THE EFFECT OF TWO-PHASE FLOW REGIME ON HYDRODYNAMICS AND MASS TRANSFER IN A HORIZONTAL-TUBE GAS-LIQUID REACTOR

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The effect of two-phase flow regime on decisive hydrodynamic and mass transfer characteristics of horizontal-tube gas–liquid reactors (pressure drop, liquid holdup, $k_L a_L$) was determined in a co-current-flow experimental unit of the length 4·15 m and diameter 0·05 m with air–water system. An adjustable-height weir was installed in the separation chamber at the reactor outlet to simulate the effect of internal baffles on reactor hydrodynamics.

Flow regime maps were developed in the whole range of experimental gas and liquid flow rates both for the weirless arrangement and for the weir height 0·05 m, the former being in good agreement with flow-pattern boundaries presented by Mandhane. In the whole range of experimental conditions pressure drop data could be well correlated as a function of gas and liquid flow rates by an empirical exponential-type relation with specific sets of coefficients obtained for individual flow regimes from experimental data. Good agreement was observed between values of pressure drop obtained for weirless arrangement and data calculated from the Lockhart–Martinelli correlation while the contribution of weir to the overall pressure drop was well described by a relation proposed for the pressure loss in closed-end tubes. In the region of negligible weir influence values of liquid holdup were again succesfully correlated by the Lockhart–Martinelli relation while the dependence of liquid holdup data on gas and liquid flow rates obtained under conditions of significant weir effect (*i.e.* at low flow rates of both phases) could be well described by an empirical exponential-type relation.

Results of preliminary $k_L a_L$ measurements confirmed the decisive effect of the rate of energy dissipation on the intensity of interfacial mass transfer in gas-liquid dispersions.

In last few years a clearly apparent trend can be observed toward diversification of reactor types used for chemical and biotechnological processes in gas–liquid and gas–liquid–solid (susp.) systems. Apparently such a development has been evoked primarily by economical reasons namely by increasing energy and investment costs. Numerous novel types of gas–liquid (gas–liquid–solid) reactors have been designed and tested lately with the aim to find the most suitable arrangements for specific reaction processes regarding the demands on maximum process efficiency (reactor productivity) at minimum energy and investment requirements. Indeed the requirement of reactor optimization becomes imperative in the case of secondary (non--productive) processes such as solid and liquid wastes treatment, air polution prevention *etc.* which have been gaining increasing importance from ecological view point.

In many cases horizontal-tube two-phase flow reactors can be recommended as the most suitable for such processes. Their advantages include simple contruction, low energy and investment requirements, flexibility of process conditions, simple pressure and temperature control, and limited foam formation even in strongly foaming systems. The most attractive feature of these reactors however seems to be the possibility of direct use of transport pipelines for contacting of phases. In such a case the costs of additional reactor vessel can be saved and the profit resulting from such arrangement apparently outweights by far the relatively low energy effectivenes (low efficiency of interfacial area formation) reported by Nagel and coworkers[1] as a drawback of horizontal-tube reactors. An interesting example of such an application was reported by Kubička and Kvapil[2] who described the use of long transport pipeline (2 000 m) for rafination of waste waters from the process of partial oxidation of mazut in chemical enterprises CHZ Litvínov (Czechoslovakia). Oxidation of hydrogen cyanide and hydrogen sulphide by oxygen from air has been carried out in the pipeline instead of subsequent oxidation by sodium hypochlorite in storage tanks.

