

COMPARISON BETWEEN TWO DIFFERENT MODELS TO PREDICT INTERFACIAL MASS TRANSFER AREA IN DISTILLATION COLUMN WITH SIEVE TRAY

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ABSTRACT

The volumetric mass transfer rate in distillation with sieve tray column is affected by fluid behavior on the trays, which influences the mass transfer coefficients and the interfacial mass transfer area between vapor and liquid. Classically, vapor is released in the form of small bubbles to yield a large surface area and also an efficient mass transfer between vapor and liquid phases. To improve the mass transfer efficiency, the interfacial mass transfer area and mass transfer coefficient have to be controlled closely.

The aim of this work is to present the distillation process on a sieve plate laboratory column and carry out a comparison between two different models to predict the interfacial mass transfer area. The first one is based on the geometric characteristics of the bubbles like sauter mean diameter (d_s), bubble formation frequency (f_B), number of bubbles in the dispersion (N_B), bubble surface (S_B), height and length of a bubble. The second model assumes to predict the interfacial mass transfer area based on the classical penetration theory.

Key words: *distillation, mass transfer coefficient, tray column*

INTRODUCTION

The influence of plate geometry on the characteristics of fluid flow and mass transfer in a laboratory column was experimentally examined using different binary blends. The interfacial mass transfer area depends on the properties of the fluid, the hydrodynamic regime, and the configuration of the gas–liquid contacting device. The intensity of interfacial mass transfer is characterized by the volumetric mass transfer coefficient ($K_{OG}a$) and determines the amount of gas transferred from bubbles into the liquid phase. Bubble size is an important design parameter which has a strong influence on the hydrodynamic behaviour and on the volumetric mass transfer coefficients [3].

Prediction of interfacial area is an important part of gas–liquid contactor design. The individual terms in interfacial area are difficult to measure directly.

MATERIALS AND METHODS

For determination of interfacial mass transfer area in distillation laboratory column with one sieve plate were used two different ways.

In the first one interfacial mass transfer area is obtained by theoretical model, based on the geometric characteristics of the bubbles like sauter mean diameter, bubble formation frequency, number of bubbles in the dispersion, bubble surface, height and length of a bubble.

Under the examined operating conditions the theoretical model cannot be applied successfully for the sake of interfacial area (a) prediction since this model is explicitly valid only for rigid spherical bubbles. For all other bubble shapes (ellipsoidal in our case) some correction term is needed since the theoretically calculated (a) values are somewhat inflated and that is why some mitigation will reflect to a greater extent the reality. In the case of stripping of carbon dioxide from the aqueous solution with air, Miller has introduced the correction factor for the ellipsoidal shape of bubbles. [5]:

$$f_c = 683d_s^{1.376} \quad (1)$$

According to the above logic the interfacial mass transfer area it is depending on the bubble shape [6]:

$$a = f_c \frac{f_b \cdot S_b}{A u_b} \quad (2)$$

Where f_c is a correction factor

The surface area S_b of an ellipsoidal bubble can be calculated as follows [1]:

$$S_b = \pi \frac{l^2}{2} \left[1 + \left(\frac{h}{l} \right)^2 \frac{1}{2e} \ln \left(\frac{1+e}{1-e} \right) \right] \quad (3)$$

Where, the eccentricity e is

$$e = \sqrt{1 - \left(\frac{h}{l} \right)^2} \quad (4)$$

The bubble formation frequency f_b (number of bubbles formed at the orifice per unit of time) is expressed as [7]:

$$f_b = \frac{Q_G}{V_b} \quad (5)$$

The Sauter mean bubble diameter is estimated by means of Wilkinson correlation which is one of the most recommended in literature [9]:

$$\left(\frac{g \rho_L d_s^2}{\sigma} \right) = 8.8 \left(\frac{u_G \mu_L}{\sigma} \right)^{-0.04} \left(\frac{\sigma^3 \rho_L}{g \mu_L^4} \right)^{-0.12} \left(\frac{\rho_L}{\rho_G} \right)^{0.22} \quad (6)$$

Bubble shape, motion and any tendency for the interface to ripple, fluctuate or otherwise deform are all related to the bubble size. In turn, bubble size is determined by the physical characteristics of the system and operating conditions. Equation implies that the bubble size decreases with the increase of both superficial gas velocity and gas density [6].

The bubble diameter is needed also for the calculation of the bubble rise velocity [4]:

$$u_b = \sqrt{\frac{2\sigma}{\rho_L \cdot d_s} + \frac{g d_s}{2}} \quad (7)$$

This equation along with equation (6) was also used to calculate the bubble Reynolds number Re_b needed for estimation of both bubble length l and height h . Terasaka derived the following equations for calculating the ellipsoidal bubble length and height [8]:

$$l = \frac{d_s}{1.14 \cdot Ta^{-0.176}} \quad (8)$$

$$h = 1.3 \cdot d_s \cdot Ta^{-0.352} \quad (9)$$

The second model assumes to predict the interfacial mass transfer area based on the classical penetration theory. The method is based on comparison between experimentally obtained overall volumetric mass transfer coefficients of a vapor phase $(K_{OG}.a)^{exp}$, with the penetration model calculations of overall mass transfer coefficient $(K_{OG})^{cal}$.

