MASS TRANSFER IN BINARY AND TERNARY DISTILLATION BY A PACKED COLUMN WITH STRUCTURED PACKING

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Key words: Distillation, Mass Transfer, Packed Column, Structured Packing, Binary System, Ternary System

A mass transfer correlation for distillation using a laboratory size wire-gauze structured packing was developed in terms of dimensionless diffusion flux based on experimental data for an acetone-ethanol system under total reflux conditions. The theoretical basis for mass and heat transfer in distillation that was previously developed for wetted-wall columns is shown to apply to structured packings as well. It is also demonstrated that the Sherwood number for distillation exhibited wide scattering due to the effect of convective flux, which is highly influenced by external heat effects and differences in the latent heat of vaporization of the transferring components.

The model obtained from the binary system, which is based on the total area of the structured packing and corrected for the effect of condensation of mixed vapors, is shown to apply to a ternary system of acetone-methanol-ethanol. Simulation results using the correlation for a short column compared well with experimental results for longer columns for binary and ternary systems.

Introduction

The replacement of standard random packings or tray internals with more modern structured packing in distillation and other gas-liquid contact operations is receiving increasing acceptance among process engineers. This is due to the many operational advantages of these packings such as: high efficiency, low pressure drop, low liquid hold-up, and high capacity. Kuk (1979) observed that Koch-Sulzer packing generates three to four times the equilibrium stages compared to conventional packing of the same height. There are various types of structured packings that appeared in the market, ranging from wire-gauze (woven cloth) sheets to corrugated metal sheets of various configurations and arrangements. Some are embossed, perforated or lanced to improve contact between vapor and liquid. Kister (1992) gives an excellent description of many of these types of packings in his recent book. Because of the large variations in the design characteristics of these packings, difficulty is encountered in establishing generalized models to define the hydrodynamic, heat and mass transfer behavior to assist process designers in predicting performance of distillation columns using this new generation of packings.

One of the early models specifically developed for distillation using structured packing is that of Bravo, *et al.* (1985). They suggested a gas-side mass transfer correlation for wire gauze packing based on the wetted-wall column relationship of Sherwood and Pigford. They extensively tested this correlation using distillation data reported in the literature. Fair and Bravo (1990) later modified this correlation for sheet metal structured packing. Many of the mass transfer models proposed for distillation using structured packing are based on data from gas absorption tests. Such models, some of which account for the estimation of effective surface area, are proposed by Weiland and Ahlgreen (1993), Billet and Shultes (1993), Nawrocki, *et al.* (1994), de Brito, *et al.* (1994), and recently by Hanley, *et al.* (1994) who related hydrodynamics with mass transfer using the percolation theory.

Many of these models, however, are expressed in terms of either the Sherwood number, which is based on total mass flux, or in terms of HETP. In our previous study (Ito and Asano, 1982) on binary distillation in a wetted-wall column, correlations for vapor-phase diffusion fluxes, corrected for partial condensation and evaporation, were proposed and applied in the simulation of the separation performance of ternary distillation (Kosuge and Asano, 1984). This study was extended to binary and ternary distillation using random packings (Kosuge, *et al.*, 1990; 1991) where new correlations considered the effect of condensation of mixed vapors under high mass flux conditions.

The purpose of this paper is to apply these concepts to structured packing of the wire-gauze type to develop mass transfer correlations based on binary distillation data under total reflux conditions using a short column with single element structured packing. The model, together with the effects of sensible heat flux and external heat flux, is compared with experimental data using longer columns for binary and ternary distillation.

* Received on May 23, 1995. Correspondence concerning this article should be addressed to K. Asano.

Table 1 Ranges of Variables

System	Concentration	N	U_G [m/s]	Re_G [-]	Re_L [-]
Acetone-Ethanol		1	0.057-0.72	35-340	0.10-1.0
	x_{Ab}: 0.02-0.35	2	0.10-0.60	60-300	0.30-0.9
		3	0.10-0.60	65-300	0.30-1.0
Acetone-Methanol-Ethanol	x_{Ab}: 0.02-0.50	1	0.06-0.36	35-220	0.15-1.0
	x_{Bb}: 0.15-0.50	2	0.09-0.50	50-250	0.20-1.1
	x_{Cb}: 0.24-0.60	3	0.10-0.20	50-100	0.20-0.4

1. Experimental Apparatus and Procedure

1.1 Experimental Apparatus

The set-up used in the experiments was similar to those described previously (Kosuge, *et al.*, 1990) except that the test section was replaced with a laboratory-size Sulzer EX wire gauze structured packing made of stainless steel. The diameter of the packing is 19.5 mm and a single element is 55-mm long with an area of 1700 m^2/m^3. The element was fitted inside a 20-mm inner diameter glass tube. The column was equipped with a conical distributor made of closely knit multilayered wire mesh to direct the liquid reflux near the center of the packing. The bottom of the element was provided with a guide to direct all bottom reflux to the sampling reservoir connected to a reflux flow meter. The flow rate of the vapor was adjusted by means of a variable transformer connected to the heaters in the reboiler. The system was properly insulated and tape heaters were installed that can be regulated to minimize heat losses and prevent condensation along the vapor lines.

