IMPACT OF STRUCTURED PACKING ON BUBBLE COLUMN MASS TRANSFER CHARASTERISTICS. PART 2. ANALYSIS OF GAS-LIQUID MASS TRANSFER MEASUREMENTS

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Abstract

The impact of the Sulzer SMV 16 packing elements on mass transfer characteristics in a packed column was studied. The gas phase (oxygen) and the liquid phase (tap water) passed the device in a concurrent upflow mode. The measurements of physical absorption of the oxygen into the liquid were taken in both hydrodynamic regimes, partly in the homogeneous and partly in the heterogeneous. Also, the experiments were doubled in an empty column for comparison. In the calculation of the volumetric gas-liquid mass transfer coefficients, $k_L a$, axial dispersion model (ADM) and plug flow model (PFM) were tested. It is systematically shown, that ADM gives far more reliable interpretation of the recorded data than PFM. For both modifications of the column the gas velocity affects the mass transfer coefficients the most. Higher values of $k_L a$ were found in the packed bed, and the impact of internals increases progressively with the gas velocity. For an empty column the correlation of Akita and Joshida[5] gives the coefficients close to ours within 18.5%.

Introduction

The design and scale-up procedure of bubble columns is still not a straightforward procedure. Simple in construction bubble columns show a very complex fluid dynamic behavior. Thus the estimation of the design parameters has mainly empirical base. Depending on the nature of processes the bubble columns are modified in various ways.[1] However, as an absorber device the bubble column must provide an efficient contact between the flowing phases, which results in high interfacial area and high mass transfer coefficient. Nowadays structured packing is used to increase efficiency by splitting the flow into subflows and then recombining them again. And this is why the performance characteristics of these packings are somewhat superior to those of the classical random packings.

Bubble columns with or without internals are primarily meant for the gas-liquid absorption processes accompanied with a slow reaction. In this case the volumetric mass transfer coefficient, $k_L a$, defines the mass transfer between the flowing phases reasonably well. Thus there is no need to split this parameter into the liquid side mass

transfer coefficient, k_L, and the specific interfacial area, a. However this is not the case for fast and instaneous reactions. Experimental determination of this combined parameter, $k_L a$, is a straightforward procedure. Most experimental studies are performed in the absence of chemical reaction. Pure physical transport of a gas component from the gas phase into the flowing liquid is evaluated at the inlet and outlet of the column. The volumetric gas-liquid mass transfer coefficient evaluation then strongly depends on the selected flow model. Mainly based on the geometry of column, the following flow patterns are imposed: for a large diameter column or a small ratio of column length to column diameter, complete mixing of the liquid phase seemed to be an adequate approximation of the actual flow pattern in the column. Under stationary condition the volumetric mass transfer coefficient is then calculated as

$$k_L a = \frac{u_L}{L} \frac{c_L - c_{L_0}}{c_L^* - c_L} . \qquad (1)$$

This model is restricted to columns whose diameters are either larger than 0.3 m or their length do not exceed three times the column diameter.[2] On the other hand, the mass balance, that is based on the ideal plug flow model (PFM), with no backmixing in the liquid phase, is only applicable to tall and narrow columns. The mass balance on the gas component transferred into the liquid results to the following expression

$$k_L a = \frac{u_L}{L} \ln \frac{c_L^* - c_{L_0}}{c_L^* - c_L} . \qquad (2)$$

PFM offers the highest possible driving force for the mass transfer between the phases and consequently, the calculated values of $k_L a$ are the lowest.

Nevertheless, a certain degree of the backmixing in the liquid phase always exists, the gas phase passes the column in the form of bubbles, which are responsible for nonideal flow of the liquid in the column, as already explained in Part 1.[3] It seems that the one dimensional axial dispersion model (ADM) is more appropriate choice in the analysis of mass transfer measurements.[2] The differential mass balance of the gas component in the liquid phase can be written [2]

$$\frac{1}{Pe_L}\frac{d^2c_L}{dz^2} - \frac{dc_L}{dz} + St(c_L^* - c_L) = 0 . \tag{3}$$

The axial dispersion coefficient, E_L, which can be taken as a measure of backmixing phenomenon in liquid, is expressed through the modified Peclet number

$$Pe_L = \frac{u_L L}{E_L(1-\varepsilon_G)} , \tag{4}$$

and the volumetric gas-liquid mass transfer coefficient, $k_L a$, is found in the Stanton number, St, defined as

$$St = k_L a \frac{L}{u_L} . \tag{5}$$

With the Dankwerts boundary conditions

$$c_L = c_{L_0} - \frac{1}{Pe_L}\frac{dc_L}{dz} \quad \text{for } z=o \tag{6}$$

and

$$\frac{dc}{dz} = 0 \quad \text{for } z=1 \tag{7}$$

the analytical solution of Eq.(3) yields the following concentration profile along the column axis

$$c_L(z) = A_1 \exp(r_1 z) + A_2 \exp(r_2 z) + c_L^* . \tag{8}$$

The coefficients A_1 and A_2 in the above equation are defined as

$$A_1 = -A_2 \frac{r_2}{r_1} \exp(r_2 - r_1), \tag{9}$$

$$A_2 = \frac{Pe_L(c_L^* - c_{L_0})}{(Pe_L + 1)\left(\frac{r_2}{r_1}\exp(r_2 - r_1) - 1\right)} ; \tag{10}$$

and the roots, r_1, and r_2, are

$$r_{1,2} = \frac{1 \pm \sqrt{1 + 4St/Pe_L}}{2/Pe_L} . \tag{11}$$

The saturation concentration of the gas component in the liquid, c_L^*, is given by the Henry's law and is taken as a constant along the column. This assumption fails in the

case of tall columns. When the axial dispersion model is involved in mass transfer analysis, the values of the gas hold-up and the liquid dispersion coefficient must be known either through the experimental measurements or predicted by the reliable correlations.

