of the stream parameters. Delivered equipment costs are presented in Table 2. Vertical separators were selected. Depreciable capital costs are presented in Table 3, and production costs are summarized in Table 4.

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References

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2. Kern, P. Q., "Process Heat Transfer," McGraw-Hill, New York, 1959
3. Goyal, O. P., Guidelines on Exchangers, *Hydrocarbon Processing*, 64, *8, 55, 1985.*

Biographical Information

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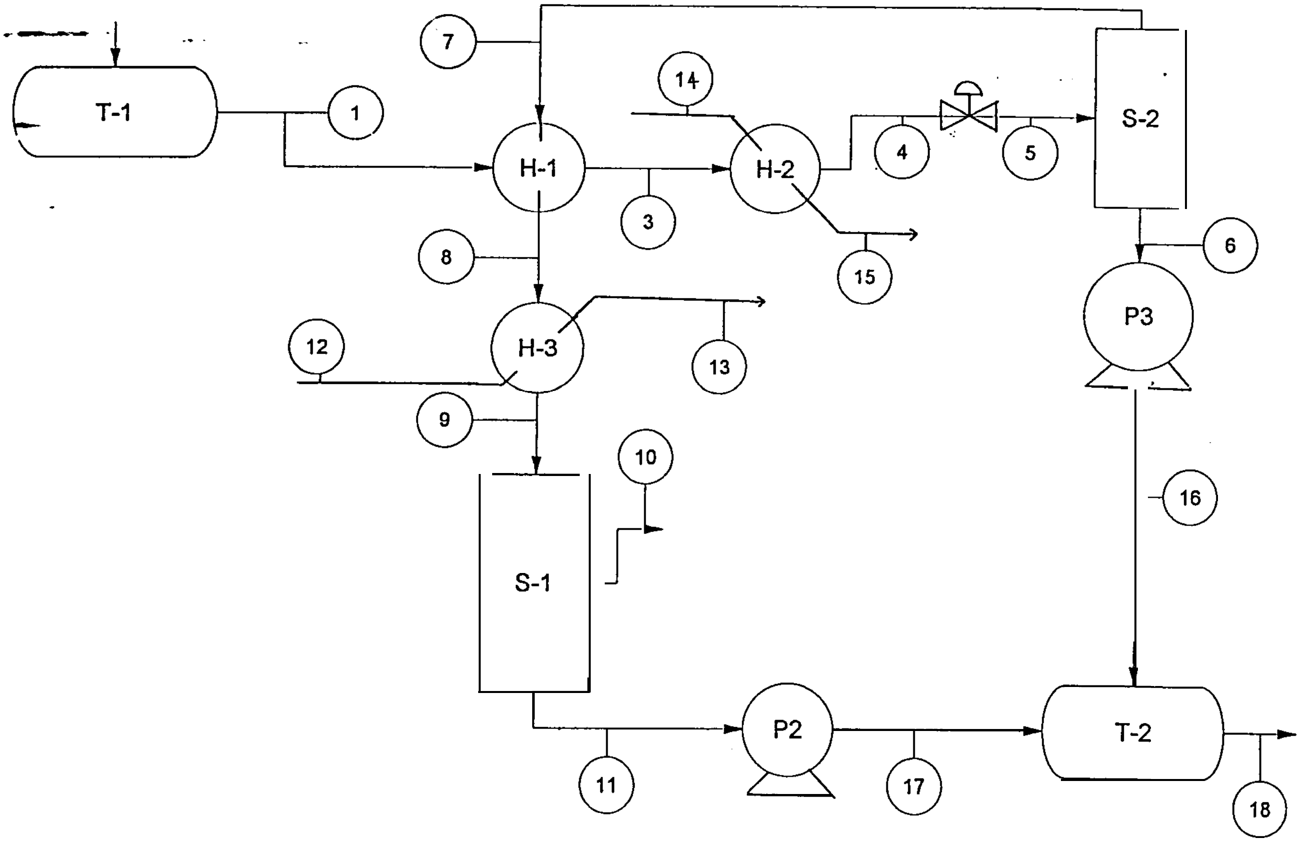