A single element was used for binary and ternary distillation runs, while two and three elements were used for both binary and ternary simulation studies. A maximum of three elements was used based on the standard 8-to-1 ratio of height-to-diameter to avoid the installation of redistributors. When two or three elements were assembled in the test section, each packing element was axially oriented 90 degrees with respect to the adjacent element.

Acetone-ethanol and acetone-methanol-ethanol systems were used for binary and ternary studies, respectively.

1.2 Measurements

The flow rates of the liquid at the top and bottom of the column were measured by precision rotameters. Temperatures at strategic points in the set-up were measured using 1-mm outer diameter sheathed chromel-alumel thermocouples. Concentrations were determined using a Shimadzu gas chromatograph of the FID type.

The average vapor phase mass fluxes in the test section $\overline{N_i}$ were calculated using the top and bottom flow rates L_t, L_b and the corresponding mass fractions of the components and are given by

$$\overline{N_i} = \frac{L_t \omega_{Lit} - L_b \omega_{Lib}}{a_t V_p} \tag{1}$$

where a_t is the total surface area per unit volume of the gauze packing assumed to be equal to the wetted or effective area. Total wetting of the packing was observed visually during the experiment due to the excellent wetting

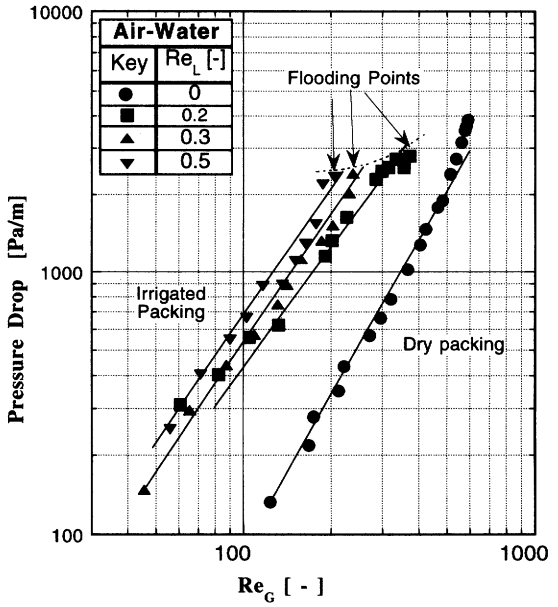

Fig. 1 Dry and irrigated pressure drop and flooding points for structured packing with air-water system

characteristics of the gauze sheets. This is in agreement with the observation of Bravo, *et al.* (1985).

The experimental convective flux $\overline{\rho_{Gs} v_s}$ and the average vapor-phase diffusion fluxes $\overline{J_{is}}$ were calculated by

$$\overline{\rho_{Gs} v_s} = \sum_{i=1}^{n} \overline{N_i} \tag{2}$$

$$\overline{J_{is}} = \overline{N_i} - \left(\sum_{i=1}^{n} \overline{N_i} \right) \omega_{Gis} \tag{3}$$

The average Sherwood number $\overline{Sh_{Gi}}$, which is normally defined as

$$\overline{Sh_{Gi}} = \overline{N_i} d_{pe} / \rho_{Gs} D_{ims} \overline{\Delta\omega_{Gis}} \tag{4}$$

is expressed in terms of the dimensionless diffusion flux by

$$\overline{Sh_{Gi}} \left(\frac{\overline{J_{is}}}{\overline{N_i}} \right) = \frac{\overline{J_{is}} d_{pe}}{\rho_{Gs} D_{ims} \overline{\Delta\omega_{Gis}}} \tag{5}$$

This in effect makes the Sherwood number independent of the mass flux and for binary system this becomes identical for both components. The $\overline{\Delta\omega_{Gis}}$ is calculated as the average of top and bottom concentration driving forces with the driving force expressed as $(\omega^*_{Gi} - \omega_{Gi})$.

1.3 Vapor-Liquid Equilibria and Physical Properties

The vapor pressures of pure components are determined by the Antoine equation while the activity coefficients of the liquid mixtures are evaluated using Wilson's equation. Antoine's constants, molar volume coefficients and Wilson's parameters are taken from Holmes and Van Winkle (1970). The viscosity and thermal conductivity of pure gas are obtained using the Hirschfelder equation and for mixtures, Wilke's method is applied (Reid, *et al.*, 1987). The binary diffusion coefficient of the vapor is estimated from the Hirschfelder equation