Wider use of horizontal-tube gas–liquid reactors has been however hampered by the lack of reliable relations for reactor design/and/or calculation of reaction processes in such type of equipments. Considerable attention has been paid up to now to studies of two-phase flow of gas–liquid mixtures in small diameter tubes ($D \leqq 0{\cdot}025$ m). Results of such studies aimed primarily at establishing a data base for heat-exchangers design are summarized *e.g.* in extensive review papers of Scott[3] and Hewitt[4]. Due to their purpose these studies have been devoted mostly to characterization of two--phase flow patterns and determination of their limits as well as to determination of pressure drop and heat transfer characteristics. Much less has been however known about the rate of mass transfer under conditions of individual two-phase flow regimes and about the relation between parameters characterizing hydrodynamics and mass transfer in horizontal-tube gas–liquid reactors. Our present work reports the results of the first part of extensive experimental program aimed at obtaining a useful data base for horizontal-tube reactors design and for the choice of optimum working conditions of these reactors for specific reaction processes. Indeed such an orientation toward reactor design had to be reflected even in design of the experimental apparatus having consequently substantial construction alterations (namely larger tube diameter and lower L/D ratio) in comparison with units used for purely hydrodynamic or heat transfer studies. It was therefore the main goal of this work to test the experimental unit to be used in the whole long-term research program and to determine fundamental hydrodynamic characteristics of the reactor fo standard air-water system. Our experimental program included determination of flow-pattern map and measurements of two-phase flow pressure drop and liquid holdup in the whole range of two-phase flow regimes studied. Special attention was paid to values of these parameters in regions of transition between individual flow regimes. In addition, values of $k_L a_L$ were measured in limited range of gas and liquid flow rates to test the experimental method and to obtain preliminary informations on the relation between hydrodynamics and the rate of interfacial mass transfer in reactors of this type. Such preliminary results should serve as starting point for subsequent detailed study of this topic.

EXPERIMENTAL

Equipment. The experimental set-up is shown in Fig. 1, detailed construction chart of the horizontal-tube reactor is given in Fig. 2. The reactor consisted of the mixing chamber and phase-separating chamber at the apparatus inlet and outlet respectively and of the contacting part made of perspex pipe 0·05 m I.D. of the overall length 4·15 m ($L/D = 83$). The contacting tube was subdivided into flow-stabilizing, measuring and terminal sections (Fig. 2). Its diameter

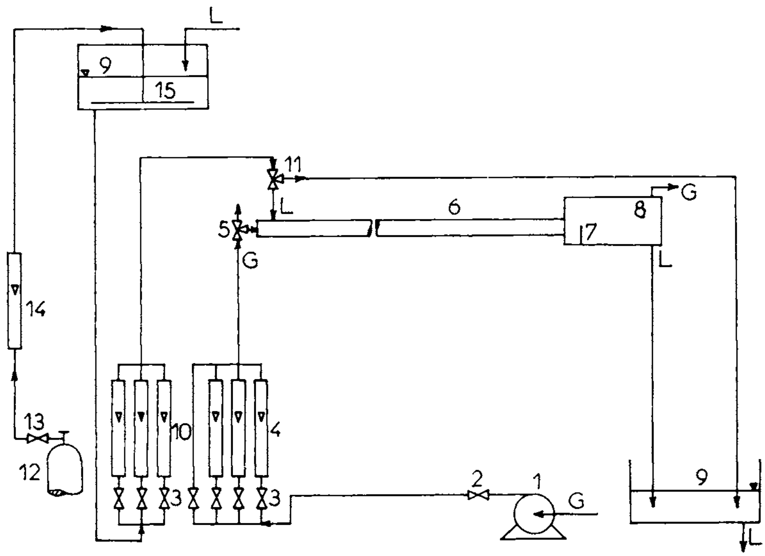

FIG. 1

Experimental set-up. *1* Gas blower, *2* reducing valve, *3* valves, *4* gas-flow rotameters, *5* three-way valve, *6* horizontal-tube reactor, *7* weir, *8* phase separator, *9* storage tank, *10* liquid-flow rotameters, *11* three-way valve, *12* pressure cylinder (nitrogen), *13* reducing valve, *14* rotameter, *15* gas distributor was chosen as a compromise between the requirement of large diameter value needed for wal effects elimination and the necessity to keep sufficiently large values of L/D ratio to ensure appropriate flow-pattern stabilization prior to measuring section inlet. It has been assumed that the diameter value 0·05 m represents a reasonable approximation of real conditions allowing for application of experimental data for reactor modeling and scale up purposes. A weir with adjustable height was built in the separation chamber to enable liquid holdup variations and stabilization of flow regimes at low flow rates of phases. The tube was suspended flexibly on the supporting construction to damp the vibrations caused by the two-phases flow, measures were also taken to prevent the transmission of pump vibrations on the tube by the water and air supply piping.

