

## Specific interfacial mass transfer area in distillation process

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**Abstract:** *The occurrence of bubbles and drops depends on the two-phase flow pattern or flow regime which develops on the tray. The two extremes are a liquid-continuous flow pattern: where gas is present as bubbles (free-bubbling regime) and a vapour-continuous flow pattern: where vapour is present as drops (spay regime). Between these extremes a large transition regime exists where both bubbles and drops are present (mixed-froth regime). The mixed froth regime, which is the mostly occurent regime in industrial sieve tray columns, is also not well studied. Finally little thought has been given to the significance of the regime concept in relation to tray performance characteristics like vapour velocity, interfacial area and mass transfer coefficient. The aim of the present study in the clarification of some of the above mentioned problems, in particular the prediction interfacial area of the mixed-froth regime.*

**Key words:** *interfacial area, mass transfer, tray column*

### INTRODUCTION

Mixed-froth regime is a regime with dynamic interfacial mass transfer area mostly occurent on sieve trays, which is difficult to define. This intermediate regime tends towards bubbling near the tray floor and towards spray near the top of the dispersion. The extremes are a liquid-continuous flow pattern: where gas is present as bubbles (free-bubbling regime) and a vapour-continuous flow pattern: where gas is present as drops (spay regime). Between these extremes a large transition regime exists where both bubbles and drops are present (mixed-froth regime). The complexity of the bubbling liquid flow on a sieve tray is attributed to a large amount of bubbles and their interactions in turbulent dispersions. The stochastic behaviors of turbulence cause many difficulties for the theoretical prediction of interfacial area. Therefore the mixed froth regime, which is the mostly accurent regime in industrial sieve tray columns is not well understood. The bubbling regime also has yet received only little attention. Finally little thought has been given to the significance of the regime concept in relation to tray performance characteristics like vapour velocity, interfacial area and mass transfer coefficient. The specific interfacial area depends on the properties of the fluid, the hydrodynamic regime, and the configuration of the gas-liquid contacting device. The aim of the present study is to propose a simple theoretical clarification of some of the above mentioned problems, in particular prediction the specific interfacial area of the mixed-froth regime [1].

### MATERIALS AND METHODS

The experiments are carried out in a quartz glass column with diameter 0.1 m that can be equipped with tree sieve trays having several fractional perforated areas with a clear section 4.66%. The number of holes in the plate is 52 with a diameter of 3mm, as the overflow limit of each plate is 15mm. The height of the gas-liquid layer on the plate is determined visually, the hydraulic resistance of the plate through the U - tube manometer. [2]. The determination was conducted within the speed range of tray orifice from 3,59 to 8,47 m/s, under atmospheric pressure and full reflux.

The object of this work is to predict values of specific interfacial area theoretically and to compare them with the experimental ones. The formula for these theoretical calculations depends on the bubble geometry. The specific interfacial area  $a$  is a function of the bubble formation frequency  $f_b$ , the bubble surface area  $S_b$ , cross-section area of the tray  $A$ , and the bubble rise velocity  $u_b$  [3]:

$$a = \frac{f_b \cdot S_b}{A \cdot u_b} \quad (1)$$

The specific interfacial area  $a$  depends on the bubble diameter [4]:

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

Equation (2) implies that the bubble size decreases with the increase of both superficial gas velocity and gas density [5].

For the calculation of the specific interfacial area we need also to calculate the surface area  $S_b$  of an ellipsoidal bubble and the bubble formation frequency  $f_b$  [6]:

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

Where, the eccentricity  $e$  determines the amount by which an ellipsoidal bubble deviates from a perfect sphere:

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

This equation along with equation (2) 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 [7]:

$$l = \frac{d_s}{1.14.Ta^{-0.176}} \quad (5)$$

$$h = 1.3.d_s.Ta^{-0.352} \quad (6)$$

The specific interfacial mass transfer area was experimentally defined according equation (9). According to this method the specific interfacial area is function of vapour velocity  $u_G$ , local point efficiencies  $E_{OG}$ , froth height  $h_f$  and the gas-side mass transfer coefficient  $K_{OG}$ . The parameters  $u_G$ ,  $E_{OG}$  and  $h_f$ , are experimentally determined, and  $K_{OG}$  is calculated according to two-film model for mass transfer [8].

$$a = \frac{u_G \left[ -\ln(1 - E_{OG}) \right]}{h_f K_{OG}} \quad (7)$$

The theoretical relationship for tray efficiencies is commonly based on the two-film model for mass transfer. This model gives the vapour-phase point efficiency  $E_{OG}$  and describes the degree of approach of the thermodynamically vapour-liquid equilibrium on local point on the tray.

The calculating of gas-phased mass transfer coefficient  $K_{OG}$  is based on two-film theory and penetration model:

$$\frac{l}{K_{OG}} = \frac{l}{\beta_G} + \frac{m}{\beta_L} \quad (8)$$

According to Higbie's penetration model, for explaining the mass transfer through the free surface, it is assumed that the turbulence brings elements of bulk liquid to the free surface. In the short time of contact, the molecules of the bulk liquid are transferred to another phase by unsteady mass transfer, also called penetration. After the time of contact the element is returned to the bulk liquid and is replaced by another one [9].

