Gas-side mass transfer coefficient of a laboratory column equipped with one sieve tray

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Abstract: *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 volumetric gas-liquid mass transfer coefficient depends on the properties of the fluid, the hydrodynamic regime, and the configuration of the gas–liquid contacting device. Prediction of mass transfer coefficient is an important part of gas–liquid contactor design. The individual terms in volumetric mass transfer coefficients are difficult to measure directly.*

The aim of this work is experimental study of the kinetic coefficients in distillation of the binary blends in laboratory column at conditions near to model of ideal mixture for liquid phase and ideal displacement for vapour phase.

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 volumetric gas-liquid mass transfer coefficient 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 [1].

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MATERIALS AND METHODS

For direct measurement of overall point efficiency is to make use of the glass lab column is complicated-modification of Oldershaw with take outward overflows [2]. The column has one sieve plate with follow geometric characteristics: diameter 32mm, number of the openings-44, diameters of which is 1.1mm. The height of the overflow is 12mm.

The overall point efficiency, which takes into account effects of mass transfer on tray and in the settling zone, is defined as [3]:

$$E_{OG} = \frac{(y_n - y_{n-1})}{(y^* - y_{n-1})} \quad (1)$$

Once the point efficiency has been deduced from the foregoing relationships it can be re-cast in the form of transfer units:

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

Mass transfer effectiveness in gas–liquid contactors is most often expressed by means of the volumetric mass transfer coefficient ($K_{OG}a$). This may be correlated, for example, with power input per unit volume and gas superficial velocity, but the resulting correlations do not achieve any degree of generality. Too many phenomena contribute to the values of the film coefficient, k_{OG} and of the specific area a and their combined effect cannot easily be predicted. Separation of k_{OG} and a in the volumetric mass transfer coefficient is thus a first step for a better understanding of the underlying phenomena [4].

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

$$K_G a = \frac{u_G \rho_G N_{OG}}{h_f M} \qquad (3)$$

The $K_{OG}a$ value can be predicted if one knows how to estimate the vapour-side mass transfer coefficient K_{OG} and the interfacial area a, individually. The pure mass transfer coefficient is found through following depends:

$$K'_{OG} = \frac{K_{OG} \cdot M}{\rho_G}, \qquad (4)$$

Where:

$$K_{OG} = \frac{K_{OG} \cdot a}{a}, \qquad (5)$$

In order to calculate the volumetric gas-side mass transfer coefficient $K_{OG}a$, one also needs to know how to calculate the specific interfacial area, a. The formula for its calculation depends on the bubble shape [5]:

$$a = \frac{f_b . S_b}{A . u_b} \qquad (6)$$

The surface area S_b of an ellipsoidal bubble can be calculated as follows [6]:

$$S_b = \pi \frac{l^2}{2}\left[1 + \left(\frac{h}{l}\right)^2 \frac{1}{2e} \ln\frac{(1+e)}{(1-e)}\right] \qquad (7)$$

Where, the eccentricity e is

$$e = \sqrt{1 - \left(\frac{h}{l}\right)^2} \qquad (8)$$

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

$$f_b = \frac{Q_G}{V_b} \qquad (9)$$

The both basic parameters gas-side mass transfer coefficient $K_{OG}a$ and specific interfacial area a depend on the bubble diameter [7]:

$$d_S = 3 g^{-0.44} . \sigma^{0.34} . \mu_L^{0.22} . \rho_L^{-0.45} . \rho_G^{-0.11} . u_G^{-0.02} \qquad (10)$$

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 [5].

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

$$u_b = \sqrt{\frac{2\sigma}{\rho_L . d_S} + \frac{g d_S}{2}} \qquad (11)$$

This equation along with equation (10) 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]:

$$l = \frac{d_S}{1.14.Ta^{-0.176}} \qquad (12)$$

$$h = 1.3.d_S.Ta^{-0.352} \qquad (13)$$

The bubble rise velocity u_b and both the bubble length l and height h of an ellipsoidal bubble take part in the calculation of the rate of surface formation R_{sf}:

