THE EFFECT OF PLATES ON SOME PARAMETERS OF A HETEROGENEOUS BED IN BUBBLE-TYPE FLOW REACTORS

J. ZAHRADNÍK and F. KAŠTÁNEK

Institute of Chemical Process Fundamentals, Czechoslovak Academy of Sciences, 165 02 Prague 6 - Suchdol

Received March 5th, 1973

The effect of several plates situated in a bubble type reactor on residence time distribution of liquid, backmixing and porosity of the bed is studied experimentally. The presence of plates improves hydraulic properties of these reactors. The operating range of plates with downcomers is wider than of those without downcomers.

The use of several distributing plates situated in a bubble-type plate reactor is advantageous as it is improving hydraulic properties of these reactors. Residence time distribution of liquid is approaching the piston flow with increasing number of plates together with better mixing of the liquid in the interplate space and with a more uniform porosity of the heterogeneous bed. This is accompanied by a perfect redistribution of bubbles with a subsequent renewal of the interfacial area without any significant coalescence of bubbles taking place. The investment cost of additional plates does not represent any significant increase in the cost of the reactor and the increase in pressure drop is, in respect to high liquid beds used in these reactor types, insignificant.

The majority of studies made so far have been oriented to single-stage reactors. Studies of multistage reactors are rare[1], therefore here the effect of several plates situated in a reactor on behaviour of the bubbled-bed is studied experimentally.

EXPERIMENTAL

The experimental unit used was already described earlier[2]. Diameter of the reactor was 292 mm. It had one distributing plate and was operated as a single-stage flow reactor having the height of the heterogeneous bed equal to 1 400 mm or with several stages operated as a three-stage bubble-type reactor with plates situated 460 mm apart. Sieve plates with two downcomer pipes of 40 mm I.D. without weirs were used. Free plate area was in the range 4–8%, hole diameter 1·6 to 5 mm. For comparison, also plates without downcomers were employed having the same geometrical parameters. Liquid flow rate was within the range of W_L $\langle 2{\cdot}5 \,.\, 10^{-3} - 7{\cdot}5 \,.\, 10^{-3}$ $m^3/m^2\,s\rangle$, gas flow rate within $\langle 0{\cdot}008$ to $0{\cdot}042$ m/s$\rangle$. Phases were flowing countercurrently.

* Part IV in the series Gas–Liquid Reactors; Part III: This Journal *39*, 1403 (1974).

The method of measurement of liquid residence time distribution, backmixing and of bed porosity was given detail in our previous studies of this series[2-4].

RESULTS

Residence time distribution of single-stage and multi-stage reactors. Dimensionless E-curves obtained at measurements in single-stage reactor (distributing plate $\varphi = 4\%$, $d_0 = 3$ mm) are compared for selected limiting phase flow rates used in these types of reactors[1], with theoretical E-curves calculated by solving the mathematical model representing a single-stage reactor as a backmix vessel with by-pass and dead region which in general can transfer mass with the bulk of the ideally mixed reactor[5] (Fig. 1). It is obvious that by this model the character of liquid mixing can be expressed in the whole range of phase flow rates. The active, backmixed part of the reactor to those total volume of the stage is in the range 0·8 to 1·0. The shape of the curves corresponds to the presence of the mass transfer between dead region and the ideally mixed part of reactor volume. The results indicate absence of by-pass. E-curves obtained in a single-stage reactor are compared with the response curves obtained for the same values of W_L and w_G in a multi-stage plate reactor where the plate parameters correspond to those of the distributing plate in a single-stage reactor. The comparison demonstrates that the use of several plates significantly decreases liquid backmixing (Fig. 2). Only an insignificant dead region has been found in multi-stage reactor at a relatively large liquid flow rate and low gas flow velocities (0·008 m/s). As in practical applications the linear gas velocities are always more likely to be above 0·01 ms^{-1} the multi-stage plate reactor is, as concerns the use of

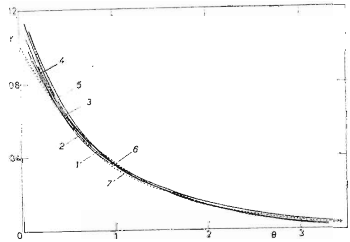

FIG. 1

Comparison of Experimental and Theoretical E-Curves for a Bubbled Bed without Internals

