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HYDRODYNAMICS OF A CONCENTRIC TUBES AIRLIFT REACTOR

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ABSTRACT:- In airlift reactors transport phenomena are achieved by pneumatic agitation and circulation occurs in a defined cyclic pattern through a loop. In the present work, the effect of geometrical relations on gas holdup and liquid velocity, and consequently on the gas-liquid mass transfer coefficient, was studied in a 15-liter airlift bioreactor with A_d/A_r = 0.224, A_d : downcomer cross-sectional area, and A_r : riser cross-sectional area. Measurements of the volumetric oxygen transfer coefficient (K_La) were taken in a raw water-air system' using a sulfite oxidation method for depleting the oxygen to zero. The unsteady state concentration of oxygen is then measured using DO probe. Different conditions were examined by varying parameters such as superficial air velocity in the riser (U_{sg}), using two types of air sparger which have a remarkable effect on K_La values. The effect is due to their influence on gas holdup and liquid velocity, consequently affecting K_La. Superficial air velocity in the riser (U_{sg}) ranged from 0.03 to 0.06 $m.s^{-1}$ and K_La varied between 0.01 to 0.07 s^{-1}. Empirical equations were obtained between K_La and U_{sg} as:

$$K_La = 0.28\ U_{sg}^{0.53} \quad \text{for cross sparger}$$

$$K_La = 0.58\ U_{sg}^{0.86} \quad \text{for o-ring sparger}$$

***Keywords*:-** Airlift, Bioreactor, hydrodynamics, oxygen mass transfer rate.

INTRODUCTION

Airlift reactors (ALRs) designs have been used widely in many sectors of industrial activities. They are mainly used as bioreactors in fermentation processes and in the biotransformation of many substances (Sánchez Mirón et al. 2002; Acién Fernández et al., 2001). In wastewater treatment ALRs are increasingly being developed (Frijters et al., 1997; Heijnen et al. 1997; Van benthum et al., 1997 and 2000; Beun et al., 2002).

Airlift reactors are also used in several chemical applications, such as leaching, high tonnage heterogeneous catalyst industries as well as in the ethylene chlorination process (Orejas, 1999).

The advantages of ALRs consuming low energy input, provide efficient mixing, avoiding destruction in shear sensitive organisms, simple construction, good heat transfer and easier scale up.

Based on the configuration of the geometry, airlift reactors are generally classified into two main categories: (1) the internal loop (IL-ALR) where what would otherwise be a simple bubble column has been split into a riser and a downcomer by an internal baffle; and (2) the external loop (EL-ALR)- or outer loop airlift reactors where the riser and the downcomer are two quite separate tubes connected by horizontal sections near the top and bottom (Chisti, 1989).

The intrinsic complicated hydrodynamic structures induced by bubble motion and associated with wake interaction, have been recognized to be the key factors responsible for heat and mass transfers. Because bubble-induced flows in the airlift reactor are identified to be dynamic in nature, the time averaged flow properties cannot well represent the dynamic governing mechanisms of flow structures.

IL-ALR and EL-ALR have been widely studied experimentally. Some of these studies focus on liquid velocity circulation, gas and solid phase hold-ups (van Benthum et al., 2000) and on mass transfer (Nicollela et al., 1999).

To design and operate ALRs with confidence, the knowledge of gas-liquid mass transfer is required to characterize the performance of the ALR. The main parameter used as an indicator for gas-liquid mass transfer rate is the gas-liquid mass transfer coefficient (KLa). A large number of researchers (Koide et al., 1983; Chisti and Moo, 1988; Choi and Lee, 1993; Merchuk et al., 1994; Shimizu et al., 2001, Zhang et al., 2006; Tongwang et al., 2006) have investigated the mass transfer performance in the ALRs together with their hydrodynamic behavior. It was found that the knowledge of hydrodynamic behavior is critical for design purposes because of their strong influence on mass transfer. The characteristics generally used to describe hydrodynamic behavior in ALR consist of gas holdup, bubble average size and velocity, as well as liquid velocity or mixing time. In addition, each of these quantities may be influenced by several independent factors, such as superficial gas velocity, cross-sectional area, ratio between downcomer and riser, gas sparger, geometry, etc. Due to the rather complex interactions between these parameters, the design of ALR to fulfill specific purposes is still extremely difficult (Porntip et al., 2003; Peter et al., 2006).

