ABSORPTION IN VERTICAL STREAM TUBES WITH SWIRL BODIES

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The absorption of oxygen from air into water has been measured in vertical tubes with swirl bodies in the inlet part. The tubes were of 70 mm I.D. and of height H with diameter ratio $H/D \leqq 22$. The two-phase flow was directed upward. Superficial gas velocity was $\bar{w}_G = 10$ to 35 m s^{-1} and specific liquid load $\dot{Q}_{LE} = 13$ to $80 \text{ m}^3 \text{ m}^{-2} \text{ h}^{-1}$. Values of the liquid-side mass transfer coefficient were determined as ratios of experimental values of volumetric mass transfer coefficient and specific interfacial areas, which were measured earlier. Murphree efficiencies were obtained with experimentally determined Peclet numbers considering the real mixing conditions of the liquid phase.

Stream tubes with swirl bodies show a possible way for increasing throughputs of the gas phase with satisfactorily mass transfer efficiency. One field of work concerns the influence of the various flow regimes on transfer processes in thin horizontal[1], helically coiled[2] or vertical tubes[3,4], without additional mixing elements. By inserting swirl bodies in shorter tubes, the separation of the liquid phase can be achieved. Multistage devices can be then formed combinating such elements for industrial-scale equipment.

Some work has been already published on mass transfer coefficients in single co-current contact elements not only in the field of absorption/desorption[5,6] but also in distillation[7,8]. Several authors based efficiency models on supposed[9,10] or also measured[11] flow conditions, even in the elements[12] themselves or in contact devices which are composed from parallel arranged elements on the single stage[12,13].

Further progress in the field of mass transfer in stream tubes with swirl bodies can be obtained, when the interfacial areas are determined separately, like in[3,4], to separate both terms of the volumetric mass transfer coefficient according to the basic equations (*1*) and (*2*) of mass transfer. This intention seems important also from the point of view that the geometrically estimated interfacial areas may differ considerably from so called active areas, determined by chemical methods[14]. This was the purpose of the present work.

THEORETICAL

The basic equations

$$K_{vL} = L(x_E - x_A)/V_0 \Delta \bar{x} \quad (1)$$

$$K_{vL} = K_{aL} a/V_0 = K_{aL} a_S \quad (2)$$

allow to convert the volumetric mass transfer coefficients related to the single stage into K_{aL}. For the system O_2-air–water we can further assume $K_{aL} \approx k_{aL}$. Gas phase ideal mixing can be supposed for higher throughputs on the basis of measured swirled flow profiles[14]. The common mass balance gives the model for the liquid-side Murphree efficiency

$$E_{OL} = \frac{x_E - x_A}{x_E^* - x_A} = \frac{(p/\lambda)\, N_{OL}}{1 + (p/\lambda)\, N_{OL}} \quad (3)$$

if representative mean K_{aL} are supposed to be valid over the whole stage[9]. λ is the absorption factor $\lambda = mG/L$, $N_{OL} = (1/P)(x_E - x_A)/(x_E^* - x_E)$ the number of transfer units. With the dependence $p = p(Pe_L, N_{OL}, \lambda)$ now it is possible to abandon the assumption of the piston flow of the liquid in Eq. (1). Eq. (3) include both the limiting cases, $Pe_L \to 0$ (ideal liquid mixing) and $Pe_L \to \infty$ (piston flow both of the phases)[15]. For comparison, the function

$$g = (E_{OL}/E_{OL}\,.\,Pe_L \to 0) - 1 \quad (4)$$

was defined which allows to estimate the deviation from the ideal mixing. With $E_{OL} \to E_{OL}$, $Pe_L \to \infty$ this function changes into its maximal values $g \to g_\infty$.

We demonstrated, using an error estimation on the influence of interfacial area, that with relative deviations of $|\Delta a_S/a_S| = 10\%$, errors in Eq. (4) may be caused of $|\Delta g/g| = 30$ to 60%, depending on p and λ. Errors in Pe_L in the magnitude $|\Delta Pe_L/ /Pe_L| = 10\%$ cause error values $|\Delta g/g|$ only maximally of the same order.

EXPERIMENTAL

The experiments were carried out in vertical tubes with swirl bodies which were described earlier (see[14], compare the variants A_L and C_L, I.D. of the tube $D = 0.07$ m, $H/D = 22$). Other variables were: superficial gas velocity $\bar{w}_G = 10$ to 35 m s^{-1}, specific liquid load $\dot{Q}_{LE} = 13$ to 80 m^3 m^{-2} . . h^{-1}, inclination angle of the swirl body blades to the horizontal $\alpha = 30$ to $90°$, liquid temperature $t_L = 12$ to $18°$C.