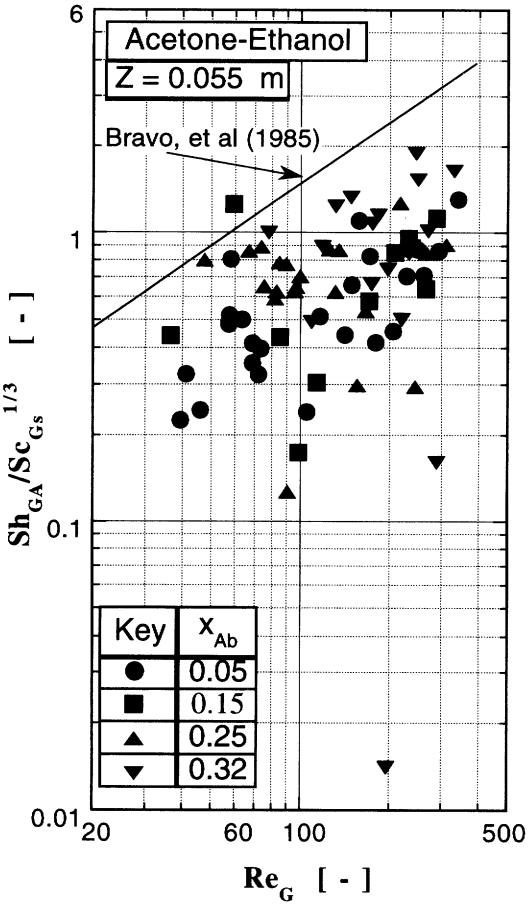

Fig. 2 Observed mass flux for acetone-ethanol system in a gauze-type structured packing compared with Bravo, *et al.* (1985) correlation

while Wilke's effective diffusion coefficient is used for the ternary system.

1.4 Ranges of Variables

Total reflux runs under atmospheric conditions were made for a single element, or two, or three elements for the acetone-ethanol system and the acetone-methanol-ethanol system at different flow rates and bottoms concentrations. The ranges of variables applied in the test runs are summarized in **Table 1**.

2. Experimental Results

2.1. Pressure Drop

The efficiency of the packing is very much affected by the hydrodynamic behavior of the liquid and vapor. We decided to measure the pressure drop directly for the actual set-up using an air-water system to define the operating range of vapor and liquid flow rates to avoid flooding conditions in our experiments. The results of the test both for irrigated and dry packing are shown in **Fig. 1**. The tests indicate that even at a liquid flow rate equivalent to Re_L = 0.2, the column tends to flood at Re_G of about 400 with a pressure drop $\Delta P/Z$ of 3 kPa/m. This pressure drop at flood point agrees closely with the experimental data for cyclohexane/n-heptane at 1.00 atm for Sulzer BX packing as

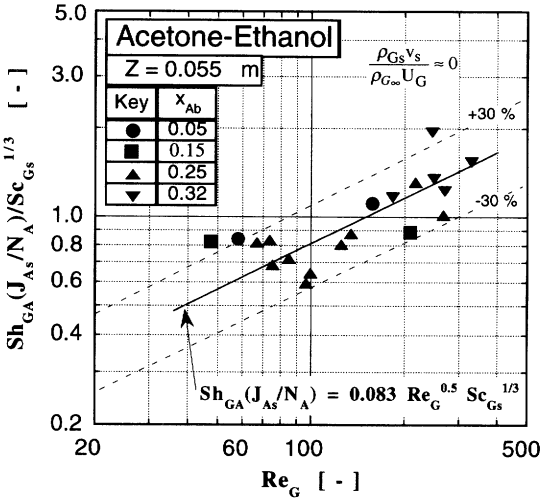

Fig. 3 Diffusion flux under low mass flux conditions for acetone-ethanol system

reported by Fair and Bravo (1990).

2.2 Mass Transfer in Binary Distillation

The effective vapor velocity in the packed column and the equivalent diameter of the packing are estimated using the equations recommended by Bravo *et al.* (1985) for gauze-type structured packings,

$$U_{Ge} = \frac{U_G}{\varepsilon \sin \alpha} \tag{6}$$

$$d_{pe} = Bh\left[\frac{1}{B+2S} + \frac{1}{2S}\right] \tag{7}$$

On the basis of our previous studies for wetted-wall columns (Ito and Asano, 1982), it was shown that the liquid-phase resistance is very small. Since the order of magnitude of the liquid-phase resistance is expected to be independent of the geometry of the column, in this study, only the gas-phase resistance is considered. The experimental values of the mass flux as evaluated from Eq. (1), together with the calculated values of the physical properties, are expressed in terms of the Sherwood number and plotted as $\overline{Sh}_{GA}/Sc^{1/3}$ versus the average gas phase Reynolds number Re_G for various bottom concentrations. The results are shown in **Fig. 2** together with the correlation recommended by Bravo *et al.* (1985) for gauze-type structured packings. It is observed that the points are highly scattered, suggesting the difficulty of obtaining a reliable correlation. To improve the correlation, $\overline{Sh}_{GA}$ is converted to a dimensionless diffusion flux as given by Eq. (5) and data points at low mass flux conditions, that is, $\overline{\rho_{Gs} v_s} / \rho_{G\infty} U_{G\infty} \cong 0$ (or $|\overline{\rho_{Gs} v_s} / \rho_{G\infty} U_{G\infty}| < 0.0005$) are chosen and plotted using similar coordinates. These data, as shown in **Fig. 3**, are correlated by

$$\left[\overline{Sh}_{GA}\left(\frac{J_{As}}{N_A}\right)\right]_0 = 0.083\, Re_G^{0.5}\, Sc_{Gs}^{1/3} \tag{8}$$