The volumetric gas-liquid mass transfer coefficient – experimental studies

Numerous experimental studies of the mass transfer phenomena in bubble columns confirmed that for a certain system of fluid phases and for a certain type of gas distributor, the gas velocity has the major impact on the mass transfer coefficient. Even the well known Kastanek´s expression for $k_L a$ evaluation

$$k_L a = \frac{u_G (u_G + C)^{13/20}}{2u_G + C},$$
(12)

which was derived on the bases of the Higbie´s mass transfer theory and the Kolmogoroff´s isotropic turbulence[1], can be reduced to a simple form

$$k_L a = b \cdot u_G^n.$$
(13)

Both parameters *b* and *n* are empirical constants. The liquid phase physical properties and the type of gas distributors strongly affect the value of *b,* whereas the exponent *n* is rather insensitive to both of them over a wide range of superficial gas velocity. Decker et al.[4] confirmed this experimentally. They fond *n* to be in the range of 0.78 to 0.82, and *b* was affected by both, the gas distributor and the liquid properties. In their analysis of the measured O_2 concentration profiles, ADM was applied. Akita and Joshida[5] assumed the total dispersion in the liquid and the plug flow behavior of the gas phase. The result of their extensive mass transfer study yields the following dimensionless correlation

$$\frac{k_L a \cdot D^2}{\mathcal{D}} = 0.6 \left(\frac{\nu_L}{\mathcal{D}} \right)^{0.5} \left(\frac{gD^2 \rho_L}{\sigma} \right)^{0.62} \left(\frac{gD^3}{\nu_L^2} \right)^{0.31} \varepsilon_G^{1.1}.$$
(14) The

authors found that the gas hold-up is proportional to the gas velocity on the exponent of 0.64, therefore Eq. (14) can be further simplified, thus

$$k_L a \propto D^{0.17} u_G^{0.7}.$$
(15)

Even though an assumption of complete backmixing gives the highest values of the volumetric mass transfer coefficients, Akita and Joshida´s correlation[5] was found to cover the data of other investigators evaluated with ADM suprisingly well. As explained

by Deckwer and coworkers[2] , in a large diameter column the concentration at the gas inlet jumps due to the liquid backmixing, which brings the column operation closer to the well mixed system.

The use of PFM in the evaluation of $k_L a$ usually results in the coefficient dependency on the liquid velocity as well.[6,7] Deckwer et al. criticized this conclusion and they have proved that the use of an inadequate flow model lead to erroneous conclusions.

Shah and coworkers[8] derived a simple correlation for $k_L a$ in a bubble column with air-water system

$$k_L a = 0.0107 \cdot u_G^{0.82} , \tag{16}$$

where the gas velocity is given in cms^{-1}.

Numerous investigators studied gas-liquid mass transfer in columns with different internals.[9,10,11] Although the affirmative role of motionless static mixers has been emphasized, yet no unified correlation for the evaluation of $k_L a$ is proposed till now.

The approach for packed columns is similar to that for empty bubble columns. For certain types of internals the functional dependence of $k_L a$ mainly relies on the gas superficial velocity (Eq. 13), whereas the two parameters, *b* and n, in Eq. (13) must be determined by experiments.

Wang and Fan[9] measured the absorption of pure oxygen into water in empty bubble column and in column packed with Koch motionless mixers. The oxygen inlet and outlet concentrations in the liquid phase were recorded, and the $k_L a$ values were calculated according to Eq. (2). They proposed the following correlation

$$k_L a = a_1 u_L^{\beta} u_G^{\gamma} . \tag{17}$$

For empty column the influence of the liquid velocity was insignificant ($\beta = 0.083$), but for packed columns β ranged from 0.457 to 0.611, depending on the mixers arrangement within the column. The exponent γ was 0.664 in the first case and a little lower in the later case (from 0.501 to 0.623).

The experimental part of this work includes extensive measurements of the physical absorption of oxygen from the gas phase into the liquid phase under different operating conditions. The aim of this work was, however, first to look for a model which is able to

give a reliable interpretation of the recorded gas-liquid mass transfer data and second, to determine the impact of the liquid and gas velocities on the volumetric gas-liquid mass transfer coefficient in both types of columns, packed and nonpacked. We also made some comparison with the existing correlations for the evaluation of k_La valid for empty columns.

Experimental

The details of the apparatus and packing material are all given in Part 1.[3] The Plexiglas column had a high length to diameter ratio (13.5). The column operated in the cocurrent upflow mode, tap water and oxygen were employed as the flowing phases. The column was packed with motionless mixing elements (Sulzer SMV 16). The superficial velocity of the liquid varied from 0.31 cms^{-1} to 5.41 cms^{-1} and the gas velocity was in the range of 0.91 cms^{-1} to 9.0 cms^{-1}. Thus, both hydrodynamic regimes, the homogeneous and heterogeneous, were partly covered in this experimental study. The perforated plate with 43 holes, each of 1.2 mm in diameter, served as the gas distributor.

The volumetric gas-liquid mass transfer coefficient - measuring technique

For the evaluation of mass transfer characteristics the measurement of oxygen concentration profiles in the liquid phase were taken under stationary conditions. At given operating conditions the liquid samples were consequently withdrawn from the column at four different axial positions (36.6 cm, 74.4 cm, 112.1 cm and 149.8 cm above the gas inlet) and then lead into a sample cell, in which the concentration of the absorbed oxygen was determined by a polarographic electrode (OXI 196; WTW).

The experiments were performed in empty and packed column. The packing configuration of the internals is shown in Figure 3 in Part 1.[3] The liquid may become saturated with the gas at some operating conditions, what is a severe drawback of this method. Indeed, we were confronted with this problem at low liquid flow rates (0.2 and 1 m^{3}h^{-1}) and high gas throughputs (5 m^{3}h^{-1}). This limited our range of operating conditions. Only those experiments were considered for mass transfer characterization,

in which the measured liquid oxygen concentration did not exceed 95% of the saturation concentration.

