

Gas Hold-Up and Volumetric Liquid-Phase Mass Transfer Coefficient In Solid-Suspended Bubble Columns With Draught Tube

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Abstract

The solid suspended-bubble columns with draught tube are widely used as a three-phase slurry reactor in industrial chemical process such as absorption, biochemical reactions, coal liquefaction. etc. To design such a column the average gas holdup (e_g) and the volumetric liquid-phase mass transfer coefficient (K_{La}) should be known. This study includes the effect of gas velocity, liquid-phase properties, solid-particles concentration and the static liquid height on both (e_g) and (K_{La}). It was found that both (e_g) and (K_{La}) increase with increasing gas velocity and decrease with increasing solid particles concentration, static liquid height, viscosity and surface tension of liquid-phase.

زمن احتجاز الغاز ومعامل انتقال الكتلة الحجمي السائل بوجود الصلب العالق في العمود الفقاعي ذو الأنبوب الداخلي

الخلاصة

لقد شاع استخدام الاعمدة الفقاعية ذو الانبوب الداخلي وذات العالق الصلب كمفاعلات ثلاثية الطور في اغلب العمليات الكيمياءوية الصناعية مثال عاي ذلك عمليات الامتصاص والتفاعلات البايوكيمياءوية واسالة الفحم وغيرها. لتصميم هذه الاعمدة يجب معرفة عاملين اساسيين هما معامل انتقال الكتلة (K_{La}) وزمن احتجاز الغاز (e_g). الدراسة تتضمن بحث تأثير كل من سرعة الغاز, خواص الطور السائل, تركيز المادة الصلبة, اضافة الى تأثير ارتفاع عمود السائل فوق الموزع على كلا من (K_{La}) و (e_g) وجد من خلال البحث انه بزيادة سرعة الغاز تزداد قيم كلا من (K_{La}) و (e_g) اما في حالة زيادة تركيز المادة الصلبة, ارتفاع عمود السائل, لزوجة السائل, معامل الشد السطحي للسائل فيؤدي الى نقصان قيم كلا (K_{La}) و (e_g).

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⁽⁴⁾. Slurry columns are similar to fluidized

bed columns, in that a gas is passed through a column containing solid catalyst particles suspended in a fluid. In slurries the catalyst is suspended in a fluid, in fluidized beds the suspending fluid is the reacting gas itself⁽¹⁾. In (sc) an attempt is made to realize intensive and intimate contact between a gas –

phase component and a finely dispersed solid. With respect to this purpose (sc) are related to packed bed columns with the different gas-liquid flow regimes that can be realized (such as trickle flow, pulsed flow, dispersed bubble flow etc....). Also there is a lot of similarity with three-phase fluid bed systems. The latter systems share many properties with (sc), but the main difference is the fact that in fluid beds with upward fluid flows the drag force acting on the solids by the gas and liquid flow is on the average balanced by the net weight of the particles, while in (sc) the overall liquid-solid slip velocities are practically zero and particles remain suspended by the action of the turbulence, in the liquid phase^(2,3).

Deckwer and Schumpe⁽⁵⁾ studied the effect of various design parameters. They found that mixing time decreases initially with increasing gas velocity and then increases in the higher range of velocity. Guerin et. al⁽⁶⁾ studied the effect of gas flow rate on mixing time. They found that the mixing time does not decrease proportionally with the increase of gas flow rate, and this is a main difference between BC and mechanically stirred columns. Solid-mixing and solid replacement are important factors in cases where the solids have a short life-time⁽⁷⁾. Fan and Chern⁽⁸⁾ studied solid-mixing in a gas-liquid-solid system. They reported three-states of solid mixing, these states are complete segregation, partial intermixing and complete intermixing.

