DISTRIBUTION OF VELOCITY COMPONENTS AND LIQUID HOLDUP IN SHORT VERTICAL TUBES WITH SWIRL BODIES IN THE INLET PART OF THE TUBE AT COCURRENT GAS–LIQUID FLOW

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The axial and tangential velocity components and local liquid holdup rates have been measured in vertical tubes with swirl bodies located in the inlet part of the tube. The tubes were of 70 mm I.D. and $H/D = 10-23$. The air–water flow was directed upward. Superficial gas velocity was $\overline{w}_G = 14-35 \text{ m s}^{-1}$ and specific liquid load $\dot{Q}_{LE} = 15-65 \text{ m}^3 \text{ m}^{-2} \text{ h}^{-1}$. In the experiments, the blade angle of the swirl body and the liquid inlet have been altered. The resulting centrifugal moments were correlated with friction factors.

The rotating gas–liquid flow has drawn a considerable attention of chemical engineers. In such flow the mass and heat transfer rates can be considerably increased by increase in the gas flow rate. This results in the increase of the intensity of operation per unit volume of the contact element.

In the hydrodynamics of pure gas phase centrifugal flow[1-7], in the presence of the liquid phase [8-12] too, a reasonable progress has been achieved. Solution of the basic flow equations mostly numerically, has been attempted[1,2]. Some studies have been devoted to experimental determination of velocity components by small probes[3,12,13] and to modeling of the flow field by use of suitable empirical parameters[6,12]. But there remains a lot of work to be done in which the flow parameters should be related to mass transfer efficiency. This problem can be attacked by relating the pressure drop to the field parameters such as velocities, effective and local phase densities and tube dimensions so that the nature of the flow resistance could be explained.

This study has been devoted to hydraulic investigation of short contact tubes ($H/D = 10-23$) with swirl bodies located in the inlet part of these tubes. Other authors[3,9] have fixed this parameter mostly at values under 10. Velocities of both phases were measured by use of the Pitot tube which was applied successfully also at the conditions of the centrifugal field measurements[4,12]. Of interest is also the effect of liquid inlet arrangement on the centrifugal field which is important for achieving a complete liquid separation. Thus some changes in the design of the liquid inlet were made both below and above the swirl body.

EXPERIMENTAL

The experiments were performed in a tube with swirl bodies situated at its entrance inlet part. The geometry of swirl bodies has been altered both in their arrangement and blade angle (Fig. 1 and Table I). Other details concerning the experimental unit were given in the preceding paper[14].

The operating variables were: superficial gas velocity $\overline{w}_G = 14-35 \text{ m s}^{-1}$, specific liquid load $\dot{Q}_{LE} = 15-65 \text{ m}^3 \text{ m}^{-2} \text{ h}^{-1}$. The experimental arrangement consisting of the adjustable Pitot tube, manometers and a liquid probe line, is also depicted in Fig. 1.

The dynamic component has been obtained from the over-all pressure. The main stream velocity vector has been calculated by the following procedure: The pressure drop is dependent on the angle of inclination of the Pitot tube[4] (see Fig. 2) so that on basis of the measured over-all pressure the static pressure can be obtained by rotation of the Pitot tube for a certain angle. Readings at two angle positions were taken, B_1 and B_2, in which the total pressure was equal to the static pressure measured at the wall of the chamber[14] at the same value $z^+ = Z/H$. The local dynamic liquid holdup $\dot{\varphi}_{LK}$ can be measured easily by fixing the position of the sampling, probe in the direction of the main stream and by sucking and measuring the liquid volume which have reached the entrance of the probe.

