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'Influence of Draft Tube Diameter on Operation Behavior of Air Lift Loop Reactors

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Abstract

The ratio of draft tube to reactor diameters is of decisive importance for the operation behavior of air lift loop reactors. The influence of draft tube geometry was investigated with respect to oxygen mass transfer and mixing time. The diameter ratio was varied between 0.33 and 0.80. The measurements were performed in two loop reactors with liquid capacities of 11.775 and 26.49 liters using aqueous with solutions of different coalescence behavior. The results show that there is no single diameter ratio which would produce most favorable conditions for the two process parameters. With respect to the more important requirements of aerobic cultures, i.e high oxygen mass transfer and efficient mixing, a diameter ratio between 0.5 and 0.6 is recommended. If high liquid velocities in the draft tube are required a ratio of 0.6 should be used.

Keywords: *Draft tube, loop reactors, mass transfer, mixing time.*

1. Introduction

Bubble column reactors belong to the general class of multiphase reactors which consist of three main categories namely, the trickle bed reactor (fixed or packed bed), fluidized bed reactor, and the bubble column reactor. A bubble column reactor is basically a cylindrical vessel with a gas distributor at the bottom. The gas is sparged in the form of bubbles into either a liquid phase or a liquid solid suspension.

These reactors are generally referred to as slurry bubble column reactors when a solid phase exists. Bubble columns are intensively utilized as multiphase contactors and reactors in chemicals, petrochemical, biochemical and metallurgical industries[1].

Airlift loop reactors are important as bioreactor for aerobic fermentation and waste water treatment on account of their simple construction and low energy consumption together with high mass, momentum and heat transfer rates. The developed bioreactors configurations include the loop principle[2]. Among the different types of loop reactors, such as those with concentric draft tube[3] or tubular loop[4, 5], the former predominate in industrial practice[6, 7]. In the majority of designs, air is introduced and dispersed at the base of the draft tube resulting in a liquid up flow within the draft tube and down flow in the annulus.

Aeration of the annular section results in a countercurrent flow, with poorer stability at high gas throughputs[8] but a better heat transfer via the column wall[10]. Although many investigations were carried out in order to achieve optimization of the loop reactor design[9, 11], little information has so far been published about the influence of the diameter ratio of draft tube to column wall (D_D/D) on fluid dynamics and mass transfer.

In many cases of application a ratio of approximately (D_D/D) =0.5 is used, which produces the least circulation resistance and a maximum circulation rate. The diameter ratio determines not only the pressure losses in the loop but also the over all reactor behavior since the draft tube divide the reactor into two zones of completely different two phase flow behavior .

Therefore, the optimum diameter ratio for a maximum circulation rate (D_D/D) = 0.5 is not the most favorable with respect to other important process parameters such as mass transfer[10].

2. Fluid dynamics of airlift loop reactors

The fluid dynamic process in airlift loop reactors such as liquid circulation, distribution and dispersion of the gas phase and also the resulting mass, momentum and heat transfer rates are determined by a complex interaction between buoyancy, inertia, friction and hydrostatic pressure.

The circulation is induced by the difference in hydrostatic pressure between the aerated draft tube and the gas – free liquid in the annular section. The hydrostatic pressure difference at the bottom of the reactor (ΔP_H) can be evaluated if the gas holdups in the draft tube (ε_D) and in the annulus (ε_A) are known:

$$\Delta P_H = \rho_L g H_L (\varepsilon_D - \varepsilon_A) \qquad ...(1)$$

At steady state, the sum of the pressure drops in the draft tube (ΔP_D), annular region (ΔP_A) and the deflection zones (ΔP_R) are equal to the hydrostatic pressured difference (ΔP_H):

$$\Delta P_H = \Delta P_D + \Delta P_A + \Delta P_R \qquad ...(2)$$

According to the laws of fluid dynamics the pressure losses in the individual reactor zones, can be expressed by the liquid flow velocity (V_{Li}) and fraction factor (F_i):

$$\Delta P_i = F_i \frac{\rho_L}{2} V_{Li}^2 \qquad ...(3)$$

Therefore the ratio of liquid flow velocities in the draft tube and annular region is determined mainly by the ratio of the cross sectional area of the two reactor zones, as a function of the diameter ratio.

The relationship between the liquid flow velocities in the annulus and the draft tube suction for several diameter ratios in on phase flow is strongly influences the distributions of the gas phase in the different reactor zones[12].

A decreasing ratio of draft tube to reactor diameter leads to two different effects. On the one hand the superficial gas velocity in the draft tube increase at constant gas through put and on the other hand increasing liquid velocity as a result from the variation of draft tube geometry .

If the effective down flow velocity of the liquid phase in the annulus exceeds the free rise velocities of the bubbles in stagnant liquid, bubbles are sucked into the annular region and swept down by the recirculation liquid.