Particle size in (sc) can be small to very small, even down to the submicron range. The effect of average

particle size (d_p) and solid density on the critical superficial gas velocity (V_{gc}) is given as⁽⁹⁾:-

$$V_{gc} a \cdot d_p^{0.47}, V_{gc} a \left[\frac{r_s - r_l}{r_L} \right]^{0.95}$$

The pressure drop in (sc) is usually more or less independent of the gas flow and close to the hydrostatic pressure. Of course there is also a pressure drop required for the gas distributor.

Slurry columns can be classified according to the phases where the reactant are present. Generally (sc) can be classified according to⁽¹⁰⁾:-

- a. The chemical system.
- b. The contacting pattern and mechanical hold-up-fractions devices.

Hydrodynamics of (sc) includes the study of mixing and the volumetric liquid side mass transfer coefficient at the gas-liquid interface (K_{La}). For the design of (sc), whether agitated only by the flowing gas or assisted by one or more stirrers, the conditions at which the particles are just suspended are very important.

Therefore, generally only a minimum suspension criteria (M.S.C) is considered⁽²⁾. Roy et. al⁽¹¹⁾ studied by applying the pressure drop technique a large variety of gas solid and liquid systems including non-aqueous systems and particles with different degree of wetting. Kato et. al⁽¹²⁾ studied the (M.S.C) for different columns heights. Koide found that the effect of solid particles on reducing e_g value in the transition regims is larger than that in heterogeneous regime. (Fuku et. al⁽¹³⁾ studied the (M.S.C) for a column with a draft tube, they

showed that with a draft-tube applied in the system, a much smaller volumetric flow rate of gas was required to suspend the same amount of solid, compared with normal column. Narayanan et.al⁽¹⁴⁾ studied the (M.S.C) in sparged vessels with a stagnant liquid medium, and gave a relation for minimum gas velocity to suspend the particles as in this equation:-

$$V_g(\text{min,actual}) = 1.25 \frac{D_c}{0.0508} \cdot \exp(-3W_s) \times$$

$$V_g(\text{min,theo}) \text{ for } W_s > 0.1$$

$$\text{When } n = 0.2 \dots \text{for } d_p < 100 \text{mm}$$

$$n = 0.5 \dots \text{for } d_p > 200 \text{mm}$$

The average hold-up fractions of gas, solids and liquid should satisfy the equation: $e_g + e_s + e_L = 1$

In contrast to three-phase fluid-beds where the relation between the three-phase hold ups can be rather complex, in slurry columns with the much smaller particles and slip velocities, the relation between (ϵ_l) and (ϵ_s) is often simple as it is fixed by the feed ratio of solids liquid phases, or liquid and solids volumes are constant (in batch systems).

The bubble hold-up is much more difficult to predict, first of all because of the different regimes that might prevail both in stirred vessels and slurry sparger columns. Shah et. al⁽¹⁵⁾ studied the regimes in three-phase flow sparger. They showed that there are three regimes. The regimes are:-
1- Uniform bubbling at low gas velocity.

2- Churn-turbulent-flow at higher-gas velocities, with a mixture of large and small bubbles.

3- Slugging in small diameter columns, where the largest bubbles are comparable to the column diameter. Koide et. al⁽¹⁶⁾ studied the three phase flow with different solid concentrations, They showed that the effect of solid particles on reducing e_g value in the transition regimes is larger than that in heterogenous regime. Akita et. al⁽¹⁷⁾ studied the gas hold up in BC and proposed a correlation for gas hold up in terms of bond and Galilo numbers.

$$\frac{e_g}{(1-e_g)^4} = K \cdot B_o \cdot G_a \cdot F_r$$

Miller et. al⁽¹⁸⁾ studied the gas hold up in three phase system and proposed for churn turbulent regime in three phase fluidization to use the Akitas equation using liquid/solid suspended bulk properties instead of liquid properties.

$$e_g = \frac{1}{8} \ln \left[1 + 8 C_g^{7/24} \left[\frac{m_B}{r_B} \right]^{-1} \left[\frac{s}{r_B} \right]^{-1} J_g \right]$$

