# Energy-Efficient Ethanol-Water Fractionation by a Carrier-Gas Method

**James R. Beckman**
Dept. of Chemical, Bio, Materials Engineering, Arizona State University, Tempe, AZ 85287
**Walter F. Albers**
Albers Technologies Corp., Tempe, AZ 85287

Standard distillation practices for the past century have consisted mainly of pressurized towers utilizing hot reboilers for energy input and cooler condensers for energy withdrawal. Analysis of this type of operation was modeled successfully by Ponchon and Savarit (see Hines and Maddox, 1985; McCabe and Thiele, 1925). These works are presented in detail in standard chemical engineering unit operations textbooks (McCabe et al., 1985; Foust et al., 1980).

One of the disadvantages of standard fractionators is the inability to reuse the cooler energy from the condenser in the reboiler. One method used to overcome this disadvantage is vapor recompression. This technique is used to increase the pressure and therefore the temperature of the condenser vapor to temperatures above that of the reboiler. In this manner, condenser energy can be reused by the reboiler. Investigators such as Becker and Zakak (1985), Meili and Striecheli (1987), Ferre et al. (1985), and O'Neill et al. (1985) have explored the vapor recompression technique. Unfortunately, vapor recompression systems tend to be capital-intensive and power-consumptive.

Another method that has been investigated recently by Beckman and Albers to effectively reuse heat energy in fractionation utilizes an inert carrier-gas to evaporate (but not boil) water and ethanol species followed by condensation of those species. This concept, which follows from the desalination work of Albers and Beckman (1988, 1989), was demonstrated at atmospheric pressure in the distillation of an ethanol/water binary mixture.

The device used nitrogen as a carrier-gas at atmospheric pressure and subboiling temperatures to cause evaporation in a cool stripper and condensation in a warm rectifier such that energy supplied at the stripper/rectifier interface was reused through a heat transfer wall.

## Carrier Gas Theory

This heat and mass transfer approach of Figure 1 employs a carrier-gas, in this case nitrogen, as a heat and mass transfer media affecting evaporation (stripping section) and subsequent condensation (rectifying section) to occur on opposite sides of a heat transferring partition thereby reusing the energy of condensation to achieve most of the evaporation. The device operates as follows: nitrogen brought into the *N*th-stage evaporation chamber is slightly cooler than the adjacent condensation chamber. The gas increases in temperature throughout the chamber length and evaporates liquid from the wetted partition thereby affecting a gas-saturated state. Energy needed for the gas heating and liquid evaporation is furnished by the condensation chamber through the wall separating the chambers. When hot, saturated gas leaves the evaporation chamber from stage 1, which is heated slightly by an auxiliary heater and then returned to the condensation chamber at stage 1. Since the vapor-laden gas in the condensation chamber is slightly hotter than the evaporation chamber directly opposite, the hot gas in the condensation chamber cools throughout the chamber length by heat transfer to the cold evaporation chamber. As the gas cools, vapors condense on the partition giving up their heat of condensation through the partition to the evaporation chamber causing further evaporation in that chamber. The chambers are subdivided into stages, and each stage has individual pumps so that liquid and composition profile integrities are maintained throughout the chamber length.

The feed stock enters either the evaporation chamber or condensation chamber in the stage that has the feed composition. Usually this entry occurs at the hot end of the device (i.e., stage 1).

As bulk evaporation in one chamber and condensation in the other chambers are occurring, so are changes to composition of the carrier-gas vapors and liquids. As the gas increases in temperature in the stripper, the most volatile component of the liquid will tend to evaporate from the liquid, preferentially to the lesser volatile liquid species. The segmented wetting serves to preserve the concentration and temperature integrity of the liquid phase relative to the gas phase wherever the liquid wets the partition. As the gas moves through the stripper at

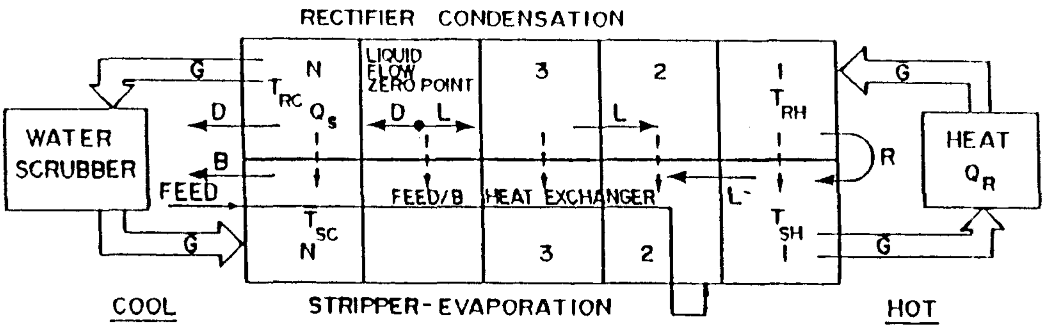

**Figure 1. Carrier gas fractionation process.**

ever-increasing temperatures, it gains volatile vapors thereby depleting the volatile species from the liquid phase. The liquid phase, migrating slowly through multiple stages, can be nearly void of the more volatile species when reaching discharge port at stage *N*. The gas, completing its passage through the stripper, then passes through the heater and enters the condensating rectifier. Upon cooling, the gas phase becomes richer in the more volatile species as: first, the lesser volatile species condenses preferentially thereby leaving the vapor phase richer in the more volatile species; and second, the gas phase remaining in equilibrium experiences a richer volatile liquid phase that is moving countercurrent to it as internal reflux. This internal reflux flows back to the stripper.

