PRESSURE DROP OF FOAM BED ON GRID TRAYS*

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An equation is derived relating the pressure drop of foam bed on slotted Grid trays without downcomers with the liquid and gas flow rates and physical properties of both phases. The coefficients of the resulting equation are expressed as functions of geometrical tray parameters and of physical properties of the used mixtures on the basis of large number of experimental data measured with different liquids on a cold hydraulic model of cross-sectional area 0·105 m^2 and on experimental distillation columns of 0·4 and 1 m diameter.

The value of pressure drop of a foam bed and with it related liquid holdup (see further) of distillation trays is not only the key point of their design but it is directly related to the problem of plate efficiency calculations, as in majority of suggested equations for efficiency it represents the major hydraulic parameter. The possibility to calculate the pressure drop of a foam bed is important as well for calculations of expansion of the gas–liquid mixture on the tray and of entrainment. The previously suggested methods discussed in our review[1] are not altogether satisfactory due to their uncompleteness and complexity. The same can be said about some newer works[2,3] even if in this case the theoretical approach employed is more fundamental.

In this paper a simple equation is derived relating the pressure drop of a foam bed on slotted grid trays without downcomers with the liquid and gas flow rates, with geometrical parameters of trays and with physical properties of the fluids employed. This paper is a contribution to the solution of the third point of our research program[4].

THEORETICAL

The foam bed on a grid tray is considered to be a dynamic system for which a simultaneous flow of both phases through the tray is possible only when the hydrostatic pressure in individual points on the tray varies, with the gas flowing through the slots above which the hydrostatic pressure is less than the average one, and on the contrary liquid flowing through the slots above which the instantaneous hydrostatic pressure exceeds the average pressure drop of the foam bed. The mentioned accidental

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variation of tray slots through which the gas and liquid are flowing is typical for operation of trays without downcomers. The liquid efflux from a tray of the grid type will take place when the instantaneous foam concentration (density) in the place considered reaches a value higher than that corresponding to the mean value of $\varrho_l g h$, which is in equilibrium with the average pressure drop across the foam bed Δp_f, which follows from the relation

$$\Delta p_f = \Delta p_T - \Delta p_d = \varrho_l g h \tag{1}$$

derived from the momentum flux and force balance across the foam bed[5] on the tray at the simplifying assumption of negligible momentum change of the gas.

This is possible as the foam height H is always higher than the clear liquid height h as part of the foam volume on the plate is occupied by the gas flowing through it. The grid tray through which the gas and liquid are passing simultaneously (countercurrently) can be considered a more complicated case of liquid efflux from a tank where against the driving force for the efflux, *i.e.* the liquid height on the tray, the pressure drop acts across the foam layer as given by Eq. (*1*) which in turn is the driving force for the upward flow of gas.

For the liquid mass flow rate as given by the efflux equation for the part of the slot area a_l occupied by the flowing liquid we can write

$$L = \varrho_l a_l u_l = \varrho_l a_l k_1 [g(h_v - h)]^{1/2}, \tag{2}$$

where $(h_v - h)$ is the driving "height for efflux" where h_v, in agreement with the above discussion, is limited by the condition

$$h < h_v < H. \tag{3}$$

According to (*3*) we assume that the foam density over the area where liquid is dumping (h_v) is higher than the average value (h) and hence that there are density waves as the efflux point moves around the tray. If we assume that for the given hydrodynamic regime (second region – see Results) the quantity h_v always reaches the same multiple of h to get the necessary flow of liquid through the tray, we may write

$$h_v = k_2 h, \tag{4}$$

which means that the relative amplitude of density waves is constant at least for a given tray, or that the potential energy of density waves $(h_v - h)$ is proportional to the pressure drop $\Delta p_f(\sim h)$. At the present state of knowledge of behaviour of the foam layer on the plate, this assumption seems to be the most logical one.

Other possibilities such as $(h_v - h) =$ const. or simple functional dependences of the "driving height" on the flow rates in a form $(h_v - h) = g(v)$ or $(h_v - h) =$

$= g(L)$ do not lead to final relations by which the course of experimental data could be expressed. Eq. (2) is then for the in time averaged random efflux given by

$$L = \varrho_l a_l k_1 [g(k_2 h - h)]^{1/2} = \varrho_l a_l k_3 (gh)^{1/2}, \qquad (5)$$

where k_3 is the orifice coefficient which should be, in view of our assumptions, only a function of the geometric tray parameters.