Depending on the coalescence behavior of the liquid phase, the bubbles in the annulus coalesce and form larger bubbles which can stagnate in the down flow zone[13].

At small draft tube diameter the liquid velocity in the annulus is too low to convey the bubbles down words while, at the same time, large bubbles are formed in the draft tube by coalescence due to the high bubble concentration.

At increased diameter ratios even larger bubbles can recalculate at the same gas through put. The recirculation of larger bubbles is of particular advantage in oxygen consuming processes, such as aerobic fermentations, because the larger bubbles are not so rapidly depleted of oxygen as the smaller bubbles.

A high down flow velocity in the annulus, which accurse only at larger diameter ratios, is therefore favorable for obtaining a better oxygen supply and shorter residence time[14].

The purpose of this study is to clarify experimentally the effects of the draft tube to reactor diameter ratio on gas mass transfer coefficient and mixing time for the case when air is sparged into the base of the draft tube.

The measurements were performed in air lift loop reactors of different volumes (11.775 and 26.49 liters) using aqueous solutions differing in coalescence behavior and viscosity.

3. Experimental

Experimental Set-up

Two air loop reactors of different sizes (11.775 and 26.49) liters were investigated. A schematic diagram of the experimental set up used in this work is shown in figures (1, 2 and 3). A Plexiglas column of (0.1, 0.15) m in inside diameter and about 1.60 m total height with draught tube dimensions of (0.05, 0.075, 0.08, 0.1) m inside diameter and 1.20 m total height was used.

The draft tube was fitted with three support legs at the upper and the lower end of the column so as to locate it in central position at any distance above the base. The column consist of two main sections namely, the gas inlet section and the liquid recycling testing section.

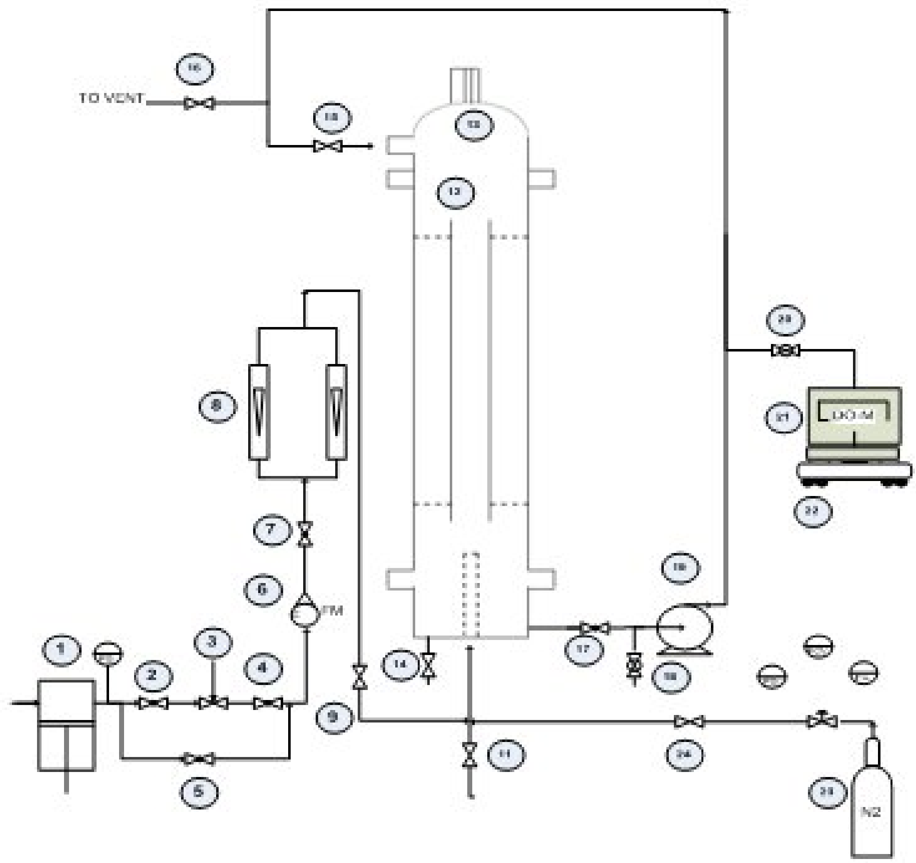


Fig.1. Experimental-Apparatus (1: Air Compressor; 2,4,5,7,17: Globe Valves; 3: Needle Valve; 6: Gas Meter; 8: Gas Rotameter; 9: Two-Way Valve; 10: Gas Distributor; 11,14,15,16,24: Gate Valves; 12: Draught Tube; 13: Column; 18,20: Ball Valves; 19: Centrifugal Pump; 21: Dissolved Oxygen Meter (Do-m); 22: Move Table; 23: Nitrogen Cylinder; PG, PCV: Pressure Control Valve Pressure Gauge).

