Mass Transfer Characteristics of a Counter Current Multi-Stage Bubble Column Scrubber

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This paper reports on the experimental investigation carried out to evaluate mass transfer characteristics of a hydro-dynamically-induced multi-stage bubble column. As conventional normal bubble column could not meet the high mass transfer rate, this novel bubble column has been conceived, designed and fabricated as a wet scrubber for the control of air pollution; in addition it has versatile use as a gas–liquid contactor in chemical processing industries. Chemical methods have been used to characterize the mass transfer of the novel multi-stage bubble column. A liquid side mass transfer coefficient and a volumetric mass transfer coefficient have been found to be effective and improved from the previous work for such an improved gas-liquid contactor.

Introduction

Bubble columns find their wide use in chemical processing industries due to their simple construction, low cost and ease of operation. However, simple bubble column operating in a single stage only cannot achieve high efficiency except for highly soluble gases in chemically reactive systems. In order to achieve high collection efficiency of pollutants bubble columns must operated in series or in multi-stages. In commercially available bubble columns, multiple stage operation has been achieved by use of perforated multi-orifice plates or multi screens.

Excellent reviews on bubble column design and performance are available in literature (Miyahara *et al.*, 1983, 1999; Deckwer and Schumpe, 1993; Miyahara and Hayashino, 1995; Yamada and Goto, 1998; Sheng and Tsui, 1999; Meikap *et al.*, 2001, 2002a). However, most of the existing bubble column design involves use of mechanical agitation, multichannel or multi-stage operation using perforated plates or screens, partition plates, draught tubes, etc. Unfortunately most of the equipment for achieving good performance efficiency result in large energy dissipation (Liu *et al.*, 1993) and lead to complex mechanical structures requiring frequent cleaning or shutdown. In

the present article an attempt has been made to focus on the conceptual design and characterization of a counter current multi-stage bubble column scrubber. This column does not suffer from the limitation, e.g. high-energy dissipation, absence of moving parts, clogging, higher capital cost, maintenance difficulty of the existing multi-stage bubble columns, but can still achieve very high efficiency (Meikap *et al.*, 2002c) with high capacity.

1. Characterization of the Counter-Current Multi-Stage Bubble Column

In the present investigation, a bubble column operating in three stages has been designed. The staging effect has been achieved through hydro-dynamically-induced continuous bubble generation, break-up and re-generation. The system has been designed (Meikap, 2000) to operate with relatively large sized bubbles. So that internal circulation can be induced in the bubbles and faster transfer of gaseous materials can take place by turbulent diffusion through the interface of the bubbles and also due to the direct rupture of the relatively large diameter bubbles. In order to logically design a bubble column used for gas–liquid mass transfer operations it is necessary to collect data on the magnitude of the individual film mass transfer coefficient. Since the values of k_L for the present new type of system are not available in the literature, detailed experiments have been conducted to determine its range of values.

When a gas bubble swarm passes liquid through the bubble generator, the pressure energy contained in the gas serves both to sub-devide it, i.e. to increase its total surface and hence its total surface energy and at the same time to agitate the liquid. The amount of agitation that can be achieved in this way is limited by the rate of the gas flow, which in turn is determined by process considerations and may be insufficient at low gas flows or with viscous liquids or slurries.

The rate of mass transfer from a dissolving gas to the absorbing liquid is dependent on the driving force for mass transfer as given by the concentration difference of the transferring components among the gas–liquid interface and the bulk liquid, the transfer area and liquid-phase mass transfer coefficient, k_L. The typical $k_L a$ or 'a' is a feature of gas/liquid contactors and power input. The energy dissipation rate rises with increasing power input into the contactor, resulting in greater turbulence, which is a major driving force for boosting the interfacial area, 'a'. A higher $k_L a$ can cut capital costs, because the same mixing duty can be achieved in a smaller device, or in a shorter batch time for a batch process or at a higher throughput rate for a continuous process. The overall mass transfer coefficient, $k_L a$, is based on the liquid side concentration differential as the driving force for the mass transfer. It is much easier to measure the product of k_L and 'a', rather than the individual terms. The value of k_L depends upon the diffusivity of the dissolved gas in the liquid and the relative velocities between the gas bubbles and the liquid. Typically k_L varies by almost 100% around a mean value of 0.02 m/s. The faster movement of bubbles encourages mass transfer, so that they tend to have a higher value of the mass transfer coefficient, k_L, but is counterbalanced by their lower values of 'a', the interfacial area (Meikap *et al.*, 2001).

## 2. Chemical Methods of Measuring Mass Transfer Coefficients

The chemical method of mass transfer coefficients is based on gas absorption followed by chemical reactions. In general, when physical absorption is followed by a chemical reaction both the physical rate of absorption and the kinetics of the reaction will govern the overall rate. The application of chemical method for the determination of interfacial area (Meikap *et al.*, 2001) and mass transfer co-efficient has been reviewed by Meikap (2000).

The mass transfer coefficient for the absorption of reacting component A between the gas–liquid phases is k_L.