Koide and Horibe⁽¹⁹⁾ studied the effect of using solids in draught tube in solid suspended bubble column. For an adequate description of mass transfer with chemical reaction in slurry columns, reliable data on the following two types

- a- parameters which are specific for slurry columns (K_G , K_L , K_{La} , K_{Ga}).
- b- parameters which are not specific for the type of reactor applied (intrinsic reaction kinetics).⁽²⁾

Akita et. al⁽¹⁷⁾ showed that a better empirical equation of ($K_{L,a}$) can be obtained by using (ϵ_g) instead of (V_g) in the BC:

Koide and Horibe⁽¹⁹⁾ studied the

$$\frac{K_{L,a} \cdot s_L}{r_L \cdot D_i \cdot g} = 2.11 \left[\frac{m_L}{r_L \cdot D_i} \right]^{0.5} \left[\frac{g \cdot m_L^4}{r_L \cdot s_L^3} \right] \dots \times e_g^{1.18}$$

bubble column with draught tube. Koide and Iwamoto⁽²⁰⁾ studied BC with draught tube and with gas dispersion into annulas, the values of ($K_{L,a}$) and (ϵ_g) is much larger than those in the bubble column without draught tube, when a liquid with forthing ability is used.

2-Purpose:-

The purpose of this study is to clarify experimentally the effects of, gas velocity and properties of the liquid and solid particles on ϵ_g and $K_{L,a}$ in the solid suspended bubble column with draught tube in liquid solid batch operation.

3- Experimental Section

A schematic diagram of the experimental set-up used in this work is shown in Figures (1, 1.1, 1.3).

Aplexiglass column of 0.15 m inside diameter and about 1.60 m total height with draught tube dimensions of 0.075 m inside diameter and 1.40 m total height was used.

The draught 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 consists of two main sections namely, the gas inlet section and the liquid recycling testing section. The gas inlet section consists of a gas distributor.

At the bottom of this section, two lines are connected together before entering the distributor section each line has a valve 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. It was supplied from a cylinder.

A gate valve 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.

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 valve 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 column was filled with water to the desired height above the distributor (0.5, 1, 1.4) m. Then the solid particles (plastic-non porous) were added to the liquid in the column. The concentration of solid particles to each level of static liquid were (30, 60, 100) kg/m³ respectively.

Compressed air at (100-150) psig was supplied using a reciprocating compressor.

The desired air flow rate was set-up using needle valve and the amount measured with a gas meter.

The dissolved oxygen concentration in the liquid phase was measured using oxygen meter, which consists of a gold metal electrode. The liquid phase (batch) consists of the following systems (only water; water and solids; water, alcohols and solids; water, NaCl and solid; water, acetic – acid and solids).

The gas distributor Fig (1.2) was constructed from a ceramic material and the type is multi hole tuyere.

The distributor has equivalent pore diameter of 0.15 mm and free section of 0.61%.

4-Results on gas hold up (ϵ_g) and solid hold up(ϵ_s)

The average gas hold up ϵ_g was calculated from equation (1) using the data of the static slurry height(H_f) and the height of the aerated slurry(H_l) which were determined by visual observation

$$e_g = \frac{H_f - H_l}{H_f} \quad \dots (1)$$

The solid hold up was calculated from equation(2) using the data of static liquid height(H) and the height of slurry after adding solid particles \bar{H}_F

$$e_s = \frac{\bar{H}_F - H_L}{\bar{H}_F} \quad \dots (2)$$

The experimental gas hold up was found by measuring the difference between initial liquid height and final

liquid height. Since it was rather difficult to read directly the level of the aerated liquid the values of gas hold up thus obtained probably involves an error of about 5%, established via repeated measurements. Fig(2) shows the effect of gas velocity for water system with and without solid particles. The gas hold up was found to increase with increasing gas velocity, because the large bubble holdup increases with increasing gas velocity leading to the increase of the overall holdup. But the effect without solid particles is larger than that with solid particles.