The liquid and vapor mass transfer coefficients are obtained by the following equations, for same time of contact [10]:

$$\beta_G = \sqrt{\frac{4D_G}{\pi\theta_G}} \quad (9)$$

$$\beta_L = \sqrt{\frac{4D_L}{\pi\theta_L}} \quad (10)$$

It is impossible to measure the time of contact for vapor phase,  $\theta_G$ , that's why it is estimated by assuming that the time of contact of the vapor is the time taken by the vapor phase to pass through the gas-liquid layer and it can be expressed as follows:

$$\theta_G = \frac{h_L}{u_G} \quad (11)$$

The contact time for the liquid phase is:

$$\theta_L = \theta_G \frac{\rho_G}{\rho_L} \quad (12)$$

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  $\theta_G$  and  $\theta_L$ . Nearly all tray performance correlation includes clear liquid height as a variable. Under total reflux conditions the correlation is given by [6];

$$h_L = 0.6 h_W^{0.5} P^{0.25} \left[ \left( \frac{\rho_G}{\rho_L} \right)^{0.5} / b \right]^{0.25} \quad (13)$$

### EXPERIMENTAL RESULTS

Figure 1 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 expected, for the binary mixture Methanol – Water the specific interfacial mass transfer area is increasing with increment the gas velocity.

Figure 2 shows that the theoretical specific interfacial area  $a_{calc}$  calculated by equation (1) values are in reasonable agreement with the experimental results equation (7). As can be seen in Figure 2, a good agreement between two models was obtained within 20% of the total relative error for binary system Methanol-Water.

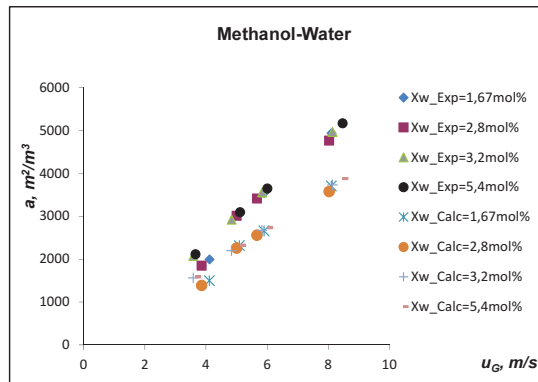


Fig.1. Effect of gas velocity  $u_G$  on the specific interfacial area  $a$

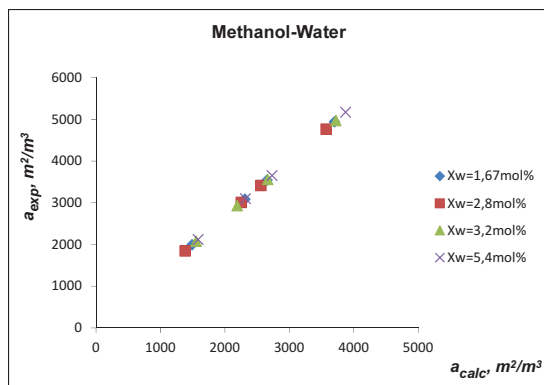


Fig.2. Comparison between specific interfacial area experimental values  $a_{exp}$  and calculated values  $a_{calc}$

### CONCLUSIONS

The classical penetration theory is applicable for predicting specific interfacial area measured in methanol-water, in a laboratory column with tree sieve trays under atmospheric pressure. The specific interfacial area between gas and liquid phases in a distillation column is a very important feature, which has been researched theoretically in this paper. According to the suggested theoretically method specific interfacial area increased with increment of the gas velocity and has a good agreement with experimental one.

- $A$  cross-sectional area of the column, [m<sup>2</sup>]
- $a$  specific interfacial area, [m<sup>2</sup>/m<sup>3</sup>]
- $d_s$  sauter mean bubble diameter, [m]
- $e$  bubble eccentricity
- $f_b$  bubble formation frequency, [s<sup>-1</sup>]
- $g$  gravitational acceleration, [m.s<sup>-1</sup>]
- $h$  height of an ellipsoidal bubble, [m]
- $l$  length of an ellipsoidal bubble, [m]
- $S_b$  bubble surface, [m<sup>2</sup>]
- $u_G$  gas velocity, [m.s<sup>-1</sup>]
- $u_b$  bubble rise velocity, [m.s<sup>-1</sup>]
- $\mu_G$  gas viscosity, [Pa.s]
- $\mu_L$  liquid viscosity, [Pa.s]
- $\rho_G$  gas density, [kg.m<sup>-3</sup>]
- $\rho_L$  liquid density, [kg.m<sup>-3</sup>]
- $\sigma$  surface tension, [N.m<sup>-1</sup>]

Morton number 
$$Mo = \frac{g \cdot \mu_L^4}{\rho_L \cdot \sigma^3}$$

Bubble Reynolds number 
$$Re_b = \frac{d_s \cdot u_b \cdot \rho_L}{\mu_L}$$

Tadaki number 
$$Ta = Re_b \cdot Mo^{0.23}$$

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**This paper has been reviewed**