Although a large number of investigations contributed to the knowledge of the effect of various parameters on hydrodynamic and mass transfer characteristics in ALRs, available information frequently showed wide variations and conflicting claims. The contradiction is regularly attributed to the difference in the reactor geometries, experimental conditions and experimental techniques. However the present knowledge suggests that this contradiction is brought about by some complicated phenomena taking place in ALR, such as the bubble size distribution, internal liquid circulation, etc (Samuel at al., 2005; Sarkar et al., 2008; Wei et al., 2008; Giovannettonea et al., 2009; Zhonghuo et al., 2010).

In the present work, a set of experiments on IL-ALR was conducted and focus on investigating the effect of U_{Gr}, sparger design on K_La; establishing empirical correlations to predict hydrodynamic behavior and mass transfer rate and investigating the effect of sparger design and superficial gas velocity on the gas hold-up.

GAS-LIQUID MASS TRANSFER THEORY

The gas-liquid mass transfer is one of the most important design considerations in bioreactor design, especially in aerobic systems which require a maximum rate of oxygen transfer. Due to a low solubility of oxygen in water, the oxygen must be continually replenished to avoid the development of anoxic conditions. The rate of mass transfer from gas to liquid phase may be expressed in terms of an overall volumetric mass transfer coefficient, K_La based on gas liquid dispersion volume. This volumetric oxygen transfer coefficient is also an important indicator for comparing the oxygen transfer capabilities of various aerobic bioreactors. The volumetric oxygen transfer coefficient is defined by the following equation (Chisti and Young, 1987):

$$K_L\,a = n_{O2}/\Delta C \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots.. (1)$$

where n_{O2} is the flux of oxygen transfer between phases, ΔC the concentration driving force between the two phases.

Generally, the determination of a mass transfer coefficient (K_L) in a system with a single bubble is much different from that in a system with a swarm of bubbles. The following subsections will give details on how to estimate the overall mass transfer coefficient from these two cases:

1. SINGLE BUBBLE SYSTEMS

The determination of the mass transfer coefficient in a single bubble system can be done theoretically, although it is more common to find empirical correlations that relate bubble characteristics to physical properties of the contacting system (Chisti, 1988).

2. SYSTEMS WITH A SWARM OF BUBBLES

In an industrial aerated reactors, air bubbles usually form swarms. The oxygen transfer coefficient correlations for air bubble swarms are different from single bubble correlations since hydrodynamics of the liquid around the bubbles are different. In addition, bubbles generally interact in some ways making the study of their behavior far more complicated than for a single bubble system.

The mass transfer coefficient in the bubble-swarm systems can also be estimated from the correlations between various dimensionless parameters (Chisti, 1988).

The gas-liquid interfacial area based on liquid volume or gas-liquid dispersion volume (a_L or a_D, respectively) need to be determined to evaluate overall mass transfer coefficient (K_La). The value of a_L and a_D can be evaluated from Eq. (2) and Eq. (3), respectively (Chisti, 1998):

$$a_L = 6\varepsilon_G / [\mathrm{d_B}\,(1- \varepsilon_G)] \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \ (2)$$

$$a_D = 6\varepsilon_G / \mathrm{d_B} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \ (3)$$

Where:

ε_G is the gas hold-up and $\mathrm{d_B}$ is the bubble diameter.

However, instead of determining K_L and a separately, the mass transfer behavior in these systems were usually presented in terms of the overall mass transfer coefficient (K_La) which was often determined using empirical correlations reported in literature.

These correlations are summarized and are listed in different literature (Chisti, 1998; Tanthikul, 2004) for both bubble column and airlift gas liquid contacting systems.

These correlations take the following form (Chisti, 1998):

$$K_La \ = \ C_1\, U_{sg}^{B} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \ (4)$$

Where:

U_{sg} superficial gas velocity ,C and B are constant and power respectively which its values differs from different resources.

HYDRODYNAMIC BEHAVIOR

Hydrodynamic behavior is essential for the understanding of the phenomena taking place in ALR. Due to their strong influence on mass transfer performance, they have received considerable attentions from most investigators. Hydrodynamic parameters of interest in design are the overall gas holdup, the gas holdups in the riser and in the downcomer, the magnitude of the induced liquid circulation and the liquid phase dispersion coefficients in various regions of the reactor.

