Scaling-up Distillation Efficiencies

This paper describes a study of the possibility of using point efficiencies, measured in a small laboratory column, in the scale-up of distillation efficiencies. The conditions on a large rectangular tray were closely duplicated in a small glass Oldershaw column with one sieve tray having similar hole size and percent free area. Point efficiencies on the large rectangular sieve tray were deduced from the composition profiles and compared with the point efficiencies measured in the small column. A simple relationship between the two columns was found in terms of the froth heights and liquid holdups. The results demonstrate that the slightly lower point efficiencies measured in the Oldershaw column could be used in conjunction with a mathematical mass transfer model based on the eddy diffusion concept, to relate the point efficiencies to the tray efficiences, resulting in a conservative design provided either that allowance is made for other factors, such as nonuniform flow patterns and stagnant zones, or that these are eliminated. A wide range of composition was covered using the two systems MeOH-nPrOH and EtOH-nPrOH operating in the froth region.

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SCOPE

The prediction of the performance of a large-scale distillation tray column is difficult, particularly if a new system is being processed. It is a problem that has attracted a considerable amount of attention over many years. However, the prediction of tray efficiency is still a very uncertain part of the design unless operating data from a similar size column distilling a similar system are available. The problem is that both mass transfer characteristics and tray hydraulic considerations are involved, and so the tray efficiency is a function of system and tray size. This paper is concerned with the mass transfer part of the problem. It describes a study of the relationship betwen point efficiencies, measured in a small glass Oldershaw Column, and the efficiencies inferred at points along a rectangular tray with a liquid flow path length of about 1 m. The tray used is a sieve type with small hole sizes typical of those used in the cryogenic distillation of air. The hole sizes and percent free area in the two trays being compared are similar.

CONCLUSIONS AND SIGNIFICANCE

The relationship between point efficiencies measured in a small laboratory column and the point efficiencies that actually exist on a large rectangular tray were studied using two systems, MeOH-nPrOH and EtOH-nPrOH, and covering a wide range of composition. The composition profiles across the tray were measured and used to infer the values of point efficiencies using a mathematical mass transfer model based on the eddy diffusion concept.

Similar conditions to those on the large rectangular sieve tray were used in a small Oldershaw column with one sieve tray and external downcomers. The results indicate that the point efficiencies obtained on the large rectangular tray were slightly higher than the efficiencies obtained from the Oldershaw column, by about 8% on average. In addition the results also indicate that the trend observed for the variation of point efficiencies with composition was similar in the two columns.

A simple relationship between the ratio of the overall

number of transfer units in the gas phase was developed in terms of the froth height and liquid hold up on the tray as follows:

$$\frac{N_{OG_2}}{N_{OG_1}} = \sqrt{\frac{h_{f_2} - h_{L_2}}{h_{f_1} - h_{L_1}}}$$

This relationship was developed based on representing the vapor-phase and the liquid-phase mass transfer coefficients using the penetration theory, and assuming an equal average bubble diameter in the two columns. The relationship was tested experimentally and good agreement was found, the difference being within 10%. By the use of this relationship, the slightly higher point efficiencies on the large tray can be explained in terms of the effect of the height of the outlet weir on the liquid holdup as well as on the froth height on the tray, resulting in shorter contact times in the small column. This is in agreement with the results of Fair et al. (1983), who studied the scale-up of plate efficiencies from a laboratory Oldershaw column with internal downcomers containing various trays, where the point efficiencies of large industrial conventional trays were deduced using an eddy diffusion model, and did not allow for the existence of liquid flow nonuniformities and stagnant zones. In the present work the large rectangular tray, previously described and used in the study of MeOH-H_2O system by Biddulph and Dribika (1986), had a small width and long flow path to avoid flow nonuniformities and stagnant zones. In addition the two columns had trays with similar hole size and percent free area.

The results presented in this work indicate that the laboratory Oldershaw column having one sieve tray and external downcomers can be used to obtain an approximation of the point efficiencies on a similar large tray, and can be used in conjunction with an eddy diffusion model (Biddulph, 1975b, 1977; Biddulph and Dribika, 1986) to relate these point efficiencies to the tray efficiencies. This would result in a conservative design provided that allowance is made for other factors, such as nonuniform flow patterns or stagnant regions, or that they are eliminated. The Oldershaw column is easily assembled and used, and thus could provide an alternative means for the prediction of point efficiencies to be used in large-scale design.

Introduction

A very important part of the design of distillation columns is the specification of the number of trays. This is usually done by first calculating the number of theoretical stages and then modifying the result by the use of a tray efficiency or an overall column efficiency. The number of theoretical plates can be calculated quite accurately, but it is still difficult for a design engineer to specify a value for the plate efficiency. This is because of the complexity of the interrelationships among the parameters that affect plate efficiency. These parameters can be summarized as system composition and properties, flow conditions and vapor-liquid flow rates, and the type and dimensions of the tray. The usual approach to predicting tray efficiency is based upon one or more of the following:

1. Field performance data with a similar column or similar system. These are often difficult to obtain from the open literature.
2. Empirical correlations such as those of Drickamer and Bradford (1943), O'Connell (1946), and Bakowski (1969). These correlations have not always been validated against large-scale data and the results are usually conservative.
3. The use of theoretical mass transfer models relating point efficiency to the tray efficiency. In this respect the prediction is divided into two unrelated parts: first, the prediction of point efficiency, and second, prediction of the relationship between point efficiency and tray efficiency taking into account such factors as liquid mixing, vapor mixing, liquid residence time, entrainment, and weeping. In this approach the AIChE method is the most widely used. However, it has been observed by Lashmet and Szezepanski (1974), Hughmark (1971), Strand (1963), and Neuburg and Chuang (1982a) that this method often overpredicts tray efficiency. In addition Swanson and Gerster (1962) have stated that the method does not apply to trays containing holes of very small diameter. This has been confirmed recently by Biddulph and Dribika (1986) using a large rectangular tray with small-diameter holes, where the AIChE method was found to underpredict tray efficiency. Over the years since the AIChE method was developed it has been found that the existence of flow nonuniformities and stagnant zones (Bell, 1972a, b; Porter et al., 1972; Weiler et al., 1973) across large distillation trays can reduce the tray efficiency. Improvement of the theoretical models to allow for such effects (Porter et al., 1972; Lockett et al., 1973; Lockett and Safekourdi, 1976; Lim et al., 1974) has achieved reasonable success. On the other hand improvement in the hydraulic behavior of large circular trays to avoid such effects could also lead to significantly higher tray efficiencies, as demonstrated by Biddulph and Dribika (1986). However the first part of the problem, the prediction of point efficiency, has received less attention. Lockett and Ahmed (1983) have pointed out some of the deficiencies in the AIChE method. Lockett and Plaka (1983) have studied the effect of nonuniform bubble size on the prediction of point efficiency. Zuiderweg (1983) has demonstrated the influence of the Marangoni effect in alcohol-water systems as compared with hydrocarbon systems. Zuiderweg (1982) has reviewed sieve tray performance and developed correlations to predict point efficiencies, taking into consideration the importance of the flow regimes. Danckwerts et al. (1960) have studied thermal distillation effects. Hughmark (1965) has developed empirical correlations for interfacial area and liquid residence time and provided an improvement in the AIChE method for the prediction of the number of transfer units, especially for liquid-phase controlled systems. Stichlmair (1978) developed a predictive method for point efficiency taking into account surface tension and flow effects, but the method does not take hole diameter into account