For the sake of accuracy much care has been devoted to the experimental procedure: Before every pressure reading, the probe was dried by passage of air delivered by the ball pump. By use of a second probe, it was determined in the distance of 20 mm above the first probe that disturbances were already damped. In accordance with the unpublished study[15–18] the ratio of the probe to tube diameters 0·05 has been found as optimal for local measurements. The effect of probe dimensions on local liquid profiles and therefore on $\dot{\varphi}_{LK}$ was negligible in the range of suction pressures $\Delta p = -(0{\cdot}5-2)$ kPa which was in agreement with the unpublished study[19]. The radial

TABLE I

Some details of the studied variants

Variant code	Liquid inlet by	Swirl body arrangement	Swirl body angles
$A_K^{a,c}$	lance, upward directed	under the	58·5°
$A_{K,3}^{a}$	lance, upward directed	liquid inlet	30°; 44·5°; 73°
$A_{K,k}^{a}$	peripher, 3 sockets	liquid inlet	30°; 44·5°; 73°
$A_{L,1}^{b}$	peripher, 3 sockets	liquid inlet	30°; 44·5°; 73°
$A_{L,2}^{b}$	lance, upward directed	liquid inlet	30°; 44·5°; 58·5°
B_K^{a}	lance, downward directed	above the	30°; 44·5°; 73°
B_L^{b}	lance, downward directed	liquid inlet	30°; 44·5°; 73°
C_K^{a}	lance, downward directed	without swirl b.	—
C_L^{b}	lance, downward directed	without swirl b.	—

[a] $H/D = 12{\cdot}9$; [b] $H/D = 22{\cdot}8$; [c] under the swirl body initial stream tube part of $H/D = 10{\cdot}3$; in all other variants without this part. For liquid inlet of variant A see Fig. 1, for variant B and C[14].

gas flow component in 70 mm tubes could have been neglected. The effect of liquid droplets on the velocity measurement was proved by solving the Pitot probe relations of Gill and Hewitt[16] and of others[17]. On the basis of varying density and impulse velocity definitions for our experimental data a negligible effect of deviations from the main flow direction for $\pm 6\%$ has been found. For all measurements the relation holds

$$\left(\dot{V}_{GE}/\pi R^2 \overline{w}_G\right) = 2 \int_0^1 \left(w_{Gz}/\overline{w}_G\right) r^+ \, dr^+ \qquad (1)$$

which should be theoretically equal to one, was always in the range 0·91 — 1·03 at a high statistical level of significance. Unfortunately, a correction of pressure values must be taken in two-phase flow for angles deviating from the main stream direction (Fig. 2). It should be also mentioned that the profiles (left part of Fig. 2) are not strongly symmetrical. For the changed variant A_L, *i.e.* at low liquid loads, the profiles are affected even at $z^+ = 15{\cdot}4$ by the arrangement of the liquid inlet tube (lance). At higher $\dot{Q}_{LE}$ this effect is not so profound.

Data Interpretation

Due to some irregularities in the velocity profiles the main results were interpreted in the form of a momentum balance analysis. By extension of the semi-empirical theory, where only the gas

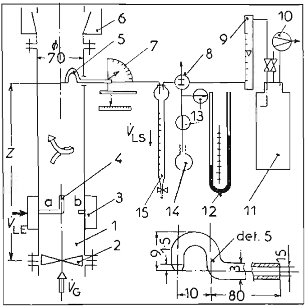

FIG. 1

Experimental device for the local velocity and hold-up measurements. *1* Flow tube, *2* swirl body, *3* liquid inlet chamber, *4* liquid inlet tube, *5* Pitot probe, *6* liquid–gas separator, *7* coordinate table, *8*, *13* valves, *9* rotameter, *10* pump, *11* bubble vessel, *12* manometer, *14* ball pump, *15* burette

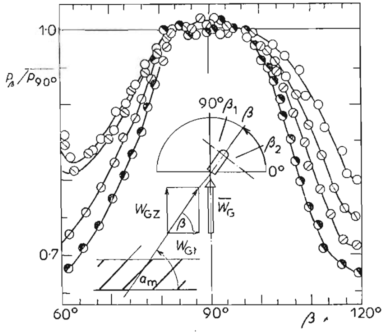

FIG. 2

Dimensionsless pressure in dependence on the flow angle β and the specific liquid load $\dot{Q}_{LE}$, $m^3 \, m^{-2} \, h^{-1}$ (○ 0, ⊗ 16·4, ⊘ 31·7, ◕ 44·7). Variant A_L, $\alpha = 58{\cdot}5°$, $Fe_G = (0{\cdot}6 - 1{\cdot}3) \cdot 10^4$, $z^+ = 15{\cdot}4$. With angle measurement techniques phase relations appear[6], the following dimensionless moments were obtained