The effect of condensation of mixed vapors or evaporation from liquids is determined next by plotting G_v defined by

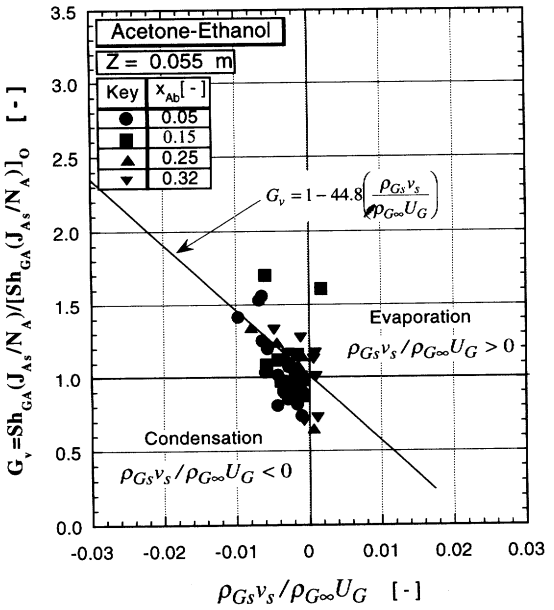

Fig. 4 Effect of condensation of mixed vapors or evaporation from liquids on diffusion flux for acetone-ethanol system using wire-gauze structured packing

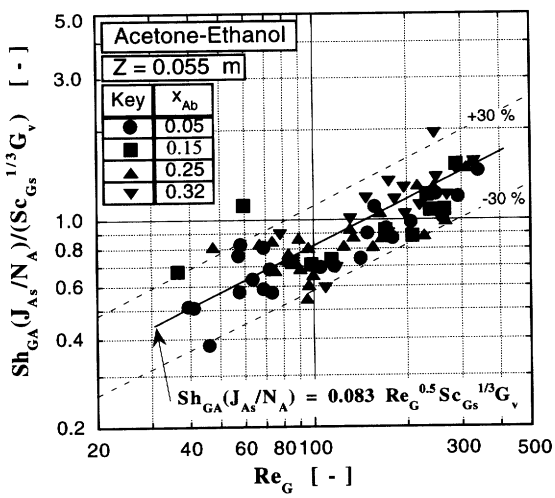

Fig. 5 Final correlation of dimensionless diffusion flux for acetone ethanol system with wire-gauze structured packing

$$G_v = \overline{Sh}_{GA}(\overline{J}_{As}/\overline{N}_A)/[\overline{Sh}_{GA}(\overline{J}_{As}/\overline{N}_A)]_0 \quad (9)$$

as a function of the dimensionless convective flux $\rho_{Gs}v_s/\rho_{G\infty}U_G$. The plot is shown in **Fig. 4.** Although the effect of G_v is not so high as indicated by the narrow range of $\rho_{Gs}v_s/\rho_{G\infty}U_G$, an increase in G_v with a decrease of $\rho_{Gs}v_s/\rho_{G\infty}U_G$ is observed. Applying the least squares method, the points are correlated by the equation,

$$G_v = 1 - 44.8\left(\frac{\rho_{Gs}v_s}{\rho_{G\infty}U_G}\right) \quad (10)$$

which is constrained by its physical meaning to pass through $G_v = 1$ at zero convective flux. This relationship is used to correct all data points and the resulting values are

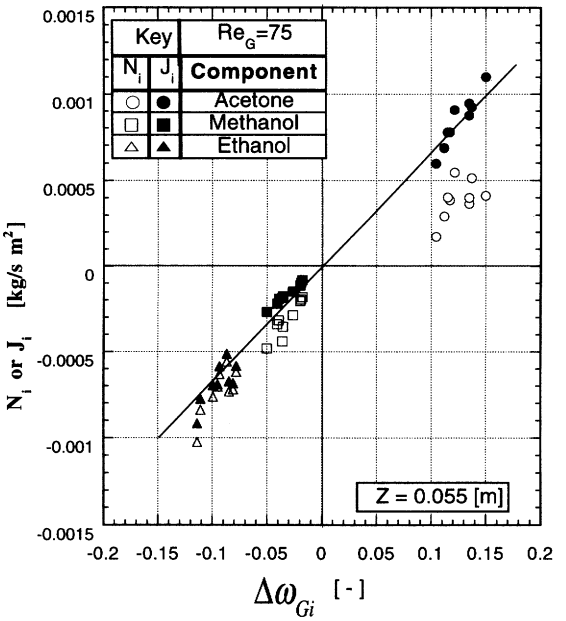

Fig. 6 Effect of average driving force on the vapor-phase mass and diffusion flux for a ternary system

plotted in **Fig. 5.** The final correlation obtained is given by

$$\overline{Sh}_{GA}\left(\frac{\overline{J}_{As}}{\overline{N}_A}\right) = 0.083 Re_G^{0.5} Sc_{Gs}^{1/3} G_V \quad (11)$$

Comparison of Figs. 2 and 5 indicates that a much improved correlation can be achieved by eliminating the effect of convective mass flux from the Sherwood number. The exponent of the Reynolds number of 0.5 suggests a flow pattern similar to that for wetted-wall columns when vapor and liquid are in laminar flow (Ito and Asano, 1982) as compared to 0.805 obtained for random packing as reported by Kosuge *et al.* (1990). This may be due to the lower range of Reynolds number considered, which is less than the flooding condition.