In the non-chemical method a highly soluble gas is absorbed in water. However, the rate of absorption depends on the residence time distribution in both the liquid and gas phases. In addition there is normally an considerable resistance to mass transfer on the liquid side and this must be taken into account. The rate of absorption can be made independent of the liquid-side residence time distribution and the liquid side resistance can be eliminated by absorbing the gas into a solution of chemical reagent with which it reacts instantaneously provided the $k_G p < k_L$Bo in the case of the instantaneous reaction or $k_G H_e \ll (Dk_2 Bo)^{0.5}$ in the case of the second order reaction of finite speed. There will be no liquid-side resistance and the rate of reaction is $R = k_L C_A$.

Carbon dioxide reacts with water to form bicarbonate and hydrogen ions:

$$CO_2 + H_2O \leftrightarrow HCO_3^- + H^+ \tag{1}$$
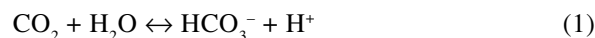
The amount of H_2CO_3 formed is known to be negligible small under all conditions. In water itself the reaction comes to equilibrium with only a very small fraction of the dissolved carbon dioxide in the form of bicarbonate ion. The values of this and other relevant equilibrium constants have been summarized by Harned and Owen (1958). The values of the stoichiometric first ionization constant at infinite dilution, $K_1 = [H^+] \times [HCO_3^-]/[CO_2]$ is a function of temperature.

A diffusion controlled slow chemical reaction is suitable for determining the liquid volumetric mass transfer coefficient $k_L a$ (Danckwerts and Sharma, 1970). The absorption of CO_2 into a sodium-carbonate-bicarbonate buffer solution is a convenient system, and is represented by a second order reaction,

$$CO_2 + OH^- \leftrightarrow HCO_3^- \tag{2}$$

In the presence of a substance, which will react with and remove hydrogen ions, however, carbon dioxide will be progressively converted to bicarbonate ion. Thus, in sodium carbonate solution there exist hydroxyl ions which react instantaneously with the H^+ formed by above reaction. The net results that CO_2 is converted to a bicarbonate ion at a rate controlled by reaction (1) and the direct reaction of CO_2 with hydroxyl ion proceeds simultaneously. The reaction is of the first order with respect to the concentration of CO_2. Values of the rate constant are a function of temperature. The pseudo-first order rate constant for the reaction of CO_2 with hydroxyl ions in the concentration present in, for instance, an equi-molar carbonate–bicarbonate mixture is about 0.85 s^{-1} at 298 K. The reaction between CO_2 and water at this temperature has a rate constant 0.26 s^{-1} and therefore usually contributes very little to the total rate of the reaction of CO_2 in such solutions. However, when the ratio of carbonate to bicarbonate is low, the direct reaction with water becomes relatively important. Further, the reaction with water can be catalyzed and can become faster than the reaction with OH^- even at high carbonate–

bicarbonate ratios.

The reaction: $HCO_3^- \leftrightarrow CO_3^{2-} + H^+$ is instantaneous. The stoichiometric ionization constant of HCO_3^- at infinite dilution can be represented as

$$K_2 = \frac{\left[H^+\right] \times \left[CO_3^{2-}\right]}{\left[HCO_3^-\right]} \tag{3}$$

Under certain conditions the above reaction is sufficiently fast, such that the concentration of CO_2 in the bulk of the liquid phase is equal to zero. The condition to be satisfied for this is,

$$k_L a << \Phi_L k_2 C_{BO} \tag{4}$$

Further, the reaction rate may be such that no appreciable amount of reaction takes place in the gas diffusion film, and the condition to be satisfied for this is given by,

$$\left[\frac{D_A k_2 C_{BO}}{k_L^2}\right] << 1.0 \tag{5}$$

When the conditions given by Eqs. (4) and (5) are satisfied, the rate of mass transfer is given by,

$$R_a = k_L a C_A^* \tag{6}$$

By measuring R_a, the rate of transfer per unit volume, the volumetric mass transfer coefficient, $k_L a$, can be determined.

The absorption of carbon dioxide in sodium–carbonate–bicarbonate solutions of certain concentrations satisfies the conditions given by Eqs. (4) and (5) and has been used for the present investigation.

3. Calculation of Physico-Chemical Constants

For calculating mass transfer coefficient, k_L, a knowledge of D_{CO_2}, k_2 and H_e is required. D_{CO_2} has been calculated from the equations suggested by Nijshing *et al.* (1959) and k_2 and H_e have been calculated as suggested by Danckwerts and Sharma (1966) and Porter *et al.* (1966), taking into consideration the ionic strength of the solution. Since it is also required to satisfy the conditions of the pseudo-first order reaction to know the value of the liquid side mass transfer coefficient k_L. As these values of k_L for this type of system are not available in the literature, the experiments were carried out to find the range of values of k_L. The theory of absorption accompanied by a slow chemical reaction was utilized to determine the values of the liquid side volumetric mass transfer coefficient $k_L a$ from which k_L was calculated. The experiments for the determination of $k_L a$ were carried out by absorbing CO_2 in a buffer solution of sodium carbonate-sodium bicarbonate. The volumetric mass transfer coefficient, $k_L a$, was calculated from the total rate of absorption of CO_2, which is true only when both the following conditions in Eqs. (7) and (8) are satisfied simultaneously,

$$k_L a < (1 - \Phi_G) k_2 C_{BO} \tag{7}$$

and

$$D_{CO_2} k_2 C_{BO} / k_L^2 < 1.0 \tag{8}$$

The physico-chemical parameters have been calculated by using the data and the equations summarized below.

