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Dynamic simulation of multi-effect evaporators

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ABSTRACT

Multi-effect evaporators are widely used for concentrating solutions in a variety of important industrial applications ranging from desalinization to biorefining. The use of multi-effects achieves very significant process intensification since energy consumption is reduced in direct ratio to the number of effects. Steady-state design has been discussed in many papers and books. Far fewer studies of their dynamic simulation and control have appeared in the open literature. This paper reviews some of the design issues and demonstrates how a dynamic simulation can be developed using standard Aspen simulation models. The dynamic model must be set up such that the vapor from each evaporator is completely condensed in the downstream lower-temperature evaporator while satisfying heat-transfer constraints.

1. Introduction

Evaporation is used to remove a very volatile component from a mixture of chemicals when the vapor generated in the single-stage flash contains very small amounts of the less volatile components. When several effects are connected in series, the vapor generated in an upstream higher-temperature vessel is used to generate vapor to the next downstream lower-temperature vessel. The multi-effect configuration reduces energy requirements very significantly. One pound of steam can, in an ideal system, vaporize five pounds of water in a five-stage configuration.

Multi-effect evaporation is a major work horse in a number of very important bulk chemical industries. Like its more studied cousin, distillation, it is a major energy consumer in paper, food and sugar manufacturing, in desalinization plants and in biorefineries for concentrating fermentation broth.

Superficially, the design of multi-effect evaporators appears deceptively straightforward. Consider the most common situation in which water is the very volatile component and the other component is a solid (or has a very high boiling point). The vapor generated in each stage is almost pure water. Starting with a given heat source (steam pressure given) and a given heat sink (cooling water), the design problem is to find the most economical number of effects that removes as much water as is economically attractive. The pressure in the last lowest-pressure stage is typically set by the temperature in the water-cooled heat exchanger that condenses the water vapor from the last stage. With typical cooling water temperatures, the last lowest-pressure stage can run at 60 °C so that the condenser can operated at 50 °C. Heat-transfer heuristics suggest that the temperature in the first high-pressure stage must be 10 to 20 °C lower than the saturation temperature of the available steam so as to have reasonable capital investment in heat-exchanger area.

If the heavy component in the water mixture is a solid, the vapor generated in all stages is essentially pure water. The concentration of solids increases as more water is removed from stage to stage, so the physical properties (boiling point temperature and viscosity) can change from stage to stage. The effect of viscosity on heat-transfer coefficients and the phenomenon of "boiling point rise" must be considered in the design.

If the heavy component is not completely non-volatile, a small amount of this component will appear in the vapor leaving each stage. In the early stages where the concentration of this component is low, negligible amounts of product will be lost in the vapor. However, as its concentration in the liquid builds up as the liquid flows through the stages, more and more product is lost in the vapor. The liquid concentration of the product in the last stage can be limited by this loss of valuable product, so an economic analysis to establish this limitation is required. Additional concentration of the mixture can be achieved in a distillation column in which the fractionation permits removing water with small loss of product. Of course distillation requires more energy per pound of water removed than does multi-effect evaporation.

The design optimization variables are the number stages and the temperature in each stage. The economic objective function can be the total annual cost, which takes into account both capital investment (vessel size and heat-transfer area) and operating cost (primarily the cost of steam). Increasing the number of stages reduces steam consumption but increases capital investment.

However, the typical design constraint of most system is not

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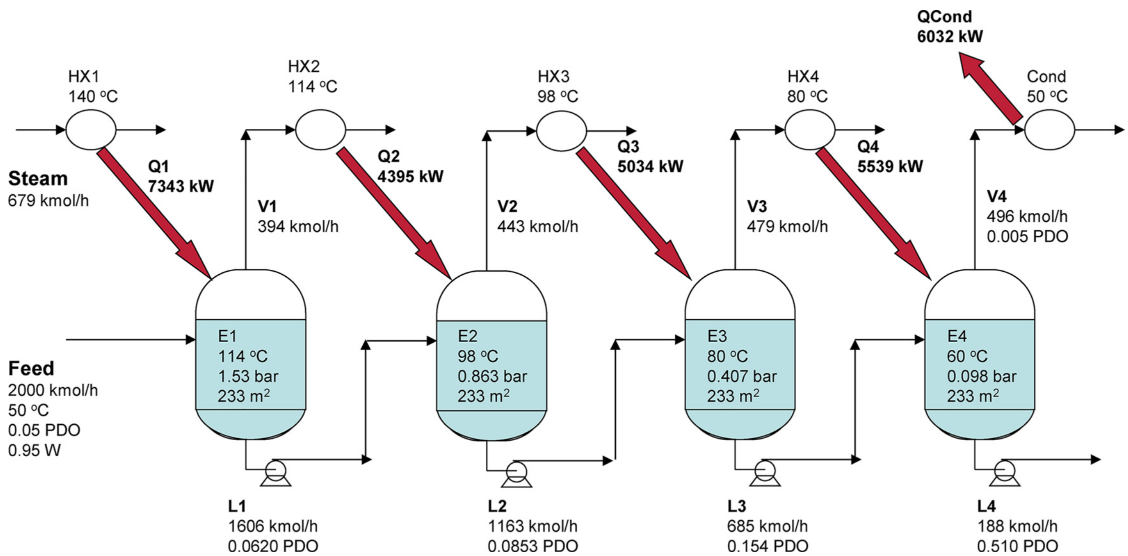

Fig. 1. Four-stage evaporator.