(Kreis and Raab, 1979; Biddulph and Dribika, 1986). Chan and Fair (1982), improving on the AIChE method, developed a correlation for the volumetric vapor-phase mass transfer coefficient in terms of vapor diffusivity, liquid holdup, and fractional approach to flooding. However the correlation was estimated to have an accuracy of ±20% in efficiency prediction. It seems the problem of predicting point efficiency still exists, and a general method covering different types of tray geometry, systems, flow regimes, and conditions is not available.

4. Utilization of small laboratory or pilot plant efficiency data to predict point efficiency provides an alternative to overcome some of the difficulties in predicting point efficiency. In this respect Fair et al. (1983), following the industrial results of Veatch et al. (1960), have shown that point efficiencies measured in a small laboratory Oldershaw column with internal downcomers were almost equal to or slightly lower than the point efficiencies deduced from large commercial trays in the range from 50 to 85% of flooding. Fair et al. recommended a method for scale-up from Oldershaw columns. Andrew (1969) also studied scale-up of efficiencies. Biddulph (1975b, 1977) and Biddulph and Ashton (1977) have inferred point efficiencies in the systems nitrogen-oxygen-argon and benzene-toluene-xylene in industrial columns and found good agreement with small-column point efficiencies of Nord (1946) and Haselden and Thorogood (1964). Garrett et al. (1977), utilizing point efficiencies measured in a 25 mm dia. Oldershaw column, showed that the tray efficiencies predicted for a 45 cm dia. sieve tray column were in close agreement with the measured tray efficiencies. Neuburg and Chuang (1982a, b) developed a mass transfer model in terms of the overall volumetric mass transfer coefficient, expressed as a function of the equivalent Sauter mean bubble diameter measured in a pilot plant sieve tray column of 0.311 m dia., and used it to predict point efficiencies in an industrial Girdler-Sulfide (GS) plant having sieve trays of 8.5 m dia. Tong-Yun and Yun-Sheng (1981) have correlated point efficiencies measured in a small-scale laboratory column of 30 mm dia. as a function of compositions and predicted the number of actual trays of an industrial column 1.7 m in dia. The predicted number of trays was in close agreement with the actual industrial column.

Although in this fourth approach the literature results discussed above are encouraging, in combination with the wealth of data available from studies in small columns for many different systems and conditions, its general validity still needs further verification. This can only be done if experimental point efficiencies that actually exist on large distillation trays, covering different types of tray geometry, dimensions, etc., become generally available. In addition, this approach is closely connected to the step of relating the point efficiency to the tray efficiency. Therefore it is intended in this work to study the relationship between the point efficiencies measured in a small laboratory column and the point efficiencies that actually exist on a larger tray. The larger tray is rectangular in shape and of small width and long flow path to avoid flow nonuniformities and stagnant zones, and in addition to avoid the high cost of building and running a large circular tray. The conditions on the rectangular tray were duplicated in an Oldershaw column of one sieve tray with external downcomers to establish the relationship between point efficiencies measured for the same composition, system, flow conditions, and with similar hole size and percent free area. It is important to stress that if hole sizes and/or free area on the two trays being compared were significantly different the froth structure might be different and the comparison might not be as satisfactory. Two systems of a polar nature were studied, MeOH-nPrOH and EtOH-nPrOH, these being nearly ideal and having almost neutral surface tension characteristics. The physical properties of the systems are quite similar except that the relative volatility of the system MeOH-nPrOH is much higher than that of the system EtOH-nPrOH.

Experimental

Large-scale equipment

The distillation equipment that was used in this work was previously described in detail by Biddulph and Dribika (1986). Briefly it consists of a rectangular distillation column having overall dimensions of 1,067 × 89 mm, a steam-heated stainless steel reboiler having a capacity of 450 L, and glass condensers providing a condensing surface area of 5.3 m^2. The rectangular distillation column incorporated three trays, with the middle one being the test tray. The test tray was designed with six equally spaced points for sampling and temperature measurement along the center line, and was provided with double-glazed glass portholes to enable easy observation of the biphase. Details of the tray are given in Table 1. The sieve tray material is typical of that commonly used in low-temperature air distillation. The equipment also was provided with temperature/sample points in the inlet and outlet downcomers as well as the reboiler. Furthermore, two other sample points were provided for the liquid leaving the condensers and the column.

The column was operated at total reflux and atmospheric pressure. About 200 L of the test mixture were used, the steam pressure being adjusted to give the required boilup rate, which was measured using a calibrated rotameter placed in the reflux line. The column was run for 4 to 5 h to establish steady state conditions. Then the boilup rate, the temperatures, and froth height were noted and the liquid samples were collected. The operation of the column was carried out at a vapor phase F factor of 0.4 m/s (kg/m^3) and covered a wide range of composition. At this F factor the column was operating in the range 45–55% flooding.

Small laboratory column

The column is 3.8 cm in diameter, with a sieve plate of the glass Oldershaw type modified by external downcomers. These replace the more usual central internal downcomers, which tend to break up the bubbling area. It is thought to be more representative of a small bubbling area on a large tray. The Oldershaw column has one tray only. This reduces the overall composition change in the column and avoids excessive averaging. The tray specifications are given in Table 2. A schematic diagram of the column is shown in Figure 1. The boiler is a 2 L flask with a

Table 1. Tray Details, Large Rectangular Column

Weir length	83 mm
Liquid flow path	991 mm
Tray spacing	154 mm
Hole diameter	1.8 mm
Free area	8 %
Outlet weir height	25 mm
Inlet weir height	4.8 mm

Table 2. Tray Details, Glass Oldershaw Column

Internal diameter	38 mm
Tray spacing	94 mm
Hole diameter	1.1 mm
Free area	8 %
Outlet weir height	2.1 mm

thermometer pocket. It is heated electrically by a 750 W heating element. Vapor passes from the boiler through a calming section and into the tray column. After leaving the column, the vapor passes through a glass elbow section to the condenser where it is totally condensed, the condensate returning to the column through a rotameter. A chimney prevents contact between vapor and liquid at the sample point *S1*. The liquid leaving the tray is sampled at *S2* and returned to the boiler. The column was contained within a heated jacket to ensure adiabatic conditions.