Parameters of the distributing plate $\varphi = 4\%$, $d_0 = 3$ mm; *1* $W_L = 2{\cdot}5 \cdot 10^{-3}$ m^3 : m^2 s, $w_G = 0{\cdot}008$ m/s; *2* $W_L = 2{\cdot}5 \cdot 10^{-3}$, $w_G = 0{\cdot}042$; *3* $W_L = 5{\cdot}0 \cdot 10^{-3}$, $w_G = 0{\cdot}025$; *4* $W_L = 7{\cdot}5 \cdot 10^{-3}$, $w_G = 0{\cdot}008$; *5* $W_L = 7{\cdot}5 \cdot 10^{-3}$, $w_G = 0{\cdot}042$; *6* $a_1 = 1$, $a_2 = 1$, $a_3 = 0$; *7* $a_1 = 0{\cdot}8$, $a_2 = 1$, $a_3 = 0{\cdot}2$.

the reactor space, more advantageous. The E-curve of liquid residence time distribution in a multi-stage reactor nearly corresponds to an ideal E-curve obtained by calculation of a serie of backmix reactors with liquid backmixing between adjacent stages where for the volume of individual backmix reactor is substituted the corresponding liquid holdup in the stage and the flow through the downcomer can be simulated by the liquid flow in a series of three backmix reactors of the overall volume equal to the volume of liquid in the downcomer[2].

Porosity in single and multi-stage reactors. From Table I where the results of comparison of porosities are given results that in a single-stage reactor the gas holdup in the considered range of variables[2] depends neither on the free plate area nor on the liquid flow rate, and its value is only slightly smaller than in a multi-stage reactor. By comparing the changes in porosities along the reactor height in both types of reactors it can be assumed that with increasing number of plates the difference in porosities of a multi-stage and a single-stage reactor will further increase as was also observed *e.g.* by Fair and coworkers[6]. In a multistage reactor for the water–air system was found that in all stages without regard to values of variables W_L, w_G, d_0, and φ the averaged porosity in the considered interplate space (stage) can be expressed by an empirical relation

$$\varepsilon_{Gi} = I_i w_G , \qquad (1)$$

where the values of constants I_i for individual stages are $I_1 = 1{\cdot}92$, $I_2 = 2{\cdot}5$, $I_3 = 3{\cdot}34$ where the stages are numbered from the bottom. The linear dependence

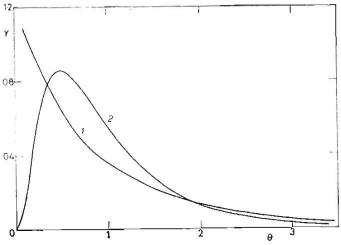

FIG. 2

E-Curves of a Single- and Multi-Stage Reactor

Plate parameters $\varphi = 4\%$, $d_0 = 3$ mm, flow rates $W_L = 7{\cdot}5 \cdot 10^{-3}$ m^3/m^2 s, $w_G = 0{\cdot}042$ m/s; *1* bubbled single-stage reactor, *2* three-stage bubbled reactor (three plates used).

of porosity on the gas velocity in the region of small gas velocities can be expected and is in agreement *e.g.* with the relation of Reith[7] derived on basis of a model on bubble motion in a single-stage reactor

$$\varepsilon_G = w_G/(2w_G + 0{\cdot}3)\,, \tag{2}$$

if for the mean gas velocity in the studied range with the assumption $2w_G \ll 0{\cdot}3$ we accept an approximatelly 12% deviation from the relation by Reith. In this case the relation holds

$$\varepsilon_G \approx 3{\cdot}3w_G\,, \tag{2a}$$

which is in agreement with the results we have obtained.