Four pressure gauges were situated along the tube length (Fig. 2) and two electric conductivity cells were placed at both ends of reactor measuring section. Experiments were carried out with air–water system at atmospheric pressure and room temperature ($t = 20°C$). Tap water was used as a liquid phase, air was supplied by a blower. Flow rates of phases ranged between 0·71 and 353 m^3/h or 0·035 and 4·6 m^3/h for air and water respectively, corresponding superficial flow rates thus varied between 0·1 and 50 m/s for gaseous phase and 0·005 and 0·65 m/s for liquid phase.

Measuring methods. Flow regimes were identified visually and from photographs of the two-phase layer in the measuring (test) section. Simple U-tube manometers were used for pressure-drop measurements. Value of total pressure drop, Δp_0, was determined over the whole tube length (*i.e.* between the pressure outlets 0 and 11, see Fig 2), values Δp_e and Δp_{RS} corresponded to the entering section (pressure outlets 0 and 1) and to the test section of the horizontal tube (outlets 2 and 8) respectively.

Values of relative liquid holdup in the test (stabilized-flow) section, $\varepsilon_L = V_L/V_{RS}$, were determined from the mean residence time of the liquid phase in this section, $\bar{\tau}_{RS}$, using the relation

$$\varepsilon_L = \dot{V}_L/V_{RS}\bar{\tau}_{RS} \qquad (1)$$

where V_{RS} denotes the overall volume of the test section and $\dot{V}_L$ volumetric liquid flow rate. Values of mean residence time, $\bar{\tau}_{RS}$, were obtained as a difference of the first moments of responses to the pulse input signal determined at the test section inlet and outlet. Conductivity method was applied for residence time distribution measurements, schema of the measuring circuit is shown in Fig. 3. A pulse input of KCl solution was introduced to the system at the liquid inlet, the system responses were determined by two semicircle electrodes placed at both ends

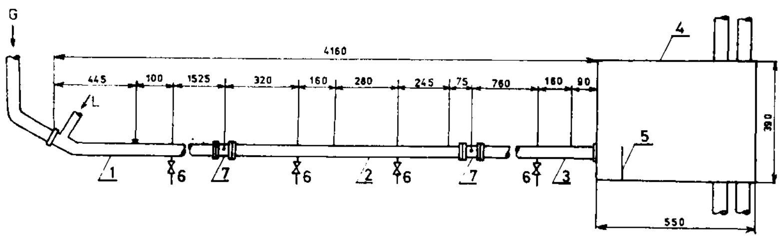

FIG. 2

Horizontal-tube reactor. *1* Flow-stabilizing section, *2* measuring section, *3* terminal section, *4* phase-separation chamber, *5* weir, *6* pressure gauges, *7* electric conductivity cells of the test section and monitored by fast speed recorders (Vareg — 2). Responses data were also digitalized and punched on tape. The first moments of response curves $\mu_1 = \bar{\tau}_1$ and $\mu_2 = \bar{\tau}_2$ were evaluated from experimental response data and the mean resdidence time of the liquid phase in the test section was calculated as a difference of both moments, $\bar{\tau}_{RS} = \bar{\tau}_2 - \bar{\tau}_1$. For comparison liquid holdup data were determined also by weighing the residual liquid in the tube after simultaneous shut off the supply of both phases (see the three-way valves in Fig. 1). Both methods were proved to yield comparable ε_L data in the whole ranges of gas and liquid flow rates used in our work.

Values of volumetric liquid-side mass transfer coefficient, $k_L a_L$, charactering the intensity of interfacial mass transfer in the system were determined by a steady state method from the balance of oxygen during its steady-state absorption into deoxygenized water. Water was deoxygenized in a storage tank (Fig. 1) by nitrogen bubbling and fed continuously into the reactor. The steady-state concentration of oxygen dissolved in the liquid phase was determined in samples taken off at both ends of the test section using a standard titration method[5]. Values of $k_L a_L$ were then calculated from relation

$$k_L a_L = (\dot{V}_L(c_{Lj} - c_{Li})/V_{RS}) \left(((c_L^+ - c_{Li}) - (c_L^+ - c_{Lj}))/\ln \frac{c_L^+ - c_{Li}}{c_L^+ - c_{Lj}} \right)^{-1}, \quad (2)$$

where c_{Li} and c_{Lj} denote oxygen concentrations at the test section inlet and outlet respectively and c_L^+ is the equilibrium oxygen concentration in water.