The column was operated at total reflux and atmospheric pressure for about 4.5 h for each run. This period was found to be more than sufficient to establish steady state conditions. After this period the flow rate, froth height, and temperature were noted and then liquid samples were collected into chilled sample tubes. The operation of the column was maintained at two different vapor phase *F* factors for each run. A wide range of composition was covered for both systems.

Equilibrium Data

Two different sets of isobaric vapor-liquid equilibrium data for the system MeOH-nPrOH and one set of equilibrium data for the system EtOH-nPrOH were found in the literature (Hill and Van Winkle, 1952; Ochi and Kojima, 1969). The boiling points as reported by these investigators were significantly different from one other in the system MeOH-nPrOH. Both sets of data were measured at 1 atm pressure. All three sets of data were shown to be thermodynamically inconsistent by Gmehling and Onken (1977), so it was necessary to measure a new set of equilibrium data.

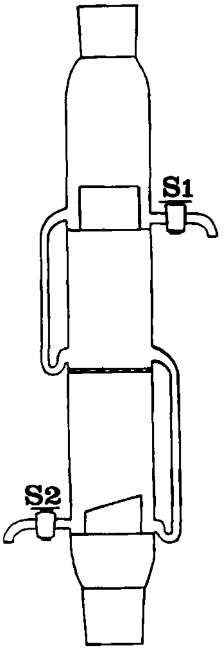

Figure 1. Laboratory Oldershaw column.

The apparatus and procedure that was previously used by Dribika et al. (1985) to measure data in the ternary system MeOH-EtOH-isoPrOH was followed closely. The data are shown in Figures 2–4. The data were tested for thermodynamic consistency using the Herington (1951) test, the activity coefficients being evaluated by using the nonidealities in the vapor and the liquid phase, since the two systems are polar. The results indicated that the two systems were nearly ideal. The data for the system EtOH-nPrOH were found to satisfy the Herington test better than the data for the system MeOH-nPrOH. This could be due to the absence of the heat of mixing term because of the lack of experimental heat of mixing data at the boiling point temperature in the system MeOH-nPrOH.

The data were then correlated using the Wilson model. The Wilson parameters were evaluated by a nonlinear least-squares method coupled with the use of the Q function as an objective function. The method is described in detail by Nagahama et al. (1971), and has been applied successfully in a similar system, MeOH-EtOH-isoPrOH, by Dribika et al. (1985). The final results for the Wilson parameters as well as the mean deviation in the vapor mole fraction predicted and the bubble point temperature are shown in Table 3.

Sample Analysis

All liquid samples were analyzed using a Varian Vista model 44 gas chromotograph with automatic injection. The chromotograph was equipped with a flame ionization detector. The column was 290 cm long, supplied ready-packed with Porapak Q, and was operated isothermally at 150°C and 3 bar column pressure. The detector and the injector temperatures were 180 and 150°C, respectively. Nitrogen was used as the carrier gas at a rate of 25 mL/min. Calibration analyses were carried out using 16 samples of known composition in the range of 0.04–0.96 mole fraction for the two systems. Two or three injections were made for each sample, area ratios of the two components being obtained directly from the digital computer output. Since the ratios of weights were known, a least-square method was

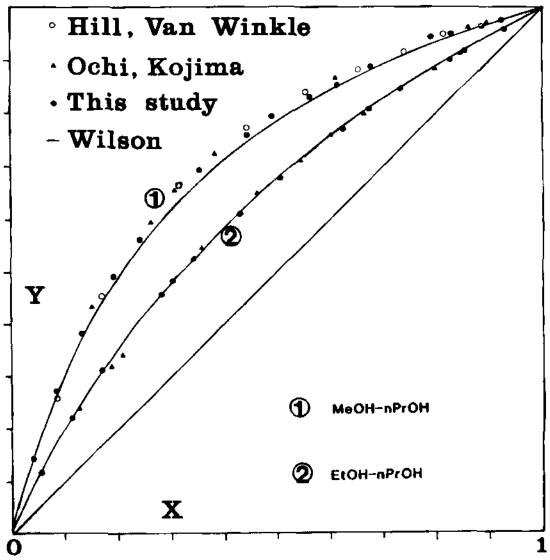

Figure 2. Vapor-liquid equilibrium data.

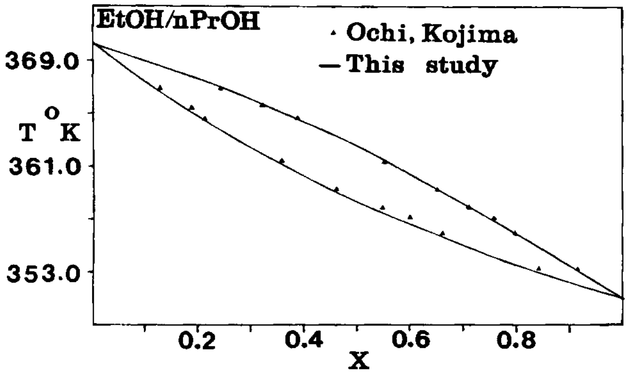

Figure 3. Saturation temperatures in EtOH-nPrOH.

used to find the best correlating equation. The accuracy of the method was determined to be ±0.0012 m.f. for EtOH-nPrOH and ±0.0025 m.f. for MeOH-nPrOH.

Theoretical Model and Mixing Studies

The concept of eddy diffusion was used in this work to model the experimental profiles across the tray and to infer point efficiencies. The model has been used previously (Biddulph, 1975b, 1977; Biddulph and Dribika, 1986). The model is based on a mass and enthalpy balance over a slice of the biphase on the tray. The resulting differential equations were solved numerically using a predictor/corrector method, stepping across the tray from the outlet weir to the inlet weir. The solution uses 50 steps across the tray and is iteratively stable; it provides predicted composition profiles across the tray for given values of component point efficiencies and predicts component tray efficiencies.

The model requires a knowledge of the extent of mixing present in the liquid phase on the tray. This was obtained directly from the same test tray by experiments with water/steam using sodium nitrate solution as the tracer injection. The results were obtained under loadings similar to those used in the experiments. Full details of the experiments are given elsewhere by Biddulph and Dribika (1986). A Peclet number of about 39 was established from the eddy diffusivity value determined from the mixing experiments. This value indicates that the conditions are approaching plug flow on this tray.

