CORRELATION BETWEEN MASS TRANSFER AND PRESSURE DROP ON VALVE TRAY

PART I.

By

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(Received March 19. 1. 1973)

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1. Introduction

In this paper the Colburn – Chilton analogy has been applied to describe the correlation between the pressure drop and mass transfer on valve tray. The following assumptions have been made in using this analogy [1]: the foamy liquid on the tray may be regarded as an irregular packed-bed, in wich the gas phase flows across little irregular channels, the average lengths of which are equal to the foam height.

2. Theoretical

2.1 *Relation between Reynolds number and interfacial area*

Supposing that the continuity law is valid for the gas flow, the velocity of gas flowing across liquid on tray is:

$$v = \frac{F_T}{F} \frac{\varrho_T}{\varrho} \frac{v_T}{\varepsilon} \tag{1}$$

where

$$\varepsilon = \frac{z_h - z_c}{z_h} \tag{2}$$

By introducing the equivalent diameter for the gas channel

$$d_e = \frac{4\varepsilon}{a} \tag{3}$$

the Reynolds number can be written for the gas flowing across liquid on the tray:

$$Re = \frac{4\,\varphi\, G_T}{a\mu} \tag{4}$$

Let the Reynolds number Re_T for the gas flowing through the net ower cross-sectional area be:

$$Re_T = \frac{G_T D}{\mu_T} \tag{5}$$

Combining the expressions (4) and (5) we get

$$\frac{Re\ a}{Re_T} = \frac{4\varphi}{D} \frac{\mu_T}{\mu} = C \tag{6}$$

If the change of gas viscosity is negligible:

$$C \cong \frac{4\ \varphi}{D} = \text{const} \tag{7}$$

For tray without downcomer ($\varphi = 1$): $C = \frac{4}{D}$

This shows clearly that C is the specific contact surface in a wetted-wall tower of diameter D

Considering relation (6)

$$Re \cdot a = Re_T\ C \tag{8}$$

So products of the Reynolds number by the specific contact surface on the tray and in a wetted-wall tower of the same diameter as the tray, resp., are equal.

Using (8) we get:

$$a = \frac{C}{K}, \quad \text{where} \quad K = \frac{Re}{Re_T} \tag{9}$$

The factor K may have the following values:

first case:	for $K > 1$, $Re > Re_T$ and $a < C$
second case:	for $K = 1$, $Re = Re_T$ and $a = C$
third case:	for $K < 1$ $Re < Re_T$ and $a > C$

From these if can be concluded that the third case corresponds to the optimal operation of tray, where the specific contact surface is greater than the specific contact surface in a wetted-wall tower, meanwhile $Re < Re_T$. The first or second case cannot occur but for very low gas velocities or in case of a poor distribution of the gas phase.

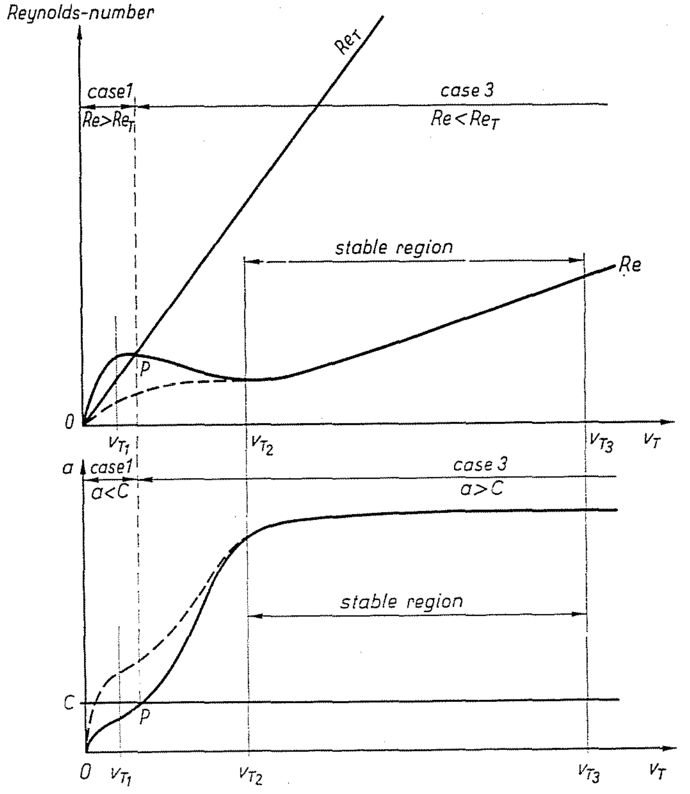

Fig. 1. The change of Reynolds numbers and interfacial area as a function of gas velocity

Fig. 1 shows the change of Reynolds numbers and specific contact surface as a function of gas velocity. For simplifying the problem, the change of the physical properties of gas is to be negligible and $\varphi = 1$. In this case the function between Re_T — number and gas velocity is nearly linear and its slope is D/ν.