When the gas velocities above 0.01m/sec and the liquid phase without solid particles, bubble size is independent of gas flow rate. The buoyant force of the bubble is opposed by viscous drag of the fluid, and if bubble size is independent of gas flow rate, the velocity of bubble rise will show similar independence.

Hence increased gas flow rate will increase the gas hold up, and ultimately the liquid will seen to be filled with bubbles. Fig(3) show, in slurries the presence of solid particles in the liquid might enhance bubble coalescence into larger bubbles and therefore reduces the value of gas hold up.

The solid particles retard the bubble rise velocity and prevent increase in bubble size.

Fig(4) show, the effect of the static liquid height on gas hold up. It is clear that increasing the static liquid height will decrease the gas hold up. As the static liquid height is increased, the bubbles have time to coal further and ultimately reduce gas hold up.

In our experimental studies the maximum static liquid height used was 1.4 m. In slurries the the presence of solid particles in the liquid might enhance bubble coalescence into larger bubbles and therefore reduce gas hold up .

Fig(5) shows the effect of using different liquid phase (alcohols and electrolytes) respectively, on gas hold up. Low electrolyte concentrations have no noticeable effect on the surface tension of the solution. However the ionic forces in the liquid bulk reduce the bubble rise velocity and the bubble coalescence. As a result, the gas hold up increase.

For high electrolyte concentration, the interfacial tension increases, resulting in increased bubble size and reduce gas-holdup.

5-Mass transfer coefficient results.

The physical absorption of oxygen in the air by the liquid was employed to determine the mass transfer coefficient. A material balance of oxygen in the liquid gives⁽²³⁾ :-

$$K_{La} = \frac{-2.303(1 - \epsilon_g - \epsilon_s)}{t} \cdot \text{Log} \frac{C_{Sa} - C_i}{C_{Sa} - C_o} \dots\dots(3)$$

Rearranging equation(3) gives

$$\text{Log} \frac{C_{Sa} - C_i}{C_{Sa} - C_o} = \frac{K_{La}}{2.303(1 - \epsilon_g - \epsilon_s)} \cdot t \dots(4)$$

Plotting the left hand side of equation (4) with(t), the average slope of the plot will give the term

$$\frac{K_{La}}{2.303(1 - \epsilon_g - \epsilon_s)}$$

The values of (ϵ_g) and (ϵ_s) were determine as mentioned in (1) and (2) respectively, then (k_{La})

can be calculated .Fig (6) shows that the mass transfer coefficient increase with increasing gas velocity. The axial dispersion coefficient (D) increases with increasing gas velocity and therefore increase (k_{La}).

Fig(7) shows the effect of solid particle concentration on (K_{La}). The presence of solid particle in the liquid will decrease the axial dispersion coefficient and it enhance bubble coalescence. The bubble size will be larger and occupying larger space in the column and therefore reduces (K_{La}). At higher gas velocities (0.1-0.2) m/sec, the effect of solid particles on (K_{La}) will be less than in low gas velocities (0.03m/sec). Fig(8) show the effect of static liquid height on mass transfer coefficient.

As the static liquid height is increased, however the bubbles have time to coalesce further and ultimately decreases the axial dispersion coefficient and the mass transfer coefficient. Fig(9) shows the effect of liquid phase properties on (K_{La}). As mentioned before, the ionic forces in the liquid bulk reduce the bubble rise velocity and the bubbles coalescence, so that the mass transfer coefficient is increased.

6-Conclusions

1. The gas hold up and mass transfer coefficient increase with increasing gas velocity, for $V_g \leq 0.2$ m/sec.
- 2-The addition of solid particles reduces both gas hold up and mass transfer coefficient.
3. Increasing the static liquid height reduces both gas hold up and mass transfer coefficient.