A richer volatile liquid phase is produced from the condensing vapor/gas mixture as the gas moves to exit from the *rectifier*. These stages become increasingly richer in the volatile component. A migratory flow of liquids is then discharged as distillate from stage *N* of the rectifier.

In the event of the gas still contains significant volatile vapors, a condenser and possible scrubber may be used to clean the carrier-gas before its return to the stripper.

Wide orifice liquid spray nozzles have been used to wet the partition with liquid droplets. Since the channels were over 7 cm (3 in.) wide, the spray induced normal velocity was much greater than the main vapor flow. Action of the spray droplets impinging on the partition disturbs the gas boundary layer adjacent to the boundary surface, generating enhanced overall heat and mass transfer coefficients.

The following equations result from molar balances around the entire fractionator and around one chamber of the fractionator. The overall material balance relates the feed rate, *F*, to the distillate rate, *D*, and bottoms rate, *B*, along with their respective species mole fractions, $x_f$, $x_d$, and $x_b$.

$$Fx_f = Dx_d + Bx_b \qquad \text{(Species balance)} \tag{1}$$

$$F = D + B \qquad \text{(Overall mole balance)} \tag{2}$$

so

$$D/F = (x_f - x_b)/(x_d - x_b) \tag{3}$$

Similar balances around the condensation chamber give the internal reflux rate at the hot end turnaround zone, *R*, in terms of mole fractions in the feed and distillate and the theoretical mole fraction in the vapor phase, $y_f$, that is in equilibrium with the feed.

$$R/F = (D/F)(x_d - x_f)/(y_f - x_f) \tag{4}$$

The carrier gas, *G*, needed to evaporate the desired distillate along with internal reflux from the evaporation chamber so that the vapors can condense in the condensation chamber is:

$$G/F = (R/F + D/F)/\delta V \tag{5}$$

The value of $\delta V$ is the change in the moles of vapor per mole of inert gas as the inert gas moves from the hot to the cold end of the condensation chamber.

From an energy balance around the hot end heater, the rectifier temperature can be calculated thereby giving the thermal driving force at the hot end of the fractionator. An energy balance around the entire fractionator gives the difference in thermal driving force at the fractionator cold end, $T_{RC} - T_{SC}$. These temperature differences yield the $\delta T_{LM}$ driving force. Given the overall heat transfer coefficient, the total partition area *A* required to perform a specified task is:

$$A = Q/(U^* \delta T_{LM}) \tag{6}$$

## Data Results

The base case operation was performed at a 76°C (168°F) turnaround temperature with a feed rate of 29.5 kg/h (65 lb/h) of 10 wt. % ethanol and an inert nitrogen carrier-gas rate of 5.1 $m^3/h$ (3 $ft^3/m$). Thermal start-up was performed by use of maximum heater energy input of 2.3 kW (8,000 Btu/h) at the turnaround point until the 76°C (168°F) operating temperature was reached. Energy input was then reduced and manually adjusted for temperature control. Figure 2 shows the temperature profiles in the liquid feed stream, stripper gas, and rectifier gas streams. Figure 3 is a liquid composition profile in the rectifier showing increasing ethanol content, while Figure 4 shows decreasing liquid ethanol composition on the stripper section.

Energy input of 1.20 kW (4,100 Btu/h) to the turnaround of the device was calculated based on the liquid flow rate (3.91