The expression of Re number is

$$Re = \frac{d_e}{\nu} \frac{v_T}{\varepsilon} \tag{10}$$

Since d_e and ε also depend on the gas velocity, the curve of Re number plotted against gas velocity is complicated. Three cases are seen in Fig. 1. The point P represents the second case. From Eq. (9) the curve "a" can be constructed utilizing the diagram of Reynolds numbers.

In order to give an explanation of the relation mentioned above, the bubbling process can be divided into three periode: initial period, forming period and stable region.

In the initial period the first case prevails. Because of the low gas velocity, the gas can not pass through an area of the tray but of the least resistance.

Of course, the valve of the porosity ε is very low, so the R_T number is somewhat higher than the Re_T number. Since this period is limited to a narrow range of the low gas velocities (V_{T_1}), in general it cannot be observed, especially when the gas phase entering the tray is well distributed, (in this case the broken line shows the change of Re number and "a").

In the forming period (from V_{T_1} to V_{T_2}) the bubbling process is extending quickly over all the active area of the tray. When the gas velocity is higher than V_{T_2}, the tray can be said to operate in the stable region belonging to the third case and its upper limit is determined by the gas velocity at flooding.

The bubbling on a tray is known to be a very complicated process and results of revelant studies are not reassuming. Starting from the idea of the equilibrium of forces caused by the pressure inside and outside the bubble supposed to somewhat decrease in the forming period and to change little in the stable one [2]. But according to earlier results [1], the porosity factor ε suddenly increases at low gas velocities and then slowly at higher ones. Relationship (10) and Fig. 1 show that the Re number does not increase during the change of gas velocity in the forming period, it may even decrease. Accordingly, the valve of "a" suddenly increases from the value lower than C to higher than C dependent on the change of K. In the stable region the Re number increases with increasing gas velocity but slower than the Re_T — number, so the value of "a" increases little and appears to be nearly constant [2].

Last but not least, as stated earlier in the discussion, the Re number is much lower than Re_T, especially in the stable region. Hence, the motion of the liquid on the tray produces the turbulence of gas flow earlier than in a wetted-wall tower, and therefore also, the critical state is reached sooner.

2.2 *Modified friction factor*

According to our assumptions, the pressure drop of gas streaming through all the "little channels" can be expressed as:

$$\Delta p_s = \xi_s \frac{z_h}{d_e} \frac{\varrho v^2}{2g_c} \tag{11}$$

This pressure drop is due to friction.

For $\varrho \cong \varrho_T$

$$\frac{\xi s}{8} a = \frac{g_c}{\varphi^2} \frac{\varepsilon^3}{z_h} \frac{\Delta p_s}{\varrho_T v_T^2} \tag{12}$$

Of course the friction-pressure drop of the gas phase flowing through the liquid on the tray is not easy to determine. Using the principle of superposition [3], the "residual" pressure drop is:

$$\Delta p_m = \Delta p_ö - (\Delta p_{st} + \Delta p_{sz}) \tag{13}$$

where Δp is composed of the pressure drops due to surface tension, friction and flow separation.

On a valve tray the effect of surface tension may be negligible. But it has been pointed out that the pressure drop of gas passing through the valve during dry running differs from that during wet running [2]. Because of this fact it is difficult to determine the "residual" pressure drop. The overal pressure drop on valve tray is:

$$\Delta p_ö = \Delta p_{st} + \Delta p_m + \Delta p'_{sz} \tag{14}$$

where

$$\Delta p'_{sz} < \Delta p_{sz}$$

Of course, $\Delta p'_{sz}$ can exist only if the valve is working. The operation of dry valve tray involves that $\Delta p'_{sz}$ is nearly constant. From relationship (14) we get

$$\Delta p_ö - \Delta p_{st} = \Delta p_m + \Delta p'_{sz} = \Delta p \tag{15}$$

For $\Delta p'_{sz} =$ const., the behaviour of functions Δp and Δp_m is obviously similar. Supposing that Δp is proportional to Δp_m, ζ, the so called "resistance factor", can be calculated in the following way:

$$\frac{\xi}{8} a = \frac{g_c}{\varphi^2} \frac{\varepsilon^3}{z_h} \frac{\Delta p}{\varrho_T v_T^2} \tag{16}$$

Therefore

$$\xi_s < \xi \tag{17}$$

This is one of the causes why the analogy between momentum and mass transfer does not or only approximately exists on a tray.