We assume the slot area to be divided into the part through which the liquid flows – a_l, to the one through which the gas flows – a_g and the area through which neither phase flows – a_b (blocked area) *i.e.* that

$$a = a_l + a_g + a_b . \qquad (6)$$

For the mass flow rate of the gas under the assumption of a negligible density change we can write

$$G = a_g u_{rg} \varrho_g . \qquad (7)$$

For the pressure drop across the foam, which is the driving force of gas flow we may write

$$\Delta p_f = k_4 \varrho_g u_{rg}^2 , \qquad (8)$$

where k_4 is the coefficient of friction across the foam, which in general will be a function of the hydrodynamic conditions which are affected by the plate dimensions.

From equation (7) by the use of Eq. (8) the slot area through which the gas flows is

$$a_g = \frac{k_5 G}{(\varrho_g \, \Delta p_f)^{1/2}} . \qquad (9)$$

In the same way the area through which the liquid is flowing can be expressed on the basis of Eq. (5) as

$$a_l = \frac{k_6 L}{\varrho_l (gh)^{1/2}} \qquad (10)$$

and the sum with a_b is the relation (6)

$$a = k_6 L / [\varrho_l (gh)^{1/2}] + k_5 G / (\varrho_g \, \Delta p_f)^{1/2} + a_b . \qquad (11)$$

As $\Delta p_f = \varrho_l g h$, we obtain

$$a(\Delta p_f)^{1/2} = k_6 L / (\varrho_l)^{1/2} + k_5 G / (\varrho_g)^{1/2} + a_b (\Delta p_f)^{1/2} \qquad (12)$$

which is the required equation relating the liquid holdup, the fluid flow rates and their physical properties. In a more convenient form it can be written as

$$(\Delta p_f)^{1/2} = k_6\, u_{0l}(\varrho_l)^{1/2} + k_5\, u_{0g}(\varrho_g)^{1/2} + (a_b/a)\,(\Delta p_f)^{1/2}\,, \qquad (13)$$

where both fluid velocities are now related to the total slot area of the tray, a. The product of velocity and square root of density is frequent and is usually called the "f factor". The coefficients k_5 and k_6 can be calculated on the basis of experimentally measured data. In view of the discussion made concerning the Eq. (5), the orifice coeff. k_6 should be a function of geometric tray parameters only.

EXPERIMENTAL

Experimental Units and Measured Data

Experimental data used for calculation of coefficients of Eq. (13) were obtained from: *1.* measurements on the cold rectangular hydraulic section described earlier[4] of cross-sectional area 0·105 m^2 with the systems water–air, glycerol–air and kerosene–air; *2.* measurements on the experimental distillation column of 1 m diameter with the methyl alcohol–water system; *3.* experimental distillation column of 0·4 m diameter with the methyl alcohol–water and acetone–water systems from data of authors[6,7]. On the mentioned experimental stations were measured: Dependence of the overall pressure drop on the flow rates of both phases, for given geometrical tray parameters

TABLE I

Number of Experimental Data used for Evaluation of Coefficients of Eq. (13) and Mean Absolute Deviations of Relative Differences of Experimental and Calculated Values According to Eq. (13)

Unit	System	Number of points	Δ, %[a]
Cold hydraulic section, d_e 300 mm	water–air	500	13·0
	water soln. of glycerol–air	70	6·0
	kerosene–air	30	18·3
Experimental distillation column, D 400 mm	methyl alcohol–water	60	15·6
	acetone–water	70	15·0
Experimental distillation column, D 1000 mm	methyl alcohol–water	40	14·3[b]

[a] Mean absolute deviation of relative differences of experimental and from Eq. (13) calculated data at use of Fig. 1 and Table IV; [b] calculated Δp_f was divided by factor 2·2.

(slot width, plate thickness and free plate area) and for given physical properties of the mixtures employed (viscosities, densities and surface tensions). The number of experimental points (an experimental point is considered a value of over-all pressure drop at given flow rates of phases and for the above mentioned parameters measured with individual systems) is given in Table I. In majority of cases it is the average value of two or three simultaneous measurements made on the neighbouring plates. Geometrical parameters of the plates used are given in Table II, physical properties of the mixtures used in Table III.

TABLE II

Dimensions of Trays Used

Tray No	Column	%	b, mm	T, mm
I	[a]	8·93	4	2
II	[a]	16·25	4	2
III	[a]	22·03	4	2
V	[a]	17·07	8	2
IX	[a]	16·12	2	2
A	[b]	23·6	4·15	4
B	[b]	14·7	4·5	4
C	[b]	10·5	4·7	4
D	[b]	14·0	8	4
E	[b]	18·2	4	4
F	[c]	12·5	4	4
G	[c]	15·0	8	4
H	[c]	16·0	4	4

[a] Cold hydraulic section, d_e 300 mm, systems: Water–air, water soln. of glycerol–air and kerosene–air; [b] distillation column, D 1000 mm, systems methyl alcohol–water; [c] distillation column, D 400 mm, systems methyl alcohol–water, acetone–water.