For the evaluation of the volumetric mass transfer coefficients the use of PFM (Eq. 2) and ADM (Eq. 3 to 11) seemed to be a reasonable choice regarding the column geometry. For both the models bottom-to-point calculations were performed. The experimental values of the integral gas hold-ups and the axial dispersion coefficients were available at the same experimental conditions (Part 1[3]) so we were able to extract the $k_L a$ data from ADM based on a single measured liquid concentration as well.[2]

Results and discussion

Gas hold-up

Gas hold-up plays an important role in the performance of bubble column. Actually it represents the integral value of all volume of bubbles throughout the column. Therefore the gas-liquid interfacial area and consequently the gas-liquid mass transfer rates are proportional to the gas hold-ups. The gas hold-up depends on the column diameter, physical properties of the system and the type of gas distributor, but mainly on the velocity of the gas phase. Introduction of a packing is expected to change the hydrodynamic behavior of the column.

The measurements of the experimental gas hold-ups for both modifications of the column were already performed in Part 1[3].The data are most evidently shown in the function of the liquid superficial velocity (Figure 1). The role of the gas velocity as a decisive factor for ε_G is clear: higher gas velocities enlarge the gas hold-up values. An increase in the liquid velocity slightly decrease ε_G for both modifications of the column, what becomes more evident at higher gas velocities. The negative impact of the liquid velocity on gas hold-up is in agreement with the work of Moustiri and his coworkers[12] and with Hill's finding[14]. The reason for a sudden increase in ε_G at the liquid velocity of 5.41 cms⁻¹ is not clear. High value of the gas hold-up at u_L=3.61 cms⁻¹ in packed bed at the highest gas velocity is rather due to an experimental error. The presence of internals becomes significant only at higher gas velocities ($u_G \geq 4 \mathrm{cms}^{-1}$) and their effect is

evaluated and represented in Figure 2. At higher gas velocities the static mixers enlarged the holdup even up to 50%. It is interesting to note, that a slight reduction in gas hold-
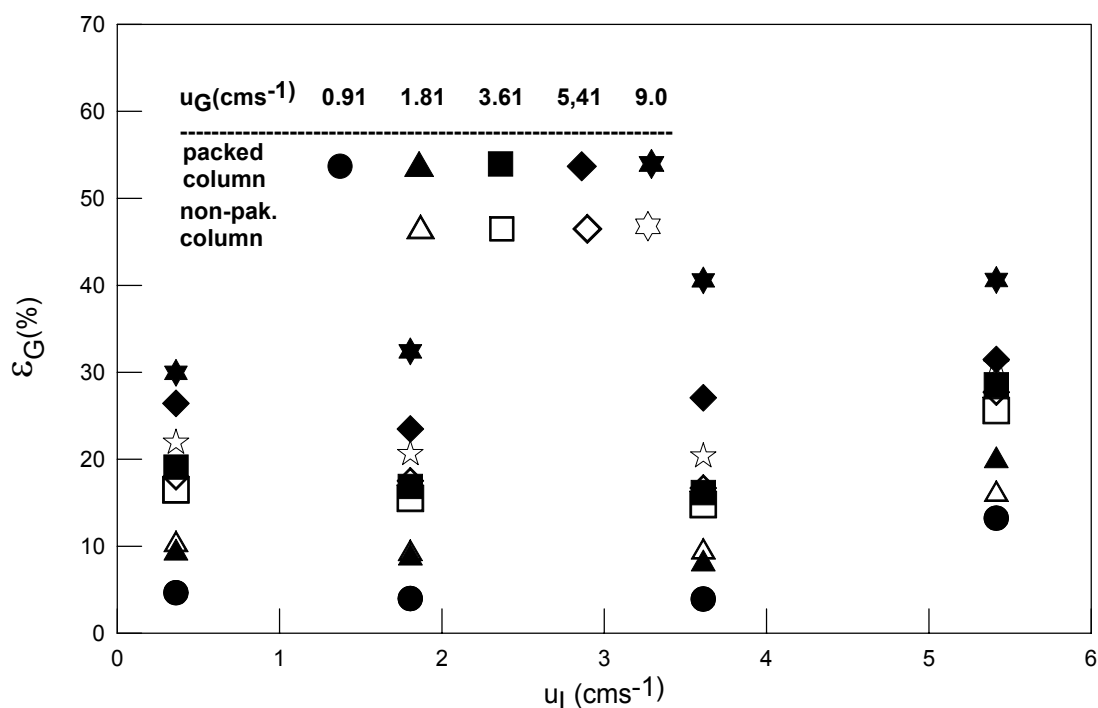

Figure 1. Gas hold-up as a function of liquid superficial velocity for both modifications of the column.

ups was observed in packed column according to those in an empty one at lower gas velocities. This is in agreement with the observation of Moustiri and his research group.[12]

The superficial gas velocity is the most crucial factor for ε_G. In order to evaluate this effect, we ignored the minor influence of the liquid velocity and represented the experimental data in a form of simple power-law functions

$$\varepsilon_G = 0.0915 \cdot u_G^{0.44} \qquad (1.81 \text{ cms}^{-1} < u_G < 9.0 \text{ cms}^{-1}) \qquad (18)$$

valid for empty column and

$$\varepsilon_G = 0.0649 \cdot u_G^{0.81}, \qquad (0.91 \text{ cms}^{-1} < u_G < 9.0 \text{ cms}^{-1}) \qquad (19)$$

which holds for column packed with SMV static mixers. In this way the mean relative errors are relatively high (e_y is 12.5% in the first case and 21% in the latter). Both relationships are shown in Figure 2. The high exponent (0.81) over the gas velocity in the case of packed column indicates, that the homogeneous regime is prolonged in the presence of internals. A weaker dependence of the gas hold-up on gas velocity in empty