Figure 1. Propylene Storage Facility

Table 1. Summary of Stream Parameters

Stream	Pressure	Temp.	Liquid Comp.		Vapor Comp.		Flow
	(psia)	(°F)	Propylene	Octane	Propylene	Octane	(kgmole/hr)
2	700	60	0.515	0.485	N/A	N/A	1003
3	695	155	0.515	0.485	N/A	N/A	1003
4	690	327	0.515	0.485	N/A	N/A	1003
5	40	235	0.039	0.961	0.7	0.3	1003
6	40	235	0.039	0.961	N/A	N/A	281
7	40	235	N/A	N/A	0.7	0.3	722
8	35	78	0.0132	0.9868	0.8321	0.1679	722
9	30	64	0.21	0.79	0.99	0.01	722
10	30	64	N/A	N/A	0.99	0.01	454
11	30	64	0.21	0.79	N/A	N/A	268
12	--	40	N/A	N/A	N/A	N/A	2677
13	--	69	N/A	N/A	N/A	N/A	2677
14	115	338	N/A	N/A	N/A	N/A	22098
15	5	164	N/A	N/A	N/A	N/A	22098
16	100	235	0.039	0.961	N/A	N/A	281
17	100	64	0.21	0.79	N/A	N/A	268
18	100	152	0.122	0.878	N/A	N/A	549

Table 2. Equipment Cost Summary

Heat Exchangers			
Unit	Size		Cost
	Area	Heat Transferred	
H-1	3682 ft^2	2.34E+07 BTU/h	$58,400
H-2	23999 ft^2	4.77E+07 BTU/h	$290,900
H-3	736 ft^2	3.28E+06 BTU/h	$11,600

Separators			
Unit	Size		Cost
	Diameter	Length	
S-1, Vertical	5 ft	22 ft	$123,000
S-2, Vertical	6.5 ft	19.5 ft	$145,000

Pumps			
Unit	Size		Cost
P-2	Capacity: Power: Efficiency:	192 gpm 90 hp 0.30	$32,700
P-3	Capacity: Power: Efficiency:	210 gpm 80 hp 0.32	$33,900

Total Equipment Cost: $695,500

Table 4. The Production Cost Summary Table

	Cost/yr	Cost/kg
DIRECT COSTS		
Utilities	$0	0
Steam	$12,623,399	0.017249794
Cooling Water	*$964,387*	0.001317829
Electricity	$41,952	5.73271E-05
Operating Labor	$123,077	0.000168184
Operating Supervision	$24,615	3.36367E-05
Quality Control	$24,615	3.36367E-05
Maintenence	$248,000	0.00033889
Land Cost	$82,832	0.000113189
INDIRECT COSTS		
Fixed Costs		
Depreciation	$5,636,250	0.007701899
Taxes & Insurance	$125,250	0.000171153
Plant Overhead Costs		
Fringe Benefits	$32,492	4.44005E-05
Overhead	$73,846	0.00010091
GENERAL COSTS		
Administrative	$37,200	5.08336E-05
Marketing	$3,451,887	0.004716981
Financing	$609,483	0.000832855
Research & Development	$1,450,248	0.001981755
TOTAL PRODUCTION COST	**$25,549,535**	**0.0349**

Table 3. Summary of Depreciable Capital Costs

	Factor	Cost
Direct Costs		
Delivered Equipment	1	$695,500
Installation	3.29	$2,288,195
Instrumentation	0.18	$125,190
Piping	0.66	$459,030
Electrical	0.11	$76,505
Buildings	0.18	$125,190
Auxillaries	0.7	$486,850
Indirect Costs		
Engineering	0.33	$229,515
Construction	0.41	$285,155
Contractor's Fee		$238,557
Contingency		$477,113
Plant Startup		$43,121
TOTAL		$5,529,921