GAS HOLDUP

The volume fraction of gas or gas holdup is an essential parameter for the design of airlift contactors. Due to the configuration of airlift contactors that allow aeration in the riser, gas holdup in riser is usually higher than the downcomer. This difference in gas holdups is the main cause of pressure difference, which creates liquid circulation pattern. Gas holdup, ε_o, is the ratio between volume of gas phase and the total volume of reactor (volume gas phase plus volume of liquid phase) or can be expressed as (Chisti, 1989):

$$\varepsilon_o = \frac{V_G}{V_G + V_L} \qquad \text{...} (5)$$

Where :

ε_o is overall gas holdup , V_G is gas volume and V_L *is* liquid volume in reactor.

EXPERIMENTAL WORK

Apparatus

The reactor apparatus was a concentric-tube airlift reactor with dimensions given in Table 1. The volume of the reactor was 15 liter and A_d/A_r=0.224, where A_d is the downcomer superficial area (m^2) and A_r is the riser superficial area (m^2). The water level in the reactor was 107 cm. The reactor was constructed of glass with the bottom and top plates made of rigid nylon. Water manometer was used to measure the pressure drop across the reactor and the distance between the manometer reading two points was 75 cm. Air spargers and other pipes were constructed of copper. Figure 1 shows the schematic arrangement of the experimental apparatus.

For determination of K_La, experiments were carried out with aqueous sodium sulfite solution and air as the gaseous phase. Air was sparged through 4.5 cm diameter ring, with 12 holes of diameter of 1 mm and a cross (+) sparger with a length of 4.5 cm and 12 holes of diameter 1 mm. Airflow rates were measured by a rotameter (max. reading 2 m^3/h) from Rota company of QVF type. All experimental runs were carried out at atmospheric pressure and a temperature of 28˚C. A series of experiments were performed by varying the superficial gas velocity (with respect to the cross-sectional area of the riser) over the range of 0.03–0.06 ms^{-1} to create a characteristic velocity curve of the airlift reactor.

MEASUREMENT OF VOLUMETRIC MASS TRANSFER COEFFICIENT

The overall volumetric mass transfer coefficient, K_La, was measured using the well-known dynamic oxygenation method (Vasconcelos et at., 2003; Vandu et al., 2004). The dissolved oxygen concentration in the batch liquid phase was measured by means of an oxygen probe inserted horizontally at 0.1 m below the exit of the riser that was connected to a DO-meter type LTH Lutron 5509. The oxygen probe signals were measured using A/D converter and recorder on a PC. In each experimental run, tap water has been first stripped of oxygen by chemical reaction. For this purpose, a small but sufficient amount of sodium sulfite solution is added at the beginning and in the liquid in the presence of cobalt or cupper sulfate as a catalyst in order to avoid the modification of the coalescent behavior of the system. The reaction between sodium sulfite and oxygen in the liquid phase is given by the stoichiometric equation .

$$Na_2SO_3 + 1/2\ \ O_2 \xrightarrow{CuSO_4} Na_2SO_4$$

METHOD OF CALCULATIONS

The overall oxygen volumetric mass transfer coefficient

The K_La is determined by using the dynamic method. The investigations of mass transfer characteristics were restricted to oxygen transfer only, and in all investigations, the ALR systems were subject to the following assumptions (Wongsuchoto, 2002):

- Gas composition is constant.
- The system is isothermal, and the effect of the dynamics of the dissolved oxygen electrode is negligible.
- For sparingly soluble gases such as oxygen, the liquid phase volumetric mass transfer coefficient (K_La) is nearly equal in value to that of the overall volumetric mass transfer coefficient (K_La).

A material balance on dissolved oxygen according to the above assumption gives the following equation (Wongsuchoto, 2002):

$$\frac{dC}{dt} = k_L a\left(C^* - C\right) = K_L a\left(C^* - C\right) \quad \text{………………………………(6)}$$

Where:

C^* : saturation dissolved oxygen concentration.

C : dissolved oxygen concentration in liquid phases .

Integrate Eq. (6) with the limits of $C = C_o$ at $t = 0$ and $C = C$ at $t = t$ results in:

$$\int_{C_0}^{C} \frac{dC}{\left(C^* - C\right)} = k_L a \int_0^t dt \quad \text{……………..………………… (7)}$$

$$\ln\left[\frac{C^* - C_0}{C^* - C}\right] = k_L a t \quad \text{…………………….………… (8)}$$

The value of $K_L a$ is obtained from the slope of the linear regression with ln [(C^* - C_0)/ (C^* - C)] with respect to time (t).