column shows that the transition from the homogeneous to heterogeneous hydrodynamic regime occurred. According to the flow chart[1], the homogeneous regime prevails up to the gas velocity of 3 cms[-1], and the heterogeneous regime develops approximately at u_G=5 cms[-1]. In the presence of internals Spicka et al.[13] experimentally proved, that the homogeneoud regime is extended, even up to u_G=9cms[-1].
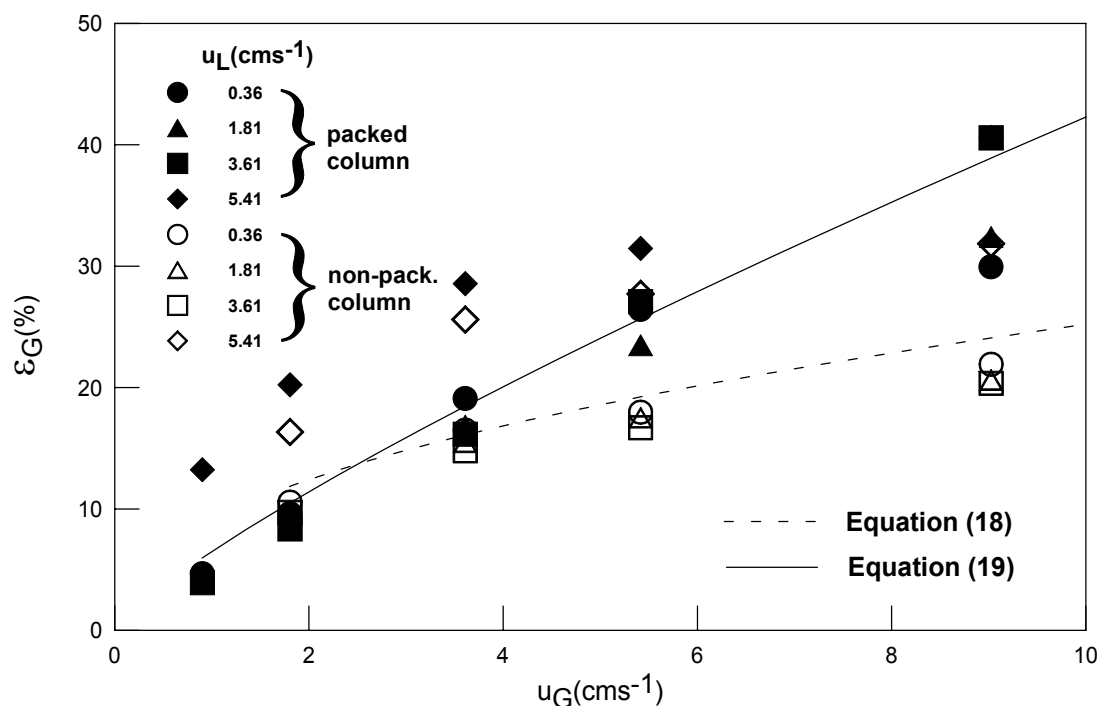

Figure 2. Gas hold-up as a function of gas superficial velocity for both modifications of the column.

For more detailed characterization of the hydrodynamic behavior of bubble column the slip velocity (or the relative velocity between the phases) is an extremely helpful parameter. For cocurrent flow the following equation holds

$$v_S = \frac{u_G}{\varepsilon_G} - \frac{u_L}{1-\varepsilon_L} . \tag{20}$$

Up to the gas velocity of 4 cms[-1] this value is roughly about 20 cms[-1] (Figure 3). At low gas velocities the internals promote the rising of gas bubbles through the channels and the slip velocity is higher than in empty column. Then at higher values of u_G the slip velocity in empty column increased sharply, even up to 40 cms[-1], due to the formation of larger bubbles in the heterogeneous regime. In the packed column the presence of the motionless mixing elements prevents the bubble coalescence, therefore the bubbles are

smaller and their rising velocities are low. This is reflected in lower slip velocities in comparison with the results obtained in an empty column.

As one may conclude from Fig.3, at higher liquid velocity even the decreasing values of v_S with the gas superficial velocity, u_G, was observed. Moustiri et al.[12] assumed that the packing also hinder the bubble rising.
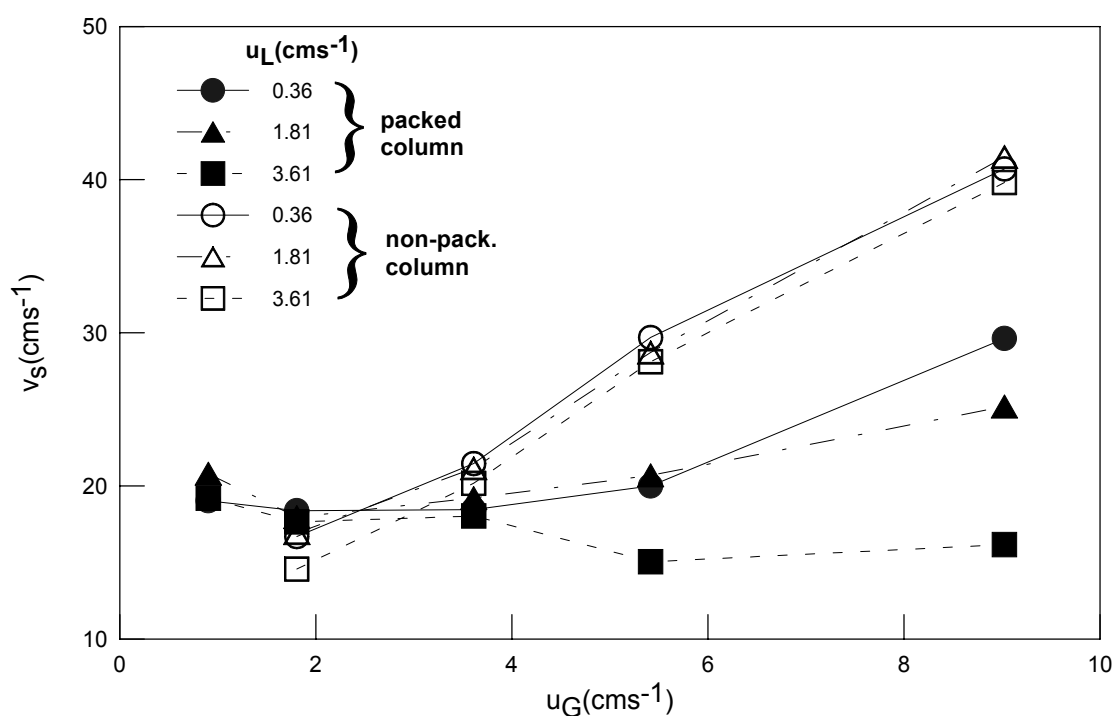

Figure 3. Slip velocity as a function of gas superficial velocity for both modifications of the column.