3.1 Diffusivity of carbon dioxide in aqueous solutions

Diffusivity of carbon dioxide in aqueous solutions, was calculated by Eq. (9),

$$\frac{D_{CO_2} \text{ solution}}{D_{CO_2} \text{ water}} = \left[\frac{\mu_{water}}{\mu_{solution}}\right]^{0.85} \tag{9}$$

which was reported by Nijshing *et al.* (1959). The diffusivity in water at any temperature was obtained from the data of Nijshing *et al.* (1959) and was used with the Nerst–Einstein equation (Yamada and Goto, 1998) as well as Wilke and Cheng equation (Sheng and Tsui, 1999),

$$\frac{D_{CO_2} \mu}{T} = \text{constant} \tag{10}$$

$$D_{CO_2} = \frac{2.946 \times 10^{-4} \left(\psi M_W\right)^{0.5} T}{\mu_L v_A^{0.6}} \tag{11}$$

3.2 Solubility of CO$_2$

The solubility of the gas in the electrolyte solution was determined by the following equations suggested by Nijshing *et al.* (1959) and Danckwerts and Sharma (1966) taking the ionic strength of the solution into consideration. The solubility parameter of carbon dioxide in water, S_W;

$$\log_{10} S_W = \frac{1140}{T} - 10.306 \tag{12}$$

Where, S_W is in kgmol/(N·m) and T in K.

3.3 Solubility in electrolyte solution S

$$\log_{10} \frac{S}{S_W} = -K_S I \qquad (13)$$

where,

$$K_S = I_+ + I_- + I_g \qquad (14)$$

and I = ionic product. The values of contribution due to cation, anion, and gas, I_+, I_-, I_g, have been obtained from the data reported by Danckewerts and Sharma (1966). The Henry's law constant, H_e, is directly obtained from the solubility.

3.4 Stoichiometric first ionization constant

The stoichiometric first ionization constant K_1, was obtained from the data reported by Harned and Owen (1958), which can be written as,

$$\log_{10} K_1 = -\frac{3404.7}{T} + 14.843 - 0.0238T \qquad (15)$$

Values of stoichiometric second ionization constant K_2 can be obtained from the following equation.

$$\log_{10} K_2 = -\frac{2902.4}{T} + 6.498 - 0.0238T \qquad (16)$$

3.5 Concentration of OH^- ions in buffer solution

The concentration of OH^- ions in carbonate–bicarbonate buffer is given by,

$$C_{BO} = \frac{K_W}{K_1} \cdot \frac{\left[CO_3^{2-}\right]}{\left[HCO_3^-\right]} \qquad (17)$$

3.6 Ionic product of water

The ionic product of water, K_W, was available from Perry and Green (1984).

3.7 Second order reaction constant

The second order reaction velocity constant, k_2 was determined by the equation suggested by Porter *et al.* (1966).