Relation (*1*) with variable I_i is expressing the increase in the porosity from stage to stage. This increase can be explained by the redistribution effect of plates due to which the accidental bubble clusters passing quickly through the bed are disappearing and the foam structure becomes more uniform with decreasing range of the bubble size, with increasing residence time of bubbles in the stage and the correspondingly greater porosity. The effect of hydrostatic pressure on porosity *i.e.* the increase in porosity along the reactor height due to the increasing volume of bubbles is in the given experimental arrangement insignificant. As porosity is increasing exponentially in dependence on the number of stages which is the result obtained

TABLE I

Comparison of Mean Gas Porosities in Single and Multi-Stage Reactors for Various φ (%) and d_0 (mm)

$W_L \cdot 10^3$ $m^3/m^2\,s$	w_G m/s	Single-stage reactor	ε_G multi-stage reactor	$W_L \cdot 10^3$ $m^3/m^2\,s$	w_G m/s	Single-stage reactor	ε_G multi-stage reactor
	$\varphi = 4$, $d_0 = 3$				$\varphi = 8$, $d_0 = 3$		
2·5	0·008	0·014	0·019	2·5	0·008	0·016	0·019
2·5	0·025	0·061	0·072	2·5	0·025	0·061	0·069
2·5	0·042	0·100	0·116	2·5	0·042	0·100	0·115
5·0	0·008	0·016	0·018	5·0	0·008	0·015	0·020
5·0	0·025	0·062	0·065	5·0	0·025	0·062	0·064
5·0	0·042	0·099	0·111	5·0	0·042	0·103	0·109
7·5	0·008	0·015	0·018	7·5	0·008	0·015	0·020
7·5	0·025	0·061	0·057	7·5	0·025	0·061	0·060
7·5	0·042	0·098	0·105	7·5	0·042	0·102	0·108

on basis of our measurements with three stages, quantitative conclusions cannot be made on the character of increase of porosity with the height of a multistage reactor without additional experimental data on a multistage reactor with changes in the height of stages. Moreover, in the top stage high porosity is the results of foaming of the liquid at the free surface, which effect does not usually appear with the inner stages.

The effect of downcomer height on residence time distribution, backmixing, and porosity of multi-stage reactors. We have found on basis of our experiments that there is no statistically significant effect between the values $\Theta_{Y_{max}}$, $Y_{\Theta} = 1$, E_{32}, E_{21}, ε_{G3}, ε_{G2} and ε_{G1} for various downcomer lengths in the range $L = 0{\cdot}1$ to $0{\cdot}42$ m. The independence of porosities and backmixing coefficients on the downcomer length is significant as the design of the downcomer does not affect the flow character inside the bed. This has verified one condition of our earlier derived theoretical model for the mechanism of backmixing[3] where we have assumed that the space below the downcomer pipe without regard to its length which is sufficiently long so that gas does not pass through it, is filled only with clear liquid. From comparison of experimental data of $\Theta_{Y_{max}}$ and $Y_{\Theta=1}$ for the downcomer lengths used in our study the residence time distribution in a three-stage reactor is approaching a serie of back-mix reactors with backmixing where in the region $W_L = 7{\cdot}5 \,.\, 10^{-3}$ m^3/m^2 s and $w_G =$ $= 0{\cdot}025$ to $0{\cdot}008$ m/s was found the existence of a completely isolated dead region ($a_1 \geqq 0{\cdot}95$).

For the zero length of downcomer pipes where in the plates were situated two circular holes having the diameter equal to that of downcomer pipes ($d_p = 0{\cdot}04$ m) the character of flow of both phases through the column was completely different from that where plates with downcomer pipes ($L \geqq 0{\cdot}1$ m) were used. Liquid does not

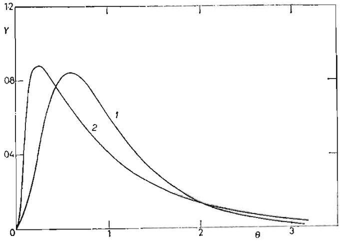

FIG. 3

Comparison of E-Curves in a Three-Stage Reactor for Plates with and without Downcomers $\varphi = 8\%$, $d_0 = 3$ mm, $W_L = 2{\cdot}5 \,.\, 10^{-3}$ m^3/m^2 s, $w_G = 0{\cdot}042$ m/s; *1* plates without downcomers, *2* plates with downcomers.