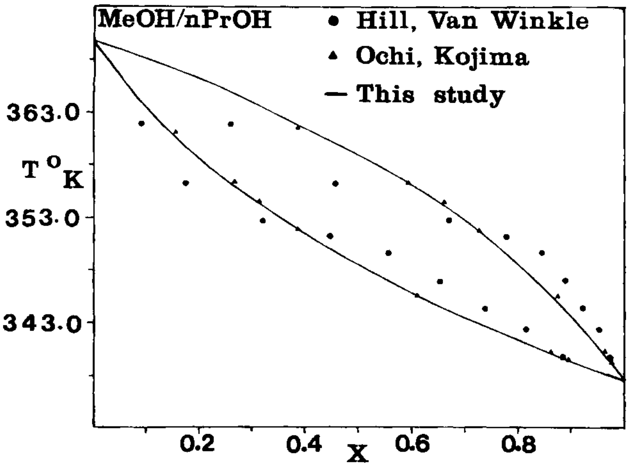

Figure 4. Saturation temperatures in MeOH-nPrOH.

Table 3. Wilson Parameters and Mean Deviation in Vapor Mole Fraction and Bubble Point Temperature

System	Wilson Parameters	ΔY	$\frac{\Delta T}{K}$
MeOH—nPrOH	421.821 and −245.909	0.010	0.82
EtOH−nPrOH	385.395 and −299.258	0.0046	0.44

Results

Rectangular column

The composition profiles predicted using the theoretical model were compared with the experimental profiles, as shown in Figures 5 and 6. The point efficiencies used in the model were varied until a good match was achieved between predicted and measured profiles, and these values were then taken as the correct point efficiencies operating during that run. The final values of the point efficiencies were plotted for the two systems, as shown in Figures 7 and 8. The sensitivity of the profiles to the assumed point efficiency is illustrated in Figure 9. The mean liquid composition on the tray was calculated from the composition profile using:

$$\bar{x} = \int_{o}^{w-1} x_w \, dw \qquad (1)$$

The predicted and experimental tray efficiencies are also plotted against the mean liquid composition in Figure 10. All the experimental runs for the two systems were made at a vapor phase F factor of about 0.4. At this F factor the entrainment, predicted using the correlations of Fair and Smith (1963), was

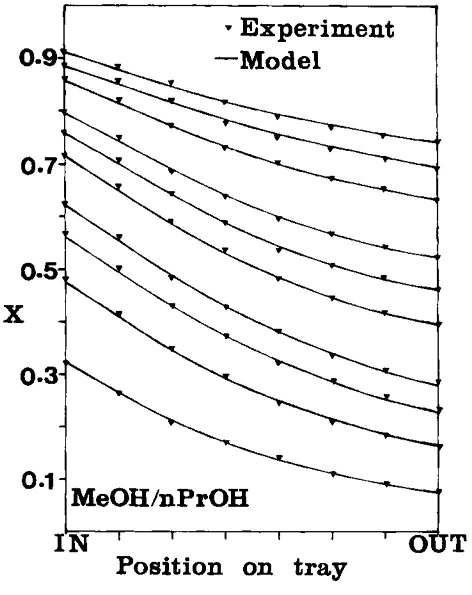

Figure 5. Liquid composition profiles along tray in MeOH-nPrOH.

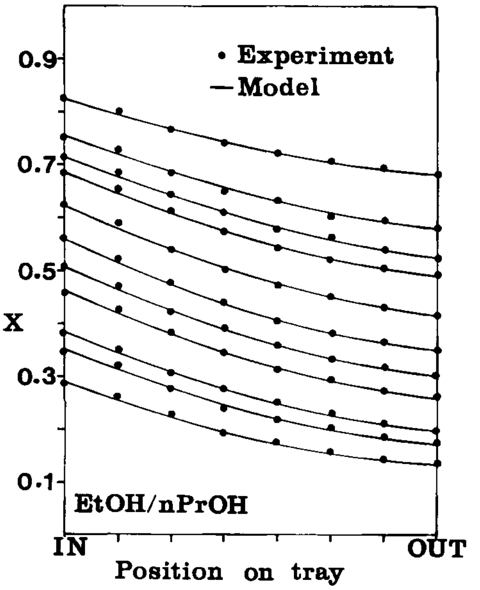

Figure 6. Liquid composition profiles along tray in EtOH-nPrOH.

found to be less than 0.02 and this value would have a negligible effect on the efficiency (Rahman and Lockett, 1981). In addition, negligible weeping was observed through the glass windows.

The experimental liquid temperature profiles across the tray are shown in Figures 11 and 12. The temperatures were compared with the saturation temperatures, as shown in Fig 13, showing very close agreement. The temperatures in the downcomers in the system EtOH-nPrOH were also very close to the bubble point temperatures. However there is slight vaporization in the downcomers in the system MeOH-nPrOH. This was probably due to the large temperature difference in this system. This observation is similar to previous results in the system MeOH-H_2O on the same tray. However the effect on the efficiency is very small.

The results of the variation of tray efficiency with composition showed, as usual, that higher tray efficiencies were obtained at the lower concentration range of the more volatile component. In addition, in the middle to higher range of composition used in the two systems, the tray efficiencies for the system EtOH-nPrOH were found to be higher than for the system MeOH-nPrOH. Since in these composition ranges the slopes of the equilibrium lines are similar in the two systems, the higher tray efficiencies were probably due to the higher point efficiencies in the system EtOH-nPrOH. The measured froth heights are shown in Figure 20.

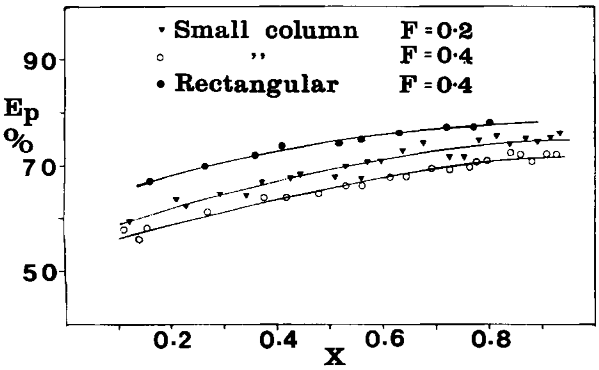

Figure 7. Point efficiencies in the system MeOH-nPrOH.

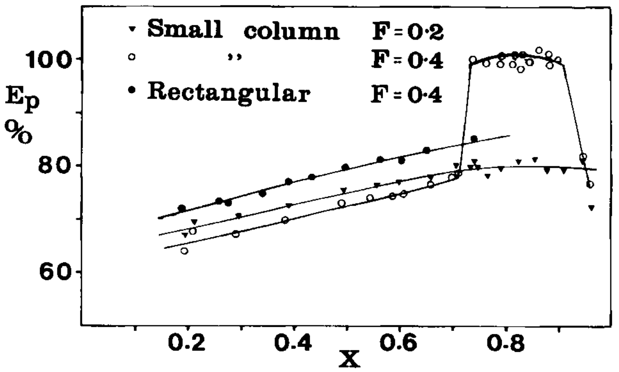

Figure 8. Point efficiencies in the system EtOH-nPrOH.