Volumetric gas-liquid mass transfer coefficient

The measured liquid oxygen concentrations were between 27% and 95% of the saturation concentration. The data recorded along the column for two sets of operating conditions are shown in Figure 4. As expected, at any operating conditions the concentration of oxygen increases along the column for both modifications. In packed column were always higher than those in empty one except near the bottom of the column. At the first measuring point (that is 36.5 cm above the gas inlet) the concentrations in empty column were always higher than those in the packed column. This confirms the fact, that the concentration jump near the gas inlet is rather due to the considerable backmixing of the liquid in the column than to an increase of mass transfer rate. Deckwer[1,2] did criticize a wrong interpretation of this phenomenon, which appears

in published works[6,7]. As it was already concluded in Part 1[3], the presence of the Sulzer motionless mixers diminishes the backmixing in column. Correspondingly the liquid axial dispersion coefficients in packed column are therefore lower than those in empty column for about 20% to 50%.
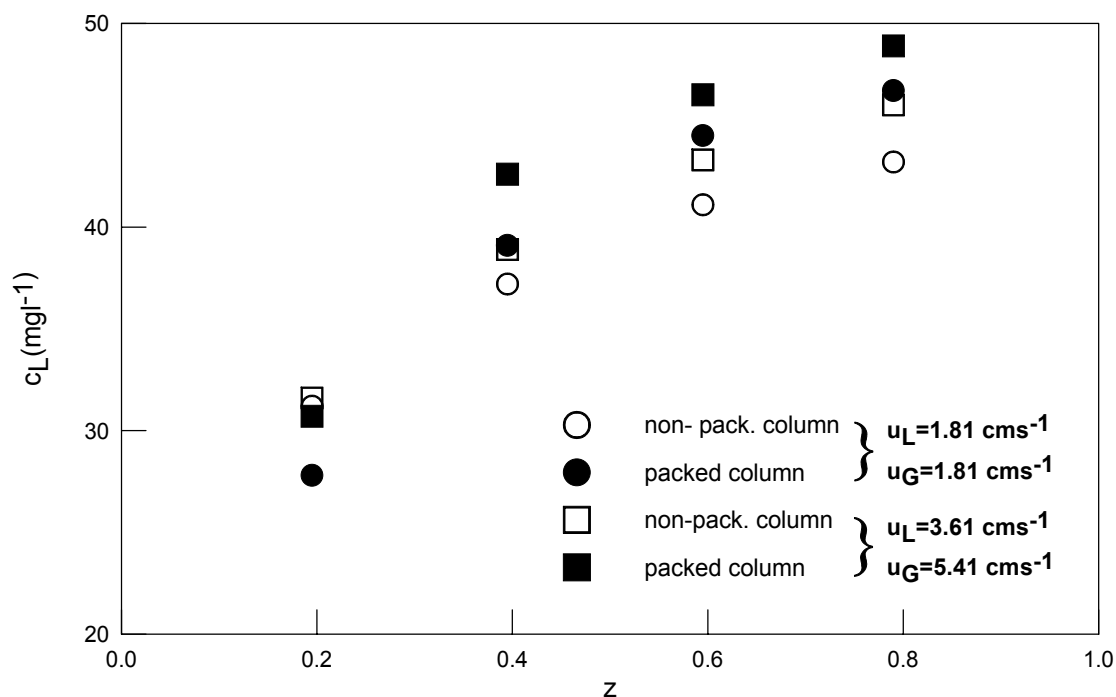

Figure 4. Oxygen liquid concentration as a function of dimensionless height for both modifications of the column at different operating conditions.

It is obvious, however, that using either PFM or ADM would lead to different $k_L a$ values. When ADM was applied the Peclet liquid numbers (Pe_L) were calculated from the experimental values of gas hold-ups and the liquid axial dispersion coefficients (Part 1[3]); they varied from 0.06 to 10. The Stanton numbers, found from fitting the measured point concentrations, took the values lower than 9 in 90% of calculation.

