Short Communication

Distillation column pressure selection

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ABSTRACT

Selection of the operating pressure of a distillation column is one of the most important design decisions. Pressure has a major impact on phase equilibrium in terms of relative volatilities, column temperatures and the existence of azeotropes, which affect energy requirements, utility costs and process configurations. Since many separations are favored by low temperatures, pressures in most columns are established by the desire to use inexpensive cooling water as the heat sink in the condenser. For components with low vapor pressures, this criterion results in pressures below atmospheric. However, there appears to be some hesitancy to use vacuum distillation columns, particularly in azeotropic separations. Numerous papers simply fix the pressure arbitrarily at 1 atm despite the fact that lower pressure could be achieved and still use cooling water.

This paper illustrates the significant economic advantages, both in capital investment and utility costs, of using vacuum distillation in some systems. The numerical example is taken from a recent paper in which an extractive distillation system to separate n-heptane from isobutanol is designed with column pressures fixed at 1 atm. Our results show that vacuum operation can reduce total annual cost by 27%, despite an 8% increase in capital investment, because of a 37% decrease in reboiler steam costs.

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1. Introduction

One of the first and probably the most important decision that must be made in designing a distillation column is what pressure should be specified. This issue is discussed in all distillation and process design text books [1–3]. There are many factors to be considered. If relative volatilities increase with decreasing temperature, which is often the case, pressure should be as low as practical. This normally translates into selecting a reflux-drum temperature of 120 °F (50 °C, 323 K) so that the heat sink used in the condenser can be inexpensive cooling water.

However, other limitations dictate other selections in some systems. High-temperature limitations due to thermal instability may dictate the maximum temperature in the column base, which together with the bottoms composition establishes the base pressure. This in turn, given the tray pressure drops, dictates the condenser pressure. Then, given the distillate composition, the condenser temperature is fixed, which may be lower than that attainable using cooling water, so expensive refrigeration is required.

Another limitation occurs in the separation of very light components. The pressure required to achieve a 323 K reflux-drum temperature is very high and can be near the critical pressure of the key components. Hydraulic problems prevent operation close to the critical pressure because the density difference between the liquid and vapor phases becomes too small. A common heuristic is to set the pressure at half the critical pressure. The resulting condenser pressure for these systems requires refrigeration or vapor recompression.

There are some chemical systems in which relative volatility *increases* as temperature increases. The water/acetic acid separation is an important example. In this situation, pressure should be increased and the 323 K reflux-drum criterion no longer gives the optimum design. In the water/acetic acid system, pressures are limited by a trade-off between equipment capital investment (more expensive materials of construction must be used as temperatures increase due to corrosion) and energy consumption.

There appears to be some hesitancy about using vacuum columns, particularly in studies of azeotropic separations. A recent paper by Wang et al. [4] is an example of this situation. A conventional two-column extractive distillation process is studied for separating n-heptane from isobutanol. The authors arbitrarily set the pressure in each of the columns at 1 atm, despite the fact that the reflux-drum temperatures are significantly greater than the 323 K minimum needed to use cooling water in the condensers. No exploration of the effect of pressure is provided in the paper.

The purpose of this paper is to use this published paper to illustrate that finding the optimum pressure can significantly impact

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the economics of the process. Our results show that energy costs can be reduced by 37% and total annual cost can be reduced by 27% when the pressure in the extractive column is set at 0.2 atm and the pressure in the solvent recovery column is set at 0.1 atm. At these pressures, cooling water can be used in the two condensers.

The numerical example used in this paper (n-heptane/isobutanol) has an azeotropic composition that changes significantly with pressure and is therefore a prime candidate for the use of pressure-swing distillation. This alternative to extractive distillation will be explored in a future paper.

2. Process studied at 1 atm pressure

The normal boiling point of n-heptane is 371.6 K and that of isobutanol is 380.8 K. This binary system at 1 atm forms a minimum-boiling homogeneous azeotrope at 364.1 K with a composition of 66.9 mol% n-heptane (see Fig. 1). Wang et al. [4] recommend the use of N-methyl-2-pyrrolidone (NMP) as the solvent, which has a normal boiling point of 477.1 K. UNIQUAC physical properties are used in the Aspen simulation.