$$\log_{10} k_2 = 13.635 - \frac{2895}{T} \qquad (18)$$

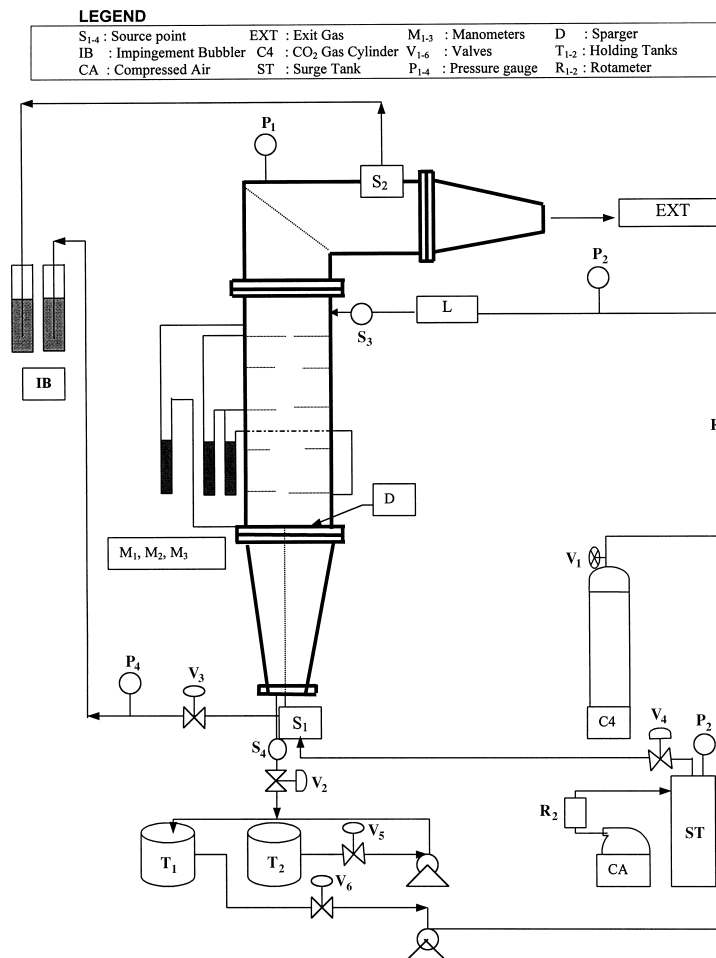## 4. Experimental Setup and Techniques

The experimental setup shown in **Figure 1** has been used for the measurement of the volumetric mass transfer coefficient. The experimental bubble column consisted of a vertical cylindrical Perspex column, 0.1905 m in diameter and 2.0 m long, fitted onto a fructo-conical bottom of mild steel. The latter had a divergence angle of 7° and a height of 0.87 m. The minimum diameter of the fructo-conical section was 0.10 m. The vertical cylindrical column was fitted with a total of five hollow disks termed as stages (3 contraction disks and two expansion disks). The expansion and contraction disks had central axial openings of 0.095 m and 0.0476 m, respectively.

At the bottom most section of the cylindrical column, just above the fructoconical cone, was fitted an antenna type sparger of 1.6 mm diameter with 144 holes for generating bubbles uniformly thoughout the entire cross section. The first contraction disk is the first stage (rupture disk) of 0.0476 m and the central opening was placed 0.26 m above the sparger. The first expansion disk is also known as the second stage (guide disk) of 0.095 m and the central opening was fitted at a height of 0.52 m above the sparger and the second contraction disk (third stage) was fitted at a distance of 0.78 m above the sparger. The column is divided into three distinct sections. Section-I consists of a sparger and a first contraction disk (up to 0.26 m). Section-II consists of a first contraction disk, first expansion disk and second contraction disk (0.26 to 0.78 m). Thus this section (section-II, 0.52 m in height) of the column consists of two contraction disks separated by an expansion disk. Section-III (from 0.78 to 1.30 m height from the sparger) and consisted of the contraction disk located at a height of 0.78 m above the bubble disperser, a second expansion disk (fourth stage) located at a height of 1.05 m and a third contraction disk (fifth stage) at a height of 1.30 m from the sparger. A 0.50-m clear space was provided above section-III, for allowing time for gas–liquid separation and also to accommodate bed expansion due to the bubbly flow. The details of the stages and sections are shown in Figure 1.

The experiments were conducted both with a constant liquid batch (without a liquid down-flow) and also with a continuous liquid down-flow (countercurrent to the upward gas flow). For conducting the experiments with a liquid down-flow, a water inlet was provided at a height of 1.80 m as shown (L) in Figure 1. All the hollow disks were supported on the column by means of threaded screws. At the bottom of the column a water outlet was provided. Four solenoid valves, at positions V_1, V_2, V_3 and V_4 were provided at the liquid and gas inlet and outlet respectively, for quick trapping of the flow when necessary. The liquid and gas flows into the column were controlled by valves.

For the determination of the mass transfer coefficient, the chemical rate of absorption of CO_2 diluted with air (6.0 to 10.0 mol% of CO_2) in aqueous solution of sodium carbonate-bicarbonate buffer solution were used. The buffer solutions of 1.0 m³ capacity was prepared and stored in a holding tank T_1 for continuous operation and outlet solution was collected in a sepa-

rate tank T_2, which is recycled again to tank T_1 after proper adjustment of desired concentration. The inlet and outlet liquid samples were collected at the points S_3 and S_4 respectively. Gas samples were also collected at sampling points S_1 and S_2, located at the inlet and outlet of the column after starting the steady state for 30 minutes. The gas and liquid phases in the samples were continuously separated in the column and analyzed by an impingement bubbler for the gaseous sample through volumetric analysis. Samples at the points S_1 and S_2 were withdrawn at an approximately iso-kinetic rate of 1 to 2×10^{-3} Nm^3/min to match the experimental gas flow rate and the conditions for iso-kinetic sampling. The pH was maintained in the range of 7.5 to 8.5 and temperature was kept constant in the range of 298–303 K. The gas samples were collected with the help of impingers containing known volume and concentration of BaCl$_2$ solution and aspirator bottles (Meikap, 2000) to find out inlet and outlet concentration of CO$_2$. Then the resulted masses from the impingers were analyzed by the volumetric method. The liquid samples were analyzed by standard methods of titration (Vogel, 1955).