2.3 Mass Transfer in Ternary System

An identical approach is followed to determine whether the findings and procedures for binary systems will equally apply for ternary system using one, two, and three elements of structured packing.

To determine the influence of the convective flux on the individual components, **Fig. 6** is plotted using a single element. This shows the variation of the component mass flux $\overline{N}_i$ and the component diffusion flux $\overline{J}_{is}$ as a function of the average driving force at nearly the same Reynolds number of 75. The figure shows that a trend for the diffusion flux to be proportional to concentration driving force exists, while for mass flux, this trend is not well observed, especially for the most volatile component as indicated by more scattered points.

The binary mass transfer model is next compared with experimental data for ternary distillation using one, two, and three elements. The observed $\overline{Sh}_{Gi}$ is converted to average dimensionless diffusion flux $\overline{Sh}_{Gi}(\overline{J}_{is}/\overline{N}_i)$ and corrected for the effect of condensation of mixed vapors G_v, as defined by Eq. (10). The resulting plot for 1, 2, and

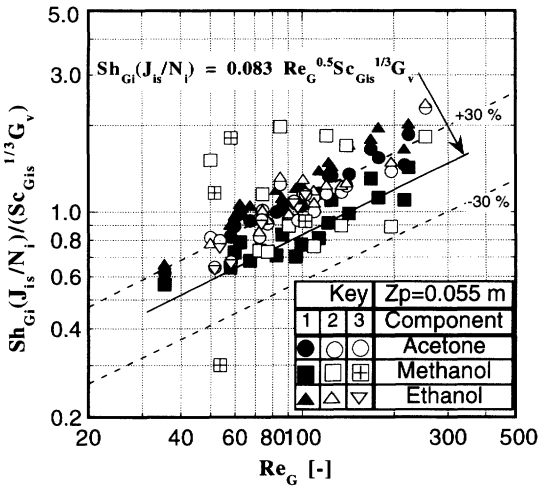

Fig. 7 Dimensionless diffusion flux for a ternary system using one, two, and three elements of packing as compared with correlation for binary system

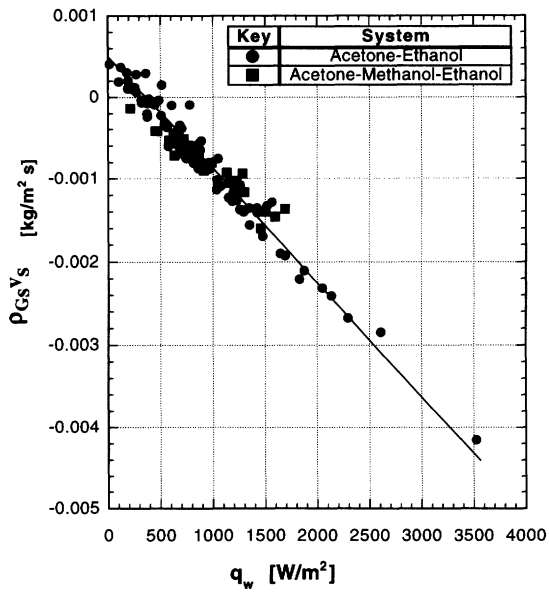

Fig. 8 Effect of wall heat flux on convective mass flux for binary and ternary systems

3 elements of packing is shown in **Fig. 7**. In general, the data for the three different lengths of packing show reasonable agreement with the binary correlation except for the intermediate component, methanol, which exhibits some scattering when longer columns are used. This suggests that, discounting the peculiar behavior of a component like methanol, the binary correlation may be applied equally well to ternary systems using a similar structured packing.

2.4. Convective Mass Flux

It is demonstrated above that for both binary and ternary systems, mass transfer in terms of mass flux in distillation, where heat effects can not be avoided, is an important factor in determining mass transfer rate. **Figure 8** shows the effect of the wall heat flux, q_w on the convective mass flux. The trend followed by both binary and ternary systems is similar. That is, as heat lost through the walls of the column increases, the convective flux moving towards the interface from the vapor phase also increases. Even under adiabatic conditions ($q_w = 0$), there is some amount of convective flux occurring. This can be explained by referring to a previously derived equation (Kosuge and Asano, 1982) given by

$$\rho_{Gs}v_s = \frac{J_{As}(\lambda_B - \lambda_A) - q_{Gs} - q_w}{\lambda_A \omega_{GAs} + \lambda_B \omega_{GBs}} \tag{12}$$

for binary system. For large values of q_w, the effect of sensible heat and the difference in latent heat becomes negligible and $\rho_G v_s$ becomes proportional to $-q_w$ while at $q_w = 0$, the sensible heat and the difference in latent heats determine the convective flux.

