# WALL FLOW IN TRICKLE BED REACTOR*

J. PRCHLÍK, J. SOUKUP, V. ZAPLETAL, V. RŮŽIČKA and P. KOVAŘÍK

*Department of Organic Chemistry,*
*Institute of Chemical Technology, 166 28 Prague 6*

Received February 26th, 1974

Wall flow is studied in dependence on parameters of the trickle bed reactor (bed height, type and size of packing, size of reactor). Empirical equation is given by which the wall flow rate can be calculated in a packed porous and non-porous bed with the central source.

The wall flow *i.e.* the ratio of liquid flowing downwards the wall to the total liquid flow rate in the reactor is one of important parameters in the design of trickle bed reactors. As the liquid flowing downwards the wall is not fully utilised for the catalytic reaction, the effort is to minimalize its quantity. Once the wall flow rate exceeds some reasonable limit (usually 5 to 10%) it is necessary to instal redistributors at certain distances in the reactor which are returning the liquid back into the packing. The wall flow rate depends on many factors: reactor size, bed height, size and type of packing, physical properties of fed liquids, type of material of which the reactor wall is made *etc.* In this study, which is related to the preceding papers[1-3], the attention is paid mainly to the qualitative description of dependence of the wall flow on the bed height and type of packing and on the size of reactor and packing. A simple equation was looked for which would enable generalization and quantitative expression of the found dependence.

## THEORETICAL

The relations resulting from mathematical models of liquid distribution in a randomly packed bed of catalyst[3] can be used for calculation of the wall flow in dependence on the bed height. The model according to Staněk and Kolář[6] which is very accurate can be applied and the relation for calculation of the wall flow rate at wetting the bed by a central point source in the form

$$W_p = \left\{\frac{1}{1 + C} - \sum_{n=1}^{\infty} \frac{2[(q_n^2/B) - 2C] \exp(-q_n^2 T)}{\{[(q_n^2/B) - 2C]^2 + q_n^2 + 4C\} J_0(q_n)}\right\} 100 . \qquad (1)$$

* Part IV in the series Liquid Distribution in Trickle Bed Reactors; Part III: This Journal *40*, 845 (1975).

is obtained. Equation (*1*) is disadvantageous though it is very accurate because of the complexity of calculation namely due to three constants which must be determined for each packing and/or the changing sizes of the reactor and packing. Thus an equation was looked for which would enable a quick and simpler calculation of the wall flow rate with a sufficient accuracy.

Several empirical equations were tested which with differing accuracy describe the experimental dependence of the wall flow on the bed height at its wetting by a central point source.

The most suitable has proved to be the exponential relation

$$y = b[1 - \exp(b_1 x)]\,, \quad b < 0 \tag{2}$$

in a dimensionless form.

The dimensionless bed height $Z = z/d_k$ and the initial bed height $z_p$ were defined. Here $z_p$ is the height necessary for the liquid to reach the wall at the given experimental arrangement (region of limited bed of random packing). On basis of the probability considerations (deviation from the vertical flow in the radial direction which is at the contact of the liquid element with the particle of the packing proportional to $d_k/d_p$ and the number of collisions proportional to $d_k$) so that it can be assumed that the initial bed height is proportional to the ratio $d_k^2/d_p$ and thus for its dimensionless form holds

$$Z_p = kd_1d_p\,. \tag{3}$$

Constant $b$ from Eq. (*2*) was for the physical interpretation substituted by the wall flow rate in the region of equilibrium liquid distribution $W_d(\infty)$.

Under these assumptions it is possible to write Eq. (*2*) for calculation of the wall flow rate at wetting the randomly packed bed by a central point source in the form

$$W_p = W_p(\infty)\,[1 - \exp(b_1 Z_1)]\,, \quad Z_1 = Z - Z_p\,. \tag{4}$$

TABLE I

Constants in Eq. (*4*) for the Studied Packings

| Packing | $k$ | $b_1$ |
|---|---|---|
| A | 0·054 | −0·6684 |
| A, B, C | 0·042[a] | −0·6393[a] |
| D | 0·101 | −0·3904 |

[a] Valid for all the specified types.

Constants of this equation can be evaluated by the method of the least squares from the linearized form of Eq. (*4*). The constants calculated for various packings are given in Table I.

## EXPERIMENTAL

The wall flow rate was measured by the apparatus described earlier[1]. The column was 0·251 m in diameter while some experimental data were measured earlier on columns of diameters 0·159 and 0·084 m. In these experiments the modified base plate was applied which made possible the measurement of the wall flow rate[4]. Other parts of the apparatus remained the same. Wetting in the majority of experiments was made by the central point source. At wetting the same procedure as in the recent studies[1,2] was applied. The fed liquid was water. The used types of packings of porous and nonporous character are given in Table II. The discussed wall flow rates represent their mean values in the region of initial wetting densities $f_0 = 1$ to $15\ m^3/m^2h$.