In endeavors to asses the $k_L a$ along the column, which are shown in Figures 5 and 6, the coefficients are based on the concentration measurements from Figure 4. As expected, the PFM evaluation always yielded lower values of the coefficients than the ADM evaluation, and the difference increased with gas velocity. Both models gave higher values of the coefficients in packed column according to those in empty one, except for the first measuring point. This phenomenon is already discussed in the previous section.
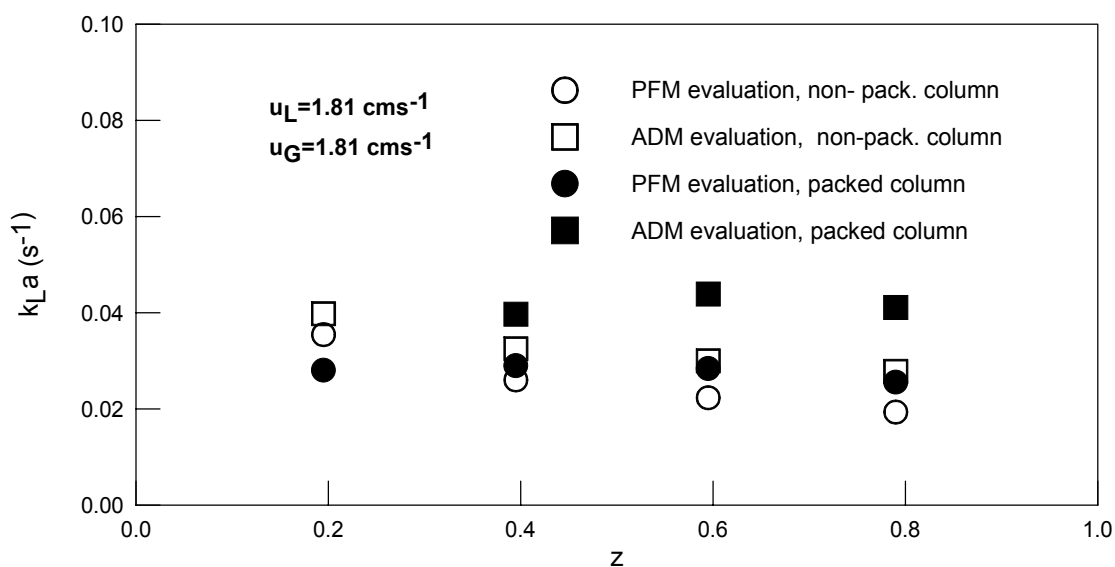
Figure 5. Volumetric gas-liquid mass transfer coefficient as a function of dimensionless height for both modifications of the column (homogeneous regime).
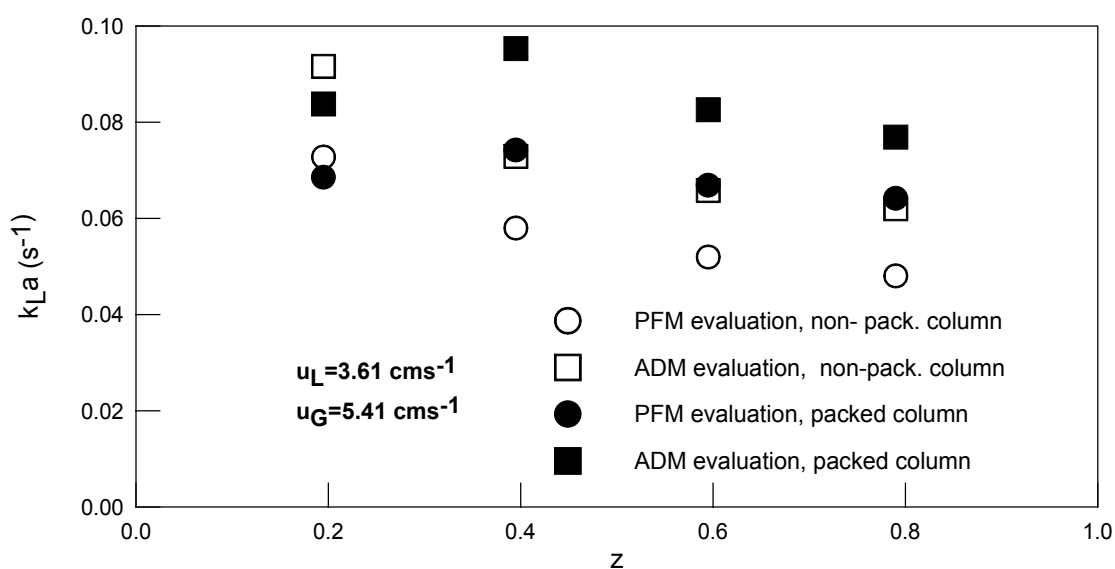

Figure 6. Volumetric gas-liquid mass transfer coefficient as a function of dimensionless height for both modifications of the column.

At low gas velocity ($u_G = 1.81 \text{cms}^{-1}$) a slight decrease in the volumetric gas-liquid mass transfer coefficients, based on PFM, was observed along the column (Figure 5), for packed and nonpacked column. For both cases the use of ADM gave nearly constant k_La values. At this operating condition the homogeneous regime exists in the column. When the gas velocity was increased to 5.41 cms^{-1}, both models resulted in decreasing k_La with the column axis, either packed or nonpacked. Based on a phenomenological model of the bubble break-up and coalescence, a simulation model of Shimizu and his coworker[15] also predicts the decrease in k_La along the column at higher gas flow rates. According to their simulations, increased numbers of larger bubbles at the top of column are responsible for decrease in k_La.

In order to validate both of the models, (PFM and ADM), only those volumetric gas-liquid mass transfer coefficients were involved, which were estimated by means of the the liquid oxygen concentrations measured on the last (i.e. fourth) position. In Figure 7 the results of k_La based on PFM, are shown as a function of the liquid superficial velocity. For the empty column the coefficients were lower than 0.08 cms^{-1}, and for the packed column lied in the range of 0.0064 cms^{-1} to 0.112 cms^{-1}. Thus, coefficients in packed
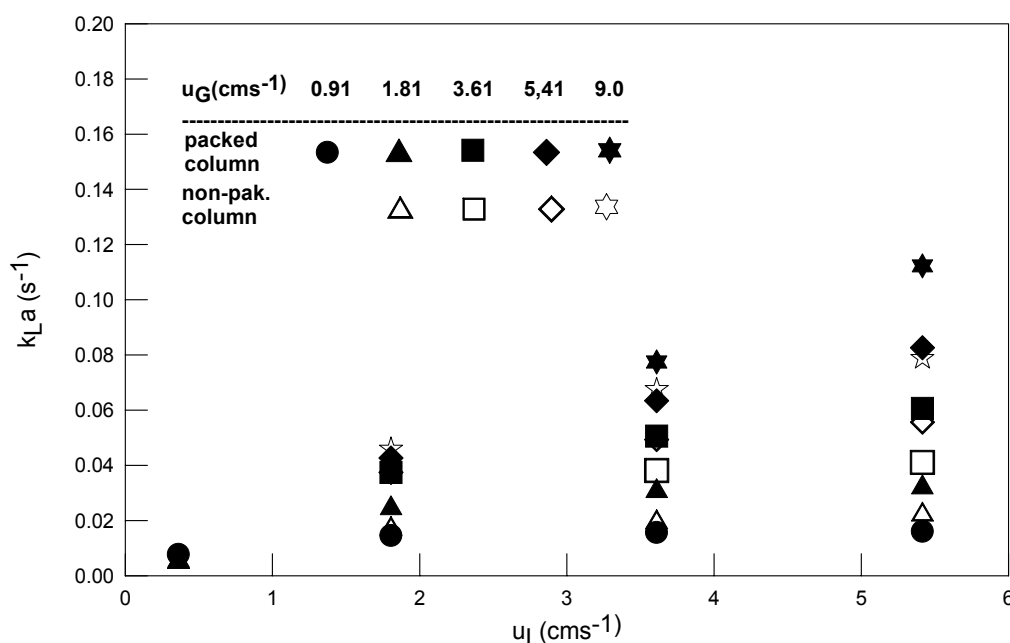