RESULTS AND DISCUSSION

Flow Pattern Identification

Values of gas and liquid flow rates corresponding to transition between individual two-phase flow regimes are plotted in Fig. 4 and compared with the flow-pattern map presented by Mandhane and coworkers[6]. It can be seen from the figure that dispersed flow regime was not observed in the experimental region of gas and liquid flow rates,

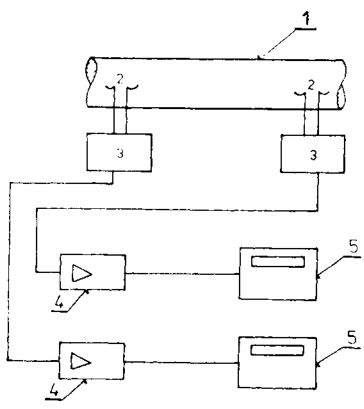

Fig. 3

Schema of the circuit for residence time distribution measurements. *1* Horizontal-tube reactor, *2* conductivity electrodes, *3* Wheatstone bridges, *4* amplifiers, *5* recorders the bubble-, plug-, slug- and annular- (annular mist) flow regimes were studied in a limited range of liquid flow rates. It is apparent from the figure that the experimental data obtained at zero height of the separation-chamber weir agreed fairly well with Mandhane's results. Such an agreement not only confirms the suitability of our unit (specific reactor configuration used) for two-phase flow studies but above all it proves that our experimental data including results of mass transfer measurements are fully compatible with other two-phase flow studies reported in literature. Consequently our conclusions on the mutual relation between character of two-phase flow and intensity of mass transfer in the reactor (see further) can be considered as generally valid for weirless horizontal-tube contactors. On the other hand data obtained at the weir height 0·05 m (Fig. 4) prove a significant effect of the weir on flow regime boundaries namely at low gas and liquid flow rates. It is obvious that the stratified-flow region was severely reduced under such conditions while wavy and plug (bubble) flow regimes developed at significantly lower liquid flow rates apparently due to "feedback" effect of weir on the character of flow in the tube. With increasing flow velocities the effect of weir ceased and it became negligible for $w_G > 5$ m/s and $w_L > 0.2$ m/s. Apparently these results suggest that in short and medium length horizontal-tube reactors (up to $L/D \approx 100$) character of flow and consequently the efficiency of phase contact at low flow rates can be positively influenced by separation chamber construction (weir height). Stratified flow regime undesirable from the viewpoint of phase contact and mass transfer intensity can be eliminated by weir installation and the plug and/or bubble flow regime can be created at significantly lower liquid flow rate. Comparable mass transfer intensity can thus be achieved at significantly lower energy input to the system. Indeed it is obvious that the

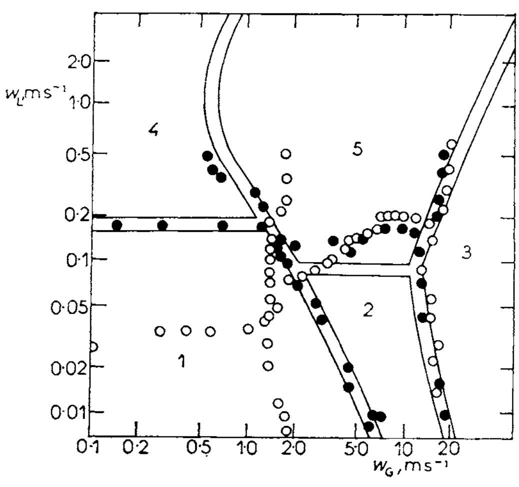

FIG. 4

Two-phase flow regime identification — comparison of experimental data with regimes transition boundaries of Mandhane[6] flow-pattern map. ● $H_w = 0$, ○ $H_w = 0.05$ m, Flow regimes: 1 stratified flow, 2 wavy flow, 3 annular, annular mist flow, 4 bubble, elongated bubble (plug) flow, 5 slug flow effect of separation-chamber weir on reactor hydrodynamics (flow patterns) decreases with increasing tube length and it would be virtually negligible in long (transport--piping type) reactors. In these cases however similar effect can be achieved by segmentation of such pipeline reactors by internal baffles.