Transfer units

The number of overall vapor phase transfer units was calculated using the usual assumption of plug flow of vapor as it passes through the froth.

$$N_{OG} = -\ln(1 - E_{OG}) \tag{2}$$

From the two-film theory at total reflux:

$$\frac{1}{N_{OG}} = \frac{1}{N_G} + \frac{m}{N_L} \tag{3}$$

The variation of $1/N_{OG}$ with m is shown in Figure 14. The mean value of the slope of the equilibrium line was calculated

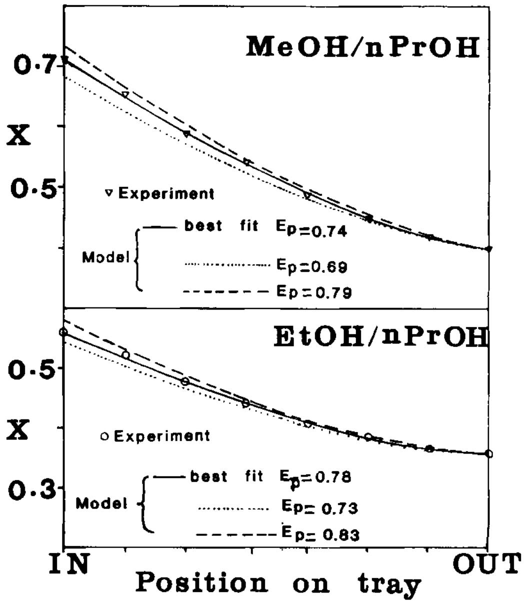

Figure 9. Sensitivity of profiles to assumed point efficiency.

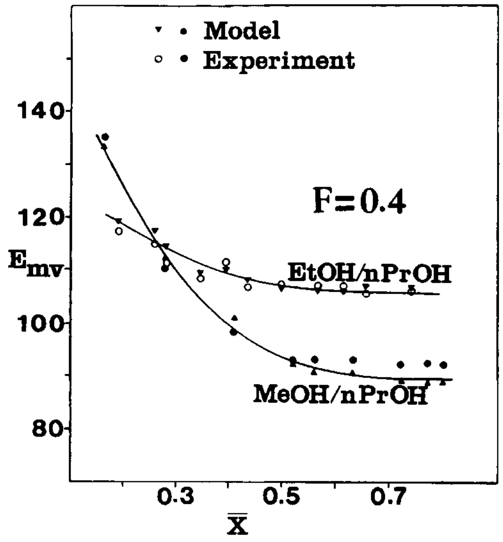

Figure 10. Tray efficiencies vs. mean liquid composition.

from the concentration profile using:

$$m = \int_o^{w-1} m_w \, dw \tag{4}$$

The values of N_G and N_L were determined using the usual intercept and slope method for the two systems and are shown in Figures 15 and 16. However this method assumes that N_G and N_L are composition-independent, which is not a realistic assumption. This is clear from the basic definitions of N_G and N_L:

$$N_G = K_G a h_f A/V \tag{5}$$

$$N_L = K_L a h_f A/L \tag{6}$$

In order to take account of the effect of all the variables in Eqs. 5 and 6, the approach normally used by many previous workers has been to assume a value of either bubble size or mass transfer coefficient and to calculate the other by fitting a mass transfer model to experimental results. The approach taken in this study has been to represent the vapor phase and the liquid phase mass transfer coefficients by the penetration theory (Higbie, 1935).

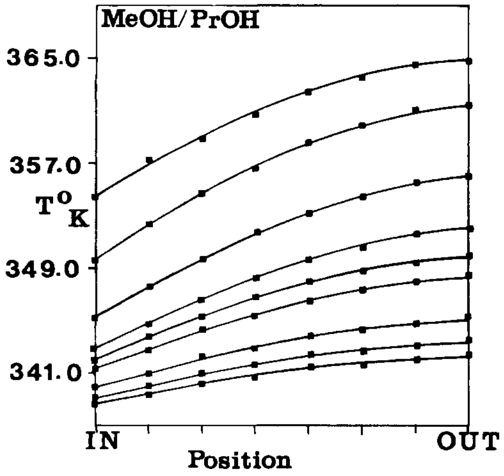

Figure 11. Measured temperature profiles along tray.

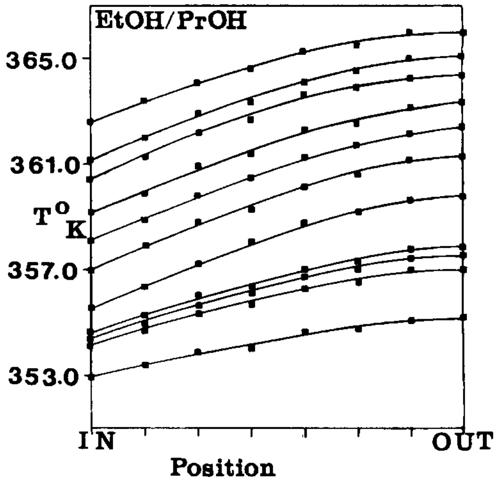

Figure 12. Measured temperature profiles along tray.

$$K_L = \sqrt{\frac{4D_L}{\pi\theta_L}} \quad \text{and} \quad K_G = \sqrt{\frac{4D_G}{\pi\theta_G}} \tag{7}$$

If it is assumed that the exposure times of the gas and liquid elements at the interface are equal, which has been assumed by many investigators (Calderbank and Pereira, 1977; Stichlmair, 1978; Neuburg and Chuang, 1982), then the vapor and liquid mass transfer coefficients are related by:

$$\frac{K_G}{K_L} = \sqrt{\frac{D_G}{D_L}} \tag{8}$$

Equation 8 has also been used by Lockett and Uddin (1980) and it is further supported by the experimental correlations of

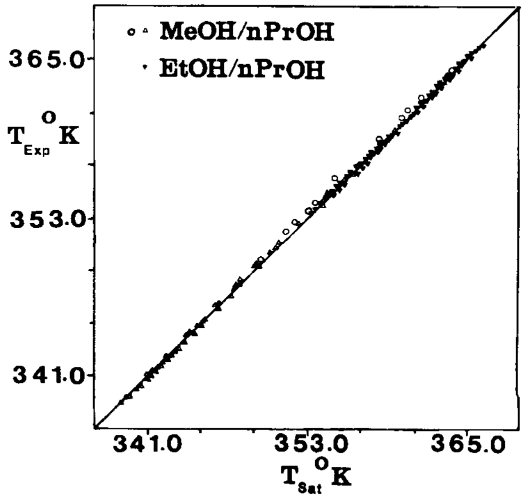

Figure 13. Comparison of measured temperatures with saturation temperatures.
0 represents downcomers