In the actual experiments, first carbon dioxide gas obtained from a pressure gas cylinder (GC) was mixed with compressed air from laboratory supply in a mixing tank. This tank was made sufficiently big so that there was undisturbed supply of a constant composition CO$_2$-air mixture. This mixture was fed to the bottom of the bubble column while buffer solutions were fed from the top, continuously. By controlling the individual rates of flow a constant liquid–gas dispersion volume was maintained in the column. After the system had attained a steady state, the flow rates of the gas and liquid, the system pressure, temperature and the readings were noted, and the gas liquid samples were collected at the sampling points for subsequent analysis. The experiments were conducted in the range of experimental conditions presented in **Table 1**. Experiments were also conducted by changing the number of stages. Stages were always removed from the top of the column, in the sequence of fifth, fourth, third and second.

Table 1 Experimental conditions for mass transfer coefficients

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Superficial liquid velocity	2.42 to 9.67×10^{-3} m/s
Superficial gas velocity	0.106 to 0.19163 m/s
Na_2CO_3 concentration for k_L measurement	0.157.6 to 0.193.8 kgmol/m^3
$NaHCO_3$ concentration for k_L measurement	0.0583 to 0.0977 kgmol/m^3
CO_3^{2-}/HCO_3^- ratio in k_L measurement	2.70 to 3.33
pH	7.5 to 8.5
Inlet CO_2 concentration	6.0 to 10.0 mol%
Diameter of the reactor	0.1907 m
Temperature	298–303 K
Height of each section	0.26, 0.52 to 0.54 m
Number of stages in the column	five

5. Results and Discussion

Experiments were carried out for the determination of a volumetric mass transfer coefficient, k_La and individual mass transfer coefficient k_L with limited objective of finding representative values for newly developed column which was used as scrubber for pollution control (Meikap *et al*., 2002a). The values of k_La was calculated from Eq. (6) for R_a in a particular CO_2 inlet and outlet concentrations and air–CO_2 gas mixture flow rate. After knowing the volumetric mass transfer coefficient and corresponding interfacial area of contact and k_L was obtained which satisfied both the conditions (5) and (6). The magnitude of k_La obtained in the modified bubble column is typically about 0.12 to 0.30 s^{-1}, which is in good agreement with the range 0.10 to 0.25 s^{-1} (Biswas, 1975). The magnitude of k_L obtained varied from 4.2×10^{-2} to 6.0×10^{-2} m/s. Kasturi and Stephanek (1972) have obtained the k_La values of 0.1 to 0.55 s^{-1}, and k_L values of 2.75×10^{-2} to 6.0×10^{-2} m/s for the vertical two-phase flow in the annular regime. Biswas (1975) have reported the values of k_L for horizontal single-jet ejectors in the range of 3.2×10^{-2} to 4.3×10^{-2} m/s. Thus, it may be seen that the values of k_L and k_La obtained in the new system are in good agreement with the values reported in the literature. The variation of the values of k_La and k_L has been shown in plots under different operating conditions.

Figure 2 shows the effect of superficial gas velocity on the volumetric mass transfer coefficient, k_La, for the multi-stage bubble column and compared with the normal bubble column. It may be seen from this plot that the volumetric mass transfer coefficient, k_La, increases with an increase in Q_G for a particular superficial liquid velocity. This may be attributed to the increase in gas hold-up (Meikap *et al*., 2002b). It has been reported by Meikap *et al*. (2001) that this dependency of interfacial area, a, on Q_G is proportional to $Q_G^{0.89}$. Earlier workers for other type contactors have shown this type of dependency and they are proportional to $Q_G^{0.7}$ (Mashelkar, 1970), which is much less
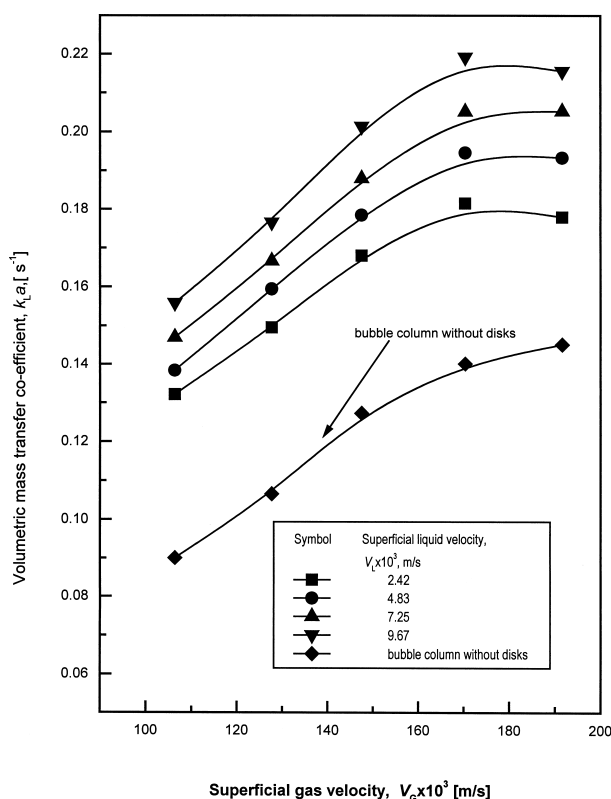
Fig. 2 Effect of superficial gas velocity on volumetric mass transfer coefficient, k_La