Two-Phase Flow Pressure Drop

In Figs 5 and 6 data of total pressure drop, Δp_0, determined over the whole tube length have been plotted in logarithmic coordinates against superficial gas velocity for the whole experimental range of liquid flow rates and for both configurations of separation chamber (*i.e.* with or without a weir). It can be seen from Fig. 6 that in the absence of weir ($H_w = 0$) the dependences $\log(\Delta p_0/L)$ *vs* $\log w_G$ were linear within the limits of individual two-phase flow regimes in the whole range of experi-

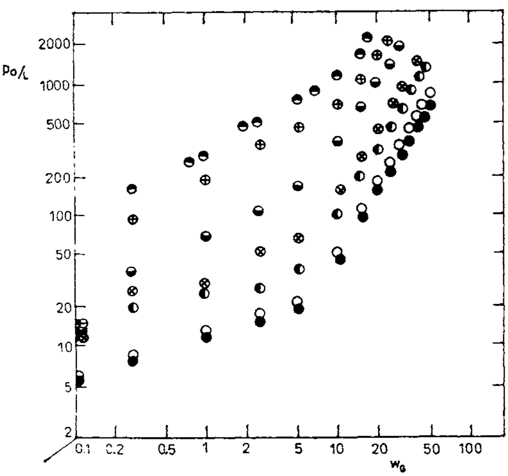

Fig. 5

Dependence of two-phase flow pressure drop on gas flow rate; $H_w = 0{\cdot}05$ m; ● $w_L = 0{\cdot}005$ m/s, ○ $w_L = 0{\cdot}01$ m/s, ◐ $w_L = 0{\cdot}05$ m/s, ⊗ $w_L = 0{\cdot}10$ m/s, ⊖ $w_L = 0{\cdot}25$ m/s, ⊕ $w_L = 0{\cdot}50$ m/s, ⊖ $w_L = 0{\cdot}65$ m/s

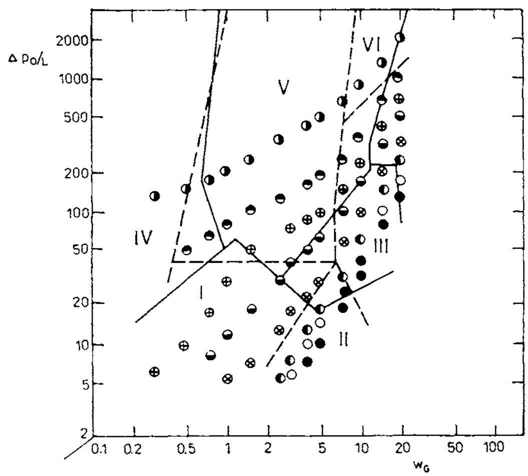

Fig. 6

Dependence of two-phase flow pressure drop on gas flow rate at zero weir height. ● $w_L = 0{\cdot}005$ m/s, ○ $w_L = 0{\cdot}01$ m/s, ◐ $w_L = 0{\cdot}025$ m/s, ⊗ $w_L = 0{\cdot}05$ m/s, ⊖ $w_L = 0{\cdot}10$ m/s, ⊕ $w_L = 0{\cdot}15$ m/s, ⊖ $w_L = 0{\cdot}25$ m/s, ◑ $w_L = 0{\cdot}50$ m/s. ----- validity limits of sets of coefficients K_1, K_2, K_3 (Table I), ——— limits of two-phase flow regimes taken from Mandhane flow-pattern map[6] mental gas and liquid flow rates. Apparently it can be concluded that the flow-pattern transitions have been solely responsible for the changes of slopes of pressure drop dependences. On the other hand distinctive non-linearities (breaks) can be observed in Fig. 5 on pressure drop dependences obtained at the presence of weir ($H_w = 0{\cdot}05$ m) even within individual flow-pattern limits. These non-linearities have been clearly apparent namely at lower gas and liquid loads in regions of superficial velocities $w_L = 0{\cdot}005-0{\cdot}15$ m/s, $w_G = 0{\cdot}28-5{\cdot}0$ m/s. Correspondingly also, significantly higher values of pressure drop were observed at the presence of weir for otherwise identical conditions (w_L, w_G = const.). Apparently such a pressure drop increase can be explained by the combined effect of the weir resistence and of the flow reflection at the weir resulting in the countercurrent flow and shock-wave formation in the tube. Understandably this additional weir effect has been more significant at lower flow rates of phases. For evaluation of the weir contribution to the overall pressure drop following relation can be used proposed[7] for calculation of the pressure loss in closed-end tubes