4. Gas hold up and mass transfer coefficient decreases with increasing the liquid phase viscosity.

Nomenclature

C_i = Concentration of dissolved oxygen at any

time p.p.m

C_0 = Intial Concentration of dissolved oxygen

p.p.m

C_{sa} = Saturated concentration of dissolved

oxygen p.p.m

C_s = Solid particle concentration K_G/m^3 .

D_c = Column diameter .

D_i = Diffusivity of oxygen in solution m^2/sec

D_L = Axial dispersion coefficient (liquid) m^2/sec

D_p = Average particle size (m)

g = Acceleration of gravity m/sec^2

H_L = Static slurry height (m)

H_F = Level of aerated slurry (m)

\bar{H}_F = Level of liquid phase+ solids (m)

K_L = Liquid phase mass transfer coefficient ($m.s^{-1}$)

K_{La} = Overall mass transfer coefficient, based on aerated slurry volume. (sec^{-1})

Sc = slurry column

t = Time (min)

V_g = Gas velocity (m/sec)

V_{gc} = Critical gas velocity (m/sec)

W_s = Mass of solid /kg liquid

Greek letters

ϵ_g = Gas hold up

ϵ_s = Solid hold up

ρ_L = Liquid phase density kg/m^3

ρ_s = Solid phase density kg/m^3

μ_L = Liquid phase viscosity(C_p)

ν_L = Kinematic viscosity of liquid phase (cm^2/sec)

σ_L = Liquid phase surface tension dyn/cm

Dimension less groups

Bo = Bond number, $(G.D_c^2.\rho_L)/\sigma$

Fr = Froud number, $Vg/(g.D_c^{0.5})$

Ga = Galilo number $Vg.D_c^3.\rho_L/ \mu_L^2$

Subscripts

G =gas

L = liquid

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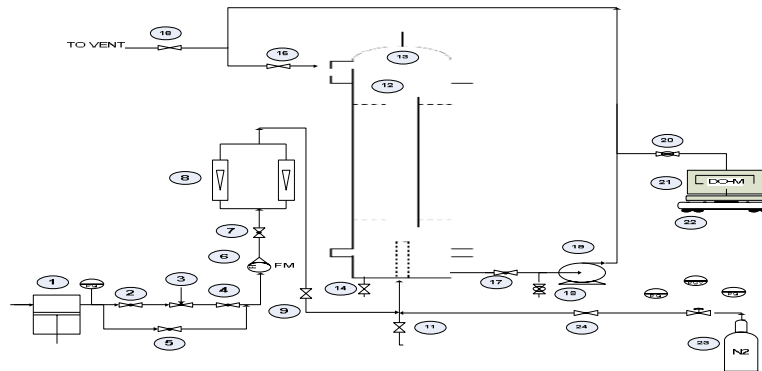
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Table (1) physical properties for pure liquids at T=25°C

Liquid phase	ρ kg/m ³	μ_{CP}	σ dyne/cm	D_i cm ² /sec	V_L cm ² /sec
Water	0.9982	0.897	22.65	0.21×10^{-4}	0.8986
Methanol	0.7920	0.547	22.61	0.16×10^{-3}	0.5660
Ethanol	0.7810	1.003	22.63	0.128×10^{-3}	0.9085s
Acetic acid	1.040	1.005	27.80	0.124×10^{-3}	0.9792
NaCl	1.1153	1.295	78.30	1.161×10^{-5}	1.2300s