Table 2 Equations Used in the Simulation

Mass Balance Equations:

$$\frac{4}{\pi d_P^2}\frac{dV}{dZ} = \frac{4}{\pi d_P^2}\frac{dL}{dZ} = \left(\sum_{i=1}^{n} N_i\right) a_t \tag{13}$$

$$\frac{4}{\pi d_P^2}\frac{d(V\omega_{Gi})}{dZ} = \frac{4}{\pi d_P^2}\frac{d(L\omega_{Li})}{dZ} = N_i a_t \tag{14}$$

Mass Fluxes

$$N_i = J_{is} + \rho_{Gs} v_s \omega_{Gis} \tag{15}$$

Interfacial Velocity (ternary system)

$$\rho_{Gs}v_s = \frac{J_{As}(\lambda_C - \lambda_A) + J_{Bs}(\lambda_C - \lambda_B) - q_{Gs} - q_w}{\sum_{i=A}^{C} \lambda_i \omega_{Gis}} \tag{16}$$

Effect of condensation of Mixed Vapors

$$G_v = 1 - 44.8\left(\frac{\rho_{Gs}v_s}{\rho_{G\infty}U_G}\right) \tag{10}$$

Local Vapor-Phase Diffusion Flux

$$Sh_{Gi}\left(\frac{J_{is}}{N_i}\right) = 0.083\, Re_G^{0.5}\, Sc_{Gs}^{1/3}\, G_V \tag{17}$$

$$Sh_{Gi} = \frac{N_i d_{pe}}{\rho_{Gs} D_{Gim} \Delta\omega_{Gis}} \tag{18}$$

Local Vapor-Phase Sensible Heat Flux

$$Nu_{Gs} = 0.083\, Re_G^{0.5}\, Pr_{Gs}^{1/3}\, G_v \tag{19}$$

$$Nu_{Gs} = \frac{q_{Gs} d_{pe}}{k_{Gs}(T_s - T_{G\infty})} \tag{20}$$

3. Discussions

3.1 Simulations

The correlations derived for binary systems for a short structured packing are applied to simulate longer columns containing two and three elements for binary and ternary distillation. The equations applied in the simulation are shown in **Table 2**. Since the mass transfer correlation, as given by Eq. (11), is for the average diffusion flux of a short column, it is assumed in this simulation study to be equivalent to the local diffusion flux. The correlation for

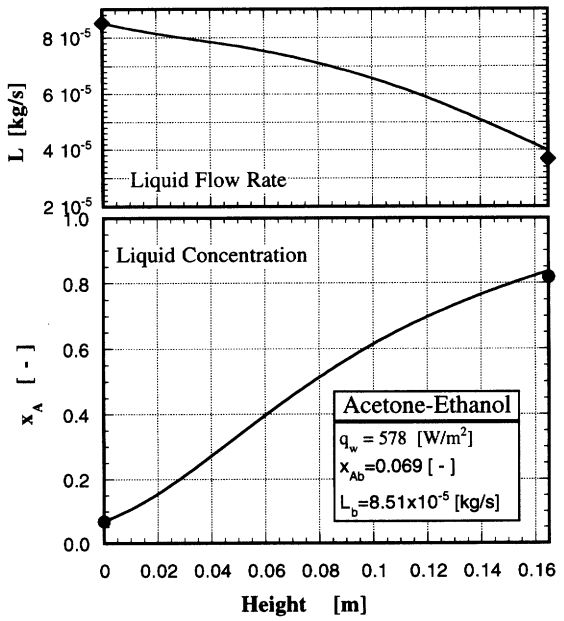

Fig. 9 Comparison amoung observed bottom and top flow rates and concentrations with simulated results using three elements of structured packing

the Nusselt number to determine sensible heat flux is derived based on the accepted principle that the mechanism of mass and heat transfer is identical. In the simulation, just like in the experimental analysis, the mass and heat transfer areas are taken as equal to the area of the packing. The other assumptions are: liquid-phase resistance is negligible (Ito and Asano, 1982); no concentration gradients in the radial direction; the wall heat flux is uniformly distributed along the length of the column; and the effect of pressure drop is negligible. The algorithm and the solution of the mass and energy balance equations followed is identical to those given previously for binary systems using random packings (Olaño *et al.*, 1994).

Figure 9 shows the comparison between the calculated liquid phase concentration and flow rate profiles for a binary system with the experimental values measured at the bottom and top of the column containing three elements. It is seen that there is good agreement in both concentration and reflux flow rate indicating that the correlation for short packing and the assumption of a fully wetted surface apply equally for three elements. The concentrations between elements were not measured due to sampling difficulties.

For ternary system utilizing two or three elements of packing, the calculated top flow rates and top concentrations are compared with the observed ones of corresponding bottom conditions as shown in **Fig. 10a** and **Fig. 10b**. There is acceptable agreement among the top flow rates both for two and three elements. For the comparison of top concentrations, acceptable agreement is also observed. For two elements, much better agreement is obtained when starting at higher bottoms concentration, while for three elements, methanol exhibits the largest deviation.