$$\Delta p_w = \varrho_L a_F w_L \tag{3}$$

where a_F denotes the velocity of disturbance propagation (backflow velocity). According to literature[7] application of Eq. (3) has been limited to cases when $w_L \ll a_F$. It was assumed on the basis of our observations that this condition was fulfilled in the whole experimental region of flow rates. Values of additional pressure drop contribution Δp_w for air-water flow ($\varrho_L = 1\,000$ kg/m^3, $w_L = 0{\cdot}15$ m/s) were evaluated from Eq. (3) for backflow velocities, a_F, ranging between 10 and 100 m/s. Calculated values $\Delta p_w/L = 3-30$ Pa/m agreed fairly well with our experimental data. Our conclusions on the effect of weirs on flow-pattern limits and pressure drop increase correspond well with results reported by Wallis and Dobson[8] who observed considerable increase of pressure drop as well as decrease of gas flow rates corresponding to the onset of slug- or plug-flow regime under conditions of flow reflection at the tube end.

It was proved that within the experimental range of flow rates ($w_L = 0{\cdot}005-0{\cdot}65$ m/s, $w_G = 0{\cdot}3-50$ m/s) our pressure drop data could be well correlated by exponential-type relation

$$\Delta p/L = K_1 w_G^{K_2} w_L^{K_3} \tag{4}$$

for both weir heights. Values of coefficients K_1, K_2, K_3 obtained by experimental data correlation as well as their validity regions are summarized in Table I. It can be seen from the table that values of coefficients varied with gas and liquid flow rates so that specific sets of coefficients corresponded to each two-phase flow regime. This is illustrated in Fig. 6 (for data obtained at $H_w = 0$) where dotted lines represent the validity limits of various sets of constants K_1, K_2, K_3 (Table I). It is apparent that these limits coincide reasonably well with limits of individual two-phase flow regimes taken from Mandhane's flow-pattern map[6].

In Fig. 7 experimental values of pressure drop obtained at $w_L = 0{\cdot}01$ and $0{\cdot}5$ m/s ($H_w = 0$ and $0{\cdot}05$ m) are compared with data calculated from Eq. (*4*) (for appropriate

TABLE I

(Values of coefficients K_1, K_2, K_3 (Eq. *4*) determined from experimental data correlation)

Weir height, m	K_1	K_2	K_3	Validity region w_L m/s	Validity region w_G m/s	Flow regime
0	4·74	1·56	0·32	0·005—0·025	2·5— 7·5	I. stratified flow
0	313	1·10	1·39	0·05 —0·15	1·0— 5·0	II. stratified-wavy flow
0	10·2	1·81	0·58	0·05 —0·25	7·5—20·0	III. wavy-annular flow
0	603	0·20	1·88	0·25 —2·2	0·3— 1·0	IV. plug flow
0	656	0·58	1·51	0·10 —2·0	1·5—10·0	V. plug — slug flow
0	211	1·03	1·27	0·50 —1·0	7·5—20·0	VI. slug-annular flow
0·05	81·0	0·316	0·385	0·005—0·10	0·3— 7·0	—
0·05	431	0·504	1·13	0·10 —0·65	0·3— 7·0	—
0·05	7·00	1·71	0·405	0·005—0·10	7·0—5·0	—
0·05	63·9	1·34	0·865	0·10 —0·65	7·0—50·0	—