The absorption experiments were performed with oxygen-free water (treatment by intensive N_2 bubbling). At the inlet part of the tube and at the tube end (film separator) probes were taken which were analysed iodometrically after Alsterberg[17]. Equilibrium data[18] for determination of the mean driving force were corrected in dependence on the temperature.

The liquid mixing was measured by the conductivity method. Therefore we have arranged in four positions along the tube in each case two ring electrode pairs at the distance of 0·285 m. The electric signals, obtained by imperfect tracer pulses from 0·05M-KCl-solutions, were converted and processed with a parametric identification technique[19] to give first and second order moment of the distribution curves which were computed from the parametric model and therefore, mean residence times and Peclet numbers. The mass transfer in the lower spray part was still neglected.

RESULTS AND DISCUSSION

Fig. 1 shows volumetric mass transfer coefficients in dependence on the gas velocity $\bar{w}_G$ and the specific liquid load $\dot{Q}_{LE}$, according to Eq. (*1*). In the intermediate regime, between the rotating slug flow (further regime I) and the swirled annular-mist flow (regime II)[14], exists a characteristic minimum. Over the whole range of $\bar{w}_G = 10$ SE 35 m s^{-1}, various types of stages which work either in the flooding regime I without additionally inserted parts (at the begining of strong entrainment generation) or in the ordered co-current flow II, seem to be equally effective. In both of the cases, the values K_{vL} are considerably higher than that which can be obtained in a coun-

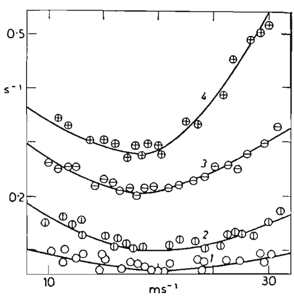

FIG. 1

Volumetric liquid-side mass transfer coefficients K_{vL} in dependence on the superficial gas velocity $\bar{w}_G$ (m s^{-1}) and the specific liquid load $\dot{Q}_{LE}$ (m^3 m^{-2} h^{-1}) ○ 13; ⦶ 22·7; ⊖ 53·1; ⊕ 75·2. Angle of swirl body blades $\alpha = 44·5°$

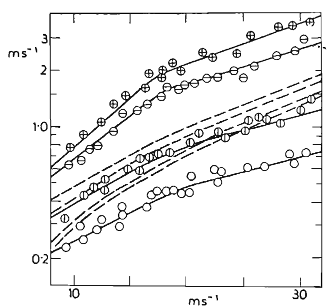

FIG. 2

Mass transfer coefficients K_{aL} related to interfacial area. For description see Fig. 1. Dotted lines: calculated by Eq. (*6*), dependence on $\dot{Q}_{LE}$ in the same succession ter-current apparatus[20]. However, when they are converted into transfer coefficients related to interfacial area, an increase of K_{aL} depending on $\bar{w}_G$ is observed. Because the interfacial area increases in the regime II only slightly with $\bar{w}_G$[14], we can conclude that in the product $K_{aL} \cdot a$ the flow energy $\Delta p_G \cdot \dot{V}_G$ affects mainly the mass transfer rate. On the other hand, every possibility is explored to increase the dispersion of the liquid by further energy dissipation. The centrifugal field confines the development of the secondary entrainment from the rotating wall film and therefore any increase of interfacial area. Due to raising stress forces on the film the surface renewal increases. In Fig. 2 we can distinguish the two regimes I and II

$$K_{aL,I} = 2{\cdot}2 \cdot 10^{-6} \bar{w}_G^{1{\cdot}5} \dot{Q}_{LE}^{0{\cdot}6} f^{0.2}\,, \tag{5}$$

$$K_{aL,II} = 1{\cdot}6 \cdot 10^{-6} \bar{w}_G \dot{Q}_{LE}^{0{\cdot}9} f^{0.5} (H/D)^{0.3}\,, \tag{6}$$

where f is the geometrical factor of the swirl body[14]

$$f = 1{\cdot}04(1 - 0{\cdot}89 \sin \alpha)\,. \tag{7}$$

The height of the slug flow layer in I is determined mainly by the force balance on the suspended liquid particles. Therefore the gas phase has in Eq. (5) stronger influence on K_{aL} than the geometry of the swirl body. The regime II is influenced by centrifugal forces more vigorously; in the film regime a certain effect of the stage becomes important. The exponents of $\bar{w}_G$ (1·0) and Q_{LE} (0·9) are the same as in[5] but not the absolute values K_{aL}. This can be explained by the influence of the used geometrical interface, which is too low, and therefore higher K_{aL} are observed.