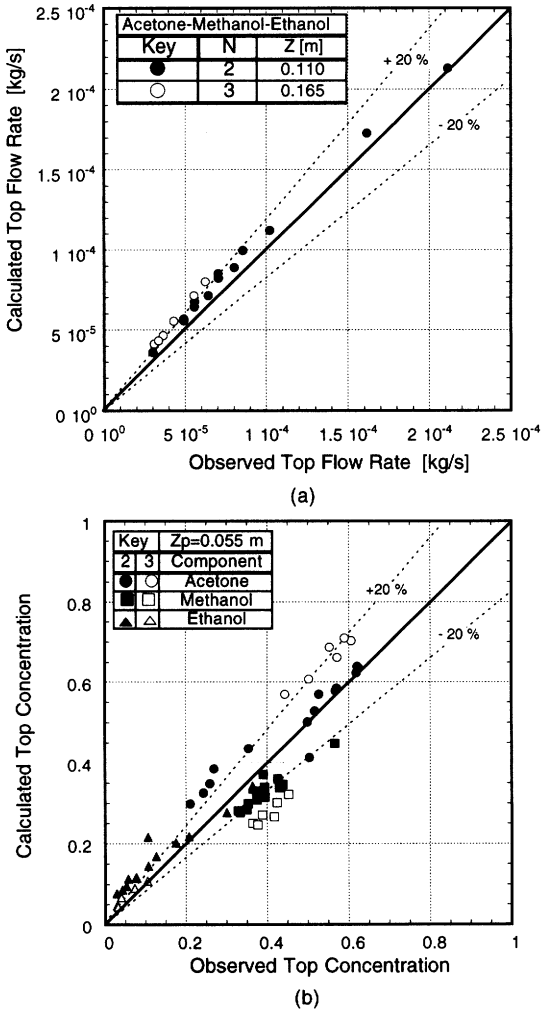

Fig. 10 Comparison between observed and calculated (a) top flow rates and (b) top concentrations for 2 and 3 elements of structured packing

3.2 Height of a Transfer Unit

The use of HTU has long been accepted as a convenient method for the design of packed distillation columns. To investigate the effect of certain operating conditions on the overall HTU for structured packings, the HTU for a binary system is plotted against Re_G for one, two, and three elements of packing at nearly the same initial bottoms concentration. As shown in **Fig. 11**, the trend is for HTU to increase with Re_G. However, the points are scattered with respect to the length of the packed column. This also shows that HTU is affected by the height and therefore, it appears that it is not a suitable parameter for packed distillation design purposes, especially under high mass flux conditions where the difference in mass flux and diffusion flux is not negligible.

For a ternary system, **Fig. 12** shows the behavior of the component HTUs with Re_G for three lengths of packing. In general, just as for a binary system, the tendency is for the HTU to increase with Re_G. It is to be noted that for a single element, no untoward scattering is obtained, unlike those exhibited by methanol for two and three elements. In our previous paper (Kosuge *et al.*, 1991), it was shown that

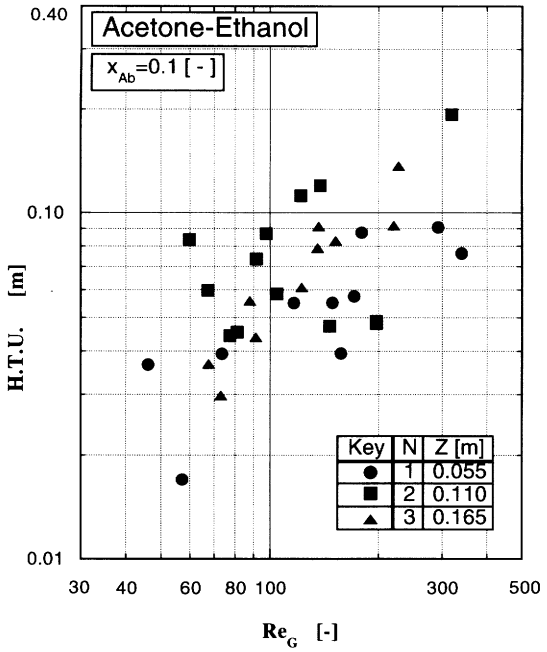

Fig. 11 Comparison of H.T.U.s for various heights of structured packing at nearly same bottoms concentration for acetone-ethanol system

concentration curves existed where methanol encounters zero concentration driving force. This is represented by curves MN and OP in the inset of Fig. 12. If the experimental concentration path passes through this curve, discontinuity in NTU is encountered and the HTU based on methanol becomes indeterminate. Figure 12 indicates that such a condition was not encountered for a single element since the initial and final concentrations are within these boundaries, that is, inside the area OPMN. However, for longer columns such a situation is difficult to avoid, generating an average HTU for methanol, calculated based on top and bottom conditions, that is meaningless and misleading. It must be reiterated, however, that the simulation technique followed is not affected by the presence of these zero-driving-force curves since local diffusion fluxes, not HTUs, are used for all components.