than that in the present system. It is interesting to note that the volumetric mass transfer coefficient, k_La, becomes almost constant beyond a superficial gas velocity of 160.44×10^{-3} m/s, this may be attributed to faster bubble coalescing than bubble formation. It can be seen from Figure 2 that the volumetric mass transfer coefficient of a normal bubble column (bubble column without any contraction and expansion disks) is always lower than that of the bubble column with disks shows that enhance bubble formation and bursting favorable for gas absorbers.
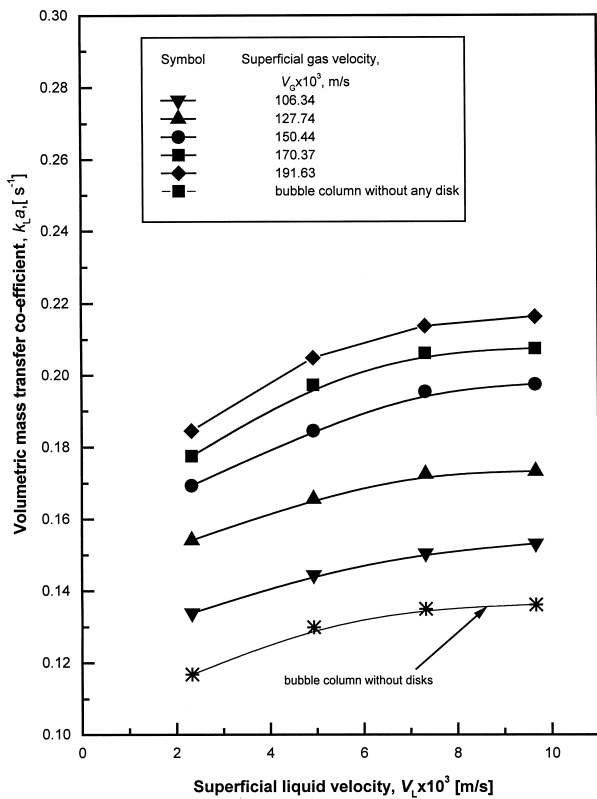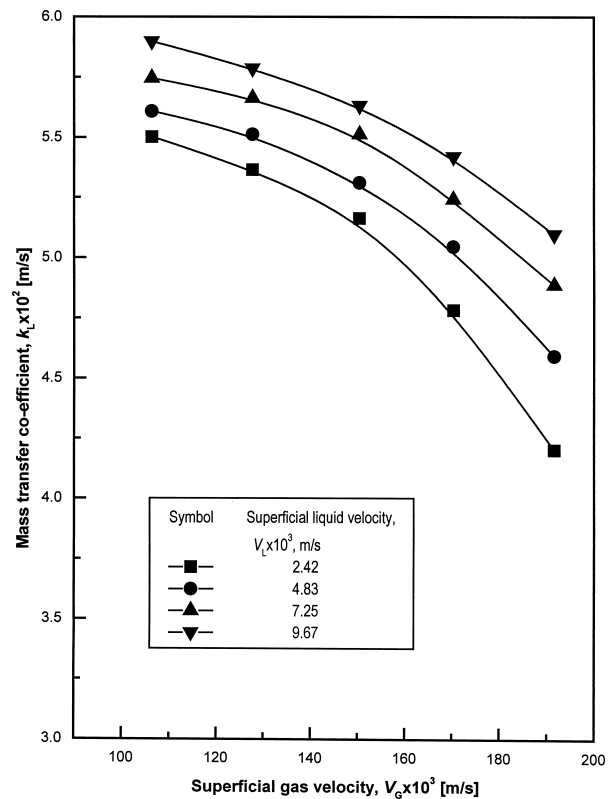
Fig. 3 Effect of superficial liquid velocity on volumetric mass transfer coefficient, $k_L a$

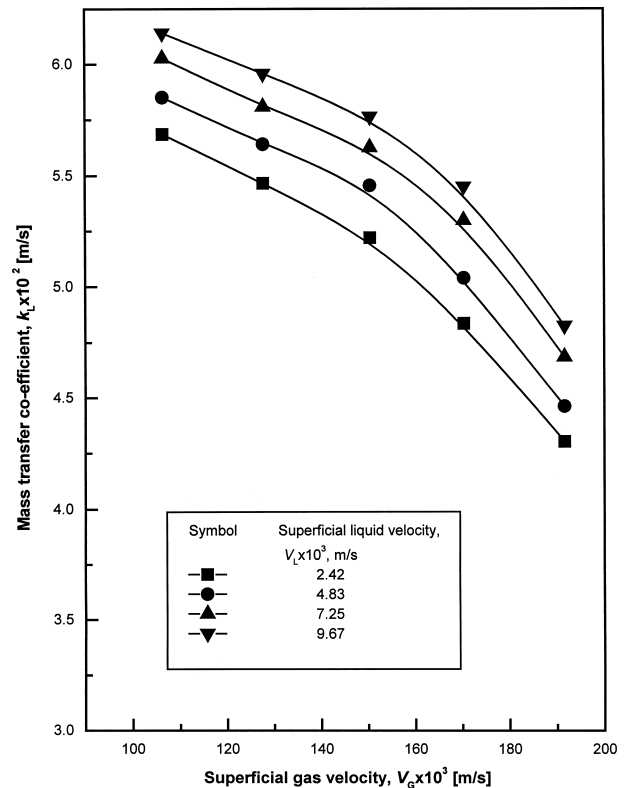**Fig. 4** Effect of superficial gas velocity on mass transfer coefficient, k_L (Section-I)