Fig. 2 gives the flowsheet for the process with the number of trays, feed and solvent tray locations and solvent flowrate (100 kmol/h) recommended in the Wang et al. paper. The pressure in the condensers of both columns is set at 1 atm. The product specifications are 99.9 mol% n-heptane in the distillate from the extractive column (C1) and 99.97 mol% isobutanol leaving in the distillate from the solvent recovery column (C2). The variables and condition shown in Fig. 2 are results from our Aspen simulation and are essentially identical to those given in the Wang et al. paper.

Values for tray pressure drop used in their simulations are not given. This parameter becomes important at low pressures. Low pressure-drop trays or packing (0.002 atm per stage) are assumed in our simulations.

Note that the reflux-drum temperatures are 371 and 381 K at the 1 atm pressure assumed in both columns. Temperatures down to 323 K could be attained and still use cooling water.

The reboiler temperature (412 K) in the extractive column at 1 atm permits the use of low-pressure steam (433 K with a cost of $7.78 per GJ). The reboiler duty is 1.852 MW. The reboiler temperature (479 K) in the solvent recovery column at 1 atm requires the use of high-pressure steam (527 K with a cost of $9.88 per GJ). The reboiler duty is 1.38 MW. Note that these reboiler duties are slightly lower than those reported by Wang et al. (1.926 and 1.393 MW), which is probably due to the use of larger tray pressures used in the published paper. As Fig. 1 shows, the n-heptane/isobutanol separation is favored by lower pressures. Thus vacuum columns would be expected to be more efficient, which is demonstrated in the next section.

3. Process studied at lower pressures

The design pressures can be economically reduced as long as cooling water can be used. We now explore the effect of vacuum operation. The same numbers of trays and feed locations are used as in the base case. Only the condenser pressure is changed. Product purities are held at their specifications. Fig. 3 shows the flowsheet when the pressure in the extractive column is set at 0.2 atm and the pressure in the solvent recovery column is set at 0.1 atm.
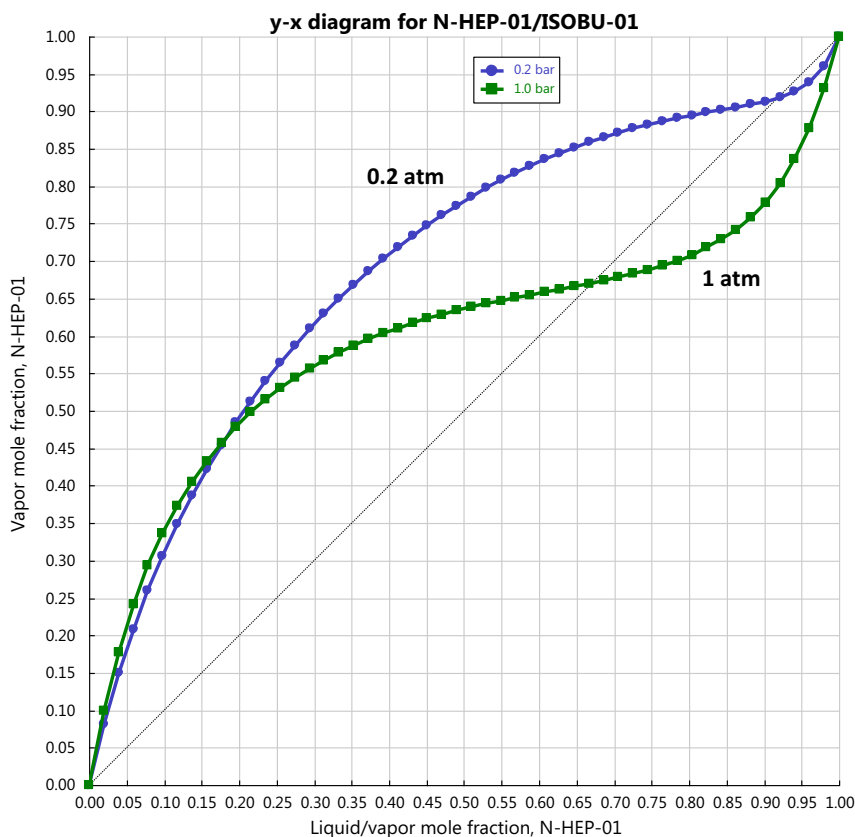

Fig. 1. n-heptane/isobutanol VLE.