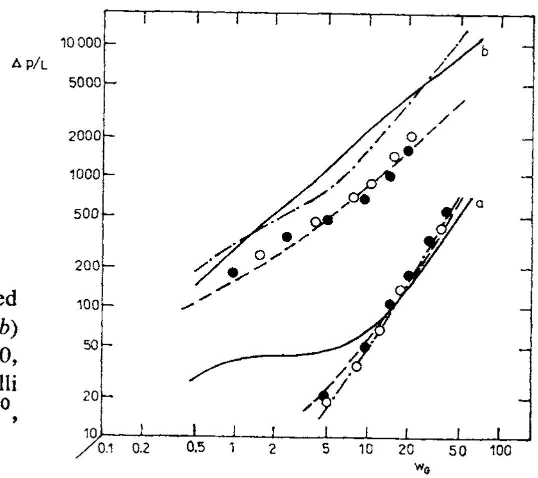

FIG. 7

Comparison of experimental and calculated pressure-drop data. *a*) $w_L = 0{\cdot}01$ m/s, *b*) $w_L = 0{\cdot}5$ m/s; experimental data: ● $H_w = 0$, ○ $H_w = 0{\cdot}05$ m ----- Lockhart, Martinelli relation[9], -·-·-·- relation of Chawla[10], ——— relation of Storek and Brauer[11] sets of coefficients K_1, K_2, K_3) and from relations proposed for correlation of two-phase flow pressure-drop data by Lockhart and Martinelli[9], Chawla[10] and Brauer[11]. Results of this comparison suggest that Eq. (*4*) can be suitably applied for calculation of air-water flow pressure drop in horizontal tubes (with or without weirs) using appropriate sets of coefficients K_1, K_2, K_3 for individual flow regimes. Surprisingly good agreement was observed in the whole range of w_G and w_L between Eq. (*4*) and general correlation of Lockhart and Martinelli[9] while systematic deviations between these two correlations and relations of Chawla[10] or Storek and Brauer[11] were apparent in some regions of gas or liquid flow rates (Fig. 7).

Liquid Holdup

In Fig. 8 experimental data of liquid holdup, e_L, are plotted against gas flow rate in the range of superficial liquid flow rates $w_L = 0{\cdot}01 - 0{\cdot}5$ m/s. All the data were obtained in the arrangement with the weir ($H_w = 0{\cdot}05$ m). The graph clearly demonstrates that within the experimental range of gas and liquid flow rates both methods used for liquid holdup determination were equivalent *i.e.* yielded practically identical results. It is apparent that liquid holdup varied significantly with flow rates of both phases ε_L values ranging between 0·01 and $\approx 1{\cdot}0$. Indeed this confirms large variability of working conditions attainable in horizontal-tube contactors. Extremely high liquid holdup values observed at low gas and liquid flow rates ($w_G < 1$ m/s, $w_L < 0{\cdot}1$ m/s) can be ascribed to the strong effect of weir on the character of flow under such conditions. Comparison of our data with values ε_L calculated from relation of Lockhart and Martinelli[9]

$$\varepsilon_L = 1 - (1 + X^{0.8})^{-0.378} \qquad (5)$$

$$(X = (\Delta p_L/\Delta p_G)^{0.5})$$

presented in Fig. 8 proved good agreement of experimental and calculated data in the region of higher flow rates ($w_G \geqq 2{\cdot}5$ m/s, $w_L = 0{\cdot}5$ m/s) where the influence of weir can be neglected. Similar comparison made at $w_L = 0{\cdot}01$ m/s with relations of Lockhart and Martinelli (Eq. 5), Chawla[12] and Hughmark[13] (Fig. 8) proved that under conditions of strong weir effect correlations based upon description of flow in weirless tubes could not describe even qualitatively the dependence of ε_L on w_G. Exponential-type relation

$$\varepsilon_L = M_1 w_G^{M_2} w_L^{M_3} \qquad (6)$$

was therefore used for correlation of our liquid holdup data obtained in the region of significant weir influence (*i.e.* at low gas and liquid flow rates). It is apparent from Fig. 8 that good agreement between experimental and calculated data was achieved for values of coefficients $M_1 = 0{\cdot}378$, $M_2 = -0{\cdot}226$, $M_3 = -0{\cdot}0932$ determined on the basis of our experimental data.

Interfacial Mass Transfer

Due to experimental method limitations values of $k_L a_L$ were in this work determined only at medium and low liquid flow rates ($w_L \leqq 0{\cdot}1$ m/s) *i.e.* under conditions of stratified, wavy and annular flow regimes. In Fig. 9 experimental data of $k_L a_L$ have been plotted in logarithmic coordinates as a function of parameter $(\Delta p_{3-10}/L_{3-10})$ representing the pressure loss due to two-phase flow in corresponding tube section. It is apparent from the graph that in the whole range of gas and liquid velocities