Figure 3 shows a typical plot of the effect of superficial liquid velocity on the volumetric mass transfer coefficient for the staged bubble column and compared with a simple bubble column where no disks were used. It can be seen from this figure that the effect of the superficial liquid velocity marginally enhances the volumetric mass transfer coefficient. However, the most interesting phenomena which was found is that the volumetric mass transfer coefficient was lower for the bubble column without disks compared with the bubble column with disks. This is, quite obviously, due to the lower interfacial area of contact than that of the staged bubble column.

In **Figures 4** and **5**, the values of the true mass transfer coefficient, k_L, are plotted against superficial gas velocity, V_G in section-I and section-II. It is seen from these figures that the film mass transfer coefficient, k_L, decreases with an increase in the superficial gas velocity for a particular superficial liquid velocity. This is obviously due to the increase in the gas hold-up and interfacial area of contact. It is observed that the k_L value distributed in the range of 4.2×10^{-2} to 5.8×10^{-2} m/s and the highest k_L was achieved at a superficial liquid velocity of 9.67×10^{-3} m/s.

Figure 6 is a typical plot of the film mass transfer coefficient, k_L, against superficial liquid velocity. It is seen from the Figure 6 that the film mass transfer

Fig. 5 Effect of superficial gas velocity on mass transfer coefficient, k_L (Section-II)

Fig. 6 Effect of superficial liquid velocity on mass transfer coefficient, k_L (Section-II)
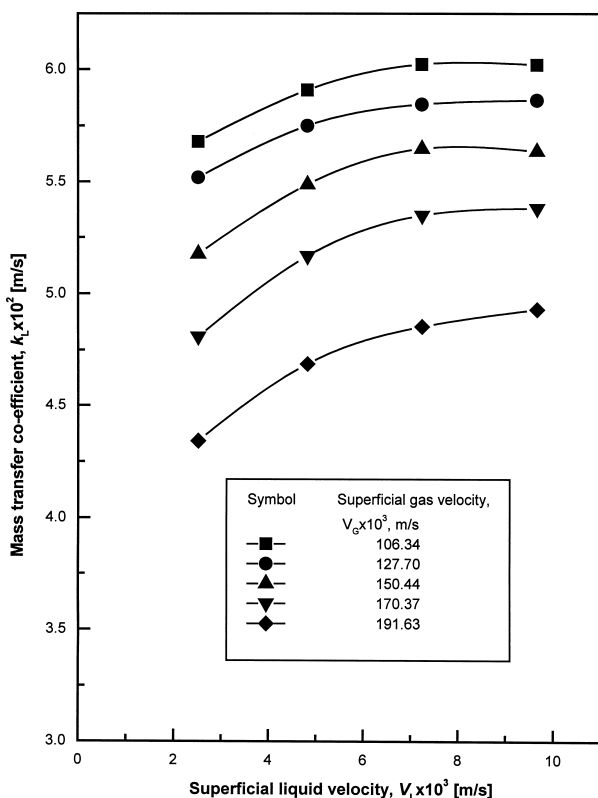

Fig. 7 Effect of number of stages on volumetric mass transfer coefficient, $k_L a$

coefficient, k_L, increases with an increase in V_L for a particular superficial gas velocity. This may be attributed to the increase in turbulence caused by superficial liquid velocity that lowers the gas liquid interface resistance. It may be seen that, in the investigated experimental range, the maximum interfacial area is obtained in the order of 250–600 m^2/m^3 (Meikap *et al.*, 2001). The specific interfacial area reported in the literature for ordinary bubble column contactors is of the order of 50 to 400 m^2/m^3 (Mashelkar, 1970; Deckwer, 1992). The higher value of the interfacial area is due to bubble generation, bursting and regeneration in the multi-stage, which does not occur in the ordinary bubble column. Though a direct comparison is not possible but it gives an idea of improvement in the present case. However, the enlargement of the interfacial area of contact contributes to surface renewal of bubbles followed by continuous bubble generation, breakup and regeneration.