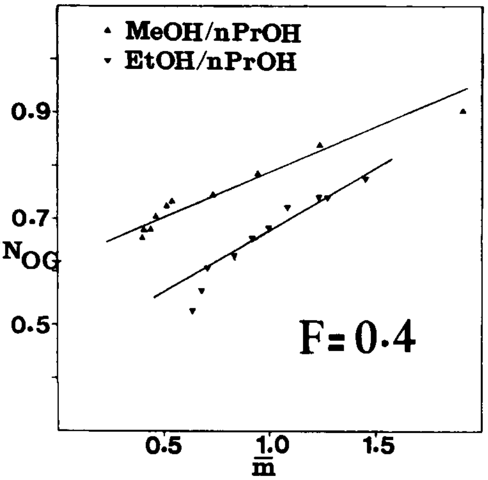

Figure 14. $1/N_{OG}$ vs. mean equilibrium line slope.

Sharma et al. (1969), which can be reduced, approximately, to the form of Eq. 8.

Rearranging Eq. 3 gives

$$N_{OG} = \frac{N_G}{1 + m\dfrac{N_G}{N_L}} \tag{9}$$

From Eqs. 5, 6, and 8 we have:

$$\frac{N_G}{N_L} = \frac{\rho_G}{\rho_L}\sqrt{\frac{D_G}{D_L}} \tag{10}$$

Substitution of Eq. 10 into Eq. 9 gives

$$N_{OG} = \frac{N_G}{1 + m\dfrac{\rho_G}{\rho_L}\sqrt{\dfrac{D_G}{D_L}}} \tag{11}$$

Table 4 shows the values of physical properties as a function of composition for the two systems used.

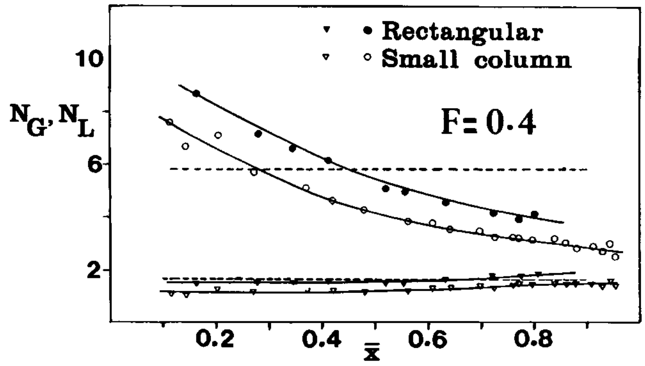

Figure 15. Variation of N_G and N_L with composition for the system MeOH-nPrOH.

. . . . .Values of N_G and N_L obtained by method of slope and intercept.

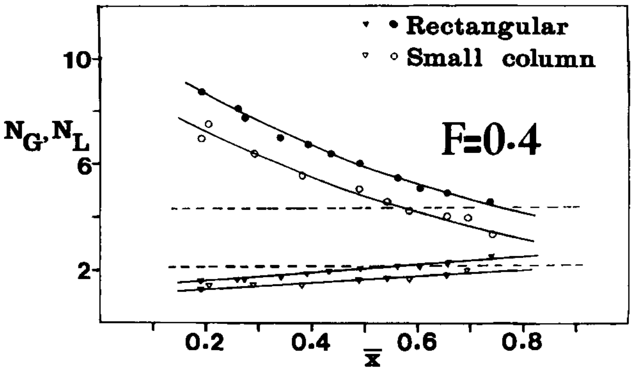

Figure 16. Variation of N_G and N_L with composition for the system EtOH-nPrOH.

. . . . .Values of N_G and N_L obtained by method of slope and intercept.

Equations 11 and 10 and Table 4 were used to calculate the values of N_G and N_L for each run independently. The final results are shown in Figures 15 and 16. The results indicate that there is a significant effect of composition, probably due to the variation of individual mass transfer coefficient and/or, in the case of N_L, interfacial area and physical properties. Using these values of N_G and N_L it was possible to estimate the extent of the liquid phase resistance (*LPR*) in the two systems by

$$\%\,LPR = \frac{mN_G}{N_L + mN_G} \tag{12}$$

The results are shown in Figure 17. In the system MeOH-nPrOH the increasing liquid phase resistance in the low methanol concentration region would give rise to lower point efficiencies in this region, an observation confirmed experimentally. This could be a result of the large value of the slope of equilibrium line. In the system EtOH-nPrOH the % *LPR* values were almost independent of composition, with a slight increase at the higher EtOH concentration. However these cannot explain completely the variations of point efficiency with composition without knowing the interfacial area. Nevertheless, these results indicate that the two systems are highly vapor-phase controlled.

Relative interfacial area

The time of contact for the vapor elements at the interface can be defined, as proposed by previous authors, to be roughly equal to the duration of the gas in the two-phase layer:

$$\theta = (h_f - h_L)/U_g \tag{13}$$

Combining Eqs. 5, 7, 11, and 13 gives:

$$N_{OG} = ah_f\sqrt{\frac{4}{\Pi}\frac{D_G}{U_g(h_f - h_L)}}\Bigg/\left(1 + \frac{m\rho G}{\rho_L}\sqrt{\frac{D_G}{D_L}}\right) \tag{14}$$

Rearranging this equation and multiplying by h_f gives the following equation for interfacial areas per unit active tray area:

$$\bar{a} = N_{OG}\left(1 + m\frac{\rho_G}{\rho_L}\sqrt{\frac{D_G}{D_L}}\right)\Bigg/\sqrt{\frac{4}{\pi}\frac{D_G}{U_g(h_f - h_L)}} \tag{15}$$

Equation 15 was used to determine the values of $\bar{a}$. The clear liquid heights could not be measured directly from the distilla-

Table 4. Physical Properties for the Two Systems

$\bar{x}$	m	ρ_G	ρ_L	D_G	D_L
		System: MeOH—n—PrOH			
0.2790	1.204	1.844	748	1.04×10^{-5}	1.46×10^{-9}
0.3420	0.940	1.798	750	1.02×10^{-5}	1.10×10^{-9}
0.4100	0.735	1.728	752	1.02×10^{-5}	8.08×10^{-10}
0.5550	0.505	1.602	755	9.93×10^{-6}	4.31×10^{-10}
0.7710	0.432	1.403	759	9.67×10^{-6}	1.75×10^{-10}
		System: EtOH—nPrOH			
0.1910	1.421	1.974	738	8.49×10^{-6}	1.84×10^{-9}
0.3420	1.086	1.926	739	8.31×10^{-6}	9.10×10^{-10}
0.4960	0.834	1.860	740	8.20×10^{-6}	4.48×10^{-10}
0.6017	0.714	1.794	740	8.22×10^{-6}	2.75×10^{-10}
0.7390	0.620	1.749	739	8.07×10^{-6}	1.49×10^{-10}

tion experiments. However a study of clear liquid heights on sieve trays has been recently reported by Air Products and Chemicals, Inc. (Bennett et al., 1983). The experimental work was based on small diameter holes of 1 and 3 mm and with outlet weir heights ranging from zero to 25 mm.