GAS HOLD-UP

The total gas holdup was determined by the expansion volume method (Chisti, 1989). This method was chosen because it was the simplest to use. The gas holdup was estimated as the percentage increase in volume of the gassed liquid compared with ungassed liquid volume. In this airlift bioreactor the variation of liquid volume can be determined by observing the height of the surface of the ungassed liquid and aerated liquid. The dispersion height was estimated by observing the position of the liquid level on a graduated stainless-steel rod suspended from the vessel top plate. At high gas flow rates the liquid surface become very turbulent, with the level changing erratically, and so a mean dispersion height was estimated (Chisti, 1989).

Because the volume of gas cannot be measured directly, we defined V_D (dispersed volume) as the total volume of gas phase plus volume of liquid phase. Then

$$\varepsilon_o = \frac{V_D - V_L}{V_D} \quad \text{……………………………………………………………(9)}$$

$$\varepsilon_o = 1 - \frac{V_L}{V_D} \quad \text{………………...…………………………………………... (10)}$$

$$\varepsilon_o = 1 - \frac{h_L A}{h_D A} \quad \text{.....(11)}$$

Finally, $$\varepsilon_o = 1 - \frac{h_L}{h_D} \quad \text{.....(12)}$$

Where:

h_D dispersed liquid height (cm) and h_L liquid height (cm).

The downcomer gas holdup was estimated by measuring the pressure difference between the two measuring ports of the column where

$$\Delta P = \Delta Z_{manometer} \quad \text{.....(13)}$$

$$(\rho_L \varepsilon_{Ld} + \rho_G \varepsilon_{Gd}) g \Delta H = \rho_L g \Delta Z_{manometer} \quad \text{.....(14)}$$

Neglecting the wall friction loss and based on the fact that $\rho_L >> \rho_G >>$, Equation (14) can be deduced to:

$$\varepsilon_{Ld} = \frac{\rho_L g \Delta Z_{manometer}}{\rho_L g \Delta H} \quad \text{.....(14)}$$

$$1 - \varepsilon_{Gd} = \frac{\Delta Z_{manometer}}{\Delta H} \quad \text{.....(15)}$$

$$\varepsilon_{Gd} = 1 - \frac{\Delta Z_{manometer}}{\Delta H} \quad \text{.....(16)}$$

Where:

ρ_G density of gas , ρ_L density of liquid, ε_{Gd} gas holdup in the downcomer , ε_{Ld} liquid holdup in the downcomer , ΔP : pressure difference of defined liquid level.

ΔZ : distance of liquid level in manometer, ΔH : distance of liquid level.

The riser gas holdup was calculated from following equation:

Firstly, $$\varepsilon_{Gr} = \frac{V_{Gr}}{V_{Gr} + V_{Lr}} \quad \text{.....(17)}$$

$$\varepsilon_{Gr} = \frac{V_{GT} - V_{Gd}}{V_{GT} - V_{Gd} + V_{LT} - V_{Ld}} \quad \text{.....(18)}$$

$$\varepsilon_{Gr} = \frac{\varepsilon_o V_T - \varepsilon_{Gd} V_d}{V_T - V_d} \quad \text{.....(19)}$$

$$\varepsilon_{Gr} = \frac{\varepsilon_o - \varepsilon_{Gd}(V_d / V_T)}{1 - (V_d / V_T)} \quad \text{.....(20)}$$

$$\varepsilon_{Gr} = \frac{\varepsilon_o - \varepsilon_{Gd}\left[A_d h / (A_d + A_r) h\right]}{1 - \left[A_d h / (A_d + A_r) h\right]} \quad \text{.....(21)}$$

Finally, $\varepsilon_{Gr} = \varepsilon_o + (A_d / A_r)(\varepsilon_o - \varepsilon_{Gd})$ ………………………………….(22)

Where :

ε_{Gr} riser gas holdup ,V_{Gr} gas volume in the riser, V_{Lr} liquid volume in the riser, V_{GT} total volume of gas , V_{Gd} gas volume in the downcomer ,V_{LT} total volume of liquid, V_T total volume, and (A_d / A_r) the ratio between cross sectional area of downcomer to riser.

RESULTS AND DISCUSSION

Effect of airlift configuration on gas holdups

Gas hold-up is an important parameter, because it determines the amount of the gas phase retained in the system at any time. The slope of the increase of hold up in the riser and downcomer with superficial gas velocity depends on the existing two-phase flow regime.