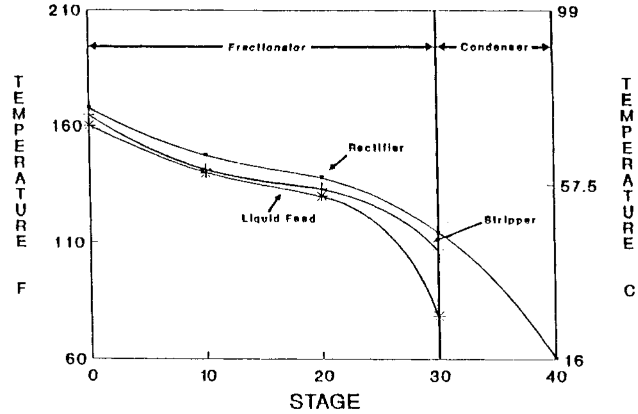

**Figure 2. Temperature profile: base case.**

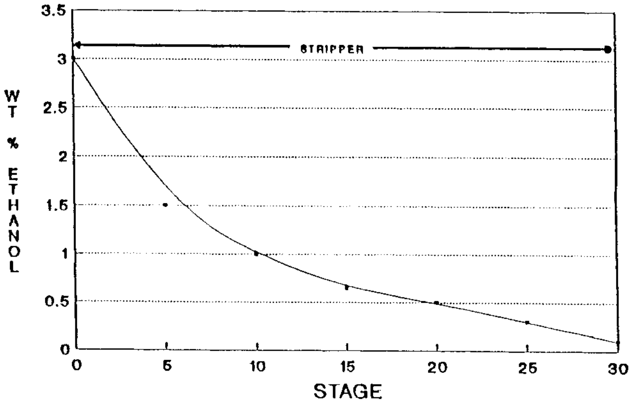

**Figure 4. Evaporation chamber ethanol composition profile.**

kg/min or 512 lb/h) to and from a heater multiplied by the temperature difference of 4.5°C (8°F) measured between the heated reinjected liquid and the basin liquids being withdrawn. An overall energy balanced on the fractionator includes energy input at the turnaround, $Q_R$, the energy difference between the entering and exiting liquid and gas streams, $Q_L$ and $Q_V$.

$$Q_R = Q_L + Q_V$$

The energy associated with the liquids, $Q_L$, was calculated to be 0.61 kW (2,090 Btu/h). The energy difference between the two gas streams was calculated to be:

$$Q_V = 0.59 \text{ kW } (2{,}010 \text{ Btu/h})$$

so

$$\begin{aligned} Q_R &= Q_V + Q_L \\ &= 1.20 \text{ kW } (4{,}100 \text{ Btu/h}) \text{ net thermal energy usage} \end{aligned}$$

Electric parasitic requirements for the base case run were calculated based on the 3.8 L/h (1 gal/h) production rate. Looking to liquid movement, the device contained a total of 168 nozzles, half of which operated with 1.91/min at 0.24 kPa and half of which operated with 1.21/min at 0.27 kPa. This results in a total pumping electrical requirement of 88 kJ/L (318 Btu/gal) of ethanol produced. Parasitic usage for gas movement was minimal, resulting 0.07 kJ/L (0.25 Btu/gal) of product (5.1 $m^3$/h at 0.05 kPa).

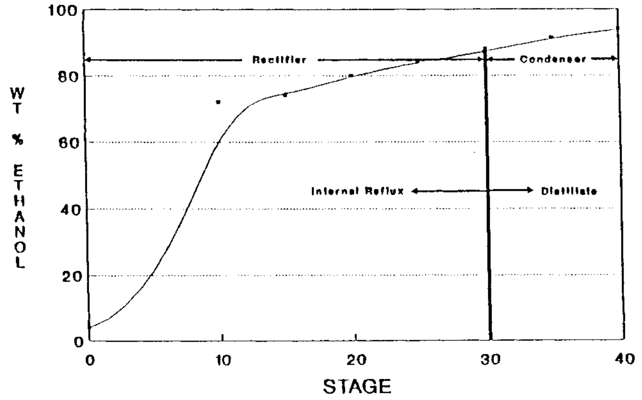

**Figure 3. Condensation chamber ethanol composition profile.**

Laboratory operation achieved separation of a 10% ethanol/water feedstock to a maximum purity of azeotrope distillate and a bottoms discharge containing 0.1% ethanol. Heat energy used was 1,100 kJ/L (4,100 Btu/gal) of distillate. This energy consumption represents a savings of 77% when compared to the standard distillation column usage of 5,000 thermal kJ/L (18,000 Btu/gal) of azeotrope ethanol. This new fractionation method might be used to separate other binary or multicomponent systems at any constant pressure and should find use in petroleum refining and chemical plant service.

## Notation

$A$ = total partition area
$B$ = molar bottoms rate
$D$ = molar distillate rate
$F$ = molar feed rate
$G$ = molar carrier gas rate
$L$ = liquid molar flow rate
$N$ = total number of stages
$n$ = number of a stage
$Q$ = total heat through partition = $N^* Q_3$
$Q_L$ = bottoms liquid energy rate relative to feed temperature
$Q_R$ = hot-end energy input rate = $Q_L + Q_V$
$Q_3$ = energy rate through a stage partition
$Q_V$ = energy difference between gas streams at stage $N$
$R$ = molar reflux rate at turnaround
$T_{RC}$ = rectifier cold-end (stage $N$) temperature
$T_{RH}$ = rectifier hot-end (stage 1) temperature
$T_{SC}$ = stripper cold-end (stage $N$) temperature
$T_{SH}$ = stripper hot-end (stage 1) temperature
$\delta T_{LM}$ = log-mean temperature difference
$U$ = overall heat transfer coefficient
$V$ = moles of vapor per mole of carrier gas
$x_b$ = ethanol mole fraction in bottom stream
$x_d$ = ethanol mole fraction in distillate stream
$x_f$ = ethanol mole fraction in feed stream
$y$ = ethanol mole fraction in vapor
$y_f$ = vapor ethanol mole fraction in equilibrium with $x_f$

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