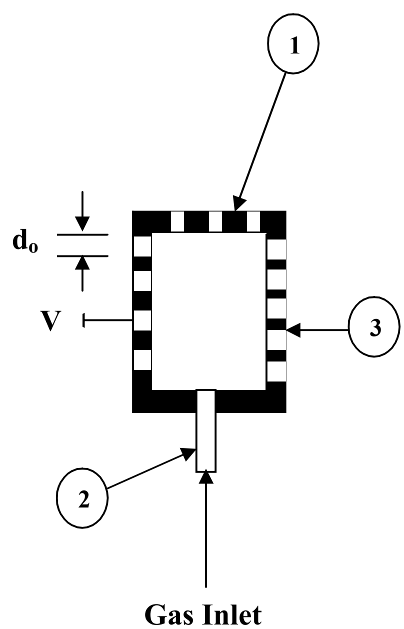
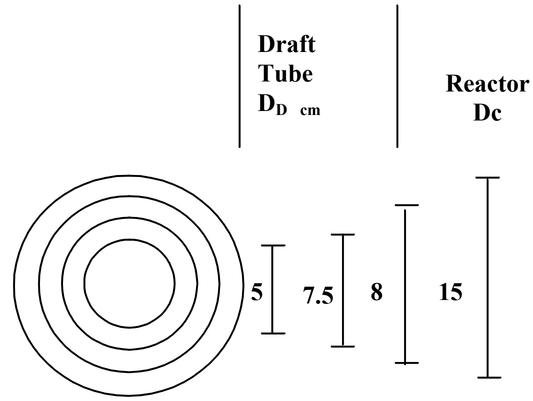


Fig.2. Gas-Distributor (1: Ceramic Material; 2: Pipe; 3: Hole).

D (cm)	D_D (cm)	D_D/D
15	7.5	0.5
10	7.5	0.75
15	8	0.533
10	8	0.8
15	5	0.33



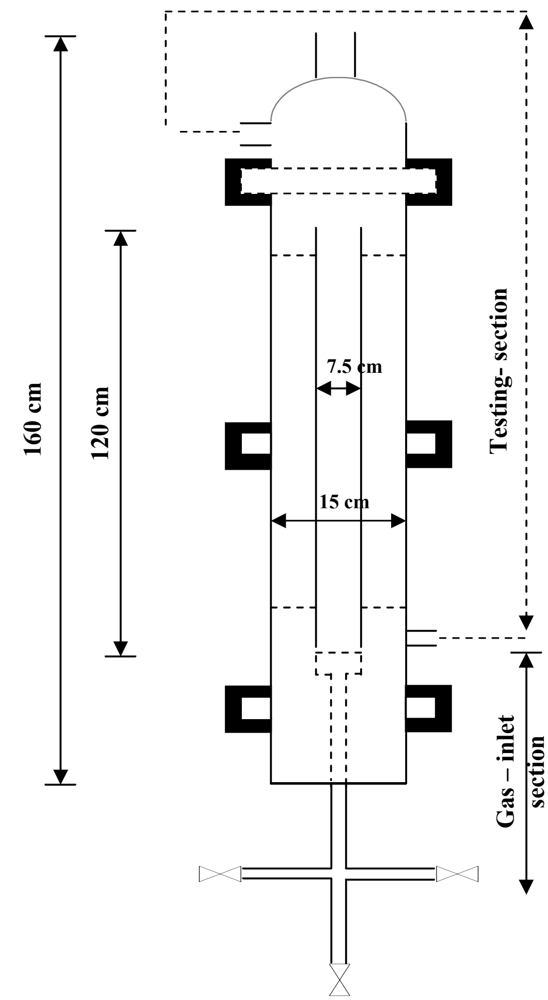


Fig. 3. Airlift Loop Reactor.

The gas inlet sections consist of a gas distributor. At the bottom of this section, two lines are connected together before entering the distributor section each line has a value to be opened or closed as required. One of these lines is the air inlet flow.

Air compressor supplied the line with the desired amount of air needed, for the experiment, the amount of air was measured using a gas meter and two calibrated rotameters connected in parallel were used to measure the air flow rate.

The other line is the nitrogen gas inlet flow. The nitrogen was supplied from a cylinder. A gate value was used in the nitrogen flow which must be shut off when air was sparged to the column, and must be opened during desorption process.

Nitrogen sparging was stopped when the dissolved oxygen meter measures (0.4 ppm O_2). The liquid testing section contains two openings, one for liquid out-flow and the other for liquid in flow.

The circulation of liquid in the column was achieved using a centrifugal pump placed in the recycling line. A ball value placed in the middle of the recycling line was used to take various samples at various times to measure the concentration of the dissolved oxygen during the operation.

The water is fed to the top of the reactor and discharged from the bottom of the reactor using a centrifugal pump.

Compressed air at (100-150) psig was supplied using a reciprocating compressor. The desired air flow rate was set up using needle value and the amount measured with a gas meter.