Figure 7. Volumetric gas-liquid mass transfer coefficients based on PFM as a function of liquid superficial velocity for both modifications of the column.

column are always higher than those in empty column, the difference increases when the gas velocity increases. The PFM evaluation actually exhibits the positive impact of the liquid velocity on k_La, which becomes more evident at higher gas flow rates. For instance at $u_G = 9.0 cms^{-1}$, a three-times higher liquid velocity almost doubles the coefficients in empty column, and causes for about 50% higher k_La in the presence of internals.According to the previous results on the axial dispersion coefficients, gas hold-ups and slip velocities, such a strong dependance of k_La on u_L can not be justified; this conclusion is equaly valid for packed and nonpacked column. Though in the packed column the axial dispersion coefficients were found to increase with the liquid superficial velocity (Part 1,[3]), this would not result in such an enlargement of k_La.

The k_La values evaluated with the help of ADM are depicted in Figure 8, as a
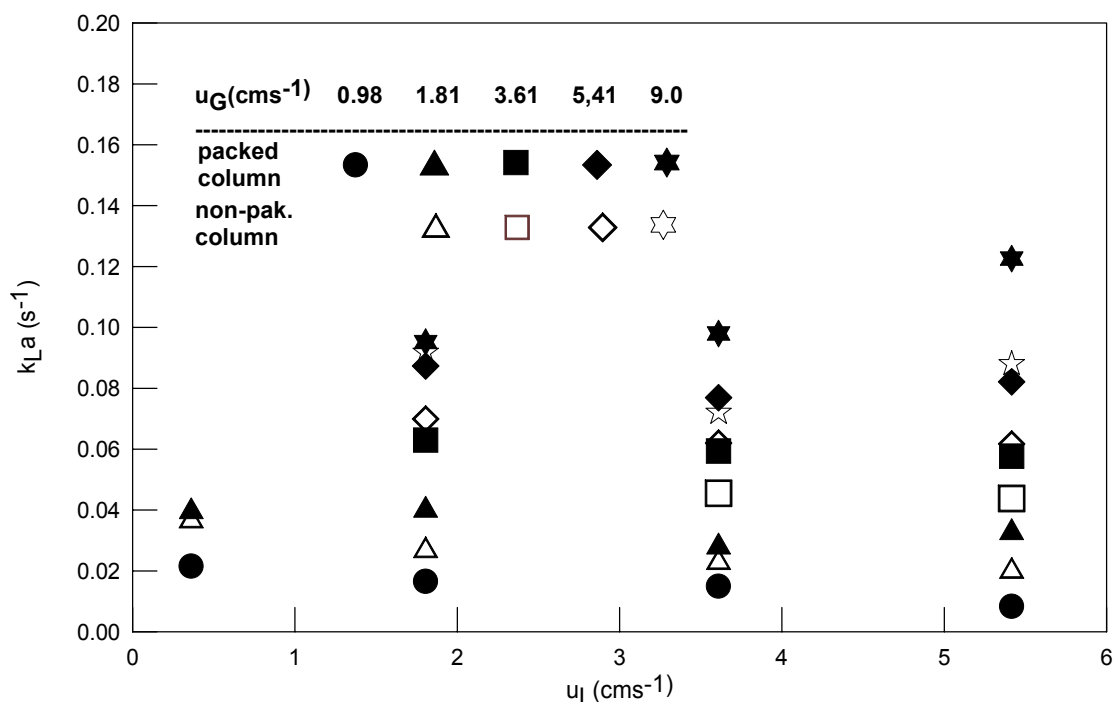

Figure 8. Volumetric gas-liquid mass transfer coefficients based on ADM as a function of liquid superficial velocity for both modifications of the column.

function of the liquid superficial velocity. The coefficients found in packed column are higher than those in empty one. The coefficients are only slightly affected by the liquid flow rate, actually they show the same trend as the gas hold-ups on Figure 1. We concluded that for an empty bubble column as well as for a packed column, the use of ADM is far more realistic for the analysis of mass transfer measurements than PFM.

This is also in agreement with Deckwer[1,2]. The use of PFM does not only lead to the wrong conclusion about the impact of liquid velocity, it also underestimates the coefficients significantly; at lower liquid flow velocities for nearly 100%.

As was already shown, the gas velocity affects the mass transfer coefficients most. The dependence of $k_L a$ on the gas superficial velocity is shown in Figure 9, for packed and nonpacked column. The coefficients are based on the ADM evaluation. The data can be reasonably well correlated with the following power-law relation, valid for empty column

$$k_L a = 0.0184 \cdot u_G^{0.70},$$ (21)

and for packed column

$$k_L a = 0.0184 \cdot u_G^{0.86}.$$ (22)