Therefore, the change of the slope in the dependence of the gas hold-up versus superficial gas velocity in the riser and the downcomer was used for the estimation of transition of the flow-patterns from the bubble flow regime to the churn flow regime.

The sectioning in airlift system rendered an uneven distribution of gas bubbles in the system, i.e. gas holdups in riser, in downcomer and overall gas holdups which are the amount of gas in riser and downcomer sections and the overall quantity of gas in the system, respectively. These parameters are affected by the superficial gas velocity and the type of air sparger as shown in Figs. 2-4. Many aspects of airlift contactors depend not only on the overall gas holdup but also on the distribution of holdups between the riser and the downcomer. Many Literatures revealed that increasing superficial gas velocity (U_{sg}) increased the gas holdup (Vial et al., 2002; Peter et al., 2006; Wei, et al., 2008; Zhonghuo et al., 2010) which is in agreement with the data represented in Figs. 2-4. Table 2, shows manometer readings for both o-ring and cross type air distributors.

VOLUMETRIC MASS TRANSFER COEFFICIENT

The relationship between the dimensionless oxygen concentration and time for cross and ring type air sparger at different superficial are shown in Figs. 5 and 6. These data were used for the estimation of the overall volumetric mass transfer coefficient (K_La) as described in the experimental section (C^*=9.1 mg/L).

Overall volumetric mass transfer coefficient (K_La) is one of the major parameters for the design of bioreactors. K_La usually depends on several design parameters such as Ad / A_r, types of sparger, liquid phase properties, and superficial gas velocity.

The relationships between the overall volumetric mass transfer coefficient (K_La) in the airlift bioreactor with various values of superficial gas velocity are shown in Figure 7. This demonstrates that K_La increased with U_{sg}. Therefore it should be reasonable to conclude that the increase in specific mass transfer area with u_{sg} in the airlift reactor was the main factor responsible for the increase in K_La . This increase in specific mass transfer area was attributed to the increase in the overall gas holdup and the decrease in the bubble size with gas throughput. This observation is in agreement with many experimental works (Carvalho et al., 2000, Peter et al., 2006; Blažej et al., 2004; Samuel et al., 2005; Zhang et al., 2006; Zhonghuo et al., 2010).

The effect of o-ring sparger on K_La appeared more effective compared to the cross sparger of the same range of condition that employed in this study and as appeared clearly in Fig. 7. This is may be attributed to the increase of gas holdup which increases the contact area between the liquid and air.The values of K_La of the two types of air spargers were correlated with equation (4) and the correlations parameters were obtained and listed in Table 3:

$$K_La = C_1 \quad U_{sg}^{B} \qquad \text{……………………………………………………..} \; (4)$$

CONCLUSIONS

The experimental study on gas holdup in a concentric internal loop airlift bioreactors revealed that the gas distributor characteristics and superficial velocity are very important factors affecting gas holdup and volumetric mass transfer coefficient. The o-ring sparger type are more effective compared to the cross sparger at the same range of condition.

The following Empirical equations were obtained:

$K_La = 0.28\, U_{sg}^{\,0.53}$ for cross sparger .

$K_La = 0.58\, U_{sg}^{\,0.86}$ for o-ring sparger.

which relates the volumetric mass transfer coefficient for two different gas spargers with U_{sg} which added to many correlations exist in literature for evaluation of hydrodynamics of airlift bioreactors.

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Table (1): Dimensions of a concentric tube airlift reactor.

	Height (cm)	Diameter (D_o) (cm)	Diameter (D_i) (cm)	Thickness (cm)
Main column	133	12	11.4	0.3
Draft tube	75	6.0	5.4	0.3

Table (2): Airlift manometer readings for both o-ring and cross type air distributors.

U_g (cm/s)	3.2	4.3	5.6	6.2
Manometer Readings (ΔZ), cm				
O-ring type air distributor	71.175	70.275	68.85	67.725
Cross-type air distributor	71.775	70.95	69.225	68.175

Table (3): Correlations constants values and the statistical analysis.