The dissolved oxygen concentration in the liquid phase was measured using oxygen meter, which consist of a metal electrode.

The liquid phase (batch) consist of the following systems (only water, water-isopropanol, water-glycerol, water CMC (carboxy methyl cellulose).

The gas distributor fig (2) was constructed from a ceramic material and the type is multinole tuyere the distributor has equivalent pore diameter of (0.15, 0.1) mm and free section of (0.5,0. 6%).

Liquid properties

The investigations were carried out with aqueous solutions of different coalescence behavior and viscosities. Physical properties of employed liquids are listed in table (1).

Table1,
Liquid Phase Properties at 25°C

Liquid phase	Density $\times 10^3$ (kg/m³)	Surface tension $\times 10^{-3}$ (N/m)	Viscosity $\times 10^{-3}$ (Pas sec)
Water	0.998	72.7	1.00
Isoproponal	0.997	70.0	0.97
Glecyrol	1.163	67.7	14.00
CMC	0.998	71.6	K=46×10⁻³ Pas.sⁿ(n=0.6)

A queans solution of isopropanal represents strongly coalescence inhibiting system , glycerol solution shows Newtonian behavior and viscosity 14 times that of water while the solution of CMC (carboxy methyl cellulose) shows non-Newtonian pseudo plastic behavior , which can be described by the power law of ostwald and deweale :

$$\tau = k\gamma^n$$

The ostwald factor k and the flow behavior index one specified in table (1).

Experimental methods

The experiments were per formed at room temperature (25°C) and the superficial gas velocities are based on the cross sectional area of the reactor at normal condition (1 bar, 30°C). The total gas hold up (ε) was calculated from equation (4) using the difference between the night of aerated and non aerated liquid phase.

$$\varepsilon_g = \frac{H_f - H_L}{H_F - (V_i / S_o)} \qquad \ldots(4)$$

(V_i / S_o) In Eq (4) is a correction term for the volume of the draft tube.

The physical absorption of oxygen in the air by the liquid was employed to be determining the mass transfer coefficient, a material balance of oxygen in the liquid gives:

$$K_{La} = \frac{-2.303(1-\varepsilon_g)}{t}\log\left(\frac{C_{sa} - C_i}{C_{sa} - C_o}\right) \qquad \ldots(5)$$

Rearranging equation (5) gives

$$\log\frac{C_{sa} - C_i}{C_{sa} - C_o} = \frac{K_{La}}{2.303(1-\varepsilon_g)}.t \qquad \ldots(6)$$

Plotting the left hand side of equation (6) with time the average slope of the plot will give the term $\left(\frac{K_{La}}{2.303(1-\varepsilon_g)}\right)$. The values of ($\varepsilon_g$) was determined as mentioned in equation (4) then (K_{La}) can be calculated[15].

Visual monitoring of acid base reaction between HCl and NaOH was chosen to measurement mixing time. Phenolphthalein was used as the indicator of the above reaction. The mixing time was taken as the time necessary to obtain complete color change from red to colorless; this technique is reliable[16].

4. Results

Mass Transfer Coefficient Results

The volumetric mass transfer coefficient (K_{La}) is a function of gas hold up and mean bubble size.

It plots in Figs (4) and (5) show a dependence on gas throughputs. But the influence of draft tube geometry is more pronounced since the bubble size distribution in the annular space is more affected by the down flow velocity of the liquid phase.