$$a = \frac{(K_{OG} * a)^{exp}}{(K_{OG})^{cal}} \quad (10)$$

The overall volumetric gas-phase mass transfer coefficient $(K_{OG}.a)^{exp}$, is calculated from the following equation,

$$K_G a^{exp} = \frac{u_G \rho_G N_{OG}}{h_f M} \quad (11)$$

The classical penetration model put forward by Higbie is the most significant mass transfer theory. It explains successfully the governing mass transfer process in the liquid phase around the bubbles. It is argued that turbulence brings elements of bulk liquid to the interface, where unsteady mass transfer occurs at time after which the element is returned to the bulk liquid and is replaced by another one.

To predict the mass transfer coefficient has traditionally used the Whitman two-film theory approach in which the resistances to mass transfer on each side of the interface are described by mass transfer coefficient. Based on the two film theory, the following equation is obtained:

$$\frac{1}{(K_{OG})^{cal}} = \frac{1}{\beta_G} + \frac{m}{\beta_L} \quad (12)$$

According to Higbie's theory the vapor-side mass transfer coefficient and liquid-side mass transfer coefficient are dependent on both the diffusion coefficient and the contact time:

$$\beta_G = \sqrt{\frac{4 * D_G}{\pi * \theta_G}} \quad (13)$$

$$\beta_L = \sqrt{\frac{4 * D_L}{\pi * \theta_L}} \quad (14)$$

The molecular diffusion coefficient of vapor and liquid phase are calculated by follow equations:

$$D_G = \frac{4.3.10^{-7} * T^{\frac{3}{2}}}{p * \left(u_A^{\frac{1}{3}} + u_B^{\frac{1}{3}} \right)^2} * \sqrt{\frac{1}{M_A} + \frac{1}{M_B}} \quad (15)$$

$$D_L = \frac{K * M_B^{\frac{1}{2}} * T}{\mu_B * (V_A * V_B)^{\frac{1}{3}}} \quad (16)$$

In Higbie's original work the contact time of an element at the vapor-liquid interface is taken as the time for bubbles to rise through its own height. For sieve tray dispersion they assumed for vapor contact time that:

$$\theta_G = \frac{h_L}{u_G} \quad (17)$$

Similarly, the liquid contact time is

$$\theta_L = \theta_G * \frac{\rho_G}{\rho_L} \tag{18}$$

where h_L is clear liquid height. The clear liquid height on a sieve tray plays an important role in the mass transfer because of its influence on θ_G and θ_L . Nearly all tray performance correlation includes clear liquid height as a variable. Under total reflux conditions the correlation is given by [2];

$$h_L = 0.6 * h_w^{0.5} * p^{0.25} * \left[\left(\frac{\rho_G}{\rho_L} \right)^{0.5} / b \right]^{0.25} \tag{19}$$

EXPERIMENTAL RESULTS

Figure 1 shows the interfacial mass transfer area obtained by the theoretical and experimental models as a function of the sauter mean diameter, which is calculated by the equation (6). According to theoretical and experimental model over the whole bubble diameter range (3-3.16 mm), the interfacial mass transfer area increase with decreasing bubble diameter.

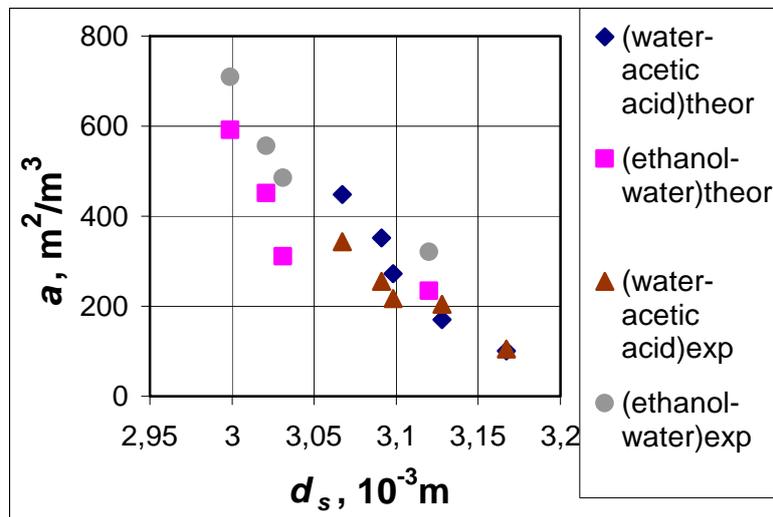


Fig. 1. Interfacial mass transfer area (a) as a function of the sauter mean diameter (d_s).