## RESULTS AND DISCUSSION

### *Dependence of the Wall Flow Rate on Type and Height of Packing*

With increasing bed height the wall flow increases too, as is obvious from Fig. 1 at wetting by a central point source. From certain bed height depending on parameters of the reactor, type and size of packing, physical properties of the fed liquid *etc.* the wall flow rate becomes constant and is independent on the type of the liquid source employed (its value is in the given case within the range of experimental error of about 7%).

The wall flow rate is also significantly dependent on the porosity of packing. This becomes obvious from comparison of results given in Table III for porous catalytic bed A (Nickel in kieselguhr) and for a non-porous packing *D* (glass spheres used in lithography) when the ratios $d_k/d_p$ are comparable (18·1 and 16·2 or 9·5 and 8·8).

TABLE II
Characteristics of Packings

| Type | No | Surface area $m^2/g$ | Porosity % | Dimension mm | Equivalent diameter[a] mm | Free volume |
|---|---|---|---|---|---|---|
| A | nickel on kieselguhr, pellets | 195·0 | 61 | 9·0 . 7·2 | 8·8 | 0·37 |
| B | kieselguhr as carrier | 2·9 | 65 | 3·9 . 11·7 | 6·4 | 0·34 |
| C | kieselguhr as carrier, pellets | 2·5 | 65 | 5·7 . 6·8 | 6·9 | 0·35 |
| D | glass spheres, lithographic | — | — | 9·8 | — | 0·36 |

[a] Equivalent diameter calculated according to Hobler[16], $d_{eq} = 1{\cdot}241\ (V_p)^{1/3}$.

TABLE III

Wall Flow Rates for Packings A and D in the Region of Equilibrium Liquid Distribution at Wetting by a Central Point Source in Columns with Diameters 159 and 84 mm

| Column ID 0·159 m | | | Column ID 0·084 m | | |
|---|---|---|---|---|---|
| bed height | wall flow rate | | bed height | wall flow rate | |
| m | packing A | packing D | m | packing A | packing D |
| 0·70 | 15·2 | 20·6 | 0·55 | 38·2 | 42·1 |
| 0·80 | 16·0 | 25·4 | 0·75 | 38·7 | 46·8 |
| 0·90 | 15·0 | 28·4 | 0·85 | 38·5 | 48·0 |
| 1·00 | 15·9 | 28·6 | 1·05 | 37·7 | 47·6 |

For the same value $d_k/d_p$ which for non-porous packings equal to 16·2 and 8·8 the equilibrium wall flow rate for porous packings can be read off from Fig. 2 (16·0 and 37·5%) which is also considerably lower than in the case of non-porous packing.

Larger wall flow rates with non-porous packings (20 to 40 rel. %) were also found in other measurements[5]. They can be also confirmed by comparison of wall flow rates for non-porous packings of various shapes and types (glass spheres, Raschig rings, packing Intalox, Berl saddles and ceramic cylinders)[6-9,13] with our own results which were performed on porous packings under similar experimental conditions.

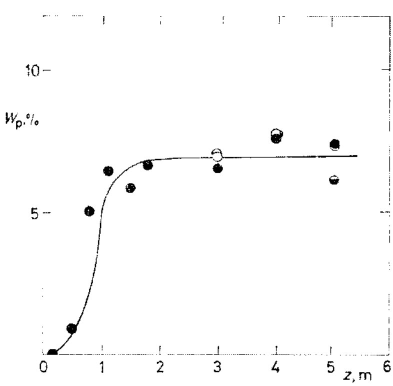


FIG. 1

Wall Flow Rate in Dependence on Bed Height A

● Central point source, ⊖ uniform source, ○ wall source.

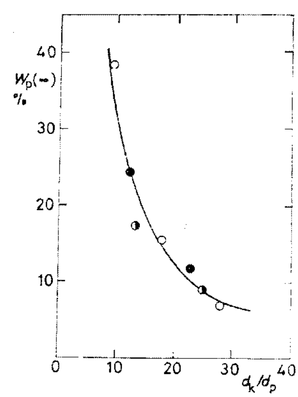


Fig. 2

Wall Flow Rate in the Region of Equilibrium Distribution in Dependence on Ratio $d_k/d_p$ for Porous Packings A to C

○ A, ◑ B, ● C.

From the wall flow rate for porous and non-porous packings also results that the region of equilibrium liquid distribution and thus of the steady state as well is reached for the porous packing in lower beds. This is in agreement with the found value of the spreading coefficient $D$ which was greater for the porous packing ($D = 0{\cdot}001939$ m) than for the non-porous glass spheres ($D = 0{\cdot}001415$).