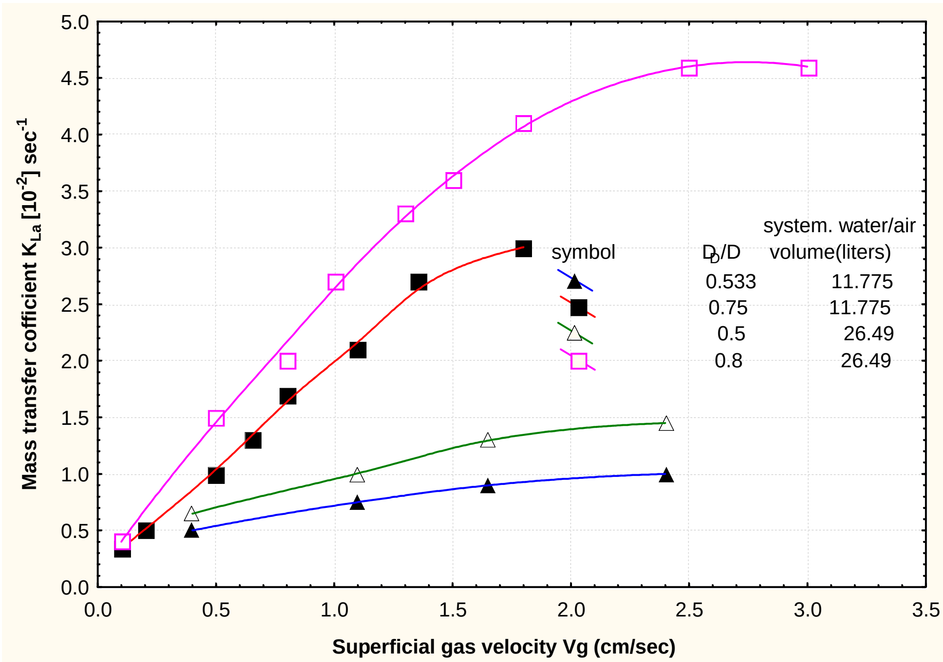


Fig.4. Effect of Gas Velocity and Draft Tube Geometry on the Volumetric Mass Transfer Coefficient in Airlift Loop Reactors of Different Sizes.

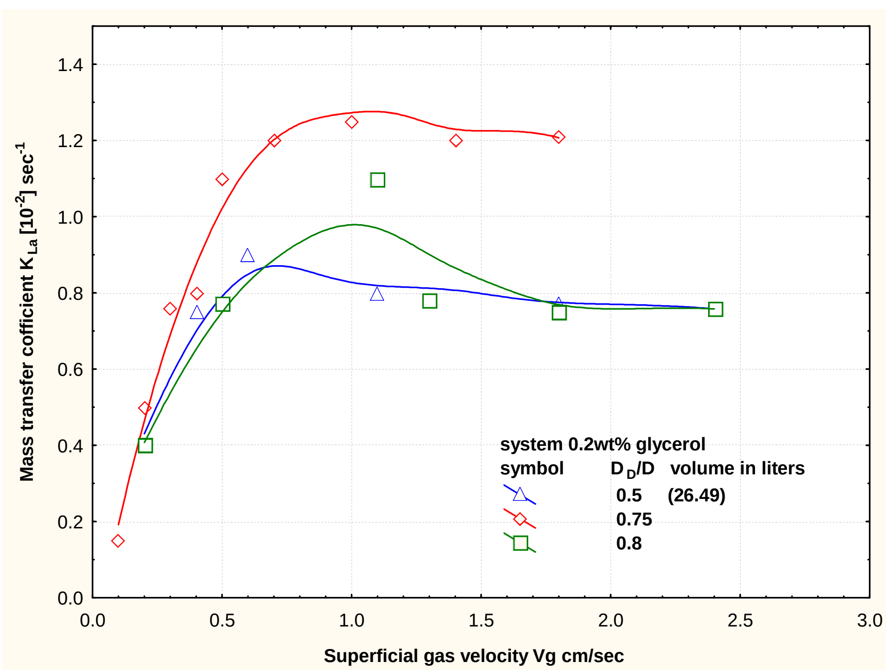


Fig.5. Influence of Draft Tube Diameter on Mass Transfer In Viscous Liquid Phase (Aqueous Glycerol).

Figure (4) shows the influence of superficial gas velocity and of draft tube to reactor diameter ratio on (K_{La}) for the (11.775, 26.44) liter air lift loop reactors. In both reactors the(K_{La}) value for water increase by about 30%, it the diameter ratio is increased from 0.5 to 0.8.

A similar dependence of (K_{La}) on gas throughput and draft tube geometry was observed for the aqueous solution of iso-propanol. One account of the strong coalescence inhibition, volumetric mass transfer coefficient reaches double the values in pure water. Opposite behavior was observed in the case of aqueous glycerol (Fig.5).

The mass transfer rate reaches its maximum value at a superficial gas velocity of about 1 cm/sec since, owing to the strong coalescence promoting properties of highly viscous liquids.

Large bubbles are formed already at very small gas throughputs. The largest volumetric mass transfer coefficient for this liquid system is obtained at a diameter ratio of 0.75 which also yields the maximum gas hold up. At this draft tube to reactor diameter ratio the gas phase distribution in the reactor is nearly uniform whereas at ratio of 0.5 the annular space is almost free of gas and large bubbles form in the draft tube.

At diameter ratio of 0.8, the two- phase flow is less ordered, in analogy to the operation without a draft tube. Thus mass transfer rate decreases at diameter ratio in excess of 0.75 .

Mixing Time Results

The mixing processes in loop reactor consist of combined effects occurring in the draft tube, annular space and in the top and bottom deflection zones. Mixing in the up and down flow zones is produced by axial dispersion which mainly results from the difference between velocities of gas and liquid phases.

The axial mixing fraction of the overall mixing process increase with the start of gas circulation since bubbles which coalesce in the annular space rise against the liquid flow and therefore considerately speed up the mixing process. The intensive mixing in the deflection zones is caused by differences between velocities in the up and down flow zones. In the top zone, mixing is intensified by the formation of a ring-vortex above the draft tube. Fig. (6) shows for the (11.775, 26.44) liter air lift loop reactors, the influence of superficial gas velocity and draft tube geometry on mixing time. Two regimes of mixing could be observed in all the investigated devices up to superficial gas velocity of about 2 cm/sec, mixing time decrease rapidly, while at higher gas velocities liquid mixing becomes less efficient. The shortest mixing time is achieved at the largest investigated ratio of draft tube to reactor diameters. The increase of the diameter ratio from 0.5 to 0.8 leads to shortening of the mixing time by about 30%.