2.3 *Expression of "j_D"*

In the general case expression of j_D is:

$$j_D = \frac{Sh}{Re\ Sc} Sc^{2/3} = f(Re)\,. \tag{18}$$

After some arrangements it can be written:

$$j_D = \frac{k_G RT}{v} \frac{p_{BK}}{P} Sc^{2/3}\,. \tag{19}$$

In complete back mixing where the main mass-transfer resistance is in gas phase, $E_{OG} \approx E_{MG}$ and $N_{OG} \approx N_G$, hence

$$N_G = -\ln(1 - E_{MG}) = \frac{k_y a}{G_M} z_h \tag{20}$$

Combining Eqs (19) and (20), after the necessary arrangements, we obtain

$$j_D a = \frac{N_G}{z_h} \frac{p_{BK}}{P} Sc^{2/3} \tag{21}$$

2.4 *Using modified Colborn – Chilton analogy for valve tray*

In general case the expressions of j_D and $\frac{\xi}{8}$ are functions of Re number:

$$j_D = A' Re^b \tag{22}$$

and

$$\frac{\xi}{8} = B' Re^c \tag{23}$$

where A', B', b and c are constant.

Introducing relationship (8) we obtain

$$j_D a = A Re_T^b \tag{24}$$

and

$$\frac{\xi}{8} a = B Re_T^c \tag{25}$$

where

$$A = A' a^{1-b} C^b \tag{26}$$

$$B = B' a^{1-c} C^c \tag{27}$$

If the coefficient "a" is constant, A and B are evidently also constant. Eqs (24) and (25) apply to bubble trays operating in stable region.

Relationships (24) and (25) were determined by an experimental method. Its results will be discussed in detail in the second part of this paper.

Notations

a — specific contact surface, m^2/m^3
C — "specific contact surface" for wetted-wall tower, 1/m
d_e — equivalent diameter of gas bubble, m
D — tower diameter, m
E_{MG} — Murphree gas-phase tray efficiency
E_{OG} — point gas-phase tray efficiency
F — active area of a tray, m^2

F_T — net tower cross-sectional area, m^2
g_c — conversion constant, 9,81 (kg/kp) (m/sec^2)
G_M — gas velocity, kmol/m^2 · sec
G_T — rate of gas through net tower area, kg/m^2 · sec
k_G — mass-transfer coefficient, kmol/m^2 · sec · atm
k_y — mass-transfer coefficient, kmol/(m^2 · sec · mole fraction)
j_D — mass-transfer dimensionless group
N_G — number of gas-phase transfer units
N_{OG} — number of overall gas-transfer units
P — total pressure, at
p_{BK} — mean partial pressure of inert component, atm
Δp — $\Delta p_{\ddot{o}} - \Delta p_{st}$, kp/m^2
Δp_m — "residual" pressure drop, kp/m^2
$\Delta p_{\ddot{o}}$ — total pressure drop, kp/m^2
Δp_s — friction-pressure drop, kp/m^2
Δp_{st} — static pressure drop, kp/m^2
Δp_{sz} — dry-plate gas-pressure drop, kp/m^2
$\Delta p'_{sz}$ — gas-pressure drop across valve when liquid is present on the tray, kp/m^2
R — universal gas constant, atm m^3/kmol · K°
Re — Reynolds number for gas through liquid on tray
Re_T — Reynolds number for gas through net tower area,
Sc — Schmidt number, dimensionless
Sh — Sherwood number, dimensionless
T — gas temperature, K°
v — velocity of gas flowing through liquid, m/sec
v_T — velocity of gas through net tower area, m/sec
z_c — clear liquid height on tray, m
z_h — foam height on tray, m

Greek letters

ε — porosity
φ — ratio of net tower area to active tray area
ξ — modified friction factor
ξ_s — friction factor
μ — viscosity of gas through liquid on tray, kg/m · sec
μ_T — viscosity of gas through net tower area, kg/m · sec
— kinematic viscosity of gas, m^2/sec
ϱ — density of gas through liquid on tray, kg/m^3
ϱ_T — density of gas through net tower area, kg/m^3

Summary

In connection with mass transfer on bubble tray the relationship between specific contact surface and Reynolds number has been dealt with of an important influence on the relationships of momentum and mass transfer process in tray towers.

Functions have been proposed for the relationship between $j_D a, \frac{\xi}{8}$ and Reynolds number. These functions were developed by both theoretical and experimental methods.

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