*Dependence of Wall Flow Rate on the Size of Reactor and of Packing*

The effect of the size of reactor and of packing on the equilibrium wall flow rate is characterized by the ratio of the reactor diameter to the packing element $d_k/d_p$. In general it holds that with decreasing ratio $d_k/d_p$ the wall flow rate increases. For the non-porous packing was earlier recommended the limiting value of the ratio $d_k/d_p = 8$ to 12 which should have ensured the minimum wall flow rate[10-12]. Recently the value 25 to 30 has been given[13-15]. For the porous catalytic packing the limiting ratio $d_k/d_p$ has not yet been published.

The dependence of the mean wall flow rate on variable ratio $d_k/d_p$ for porous packings A to C in the region of equilibrium liquid distribution is given in Fig. 2.

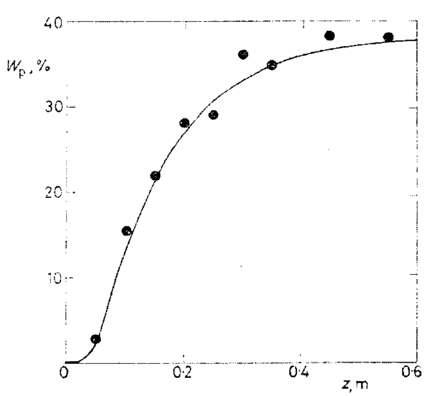


FIG. 3

Comparison of Theoretical Curve for Wall Flow Rate Obtained from Eq. (*4*) with Experimental Data

Packing A, central point source, column ID 0·084 m.

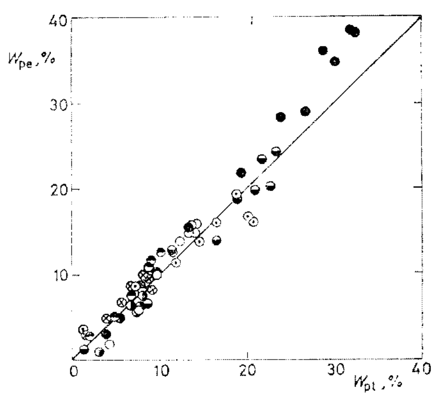


FIG. 4

Comparison of Calculated (Eq. (*4*)) and Experimental Wall Flow Rates for Columns with ID 0·251, 0·159 and 0·084 m with Packings A to C at Wetting by a Central Point Source

Packing A ◑ ID 0·251 m, ○ ID 0·159 m, ● ID 0·084 m. Packing B ⊗ ID 0·159 m, ⊙ ID 0·084 m. Packing C ◓ ID 0·159 m, ◒ ID 0·084 m.

It was obtained by changing the diameter of the column while the size of the packing elements was kept constant. It is obvious from this dependence that for the wall flow rate lower than 10% of the over-all liquid flow rate it is necessary to keep the ratio $d_k/d_d > 25$ for porous packing of similar properties like those of A to C. It can be expected that the wall flow rate for $d_k/d_d > 30$ will be independent on this ratio.

For non-porous packing the limiting ratio $d_k/d_p = 8$ to 12 is also quite insufficient which has been confirmed by our own data as well as by literature data (Table IV).

### *Calculation of Wall Flow Rate*

Equation (*4*) was verified by comparison of calculated wall flow rates with the experimental data obtained by wetting the packings A to D in columns with the ID 0·084 to 0·251 m. As is obvious from Fig. 3, the wall flow rates calculated in dependence on the bed height are in a very good agreement with the experimental data obtained at wetting columns with different diameters. Relatively worse agreement of calculated and experimental wall flow rates is obtained at wetting the non-porous packing ($\bar{\sigma}$ max. 21·3%) and can be explained by a greater scatter of experimental data in the measurements made with this type of packing[4].

Equation (*4*) was also modified for generalised calculations of the wall flow rates at wetting the porous packings of various types which have similar properties. The constants substituted into this equation were always the same for all the studied

TABLE IV

Equilibrium Wall Flow Rate in Various Non-Porous Packings in Dependence on Ratio $d_k/d_p$

| Type of packing | $d_k/d_p$ | Equilibrium wall flow rate % | Reference |
|---|---|---|---|
| Raschig rings | 4·0 | 87·0 | 8 |
| Raschig rings | 6·0 | 68·0 | 8 |
| Raschig rings | 6·0 | 51·0 | 7 |
| Raschig rings | 8·0 | 54·0 | 8 |
| Glass spheres | 8·8 | 47·5 | this study |
| Raschig rings | 10·0 | 37·7 | 7 |
| Intalox | 10·0 | 41·9 | 7 |
| Glass spheres | 10·8 | 30·0 | 6 |
| Raschig rings | 12·0 | 40·0 | 8 |
| Raschig rings | 12·0 | 36·8 | 9 |
| Glass spheres | 16·2 | 28·5 | this study |
| Raschig rings | 16·7 | 25·0 | 7 |

porous packings (A to C) while the equilibrium wall flow rate was for variable $d_k/d_p$ calculated from the curve plotted in Fig. 2. The dependence on Fig. 4 proves a considerable agreement of calculated and experimental values. The difference appearing in regions of larger wall flow rates is, first of all, the result of an error with which the equilibrium wall flow rate is estimated for small values of ratios $d_k/d_p$.