As shown in fig. (7), the total mixing time depends to a large extent on reactor size. It increases with increasing reactor size. In contrast to the total mixing time, the effective mixing time, defined as the mixing time per unit reactor volume decreases with increasing reactor volume see fig. (8).

The influence of the draft tube diameter on effective mixing time decreases with reactor volume. Mixing times in bubble columns increase with increasing coalescence inhibiting properties of the aerated liquid. The more uniform cross sectional distribution of the gas bubbles in such liquids prevents the formation of local recirculation vortices.

Fig. (7) shows the different mixing behavior of bubble columns and loop reactors for the strong coalescence inhibited aqueous solution of iso-propanol. Above a superficial gas velocity of about 1.4cm/sec, bubble columns show the poorest mixing behavior. In contrast to the operation without draft tube, the influence of the liquid phase properties on mixing time in airlift loop reactors is insignificant as can be seen from fig. (9).

The mixing time is largely affected by the ring vortex. Which forms above the draft tube at the top of the reactor due to the liquid flow deflection. Therefore mixing time is strongly dependent on the liquid head above the draft tube.

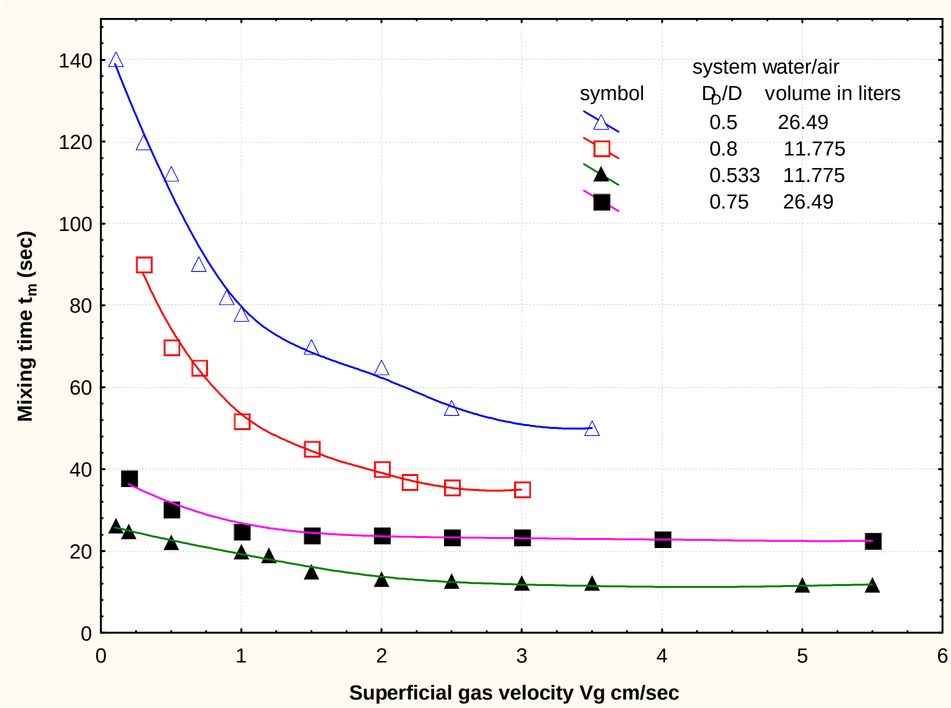


Fig.6. Effect of Gas Through Put and Draft Tube Geometry on Mixing Time in Airlift Loop Reactors of Different Sizes.