Cross type air sparger			
C_1	B	R(residual value)	Variance
0.28	0.53	0.9898	99.25%
	Mean	st. dev.	
K_La	0.032750	0.016439	
U_{sg}	0.045000	0.012910	
O-ring type air sparger			
C_1	B	R(residual value)	Variance
0.58	0.86	0.996	99.25%
	Mean	st. dev.	
K_La	0.050500	0.016823	
U_{sg}	.045000	0.012910	

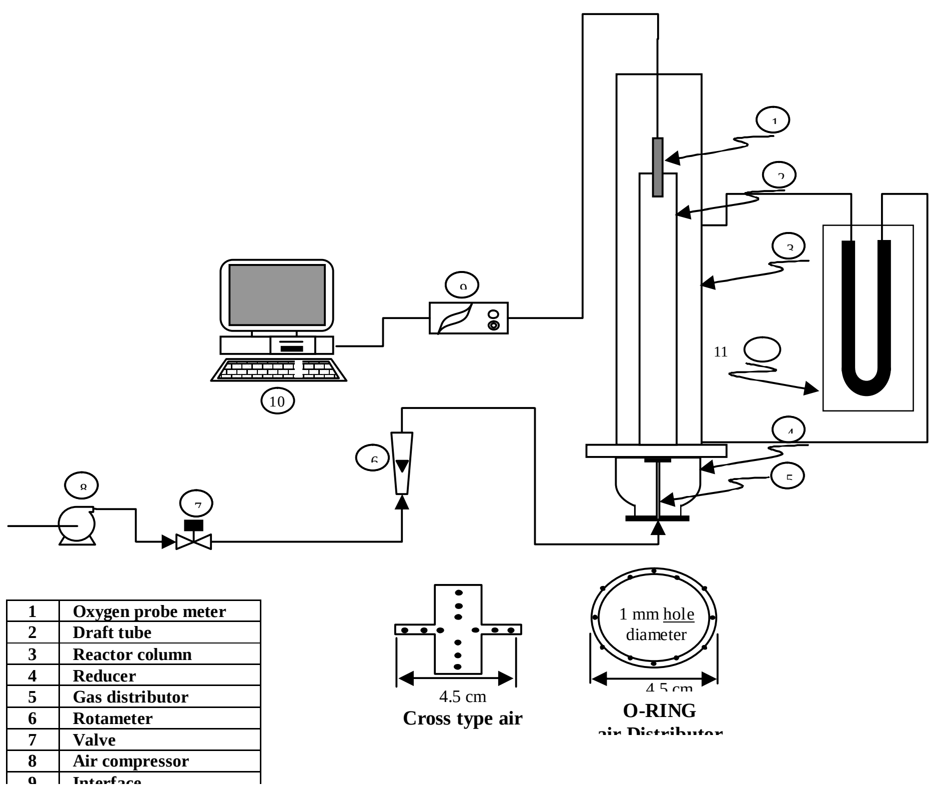

Fig.(1): Schematic diagram of experimental apparatus.

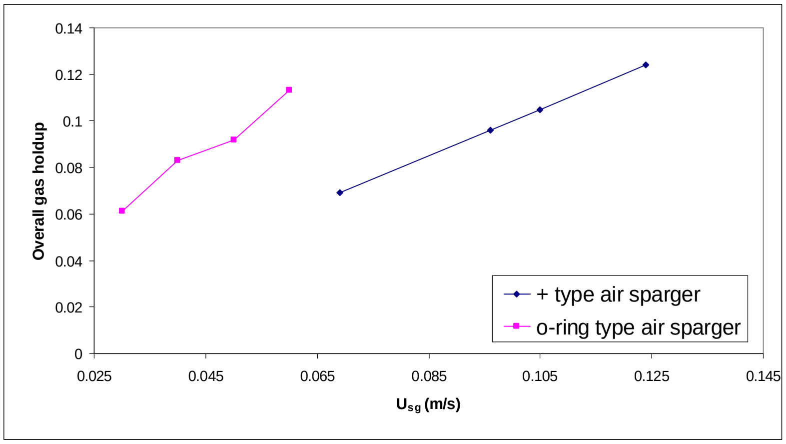

Fig. (2): The relationship between the overall gas holdup and superficial gas velocity for cross and o-ring type sparger.