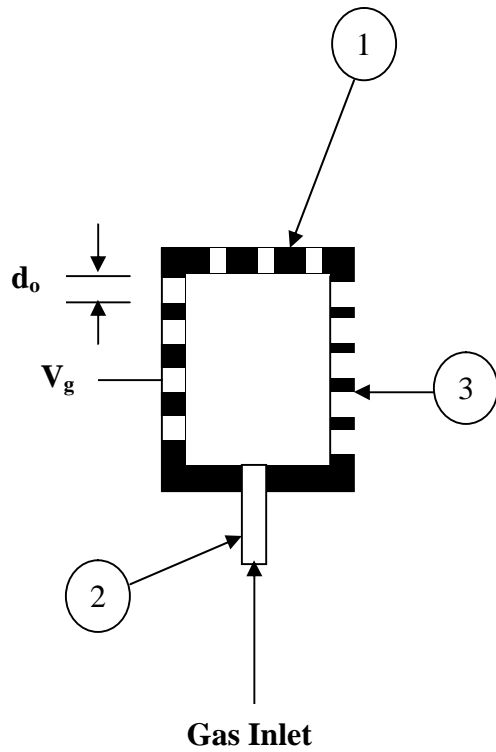
Table (2) Physical properties for mixtures used with concentration 20% at T=25°C

Liquid phase	ρ kg/m ³	μ_{CP}	σ Dyne/cm	$D_i \times 10^5$ cm ² /sec	V_L cm ² /se
Water+methanol	0.9666	0.7952	22.63	16	0.8226
Water+ethanol	0.9686	0.9105	22.64	12.8	0.9400
Water+aceticacid	1.0263	0.91675	25.225	12.4	0.8932
Water+NaCl	1.0216	0.9247	48.375	1.161	0.9051



1	Air compressor	8	Gas rotameter	12	Draught tube	21	Dissolved oxygen meter (Do-m)
2,4,5,7,17	Globe valves	9	Two-way valve	13	Column	22	Move table
3	Needle valve	10	Gas distributor	18, 20	Ball valves	23	Nitrogen cylinder
6	Gas meter	11,14,15,16,24	Gate valves	19	Centrifugal pump	PG, PCV	Pressure control valve pressure Gauge

Figure (1) Experimental-Apparatus



- 1- Ceramic material.
- 2- Pipe.
- 3- Hole.