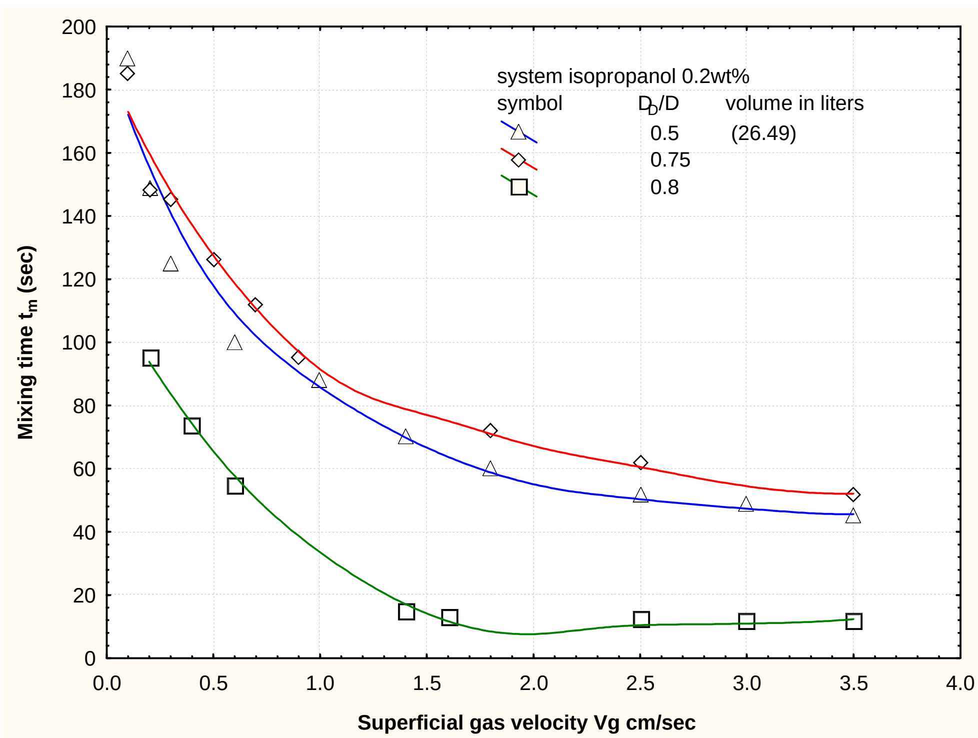


Fig.7. Mixing Time in Aqueous Solution of Isopropanol.

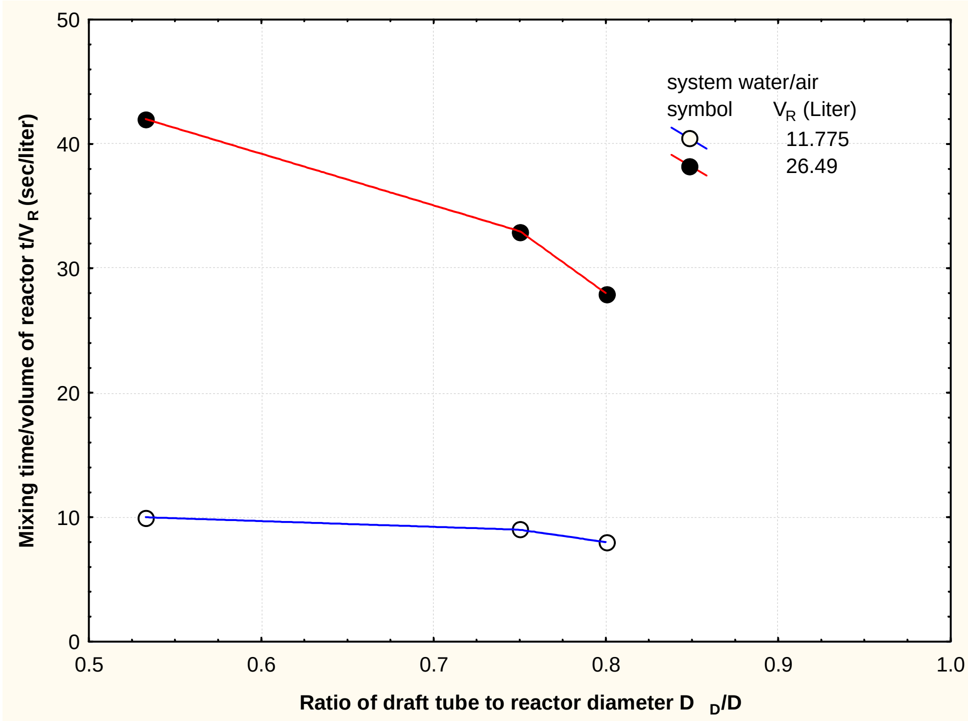


Fig.8. Mixing Time Per Unit Volume as a Function of Draft Tube to Reactor Diameter Ratio.