TABLE III

Densities of Liquids Employed

Liquid	Density, kg m^{-3}
Water	1 000
Kerosene	805
Glycerol[a]	1 174
Methyl alcohol–water[b]	780—950[c]
Acetone–water[b]	760—950[c]

[a] Soln. of glycerol in water (68%), [b] at distillation, [c] at the boiling point of mixture.

Evaluation of Experimental Data

If we plot in accordance with Eq. (*13*) the square root of pressure drop across the foam bed, calculated from the over-all pressure drop by subtracting the dry plate pressure drop according to Eq. (*1*) against the gas velocity in slots at the constant liquid flow rate, instead of the very complex curves of such graphs of $\Delta p_T \sim$ $\sim u_{0g}$, three regions of linear plots appeared (similar to the plots of $\Delta p_T \sim u_{0g}$ in logarithmic coordinates): *1*. Initial region of rapid increase of $(\Delta p_f)^{1/2}$ starting from the lower critical velocity of the tray up to the second region. *2*. Region of steady tray operation, with a very slow increase of $(\Delta p_f)^{1/2}$ (in agreement with Eq. (*1*) it is the region of about constant liquid holdup). *3*. Region of rapid increase of $(\Delta p_f)^{1/2}$ (region of visible oscillations or waving of the foam bed, described earlier in our review[1]). From practical point of view we expect the second region to be of the greatest importance, as the most stable tray operation can be expected there, *i.e.* small (or no) changes of pressure drop across the foam bed as the result of momentary fluctuation of fluid flows and as it includes the velocity ranges usual with tray operation at atmospheric pressure.

On the basis of all available experimental data measured with the cold hydraulic section with the water–air system were calculated the coefficients of Eq. (*13*) for this second region. From these calculations follows, that the third term of the right-

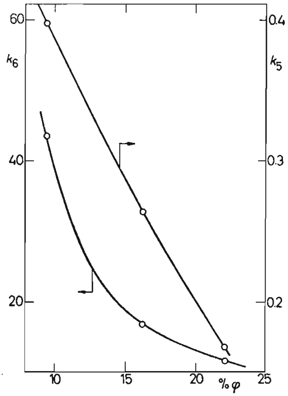

FIG. 1

Dependence of Dimensionless Coefficients k_5 and k_6 of Eq. (*14*) on Geometrical Free Plate Area φ (%)

hand side of Eq. (*13*) is, for the given gas–liquid system a universal constant *i.e.* that the term $(a_b/a)(\Delta p_f)^{1/2} = c_1$. Its dependence on the gas–liquid system employed was calculated from the data measured with the hydraulic section with the kerosene–air and glycerol–air and at distillation on a column of 0·4 m diameter with the systems methyl alcohol–water and acetone–water. The calculated values of coefficients c_1 of pure components are given in Table IV. For mixtures is used a linear interpolation of coefficients c_1 of both components on their composition in weight %.

The dependence of coefficients k_5 and k_6 on geometrical parameters of plates was expressed graphically for simplicity as a function of free plate area in Fig. 1. The plot of coefficient k_6 as a function of geometrical free plate area is considered in agreement with the earlier discussed significance of this coefficient as correct, with coefficient k_5 would be in accordance with Eq. (*8*) necessary to consider also the hydrodynamic conditions, which will be according to our opinion just mostly affected by the main geometrical plate parameter, *i.e.* by its geometrical free plate area.

Empirical limits of Eq. (*16*) are

$$7\ \mathrm{m\,s^{-1}} < u_{0g} < 17\ \mathrm{m\,s^{-1}}\,; \qquad (14)$$

and

$$(u_{0l})_{max} = 0{\cdot}16\,.\,\varphi[\mathrm{m\,s^{-1}}]\,. \qquad (15)$$

Effect of Column Diameter on Pressure Drop across the Foam Bed

As follows from a number of studies made on trays without downcomers[8] the increasing of the column diameter has a significant effect on the pressure drop of plate with a foam bed (thus in agreement with Eq. (*1*) also on the pressure drop across the foam bed). With increasing of the column diameter the pressure drop (over-all *i.e.* across the foam bed as well) significantly decreases. For expressing this effect the data measured on the experimental distillation column[6] of 1 m diameter were used, operating with the methyl alcohol–water system. These results were