The constants substituted into Eq. (*4*) are valid for packings with similar properties as those of A to C but it can be expected that they will be also suitable for other types of catalytic porous packings.

The advantage of the proposed Eq. (*4*) is in its simplicity and that it can be applied to systems with different ratios $d_k/d_p$ without the need to determine the values of constants.

### *Redistributors*

For the equilibrium wall flow rate exceeding the required limit (at $d_k/d_p < 25$), it is necessary to install redistributors in the reactors which are reversing the liquid reaction mixture back into the packing. The distance at which the redistributors should be located at wetting the given types of catalytic packings by a central point source can be estimated on basis of the discussed equation for calculation of the wall flow rate (*4*) from the relation

$$z_r = 0{\cdot}042 \frac{d_k^2}{d_p} - \frac{d_k}{0{\cdot}064} \ln \left( \frac{W_p(\infty) - W_{p,1}}{W_p(\infty)} \right). \qquad (5)$$

#### LIST OF SYMBOLS

| | |
|---|---|
| $B$ | number for transfer of liquid into the wall |
| $b, b_1$ | constants |
| $C$ | distribution number |
| $D$ | spreading factor (m) |
| $d_k$ | diameter of reactor (m) |
| $d_p$ | diameter of packing element (m) |
| $f_0$ | initial wetting density ($m^3m^{-2}h^{-1}$) |
| $J_0$ | Bessel function, first type, zero order |
| $k$ | constant |
| $q_n$ | roots of transcendent equations |
| $T = Dz/a^2$ | dimensionless spreading factor |
| $V_p$ | volume of element |
| $W$ | wall flow rate ($m^3h^{-1}$) |
| $W_p$ | ratio of wall flow rate to over-all flow rate (%) |
| $W_p(\infty)$ | ratio of wall flow rate to the total flow rate in the region of equilibrium distribution (%) |
| $W_{p,e}$ | experimentally determined wall flow rate (%) |

| | |
|---|---|
| $W_{p,t}$ | calculated wall flow rate (%) |
| $x$ | dependent variable |
| $y$ | independent variable |
| $Z = z/d_k$ | dimensionless bed height |
| $Z_p$ | dimensionless initial bed height |
| $z$ | bed height (m) |
| $z_p$ | initial bed height (m) |
| $z_r$ | bed height for location of redistributors (m) |
| $\bar{\sigma}$ | relative mean deviation in calculations of wall flow rates (%) |

## REFERENCES

1. Soukup J., Kolomazník K., Zapletal V., Růžička V., Prchlík J.: This Journal *38*, 3742 (1973).
2. Kolomazník K., Soukup J., Prchlík J., Zapletal V., Růžička V.: This Journal *39*, 216 (1974).
3. Prchlík J., Soukup J., Zapletal V., Růžička V.: This Journal *40*, 845 (1975).
4. Prchlík J.: *Thesis*. Institute of Chemical Technology, Prague 1973.
5. Kovařík P.: *Thesis*. Institute of Chemical Technology, Prague 1973.
6. Staněk V., Kolář V.: This Journal *33*, 3235 (1968).
7. Dutkai E., Ruckenstein E.: Chem. Eng. Sci. *23*, 1365 (1968).
8. Templeman J. J., Porter K. E.: Chem. Eng. Sci. *20*, 1139 (1965).
9. Jones C. M.: *Thesis*. University of Birmingham 1961.
10. Tour R. S., Lerman F.: Trans. Am. Inst. Chem. Eng. *35*, 719 (1939).
11. Kirschbaum E.: *Destillier und Rektifiziertechnik*. Springer Verlag, Berlin 1960.
12. Kirschbaum E.: Brit. Chem. Eng. *2*, 426 (1957).
13. Dutkai E.: Rev. Chim. (Bucharest) *21*, 553 (1970).
14. Jameson G. J.: Trans. Inst. Chem. Eng. *44*, 198 (1966).
15. Schnell H.: Allgem. Wärmetechn. *12*, 124 (1964).
16. Hobler T.: *Absorpce*. Published by SNTL, Prague 1967.

Translated by M. Rylek.