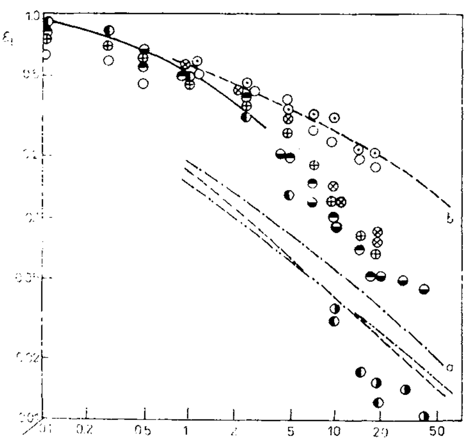

FIG. 8

Dependence of liquid holdup on gas flow rate, H_w = 0·05 m. Holdup data determined by the feed-shut off method (A) and from the mean residence time of the liquid phase (B). w_L = 0·01 m/s: ◐ A, ◑ B; w_L = 0·05 m/s: ◓ A, ◒ B; w_L = 0·10 m/s: ⊖ A, ⊘ B, w_L = 0·50 m/s: ○ A, ⊙ B calculated data: *a*) w_L = 0·01 m/s, *b*) w_L = 0·5 m/s. ----- Lockhart, Martinelli relation[9], --··-- relation of Chawla[12], -·-·-·- relation of Hughmark[13], ——— our correlation (Eq. 6)

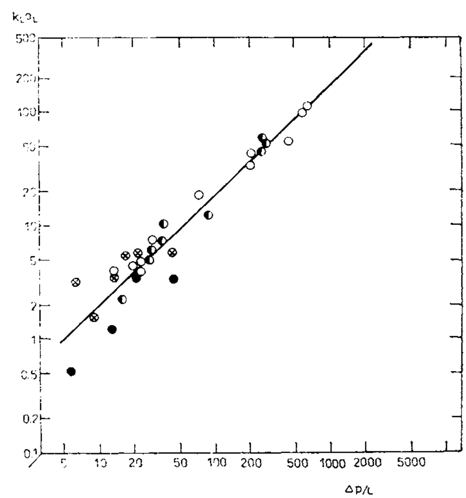

FIG. 9

Dependence of $k_L a_L$ on two-phase flow pressure drop. ● w_L = 0·005 m/s, ⊗ w_L = = 0·01 m/s, ○ w_L = 0·05 m/s, ◐ w_L = = 0·10 m/s, ——— data calculated from Eq. 7 for A = 0·225, b = 0·95 *i.e.* for all two-phase flow regimes studied a single dependence $k_La_L = f(\Delta p/L)$ was obtained which was succesfully described by exponential-type relation

$$k_La_L = A(\Delta p_{3-10}/K_{3-10})^B \qquad (7)$$

for values of coefficients $A = 0{\cdot}225$ and $B = 0{\cdot}95$ determined by experimental data correlation. Apparently such results confirm that in analogy with other gas–liquid contactors the rate of interfacial mass transfer in horizontal-tube reactors has been unambiguously determined by the rate of energy dissipation in the gas–liquid layer. Regarding our limited data base we have not attempted to proceed any further toward quantitative description of the mechanism of energy dissipation and to consequent analysis of the relation between this rate and k_La_L representing the intensity of interfacial mass transport. Thorough experimental study devoted solely to these topics is being prepared and the results will be reported later.

LIST OF SYMBOLS

A, B	empirical coefficients (Eq. (*7*))
a_F	disturbance propagation velocity (Eq. (*3*))
a_L	specific interfacial area related to a unit of liquid phase volume
c	concentration
c^+	equilibrium concentration
D	tube diameter
H_w	weir height
k_L	liquid-side mass transfer coefficient
K_1, K_2, K_3	empirical coefficients (Eq. (*4*))
L	tube length
M_1, M_2, M_3	empirical coefficients (Eq. (*6*))
Δp_0	total pressure drop
t	temperature
V	volume
$\dot{V}$	volumetric flow rate
w	superficial velocity
X	parameter of Lockhart and Martinelli correlation ($X = (\Delta p_L/\Delta p_G)^{0.5}$)
ε	holdup
ϱ	density
$\bar{\tau}$	mean residence time
μ	moment of response curve

Subscripts

G	gas
e	entering section
L	liquid
RS	test section of the tube
w	weir

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