The correlation reported was:

$$h_L = \Phi_e\left[h_w + C\left(\frac{Q_L}{\Phi_e}\right)^{2/3}\right] \tag{16}$$

where

$$C = 0.0327 + 0.0286 \exp(-137.8\, h_w) \tag{17}$$

and

$$\Phi_e = \exp[-12.55(K_s)^{0.91}] \tag{18}$$

and

$$K_s = U_g\left(\frac{\rho_g}{\rho_L - \rho_g}\right)^{1/2} \tag{19}$$

The final results were plotted as a function of composition as shown in Figures 18 and 19. For the system EtOH-nPrOH, the increase in the interfacial area with increase in composition would give an increase in point efficiency, and this was confirmed by the experimental results. The froth height also increased slightly with composition. For the system MeOH-nPrOH, the increase in the interfacial area was not as sharp as in the system EtOH-nPrOH. This could be a result of the observation that the froth was changing its structure, showing irregular surfaces and also showing more droplets at the lower methanol concentrations. It seems that there is an experimental indication of a variation in the character of the froth with increasing composition.

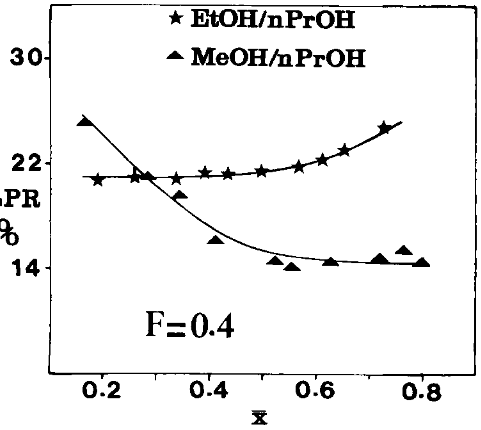

Figure 17. Variation of % LPR with mean liquid composition.

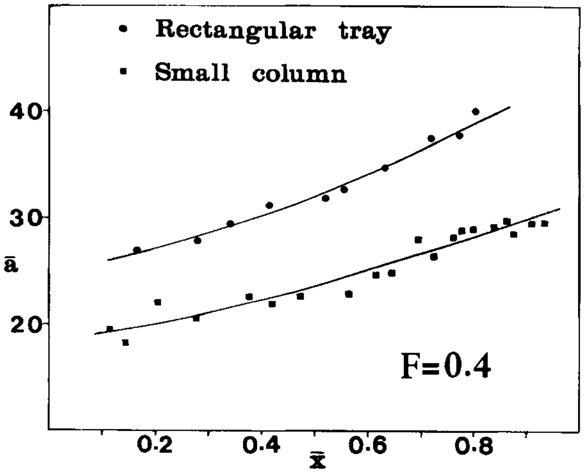

Figure 18. Variation of $\bar{a}$ with composition in the system MeOH-nPrOH.

These results indicate that the slight increase in the values of point efficiency in the system EtOH-nPrOH as compared with the system MeOH-nPrOH is primarily due to the interfacial area changes.

Oldershaw column

In a small-diameter Oldershaw column, the point efficiencies are determined directly since it can be assumed that the liquid phase is completely mixed and thus the tray efficiency equals the point efficiency.

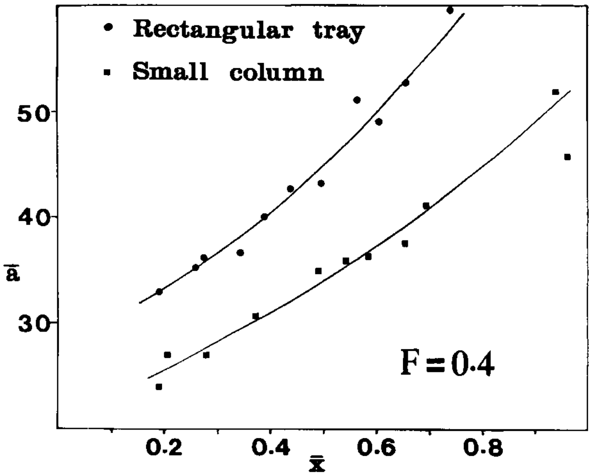

Figure 19. Variation of a with composition in the system EtOH-nPrOH.

$$E_{mv} = E_{op} = \frac{Y_o - Y_i}{Y_o^* - Y_i} \tag{20}$$

Figures 7 and 8 show the variation in point efficiency as a function of the liquid composition leaving the tray. The results of the individual number of transfer units in the liquid and vapor phase as well as the interfacial area per unit bubbling area are also shown in Figures 15, 16, 18, and 19. These results were obtained using the same equations developed for the larger rectangular tray. The same observation of a slight increase in point efficiency with increase in composition can be explained in the same way.

The efficiency results at the lower F factor of 0.2 are also shown in Figures 7 and 8. The point efficiencies were found to be slightly higher, especially in the system MeOH-nPrOH.

An interesting observation in this column was noted in the system EtOH-nPrOH. A sudden increase in the froth height, as shown in Figure 20, was observed at compositions of 0.71 and above, and this results in the sudden increase in point efficiency, some in excess of 100%, shown in Figure 8. This phenomenon was not observed in the large rectangular column at a mean composition of 0.74, nor at the lower F factor in the Oldershaw column runs.

The apparatus was carefully cleaned to remove any possible impurities, but the observation was reproducible. It appears to be a function of the design of the column. As the composition of ethanol is increased, the froth height increases and reaches a critical value when the top of the froth reaches the top of the outlet downcomer tube. The wall of the column then appears able to support the froth. The wall-supported froth had a cellular appearance. This wall effect has also been mentioned by Young and Weber (1972), Todd and Van Winkle (1972), Standart (1974), Thomas and Hag (1976), and Sawistowski et al. (1964). The effect is a drawback of this type of small column design, and one that is the subject of further study.

Discussion

The results shown in Figures 7 and 8 indicate that the point efficiencies obtained in the large rectangular tray are slightly higher than the efficiencies obtained from the Oldershaw column. The difference is 8% on average. The trend observed for the variation of point efficiency with composition is similar in the two columns. Furthermore the results showed similar trends for the number of individual transfer units and the interfacial areas in the two columns; qualitatively the froth character observed is the same in both columns.