$$M_K^+ = 4(\bar{\varrho}/\varrho_G) \int_0^{1-\delta_F/R} (w_{GzK}/\bar{w}_G)(w_{GtK}/\bar{w}_G)(r^+)^2 \, dr^+ \tag{2}$$

$$M_F^+ = 4(\varrho_L/\varrho_G)(\bar{w}_F/\bar{w}_G)(\delta_F/R)\cos\beta_F \tag{3}$$

$$M_D^+ = 2[1 + (F/G)][\cos\alpha/(A_D/A_0)]\,\bar{r}^+ \tag{4}$$

with

$$\bar{r}^+ = \sqrt{[1 - (r_N^+)^2]/2} \tag{5}$$

$$A_D/A_0 = [1 - r_N^+][(1 + r_N^+)\sin\alpha - (md/\pi R)] \tag{6}$$

The still unknown film thickness δ_F was estimated on the basis of the next assumptions: By introduction of the film parameter $\gamma_F = V_F/V_R = 2(\delta_F/R)$ and both the measured[15] film separation degree $\dot{\varepsilon}_F = \dot{V}_F/\dot{V}_{LE}$ and static liquid hold-up $\bar{\varphi}_L = V_L/V_R$ the corresponding core values were obtained

$$\bar{\varphi}_{GK} = V_G/V_K = (1 - \bar{\varphi}_L)/(1 - \gamma_F) \tag{7}$$

$$\dot{\varepsilon}_{GK} = \dot{V}_G/[\dot{V}_G + (1 - \dot{\varepsilon}_F)\dot{V}_{LE}] \tag{8}$$

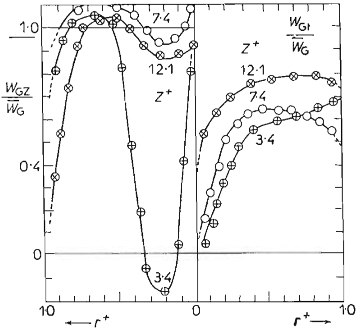

FIG. 3

Dimensionless axial ($w_{Gz}/\bar{w}_G$) and tangential ($w_{Gt}/\bar{w}_G$) velocity components of the pure gas in dependence on the dimensionless radius r^+. Variant A_L, $\alpha = 58{\cdot}5°$, $Re_G = 8{\cdot}8 \cdot 10^4$

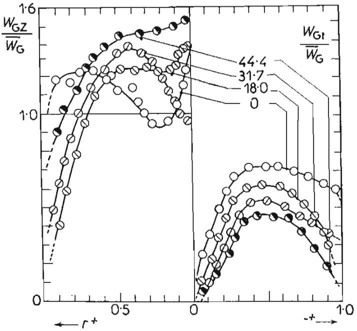

FIG. 4

Dimensionless axial and tangential velocity components in dependence on the dimensionless radius r^+. Variant A_L, $\alpha = 58{\cdot}5°$, $z^+ = 7{\cdot}4$ and specific liquid load $\dot{Q}_{LE}$ as parameter

At the assumption of equal gas and liquid core velocities, Eqs (7) and (8) must be identical. As the first approximation the film thickness is obtained

$$(\delta_F/R) = 0{\cdot}5[\bar{\varphi}_L + (1 - \bar{\varphi}_L)(1 - \bar{\varepsilon}_F)(\dot{V}_L/\dot{V}_G)] \quad (9)$$
$$(0{\cdot}5 < 10^2\delta_F/R \lesssim 3)$$

which was substituted in Eqs (2) and (3). The film velocity (always in the order of $\bar{w}_G \approx 1 \text{ m s}^{-1}$) which is needed in Eq. (2) was measured experimentally[15].