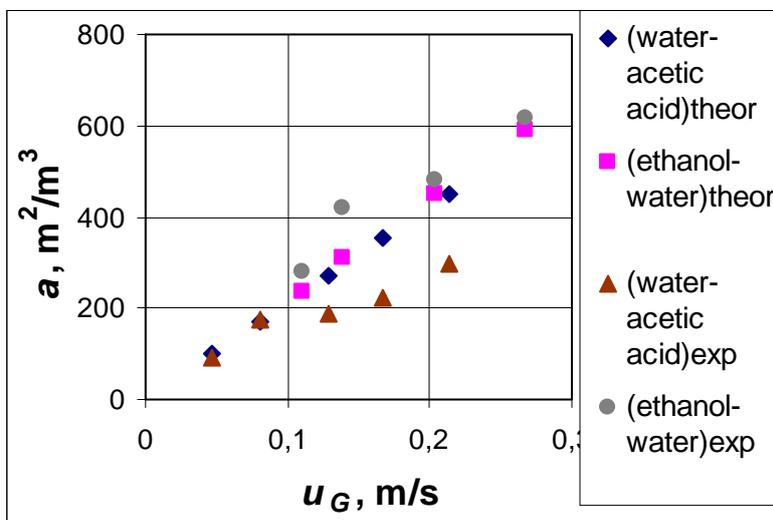


Fig. 2. Interfacial mass transfer area (a) versus superficial gas velocity u_G .

Figure 2 shows the comparison between interfacial mass transfer areas obtained by the theoretical model and by the experimental model as a function of the superficial gas velocity. As can be seen in Figure 2, a good agreement between two models was obtained within 20% of the total relative error for binary systems water-acetic acid and ethanol-water.

CONCLUSIONS

New models to predict the interfacial mass transfer area developed in this study. The classical penetration theory is applicable to predict interfacial mass transfer area and compare with theoretical one obtained for binary systems water-acetic acid and ethanol-water, in a laboratory column with one sieve tray under atmospheric pressure.

NOMENCLATURE

A	cross-section area of the column [m^2]
a	specific interfacial area [m^2/m^3]
d_s	sauter mean bubble diameter [m]
e	bubble eccentricity
f_b	cross-sectiona bubble formation frequency [s^{-1}]
f_c	correction factor
g	gravitational acceleration [$\text{m}\cdot\text{s}^{-1}$]
h_f	aerated liquid height [m]
h	height of an ellipsoidal bubble [m]
h_w	outlet weir height [m]
K_{OG}	gas-side mass transfer coefficient [$\text{m}\cdot\text{s}^{-1}$]
K_{OGa}	overall mass transfer coefficient [$\text{kmol}/\text{m}^3\cdot\text{s}$]
l	length of an ellipsoidal bubble [m]
M	molecular weight [kg/kmol]
m	slope of vapor-liquid equilibrium line
N_{OG}	number of overall vapour phase transfer units
Q_G	gas flow rate [$\text{m}^3\cdot\text{s}^{-1}$]
S_b	bubble surface [m^2]
u_G	superficial gas velocity [$\text{m}\cdot\text{s}^{-1}$]
u_b	bubble rise velocity [$\text{m}\cdot\text{s}^{-1}$]
μ_G	gas viscosity [Pa.s]
μ_L	liquid viscosity [Pa.s]
ρ_G	gas density [$\text{kg}\cdot\text{m}^{-3}$]
ρ_L	liquid density [$\text{kg}\cdot\text{m}^{-3}$]
σ	surface tension [$\text{N}\cdot\text{m}^{-1}$]
β_G	vapour-phase mass transfer unit
β_L	liquid-phase mass transfer unit
Tadaki number	$Ta = \text{Re}_b \cdot \text{Mo}^{0.23}$

REFERENCES

1. Fan, L. S., K. Tsuchiya, 1990. Bubble Wake Dynamics in Liquids and Liquid-Solid Suspensions, Butterworth-Heinemann Series in Chemical Engineering, Stoneham, U.S.A
2. Hofhuis, P. A. M., 1980. Thesis, TU Delft
3. Koichi, A., 2006. Mass Transfer from Fundamentals to Modern Industrial Application, Wiley-VCH, Weinheim

4. Mendelson, H. D., 1967. The prediction of bubble terminal velocities from wave theory, *AIChE J.*, 13, 250-253
5. Miller, D. N., 1974. Scale-up of Agitated vessels gas-liquid mass transfer, *AIChE J.*, 20, 445-453
6. Nedelchev, S., U. Jordan, 2006. A New Correction Factor for Theoretical Prediction of Mass Transfer Coefficients in Bubble Columns, *Journal of Chemical Engineering of Japan*, 39, 1237-1242
7. Painmanakul, P., Loubiere, K., 2004. Study of different membrane spargers used in waste water treatment: characterization and performance, *Chemical Engineering and Processing*, 1347-1359
8. Terasaka, K., Y. Inoue, M. Kakizaki and M. Niwa, 2004. Simultaneous Measurement of 3-Dimensional Shape and Behavior of Single Bubble in Liquid Using Laser Sensors, *Journal of Chemical Engineering of Japan*, 37, 921-926
9. Wilkinson, P. M., H. Haringa, 1994. Mass Transfer and Bubble size in a Bubble Column under Pressure, *Chem Eng Sci*, 49, 1417-1427

Acknowledgement

The authors would like to acknowledge for the financial support provided by Bulgarian Ministry of Education and Science, Fund "Scientific Investigations"