The gas velocity in the above equations is in cms^{-1}. The mean relative error (e_y) for Equation 20 is 12.3%, and for Equation 21 slightly higher, that is 17.1%. The presence of internals results in higher exponent on u_G, due to the extended range of the homogeneous regime.
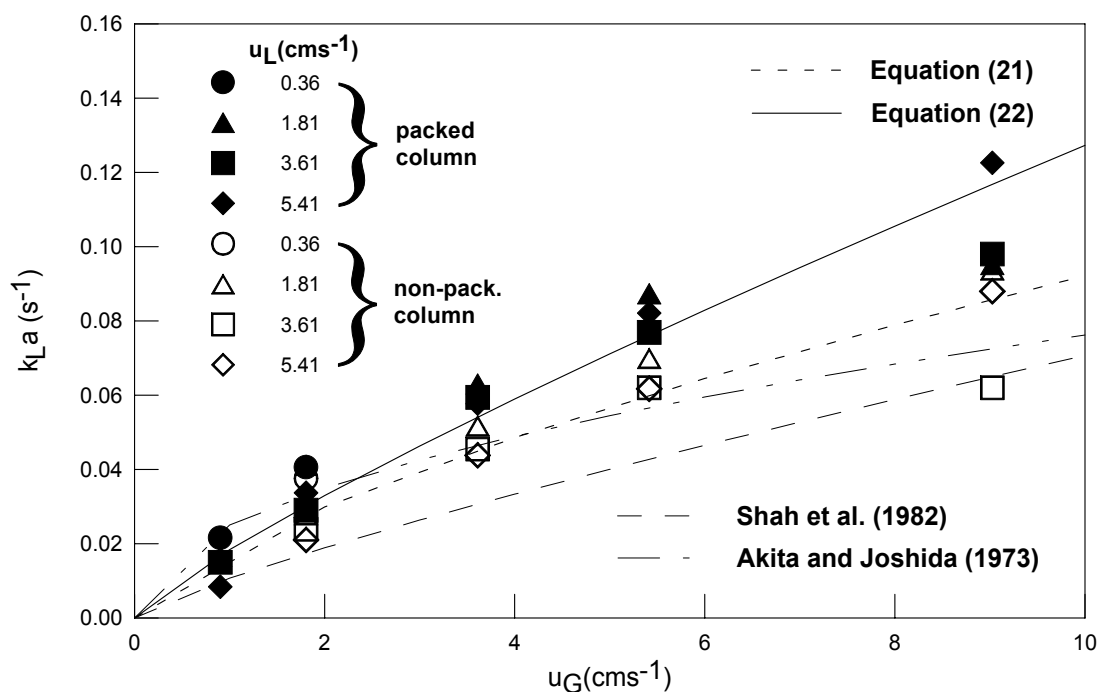

Figure 9. Volumetric gas-liquid mass transfer coefficients based on ADM as a function of superficial velocity for both modifications of the column.

It is interesting to note, that in both correlation (Eq. 20 and Eq. 21) the constants are equal, i.e. 0.0184. Higher values of k_La were found in the packed bed, and the impact of internals was found to increase progressively with the gas superficial velocity.

For empty bubble columns the correlation of Shah et al.[8] (Eq. 16) gives much lower values for k_La compared to our experimental data ($e_y =31\%$). The coefficients predicted from the correlation of Akita and Joshida[5] are close to ours. This is not surprising since we have used our own experimental gas hold-ups (Part 1,[3]) in their correlation (Eq. 14); and the mean relative error was 18.5%.

Conclusion

In the analysis of gas-liquid mass transfer measurements the use of ADM gives the volumetric gas-liquid mass transfer coefficients, which are more consistent with the complex hydrodynamic behavior of the column than those evaluated from PFM.

In both modification of the column, packed and nonpacked, the values of k_La increase with the gas superficial velocity, while the effect of the liquid velocity is negligible. At the same operating conditions the coefficients in packed column are higher than those in an empty column, and this difference increases with the gas superficial velocity.

At lower gas velocities the use of ADM gives nearly constant values of k_La along the column axis, while at higher gas flow rates the coefficients decrease with the column height, either in packed or nonpacked column.

Acknowledgements

Nomenclature

A_1	defined with Eq. 9, /
A_2	defined with Eq. 10, /
a	specific surface area of internals, m^2m^{-3}
a_1	constant in Eq. 17
b	constant in Eq. 13
C	constant in Eq. 12
c	liquid oxygen concentration, mgl^{-1}
D	column diameter, m
$\mathcal{D}$	diffusivity, m^2s^{-1}

E_L	dispersion coefficient, m^2s^{-1}
e_y	mean relative error, $\left(= \dfrac{100}{N} \sum_{i=1}^{N} \left\| \dfrac{y_{MES}(i) - y_{PRED}(i)}{y_{MES}(i)} \right\| \right)$, %
k_L	mass transfer coefficient in the liquid phase, ms^{-1}
$k_L a$	volumetric gas-liquid mass transfer coefficient, s^{-1}
g	acceleration due to gravity, ms^{-2}
L	column length, m
N	number of experimental data
n	exponent over superficial gas velocity (Eq. 13)
Pe_L	modified Peclet number in the liquid phase, $\dfrac{u_L L}{E_L (1-\varepsilon_G)}$
$r_{1,2}$	defined with Eq. 11
St	Stanton number, $k_L a \dfrac{L}{u_L}$, /
u	superficial velocity, ms^{-1}
v_S	slip velocity, ms^{-1}
z	dimensionless length of a column, /
u	superficial velocity, ms^{-1}

Greek letters

β	exponent over the liquid superficial velocity in Eq. 17
ε	phase hold-up, /
γ	exponent over the gas phase superficial velocity in Eq. 17
ν	viscosity of the phase, $kgm^{-1}s^{-1}$
ρ	density of the phase, kgm^{-3}
σ	surface tension, kgs^{-2}

Subscripts

G	gas phase
L	liquid phase
MES	measured value
o	entrance of the column
PRED	predicted value
*	in equilibrium

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Povzetek

Prikazani so rezultati eksperimentalne študije vpliva strukturiranega polnila (Sulzer SMV 16) na volumetrični koeficient snovnega transporta plin-kapljevina v kolonah z mehurčki. Za primerjavo so bili vsi poskusi izvedeni še v koloni brez polnila. Pri analizi eksperimentalnih merjenj se je aksialno disperzno model (ADM) pokazal, da daje realnejšo sliko dogajanja v koloni kot model čepastega toka (PFM). Na osnovi ADM izračunanih koeficientov je razvidno, da je $k_L a$ odvisen od hitrosti plinaste faze, medtem ko lahko vpliv hitrosti tekočine zanemarimo, in to v obeh tipih kolon. Prisotnost polnilnih elementov poveča koeficiente snovnega transporta, njihov vpliv je izrazitejši pri višjih hitrostih plina. Z uporabo ADM izračunane točkovne vrednosti koeficientov so pri nižjih hitrosti plina konstantne vzdolž kolone, medtem ko pri visokoh plimskih hitrosti le-ti z višino kolone padajo.