Fig. 2 shows also K_{aL} (dotted lines) which were calculated by Eq. (8) based on the theoretical analysis[21] of turbulent mixing fluctuations in the film part

$$k_{aL} = 5{\cdot}16 \left(\frac{\mathscr{D}_L \varrho_L}{\sigma}\right)^{0.5} \left(\frac{\Delta p}{4(H/D)\,\varrho_L}\right)^{0.75}. \tag{8}$$

As the friction pressure drop we have used the measured wall pressure drop[14]. The slopes agree well.

The K_{aL} values calculated by means of Eqs (5) and (6) fit into the survey[20] of various mass transfer apparatus, which are arranged according to their intensity of phase contact.

Fig. 3 represents the Murphree efficiencies calculated by Eq. (3). The assumption of mean stage K_{aL} seems to be still reasonable with $(H/D)^{0.3}$ in Eq. (6). Even at higher $\dot{Q}_{LE}$, $R_{OL} \approx 80\%$ are attained. The corresponding Pe_L are in the range of 2 to 12, more detailed analysis will be published elsewhere. According to Fig. 4, the deviation from the ideal mixing model tends to be smaller with increasing liquid load. Lower gas velocities cause necessarily higher mixing degrees with smaller mass transfer efficiency because the energy for the transport of the liquid phase is not sufficient to prevent recirculation in the film. At high $\bar{w}_G$, the g values decrease according to Eq. (*4*) owing to the smaller differencies in the high stage efficiencies.

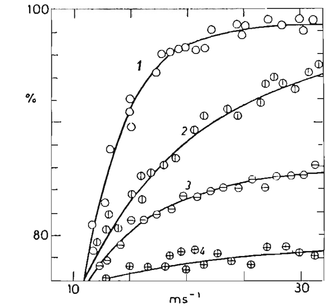

FIG. 3

Liquid-side Murphree efficiencies E_{OL}. Descriptions see Fig. 1

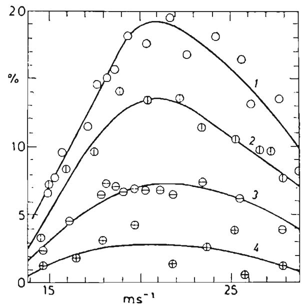

FIG. 4

Function g in dependence on $\bar{w}_G$ (Eq. (*4*)). $\alpha = 30°$. Other description see Fig. 1

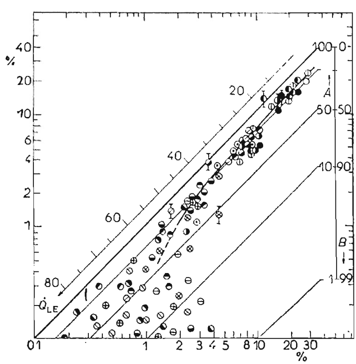

FIG. 5

Function g in dependence on g_{∞} (Eq. (*4*)). *A* % from piston flow conditions, *B* % from ideal mixing conditions

α	$\dot{Q}_{LE}$, $m^3\ m^{-2}\ h^{-1}$			
	13	22·7	53·1	7·52
30°	⦶	○	⊖	⦸
44·5°	◐	◕	◓	◔
73°	◑	◒	◒	◕
90°	□	⧄	⊟	⧅

In Fig. 5 the deviations of g from ideal mixing conditions are shown. The values g_∞ are also included. Even at high $\bar{w}_G$ and low $\dot{Q}_{LE}$, the piston flow conditions in the film are not attained. Therefore such conditions result which are characterized by about 82% realized piston flow (18% backmixing). This demonstrates that a certain internal recirculation always exists which reduces the stage efficiency. The latter increases still at higher H/D due the degradation of swirl forces.

LIST OF SYMBOLS

a	interfacial area m^2
$\mathscr{D}$	diffusion coefficient $m^2 s^{-1}$
D	diameter m
f	geometrical factor, Eq. (7) —
g	function, def. Eq. (4) —
$G, (L)$	gas (liquid) throughput rate $m^3 s^{-1}$
H	height m
$(k), K$	(partial) mass transfer coefficient $m s^{-1}$
N	number of transfer units —
P	function, defined by Eq. (3) —
Pe	Peclet number —
Δp	pressure drop $N m^{-2}$
$\dot{Q}$	specific load $m^3 m^{-2} h^{-1}$
t	temperature degree
V	volume m^3
w	velocity $m s^{-1}$
x	relative concentration —
Δx	mean integral driving force —
α	blade angle of swirl body degree
λ	absorption factor —
σ	interfacial tension $N m^{-1}$

Indices

A	outlet
a	interfacial
E	inlet
G	gas
L	liquid
S	specific
v	volume
o	over all
*	equilibrium
—	mean
·	time related
∞	related to piston flow

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