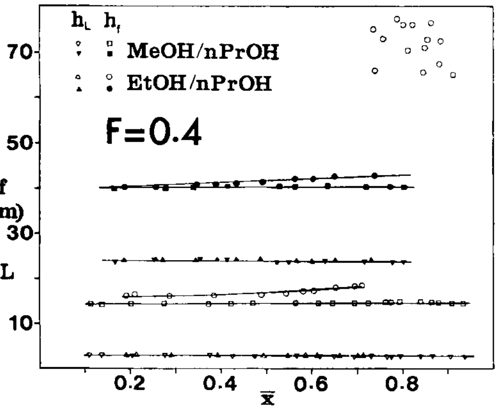

Figure 20. Variation of froth height and liquid holdup with composition.

●■▼▲ Rectangular column

○□▽△ Oldershaw column

In order to explain this slightly greater point efficiency, it is important to note that the vapor phase F factor and system properties at the mean composition for the two columns are similar. For sieve trays operating in the froth region a large proportion of the mass transfer is taking place at the bubble formation zone on the tray (Garner and Porter, 1960; Lockett et al., 1979). Thus a significant increase in froth height is likely to increase the point efficiency, but only marginally. Earlier results (Biddulph and Dribika, 1986) on the system MeOH-water support this view. The results of Haselden and Thorogood (1964) in a small apparatus using the system O_2-N_2 gives further support to the importance of mass transfer in the bubble formation zone. Therefore, since the hole diameters of the two trays are similar, it is reasonable to assume that the contribution to the mass transfer in this zone is similar. However there are differences in the froth height and the liquid holdup on the two trays, resulting in a difference in the gas residence time. It may be concluded that this difference is the major factor in the slight difference in point efficiency between the two columns.

Applying Eq. 14 to the two columns and dividing for the same conditions and composition gives:

$$\frac{N_{OG_2}}{N_{OG_1}} = \frac{a_2 h_{f_2}}{a_1 h_{f_1}} \sqrt{\frac{h_{f_1} - h_{L_1}}{h_{f_2} - h_{L_2}}} \tag{21}$$

where subscript 1 refers to the Oldershaw column and subscript 2 refers to the large rectangular column.

The interfacial area per unit volume of the froth can be defined as:

$$a = \frac{6}{d_B}\left(\frac{h_f - h_L}{h_f}\right) \tag{22}$$

Burgess and Calderbank (1975) have shown that the Sauter mean bubble diameter is almost equal at the various froth heights of interest in this work and is a weak function of the froth height at the same velocity. Therefore we may assume that the mean average bubble diameters produced on the two trays at the same conditions and composition are similar due to the similarity of the hole size. Therefore Eq. 21 becomes

$$\frac{N_{OG_2}}{N_{OG_1}} = \sqrt{\frac{h_{f_2} - h_{L_2}}{h_{f_1} - h_{L_1}}} \tag{23}$$

The experimental results are represented by this equation to within ±10%. This is achieved by defining the contact time for the mass transfer coefficients by Eq. 13. The alternative approach of defining contact time in terms of bubble diameters was found to be less satisfactory. Furthermore, in Eq. 23 h_{L_2} and h_{L_1} could be replaced by the appropriate outlet weir heights as an alternative, as suggested by Fell (1986).

It is not known how much mass transfer is concentrated in the formation zone compared with the froth zone. However, formation conditions would have an influence on the bubble diameter and hence on the interfacial area. In our study the two trays were similar and this part of the process can be assumed to have been similar. It should be noted that this form of equation, shown as Eq. 23, might not apply as well if the two trays being compared were significantly different.

For the two efficiencies to be equal it is important to note that the froth height and the liquid holdup are functions of the outlet weir height. The effect of the weir height has been extensively reported in the literature for small columns and large columns. The efficiency has been found to increase with an increase in the weir height (Haselden and Thorogood, 1964; Fane and Sawistowski, 1969; Kuehler and Van Winkle, 1969; Kreis and Raab, 1979; Neuburg and Chuang, 1982; Shore and Haselden, 1969). The weir height in the larger rectangular tray was 12 times the weir height in the Oldershaw tray. At lower weir heights on the larger tray the point efficiencies are likely to approach each other. This agrees with the results of Chan and Fair (1982) and Fair et al. (1983), who emphasized that better agreement between the point efficiencies deduced from the FRI (Fractionation Research Incorporated) data of Sakata and Yanagi (1979) and Yanagi and Sakata (1982) and an Oldershaw column with internal downcomers could be obtained at lower weir heights on the commercial scale tray.

From a design point of view, the slightly lower point efficiencies deduced from the Oldershaw column could be used (Biddulph, 1975) in a mathematical mass transfer model to relate the point efficiencies to the tray efficiencies, resulting in a conservative design, provided that other factors such as nonuniform flow patterns or stagnant regions are either eliminated or allowed for.

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Notation

A = bubbling area, m^2
a = interfacial area per unit volume of dispersion, m^{-1}
$\bar{a}$ = interfacial area per unit bubbling area
d_B = mean average bubble diameter, m
D_G = vapor phase diffusivity, m^2/s
D_L = liquid phase diffusivity, m^2/s
E_{mV} = Murphree tray efficiency
E_{OG} = overall vapor phase point efficiency
E_{OP} = overall vapor phase point efficiency of the small column
h_f = froth height, m
h_L = liquid holdup, m
h_w = weir height, m
K_G = vapor phase mass transfer coefficient, m/s
K_L = liquid phase mass transfer coefficient, m/s
K_s = defined by Eq. 19
L = liquid flow rate, m^3/s
LPR = fraction liquid phase resistance
m = mean slope of equilibrium line across tray
m_w = slope of equilibrium line at a point w on tray
N_G = number of vapor phase transfer units
N_L = number of liquid phase transfer units
N_{OG} = number of overall vapor phase transfer units
Q_L = flow rate of liquid per unit length of weir, m^3/min · m
U_g = vapor velocity based on bubbling area, m/s
V = vapor flow rate, m^3/s
W = dimensionless flow path length
X = average liquid mole fraction across rectangular tray
X_w = liquid mole fraction at a point w on tray
y_i = vapor mole fraction entering the tray
y_o = vapor mole fraction leaving the tray
y^* = equilibrium vapor mole fraction with liquid leaving tray

Greek letters

ρ_G = vapor phase density, kg/m^3
ρ_L = liquid phase density, kg/m^3
ϕ_e = relative froth density
θ_G = exposure times of vapor elements at interface, s
θ_L = exposure times of liquid elements at interface, s

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