Conclusion

Mass transfer correlations for wire gauze-type structured packing have been derived from binary distillation data based on the total area of the packing and flow rates below flooding conditions. The correlations are highly improved if a dimensionless diffusion flux is used since it accounts for the effects of heat fluxes on the convective mass flux. The binary correlations obtained under low mass flux conditions are also applicable for ternary systems. Simulation studies showed that the mass transfer model can predict with acceptable results the distillation performance of longer columns with the same packing size under nearly similar conditions. The over all HTUs are shown to be affected by column height.

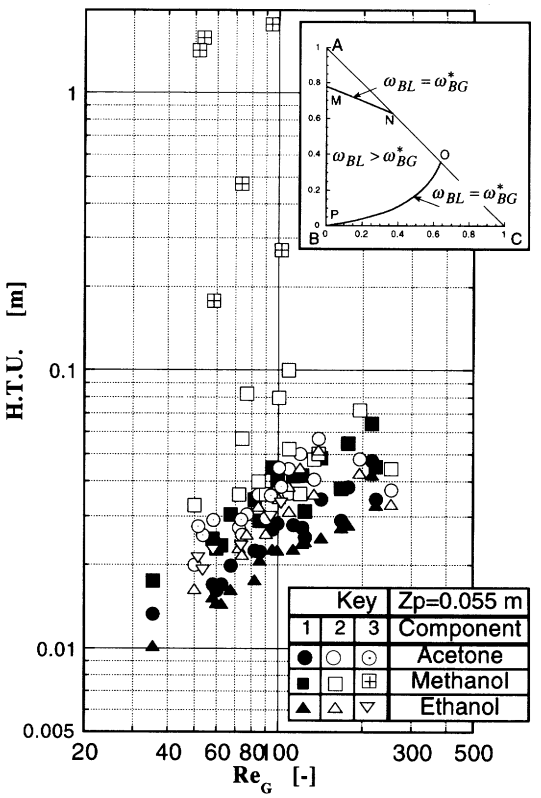

Fig. 12 Comparison of H.T.U.s for ternary system using one, two and three elements of structured packing

Nomenclature

a_t	=	total packing surface	[m^2/m^3]
B	=	channel base of packing corrugation	[m]
c_p	=	specific heat at constant pressure	[J/(kg·K)]
D_{im}	=	diffusion coefficient	[m^2/s]
d_p	=	diameter of packing	[m]
d_{pe}	=	equivalent diameter of packing defined by Eq. (7)	[m]
G_v	=	correction for mixed vapor condensation	[-]
h	=	crimp height	[m]
J_i	=	diffusion flux of component i	[kg/(m^2·s)]
L	=	mass flow rate of liquid	[kg/s]
N	=	number of packing elements	[-]
n	=	number of components	[-]
N_i	=	mass flux of component i	[kg/(m^2·s)]
Nu_G	=	vapor-phase Nusselt number $\{= q_G d_{pe}/k_G (T_s - T_{G\infty})\}$	[-]
P	=	pressure	[Pa]
Pr_G	=	vapor-phase Prandtl number $\{= c_{pG}\mu_G/\kappa_G\}$	[-]
q_G	=	vapor-phase sensible heat flux	[W/m^2]
q_w	=	wall heat flux	[W/m^2]
Re_G	=	vapor-phase Reynolds number $\{= \rho_{G\infty} U_{Ge} d_{pe}/\mu_{G\infty}\}$	[-]
Re_L	=	liquid-phase Reynolds number $\{= \rho_L U_L/\mu_L a_t\}$	[-]
S	=	length of corrugation side	[m]
Sc_{Gi}	=	Schmidt number of component i $\{= \mu_G/\rho_{Gi} D_{Gim}\}$	[-]
Sh_{Gi}	=	Sherwood number defined by Eq. 4	[-]
T	=	temperature	[K]
U_G	=	vapor-phase superficial velocity	[m/s]
U_L	=	liquid-phase superficial velocity	[m/s]
V	=	mass flow rate of vapor	[kg/s]
V_p	=	volume of packing	[m^3]
v	=	normal component of interfacial velocity	[m/s]

x	=	liquid phase mole fraction	[-]
y	=	vapor-phase mole fraction	[-]
Z	=	height of packing	[m]
Z_p	=	height of a single packing element	[m]

<greek symbols>

α	=	angle of flow channel from horizontal	[deg]
ε	=	void fraction of packing	[-]
κ	=	thermal conductivity	[W/(m·K)]
Δ_ω	=	mass fraction concentration driving force	[-]
λ	=	latent heat of vaporization	[J/kg]
μ	=	viscosity	[Pa·s]
ω	=	mass fraction	[-]
ρ	=	density	[kg/m^3]

<subscripts>

A	=	more volatile component or acetone
B	=	less volatile component or ethanol for binary system or methanol for ternary system
C	=	ethanol for ternary system
G	=	gas` or vapor phase
L	=	liquid phase
b	=	bottom condition
i	=	component i
s	=	vapor-liquid interface
t	=	top condition
0	=	zero or very low mass flux condition
∞	=	vapor free-stream condition

<superscripts>

*	=	equilibrium condition
—	=	average value

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