$$R_{sf} = \pi\sqrt{\frac{l^2+h^2}{2}+\frac{(l-h)^2}{8}}u_b \qquad (14)$$

Under the examined operating conditions the classical penetration theory cannot be applied successfully for the sake of (K_{OG}) 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 (K_{OG}) 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, has introduced the correction factor to the classical penetration theory to account for the ellipsoidal shape of bubbles. [9]:

$$K_{OG} = f_C\sqrt{\frac{4D_L R_{sf}}{\pi S_b}} \qquad (15)$$

Where the correction factor is

$$f_C = 683 d_S^{1.376} \qquad (16)$$

EXPERIMENTAL RESULTS

Figure 1 shows the profile of the gas-side mass transfer coefficient K_{OG} as a function of the superficial gas velocity obtained by experimental and theoretical methods. It is to be noted that the mixture Methanol-Ethanol the maximum difference between calculated and experimentally obtained Higbie values of the coefficients is no more than 14-15%, for mixture of Propanol-Water this difference is even smaller than 7%. This can be seen even better in Figure 2, where this comparison is made.

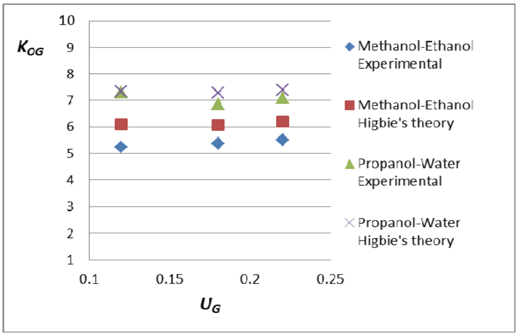

Fig.1. Effect of gas velocity and gas-side mass transfer coefficient

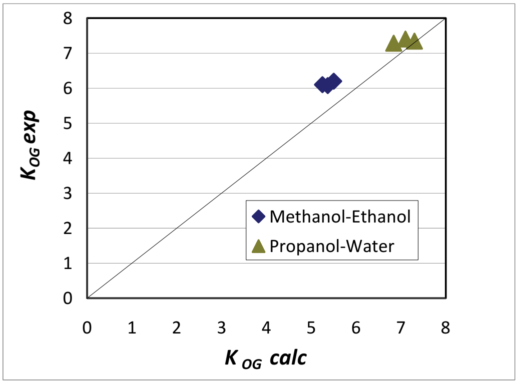

Fig.2. Comparison between experimental K'_{OG} values and calculated K_{OG} values

Figure 2 shows that the theoretical K_{OG} calculated by equation (15) values are in reasonable agreement with the experimental results equation (4).

CONCLUSIONS

The classical penetration theory is applicable for predicting gas-side mass transfer coefficient measured in methanol-ethanol and propanol-water, in a laboratory column with one sieve tray under atmospheric pressure.

NOMENCLATURE

A	cross-sectional area of the column, [m^2]
A	specific interfacial area, [m^2/m^3]
d_s	sauter mean bubble diameter, [m]
E_{OG}	overall point efficiency, [%]
e	bubble eccentricity
f_b	bubble formation frequency, [s^{-1}]
f_c	correction factor
g	gravitational acceleration, [m.s^{-1}]
h_f	aerated liquid height, [m]
h	height of an ellipsoidal bubble, [m]
K_{OG}	gas-side mass transfer coefficient, [m.s^{-1}]
$K_{OG}a$	overal mass transfer coefficient, [kmol/m^3.s]
l	length of an ellipsoidal bubble, [m]
M	molecular weight, [kg/kmol]
N_{OG}	number of overall vapour phase transfer units
R_{sf}	surface formation, [m^2.s^{-1}]
S_b	bubble surface, [m^2]
u_G	superficial gas velocity, [m.s^{-1}]
u_b	bubble rise velocity, [m.s^{-1}]
μ_G	gas viscosity, [Pa.s]
μ_L	liquid viscosity, [Pa.s]
ρ_G	gas density, [kg.m^{-3}]
ρ_L	liquid density, [kg.m^{-3}]
σ	surface tension, [N.m^{-1}]

Morton number $Mo = \frac{g.\mu_l^4}{\rho_l.\sigma^3}$

Bubble Reynolds number $\mathrm{Re}_b = \frac{d_S.u_b.\rho_l}{\mu_l}$

Tadaki number $Ta = \mathrm{Re}_b.Mo^{0.23}$

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