pass through the downcomer holes at all and is passing downward from one stage to the other nearly exclusively through the plate holes while the gas is passing across the plate through two downcomer holes so that it is not distributed by the plates and is rising through the reactor in a form of a S curve in a narrow stream formed by large clusters of bubbles. The bed with this type of plates used was not bubbled in the whole volume and, unlike the expected advantages resulting from the use of ordinary plates, they did not prevent formation of clusters of bubbles and their coalescence and thus they did not fulfill their ordinary function concerning the renewal of the interfacial area. Use of this type of plates in bubble-type reactors would thus be obviously disadvantageous. But what was interesting with such multistage column was that through the holes of downcomer pipes did not pass nearly any back flow of liquid though, at the first sight, it would be possible to expect that due to the whole gas flow passing thorough these holes validity of entrainment of liquid through these holes could be expected. The experiments have positively proved the model we have proposed earlier for formation of backmixing[3] as there is not developed the regime fulfilling the assumption on formation of the back flow as the result of hydrostatic pressure difference of the clear liquid in the downcomer pipe or in the space below it and the bubbled bed outside the downcomer.

For plates without downcomers with holes equally distributed as with the plates with downcomers we succeeded in stabilizing the regime in the column only for plates having $\varphi = 8\%$, $d_0 = 3$ or 5 mm at the liquid flow rate $W_L = 2{\cdot}5 \,.\, 10^{-3} m^3/m^2 \,.\, s$ and gas velocities $w_G = 0{\cdot}042$ and $0{\cdot}050$ m/s. For other values of liquid flow rates and plate parameters used it was not possible to get a steady bubbled bed in the reactor. Under conditions of stable bubbling a pulsating air layer 0·05 to 0·1 m high was formed below the plates which significantly limited the size of the back liquid flow between the stages. At $w_G = 0{\cdot}042$ m/s, $W_L = 2{\cdot}5 \,.\, 10^{-3} m^3/m^2$ s the values of the backmixing coefficient for plates without downcomers where $\varphi = 8\%$, $d_0 = 3$ mm were $E_{32} = 0{\cdot}053$, $E_{21} = 0{\cdot}118$, $\bar{E} = 0{\cdot}085$ while for plates with downcomer pipes ($L = 0{\cdot}42$ m) at otherwise same conditions $E_{32} = 3{\cdot}202$, $E_{21} = 3{\cdot}847$, and $\bar{E} = 3{\cdot}525$.

In Fig. 3 a comparison of E-curves is made obtained at otherwise same conditions ($w_G = 0{\cdot}042$ m/s, $W_L = 2{\cdot}5 \,.\, 10^{-3} m^3/m^2$ s, $\varphi = 8\%$, $d_0 = 3$ mm) for both types of plates. From this figure is obvious a difference of the residence time distribution of the liquid phase resulting from the decrease in backmixing when plates without downcomers are used. But the disadvantage of these plates is in their small stability of conditions on the plate which significantly reduces and complicates their application. In case of using plates without downcomers it would be necessary for the sake of stability of the regime to use larger free plate areas which would be disadvantageous for formation of a well bubbled bed. For these reasons the plates with downcomers appear to be most preferable.

LIST OF SYMBOLS

a_1	volume ratio of active ideally mixed part to over-all stage
a_2	ratio of liquid flow rate into the active part of the stage to net liquid flow rate into the system
a_3	ratio of flow rate between the active and inactive part of the stage to the net liquid flow rate into the system
d_0	diameter of plate holes (L)
$E = Q/\dot{V}_L$	backmixing coefficient
$\bar{E}$	mean value of the backmixing coefficient
L	length of downcomer pipe
Q	absolute back flow (L^3/T)
$\dot{V}$	volumetric flow rate (L^3/T)
w	superficial velocity
W_L	liquid feed rate (L^3/L^2T)
Y	dimensionless concentration
$Y_{\Theta=1}$	dimensionless concentration corresponding to the mean dimensionless time $\Theta = 1$ on the E-curve
ε_G	porosity of bubbled bed
Θ	dimensionless time
$\Theta_{Y\max}$	dimensionless time corresponding to the maximum dimensionless concentration $Y_{\max}$ on the E-curve
φ	relative free plate area

Subscripts

G	gas	3	bottom stage
L	liquid	i	i-th stage
1	top stage	21	between the 1st and 2nd stages
2	middle stage	32	between the 2nd and 3rd stages

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Translated by M. Rylek.