The disadvantage of the majority of hydraulic models for flow in tubes is that the pressure drop cannot be obtained directly from operation variables, but that a preliminary determination of the friction factor ξ and of additional parameters is needed. To overcome this problem the friction factor will be represented as a function of the centrifugal force moments.

By neglecting, in the momentum balance for the heterogeneous flow model, the smaller[15] tangential stress forces and at larger velocities also the gravitational forces (*i.e.* see[20]) the acceleration terms remain as variables. With the friction factor calculated according to[14], which is measured on the tube wall, the relation is obtained

$$\xi = 2\,\Delta p/(\varrho_G \bar{w}_2^2) \sim \sum M^+ \quad (10)$$

with

$$(\sum M^+/M_D^+) = 1 - (M_K + M_F)/M_D \quad (11)$$

Therefore it seems reasonable to express ξ as a function of Eq. (11). On basis of the plot of experimental data for pure gas phase experiments the upper and lower tube moments should be equal on basis of the momentum conservation law, *i.e.* $M_G = M_D$ with $\xi = \xi_0$. As the consequence of this procedure the correlation of the structure (14) is obtained.

RESULTS AND DISCUSSION

Axial velocity (left parts) and the characteristic tangential velocity (right parts) distributions are demonstrated in Figs 3, 4 and 5. For a short distance above the swirl body a strong influence of the core flow is noticed, especially by the liquid inlet tube of variants of the type *A*. With increasing z^+, the back flow zone vanishes particularly but can be observed even at $z^+ \approx 20$ in all cases. On the other hand the development of potential whirls is always disturbed. This is a typical picture for high friction forces in shorter and narrower tubes. As expected at high values z^+ for the *B*-type variants, the disturbance of profiles is smallest as the effect of the inlet zone is calmed by the swirl body.

More regular is the behaviour of the experimental local dynamic liquid hold-ups as can be seen from Fig. 6. At small z^+ the liquid stream is strongly distributed in the whole cross sectional area. However, at higher gas velocities the largest part of liquid is concentrated in the wall region. For medium values of z^+, the spray content is larger in the neighbourhood of the wall. At $z^+ \gtrsim 7$ most of the liquid phase is concentrated in the region which accounts for about 15% of the tube cross section which means that the largest part of liquid flow is in the film on the wall.

For all conditions considered, the experimental data can be correlated for larger distances ($z^+ \gtrsim 8$) by the relations (*1*),

$$w_{Gt}/\bar{w}_G = \left[0{\cdot}096 \operatorname{tg}(90 - \beta) + 0{\cdot}01\right]\left[1 - 1{\cdot}21 \cdot 10^{-4}(F/G)\right] \cdot \left[5{\cdot}7r^+/(1 + 8{\cdot}2(r^+)^2)\right] - 0{\cdot}91 \operatorname{tg}(90 - \beta) + 2{\cdot}50 \qquad (12)$$

$$\dot{\varphi}_{LK} = 8{\cdot}20 \cdot 10^{-4}(F/G)^{0{\cdot}3}(r^+)^{3{\cdot}5}\left[1{\cdot}81 + \operatorname{tg}(90 - \beta)\right] \qquad (13)$$

with $30° \leqq \alpha < 90°$, the correlation coefficient equals to 0·95.

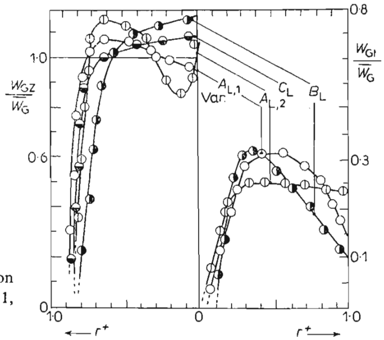

FIG. 5

Dimensionsless velocities in dependence on r^+ for some variants (Table I). $z^+ = 20{\cdot}1$, $Re_G = 1 \cdot 10^5$, $\dot{Q}_{LE} = 65\ m^3\ m^{-2}\ h^{-1}$