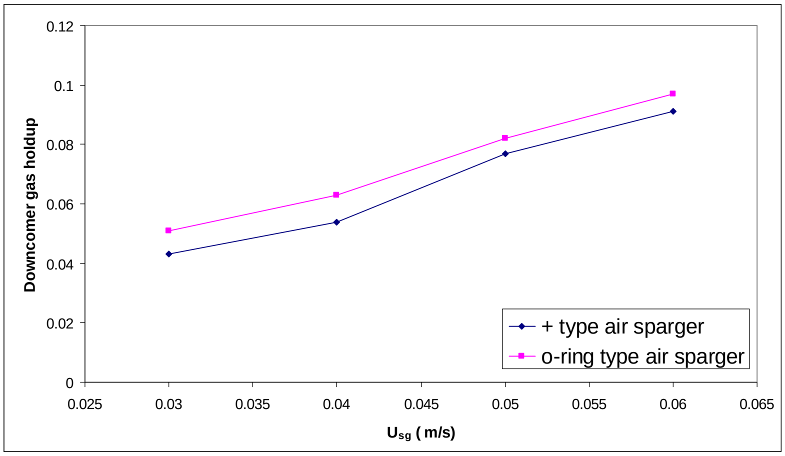

Fig.(3): The relationship between the downcomer gas holdup and superficial gas velocity for cross and o-ring type sparger.

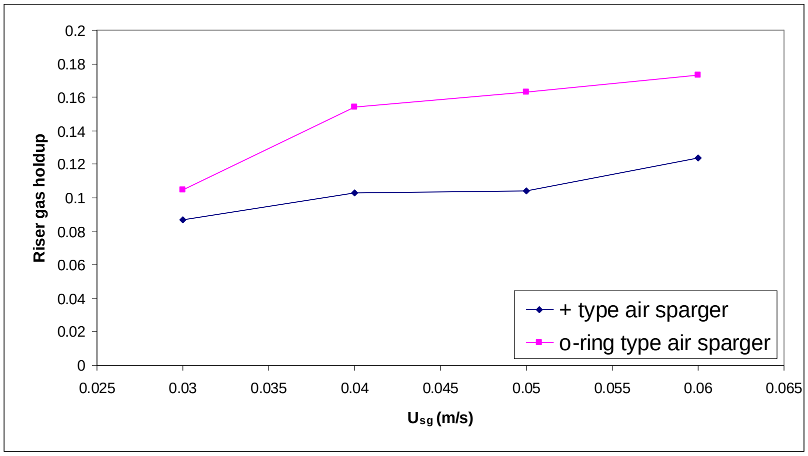

Fig.(4): The relationship between the riser gas holdup and superficial gas velocity for + and o-ring type sparger.

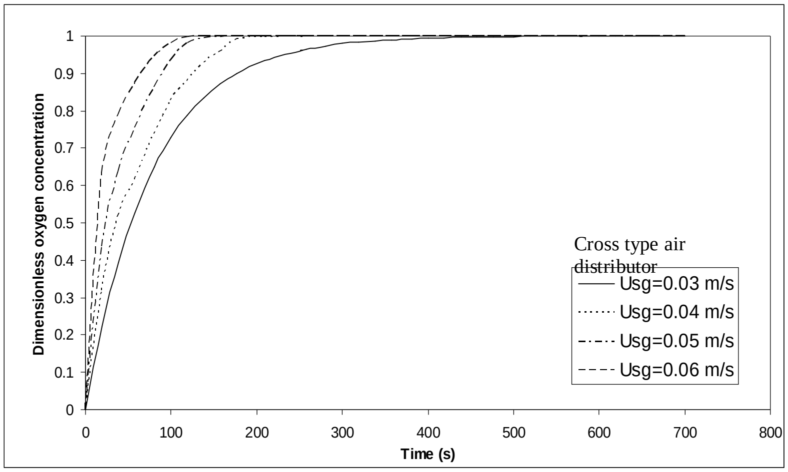

Fig. (5): The relationship between the dimensionless oxygen concentration and time for cross type air sparger.