Figure 7 shows the effect of stages on the volumetric mass transfer coefficient. It is interesting to note that the volumetric mass transfer coefficient increases as the number of stages increases up to four stages as high as the column height of 1.05 m. Beyond this the change in the increase of the volumetric mass transfer coefficient becomes negligible. This may be attributed
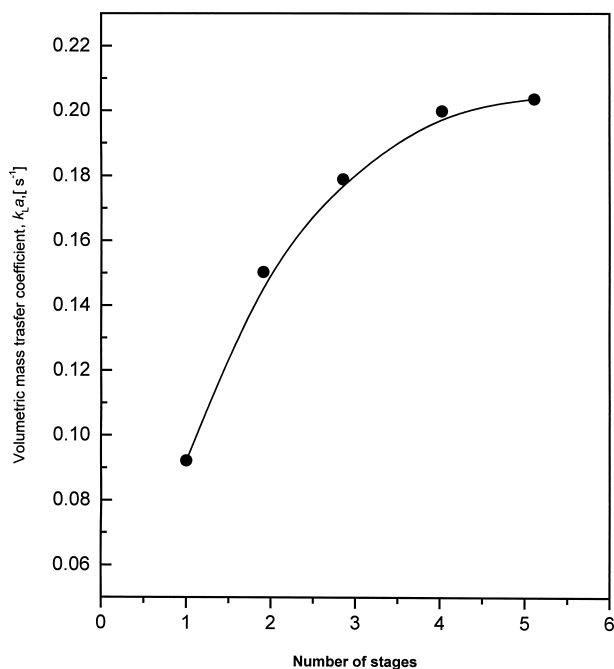
to the increase of bubble coalescence, rather than bubble bursting, predominates at this stage. This indicates that the multi-stage bubble column possibly maintain the good mass transfer (absorption) rate through out the column.

Conclusion

In the present investigation a bubble column with hydro-dynamically induced continuous bubble generation, break-up and regeneration was conceived, designed and fabricated. The system was designed to operate with relatively large sized bubbles, so that internal circulation can be induced in the bubbles and faster transfer of gaseous pollutants can take place from the bubble to the liquid through the interface of the bubbles and also followed by the direct rupture of the relatively large diameter bubbles. In the present investigation a detailed study of the liquid side mass transfer coefficient and volumetric mass transfer coefficient has been carried out for the design and characterization of the counter-current multi-stage bubble column scrubber. The obtained results are very effective and quite an improvement from the previous work reported earlier. The present system has wide applications in the field of gas–liquid reactors, aerator and wet scrubber for pollution control. The present system can also combat gaseous pollutants including fly ash and SO$_2$ in a technologically-economically-environmentally-friendly manner (Meikap *et al.*, 2002a).

Nomenclature

a	=	specific interfacial area per unit volume	[1/m]
C	=	liquid phase concentration	[kgmol/m^3]
D	=	diffusivity	[m^2/s]
D_C	=	diameter of bubble column	[m]
D_{CO_2}	=	diffusivity of CO_2 in liquid	[m^2/s]
D_H	=	diameter of expansion, contraction disks	[m]
D_L	=	dispersion coefficient, liquid phase	[m^2/s]
G	=	molar flow rate of reacting diluent gas	[kgmol/s]
H	=	height of the bubble column	[m]
H_e	=	Henry's law constant	[N·m/kgmol]
I	=	ionic product	[kg ion/m^3]
I_+, I_-, I_g	=	contribution to K_S	[l/kg ion]
J	=	moles of B reacting with one mole of A	[kgmol]
K_W	=	ionic product of water	[kgmol^2/m^6]
K_1	=	first ionization constant	[kgmol/m^3]
K_2	=	second ionization constant	[kgmol/m^3]
k_G	=	gas film mass transfer coefficient	[m/s]
k_L	=	liquid film mass transfer coefficient, for physical absorption	[m/s]
$k_L a$	=	volumetric mass transfer coefficient	[1/s]
k_1	=	first order reaction rate constant	[1/s]
k_2	=	second order reaction rate constant	[m^3/kgmol·s]
M_W	=	molecular weight of water	[kg/kgmol]
Q_G	=	volumetric flow rate of gas	[m^3/s]
Q_L	=	volumetric flow rate of liquid	[m^3/s]
R	=	rate of absorption per unit area	[kgmol/(m^2·s)]
R'	=	gas constant	[N·m/(kgmol·K)]
R_a	=	rate of absorption per unit volume	[kgmol/(m^3·s)]
S_W	=	solubility parameter of CO_2 in water	[kgmol/(N·m)]
T	=	temperature	[K]
V	=	operating scrubber volume	[m^3]
V_G	=	superficial gas velocity	[m/s]
V_L	=	superficial liquid velocity	[m/s]
v_A	=	solute molal volume at normal b.p.	[m^3/kmol]
μ_{eff}	=	effective viscosity of liquid	[kg/m·s]
μ_L	=	liquid viscosity	[kg/(m·s)]
Φ_G	=	fractional gas phase hold-up	[—]
Φ_L	=	fractional liquid phase hold up, $\Phi_L = 1 - \Phi_G$	[—]
ψ	=	association factor	[—]

<Subscript>

$-$	=	anion
$+$	=	cation
A	=	component A used for gaseous constant
aq	=	aqueous solution
B	=	component B used for the liquid
CO_2	=	carbon dioxide
G	=	gas
I	=	ionization
L	=	liquid

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