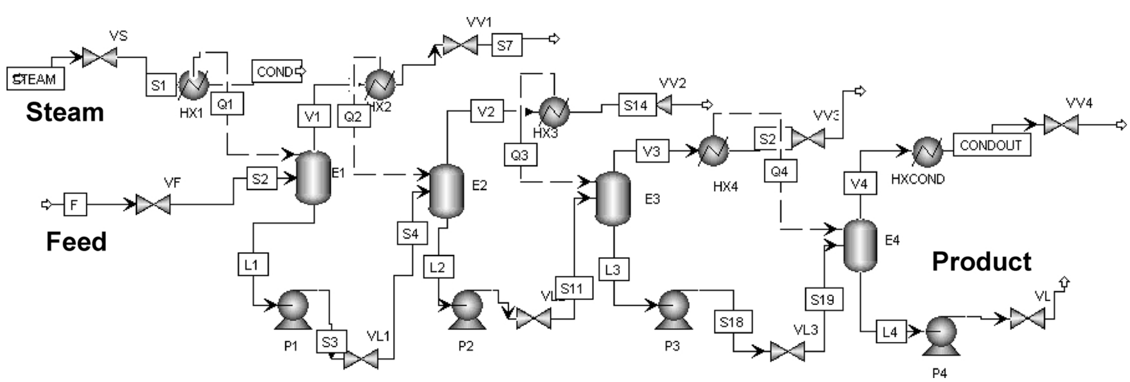

Fig. 2. Aspen Plus process flow diagram.

Table 1
Sizing and Economic Relationships.

Heat Exchangers:
Installed Cost (\$) = 7296(*area* m^2)$^{0.65}$
Vessels:
Diameter calculated to give a maximum vapor velocity in the last lowest-pressure stage (0.098 bar) of 2 m/s.
Aspect ratio L/D = 2 (L and D in meters)
Installed Cost (\$) = 17, 640$(D)^{1.066}(L)^{0.802}$
Steam Cost: \$7.78 per GJ

Table 2
Operating and Economic Results.

		Two	Three	Four	Five
Vessel Diameter	(m)	6.8	5.7	5.0	4.6
HX Area per Stage	(m^2)	233	233	233	233
Condenser Area	(m^2)	1138	796	609	510
Steam	(kmol/h)	1134	818	663	593
	(MW)	12.15	8.923	7.343	6.414
Tsteam	(°C)	140	140	140	140
T1	(°C)	96	108	114	118
T2	(°C)	60	85	98	106
T3	(°C)	–	60	80	92
T4	(°C)	–	–	60	78
T5	(°C)	–	–	–	60
Q1	(MW)	12.15	8.923	7.343	6.414
Q2	(MW)	10.07	6.270	4.395	3.253
Q3	(MW)	–	7.082	5.034	3.759
Q4	(MW)	–	–	5.539	4.257
Q5	(MW)	–	–	–	4.651
VN	(kmol/h)	931	640	496	418
LN	(kmol/h)	188	188	188	188
xN	(mf DPO)	0.51	0.51	0.51	0.51
Capital HXs	(10^6 \$)	0.5045	0.7578	1.009	1.261
Condenser	(10^6 \$)	0.7073	0.5574	0.4721	0.4198
Vessels	(10^6 \$)	1.485	1.646	1.753	1.899
Total	(10^6 \$)	2.697	2.960	3.232	3.580
Steam Cost	(10^6 \$ /y)	2.988	2.195	1.806	1.578
TAC	(10^6 \$ /y)	3.887	3.182	2.884	2.771

minimum TAC. In a practical system is makes sense to have all of the vessels the same size. This means that the heat-transfer area is the same in all vessels. If the overall heat-transfer coefficient U is the same, the ratio of the heat-transfer rate (Q) to the differential temperature driving force (ΔT) must be the same in all effects. The temperature difference in the heat exchanger in Stage n is $\Delta T_n = T_{n-1} - T_n$ where T_n is the temperature in Stage n and T_{n-1} is the stage temperature in the upstream higher-temperature stage. The stage temperatures must be selected such that $Q_n/\Delta T_n$ is the same in all stages. Of course this ratio is the product of the heat-transfer area in each stage A_n and the overall heat-transfer coefficient U. For example in the flowsheet shown in Fig. 1, the area is 233 m^2 in each stage and U is 1.2 kW m^{-2} K^{-1}.

It is important to emphasize that the heat-transfer rates Q_n are **not** the same in all stages. The stage temperatures must be found such that the $Q_n/\Delta T_n$ ratio is the same in all stages. For example, the flowsheet shown in Fig. 1 gives a heat transfer rate of 4395 kW in Stage 2 with a