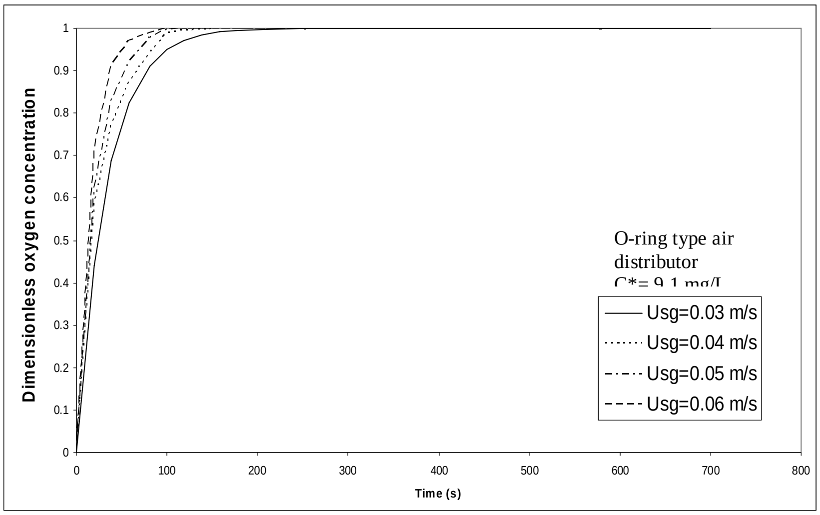

Fig.(6): The relationship between the dimensionless oxygen concentration and time for o-ring type air sparger.

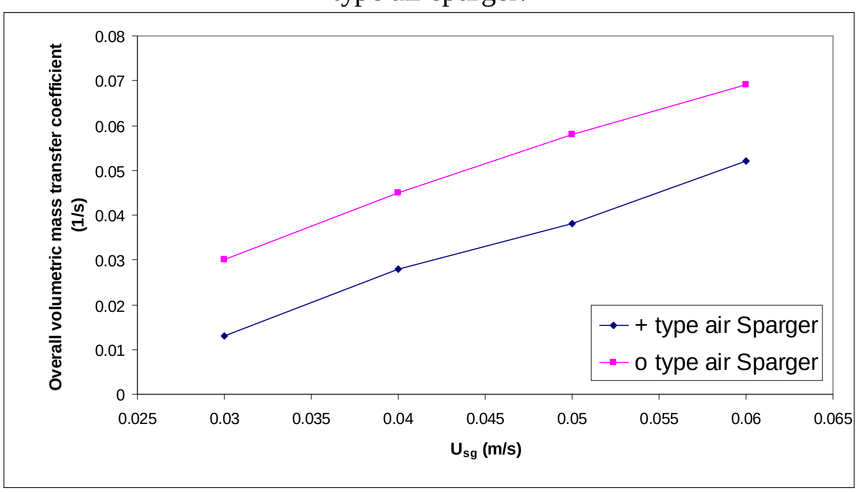

Fig. (7): The relationship between the overall volumetric mass transfer coefficient and superficial gas velocity for cross and o-ring type sparger.

هيدروديناميكية المفاعل الأنابيب المتمحور ة من نوع Airlift

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الخلاصة

تتحقق ظاهرة الانتقال في المفاعلات الحيوية من نوع (airlift) بواسطة الخلط الهوائي والتدوير الذي يحدث في دوائر محددة خلال حلقة معينة.تمت دراسة الخصائص الهندسية وتأثيراتها على الغاز المعلق وسرعة السائل وبالتالي على معامل انتقال الكتلة بين السائل والغاز في مفاعل حيوي من نوع (airlift) ذو حجم (15 ليتر) و (A_d/A_r=0.224) حيث ان A_d : مساحة المقطع للسائل النازل و A_r : مساحة المقطع للغاز المتصاعد . أخذت قياسات المعامل الحجمي لانتقال الأوكسجين في نظام ماء-هواء باستعمال طريقة أكسدة الكبريتيد لخفض الأوكسجين المذاب إلى الحدود الدنيا (الصفرية). تم اخذ قياسات الأوكسجين المذاب للحالة الغير المستقرة باستعمال متحسس الأوكسجين المذاب. أجريت التجارب بظروف مختلفة من خلال تغيير عدة عوامل مثل سرعة الغاز الظاهرية في ممر الغاز المتصاعد باستعمال نوعين من موزعات الغاز والتي كان لها تأثير واضح على المعامل الحجمي لانتقال الأوكسجين المذاب. إن التغيير في نوع الموزع آثر على حجم الغاز المعلق وسرعة السائل وبالتالي على مقدار المعامل ألحجمي لانتقال الأوكسجين المذاب. تراوحت قيم السرعة الظاهرية للغاز في ممر الغاز المتصاعد بين (0.03-0.06) م/ثا وتراوحت قيم المعامل الحجمي لانتقال الأوكسجين المذاب بين (0.01-0.07) 1/ثا . تم التوصل إلى المعادلة التطبيقية التالية بين K_La و U_{sg} :

$$K_La = 0.28\ U_{sg}^{0.53} \text{ for cross sparger}$$
$$K_La = 0.58\ U_{sg}^{0.6} \text{ for o-ring sparger}$$