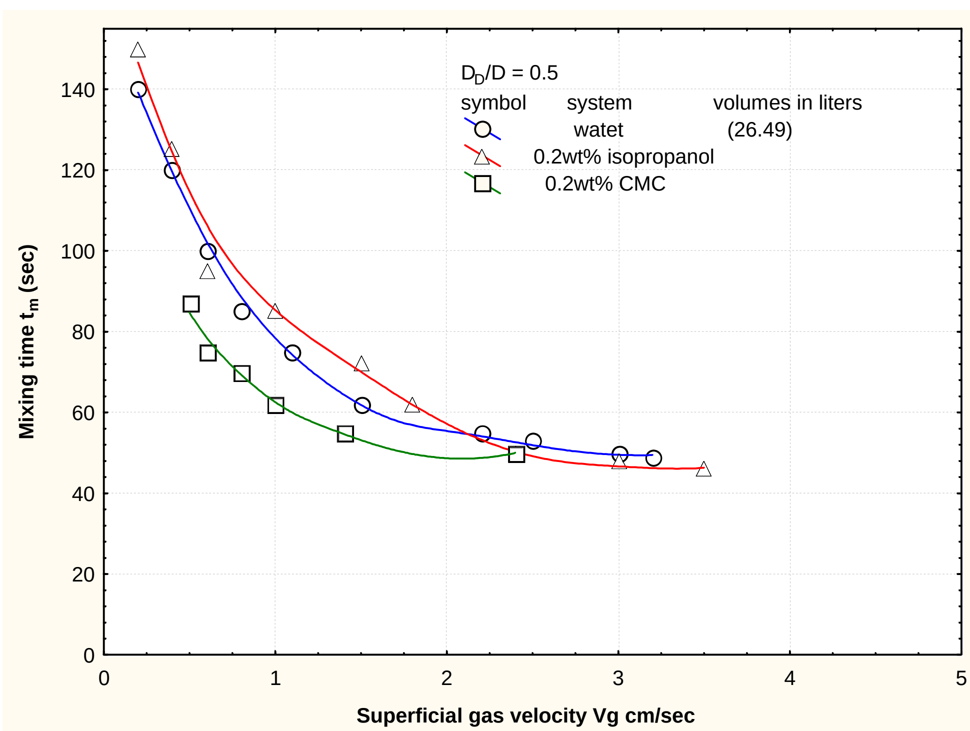


Fig.9. Mixing Times in Different Liquids as a Functions of Super Ficial gas Velocity.

5. Conclusions

The result of present experiments show that optimal values of various important process parameters (K_{la},t_m) do not correspond to the same diameter ratio. Therefore, the draft tube diameter must be selected with reference to the parameter which is most important for the process. Therefore optimum design of the draft tube is only possible if the properties of the liquid phase are known.

Nomenclature

C_i	concentration of dissolved oxygen at any time (ppm).
C_o	initial concentration of dissolved oxygen (ppm).
C_{sa}	saturated concentration of dissolved oxygen (ppm).
D	reactor diameter (cm).
D_o	Draft tube diameter (cm).
d_o	Hole diameter of gas sparger (cm).
D_L	Liquid phase axial dispersion coefficient (cm^3/sec).
F_i	Friction factor
g	Acceleration due to gravity (cm/sec^3).
H_f	Level of aerated liquid during operation (cm).
H_L	Clear-liquid height (cm).
K_{La}	over all mass transfer coefficient (sec^{-1}).
K	Fluid consistency index, $pa.s^n$
ΔP_A	Pressure loss in annulus ($kg/m.s^2$)
ΔP_D	Pressure loss in draft tube ($kg/m.s^2$)
ΔP_H	Hydrostatic Pressure difference ($kg/m.s^2$)
ΔP_R	Pressure loss due to flow deflection ($kg/m.s^2$)
t	time (sec).
V_G or V_g	Superficial gas velocity (cm/sec).
V_L	Effective liquid velocity (cm/sec).
V_R	Liquid volume of reactor (liter).

Greek letters

G	Holdup (Fractional volume).
δ	shear rate, s^{-1}
ρ_L	liquid phase density kg/m^3
ε	gas hold-up
ε_D	gas hold-up in draft tube
ε_A	gas hold-up in the annulus
μ	Dynamic viscosity (pa.s).
γ_i	Kinematic viscosity (cm^2/sec).
τ	Shear stress, (pa).

Subscripts

A	Annular space.
D	draft tube.
G	gas phase.
L	liquid phase.

6. References

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تأثير قطر الأنبوب الداخلي على سلوك تشغيل مفاعلات تدوير الهواء

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الخلاصة

ان النسبة بين القطر الداخلي الى القطر الخارجي ذو اهمية كبيرة جدا في السلوك التشغيلي الديناميكي لمفاعل التدوير الهوائي. ان تاثير الابعاد الهندسية للعمود الداخلي (draft) قد تم دراسته نسبة الى العوامل التشغيلية والتصميمية التالية: انتقال الكتلة للاوكسجين, زمن الخلط وبنسبة قطرية (D_D/D) تتراوح بين 0.45 الى 0.59 حجم السائل المستخدم فى المفاعلين هو 11.775 الى 26.49 لترتم استخدام محاليل ذات قابلية مختلفة لتجميع الفقاعات. اظهرت النتائج انه لاتوجد نسبة قطرية محدودة ثابتة هي المفضلة. لجميع العوامل التصميمية بقدر ماتوجد نسبة معينة هي المفضلة لعامل واحد او عاملين دون العوامل الاخرى فمثلا اظهرت النتائج ان النسبة القطرية بين ٠.٥ الى 0.6 هي الاكثر ملائمة لعاملي انتقال الكتلة وزمن الخلط اما فى حالة الحصول على سرعة عالية للسائل في الانبوب الداخلي (draft) فكانت النسبة القطرية (0.6) هي الامثل.