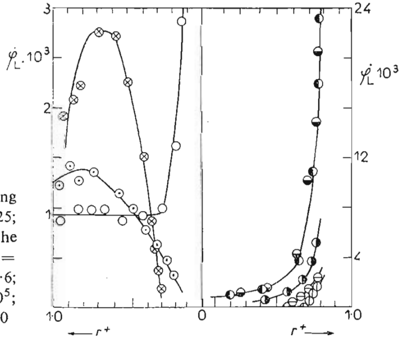

FIG. 6

Local dynamic liquid hold-up $\dot{\varphi}_L$ depending on r^+ at the left side: $z^+ = 3{\cdot}4$; $F/G = 0{\cdot}25$; $Re_G = 10^{-4}$; ○ 6·6; ⊙ 8·7; ⊗ 11·0. At the right side: $F/G = 0{\cdot}28$; *a*) $z^+ = 7{\cdot}4$; $Re_G = 6{\cdot}6 \cdot 10^4$; $\dot{Q}_{LE}$; $m^3\ m^{-2}\ h^{-1}$: ⦶ 21·6; ⊖ 31·7. *b*) $z^+ = 12{\cdot}1$; $Re_G = 1{\cdot}3 \cdot 10^5$; $\dot{Q}_{LE}$, $m^3\ m^{-2}\ h^{-1}$; ◑ 21·6; ◐ 31·7; ◒ 65·0

From Eqs (*9*) and (*10*), the relation is obtained (Fig. 7)

$$(\xi/\xi_0) = 2{\cdot}32[M_D/(M_K + M_F)]^{0.8} \qquad (14)$$

with correlation coefficients $0{\cdot}95(44{\cdot}5° \leqq \alpha < 90°)$ or $0{\cdot}89(30° \leqq \alpha < 90°)$.

By use of Eq. (*14*) the flow properties are directly related to the friction factor. With reference to Karpenkov[6], the proposed model can be extended, so that the dimensionless velocities $(w_{Gzmax}/\bar{w}_G)$ and $(w_{Gtmax}/\bar{w}_G)$ for pure gas flow are directly related to geometrical variables of the swirl body and the tube. By use of the relation $M_G = f(\xi/\xi_0)\, M_D$ of Eq. (*14*), it is now possible to estimate the effect of the liquid phase flow rate on gas velocity components in the swirled two-phase flow and thus on the friction factor.

LIST OF SYMBOLS

A area m^2
D diameter m
d thickness of swirl blades m
F/G liquid-gas mass flow ratio —
H height m
M centrifugal force moment Nm
m number of swirl blades —
P pressure Nm^{-2}

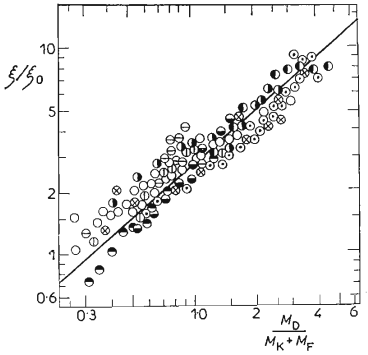

FIG. 7

Friction factor related to pure gas phase in dependence on the quotient of the centrifugal moments for all geometrical variants

Q specific volume $m^3\ m^{-2}$
r, R radius m
V volume m^3
w velocity $m\ s^{-1}$
α blade angle of swirl body degree
β angle of flow vector degree
γ (film) parameter —
δ (film) thickness m
ε separation rate —
ϱ density $kg\ m^{-3}$
ξ resistance coefficient —
φ hold-up —

Dimensionless quantities

$M^+ = 2M/\pi R^3 \varrho_G \bar{w}_G^2$
$\mathrm{Re} = \bar{w}D/\nu$
$r^+ = r/R$
$z^+ = z/H$

Subscripts

D swirl body
E inlet
F film
G gas
K gas core
L liquid
N nave
R reactor
t tangential
z axial
o without liquid
. time releted
— mean

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