TABLE IV

Values of Coefficient c_1 for Individual Components of the Liquid–Gas System (for foam formed by a gas (air) passing through the liquid the given value is used, for mixtures the interpolated value of coefficients c_1 of both components in dependence on composition in weight %)

Component	c_1	Components	c_1
Water	7·0	Kerosene	6·0
Water soln. of glycerol (68w %)	8·0	Methyl alcohol	12·0
		Acetone	11·0

compared with values of Δp_f calculated from Eq. (*13*) by use of coefficient c_1 calculated on basis of above given experimental data measured on the column of 0·4 m diameter with the same system. The pressure drop across the foam bed is in this case in average for a factor of 2·2 less. Due to lack of additional data obtained from columns of greater diameters generalization of this result is not at present possible. The mean modulus deviations of relative difference of all experimental data from that calculated from Eq. (*13*) at the use of coefficients given in Table IV within the range of velocities limited by empirical equations (*14*) and (*15*) are given for individual systems in Table I.

DISCUSSION

The method presented along with the theoretically derived and experimentally verified equation[5] relating the pressure drop across the foam bed and the liquid holdup (*1*) represents a simple method of calculation of the pressure drop across the foam bed and of liquid holdup of slotted grid plates without downcomers. It is interesting to note the similarity of Eq. (*15*) with the empirical equation used in the AICHE[9] method for the holdup calculations of bubble-cap and sieve trays with downcomers. The shortcoming of the final equation (*13*) can be seen in the insufficient verification of the effect of column diameter on pressure drop caused by insufficient data available with columns of large diameters. As it was experimentally determined that the term $(a_b/a)\ (\Delta p_f)^{1/2}$ is for the given tray dimensions constant, it means that at the given liquid flow rate, with the increasing gas flow rate the value of (a_b/a) decreases (as $(\Delta p_f)^{1/2}$ increases) which is qualitatively correct as the area of the slots "blocked" decreases as the area for gas flow with larger gas flow rates increases. This conclusion is important as by the use of Eq. (*13*) the free area distribution might be indirectly determined, which has not yet been possible. The direct experimental measurement of this quantity is not easy and yet published results[2] are not quite convincing. According to Eq. (*13*) for the given gas–liquid system at different but constant liquid flow rates, a plot of $(\Delta p_f)^{1/2}$ against the gas slot velocity should give a set of parallel straight lines. With trays of small free area, these straight lines are not exactly parallel even if the measured values differ from the calculated ones within the above mentioned limits. With trays of larger free areas, these straight lines are parallel. The usually observed oscillation region mentioned at the beginning of this paper is definitely identical with the third region (see Results). Till now, we have not proved how much these oscillations interfere with the second region, from the side of larger gas velocities. Explanation of this question in the future will help the choice of the fluid flow rates suitable for the best operation of such columns.

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LIST OF SYMBOLS

a area of tray slots, m^2
A column cross-sectional area, m^2
b slot width, mm
$c = (a_b/a)(\Delta p_f)^{1/2}$ empirical coefficient of Eq. (*13*)
d hole diameter, mm
D column diameter, mm
g acceleration of gravity, $m\,s^{-2}$
F F-factor, $kg^{-1/2}\,m^{1/2}\,s^{-1}$
G mass flow rate of gas, $kg\,s^{-1}$
h clear liquid height (height of liquid without gas bubbles covering the whole column cross-sectional area), m
h_v liquid height above the tray slots where liquid is dumping, m
H foam height, m
k coefficients of Eq. (*2*) — (*13*)
K^+ total dry plate pressure drop coefficient* (Eq. *14*)
L mass flow rate of liquid, $kg\,s^{-1}$
Δp pressure drop, $N\,m^{-2}$
T tray thickness, mm
$u = u_0\varphi$ linear velocity, $m\,s^{-1}$
ϱ density, $kg\,m^{-3}$
$\varphi = a/A$ free tray area
μ viscosity, cP

Indices

b blocked
d dry plate
e equivalent (for slots $d_e \approx 2b$)
f foam
g gas
l liquid
o slot (related to the total slot area)
r actual in slots
T over-all

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* In agreement with the above paper[4], the total dry plate pressure drop coefficient K^+ is given by $K^+ = \{K[0{\cdot}4(1{\cdot}25 - \varphi) + (1 - \varphi)^2]/2\varphi^2\}^{1/2}$, where K is a function of the ratio T/d_e only (for $T/d_e < 2{\cdot}3$).