Schematic diagram

Figure (1.2) Gas-distributor

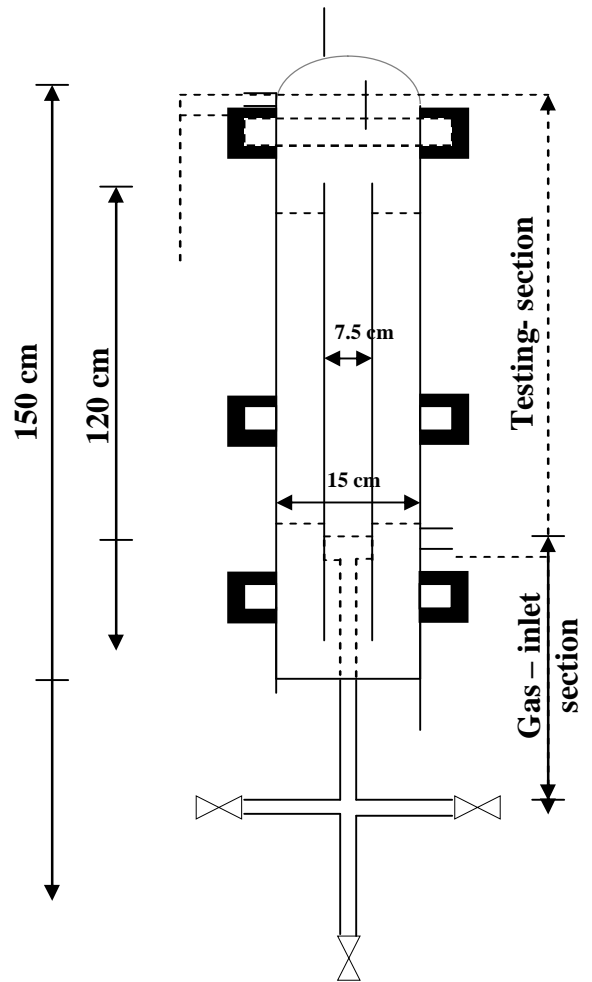


Figure (1.1) Column

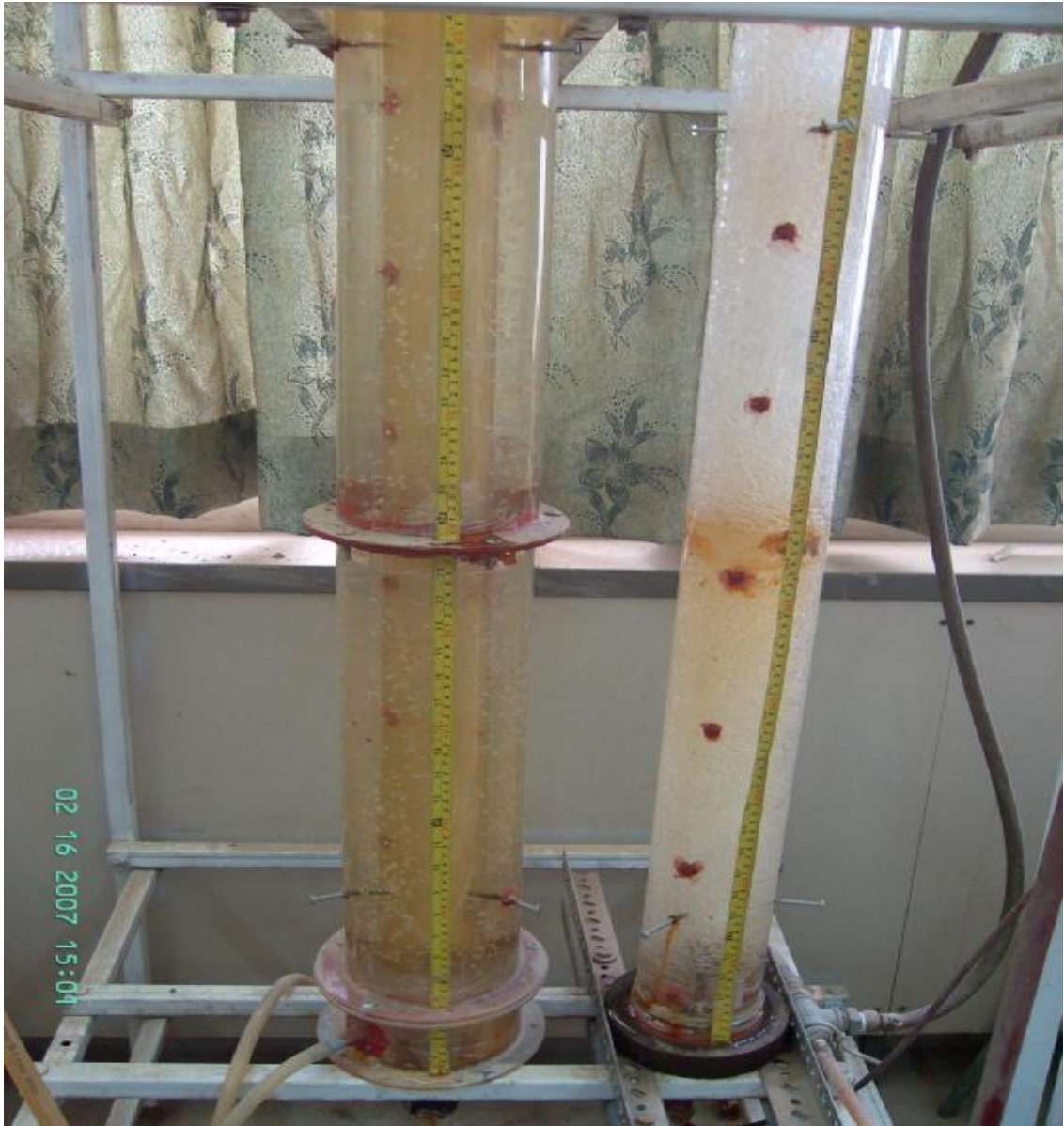


Figure (1.3) Columns

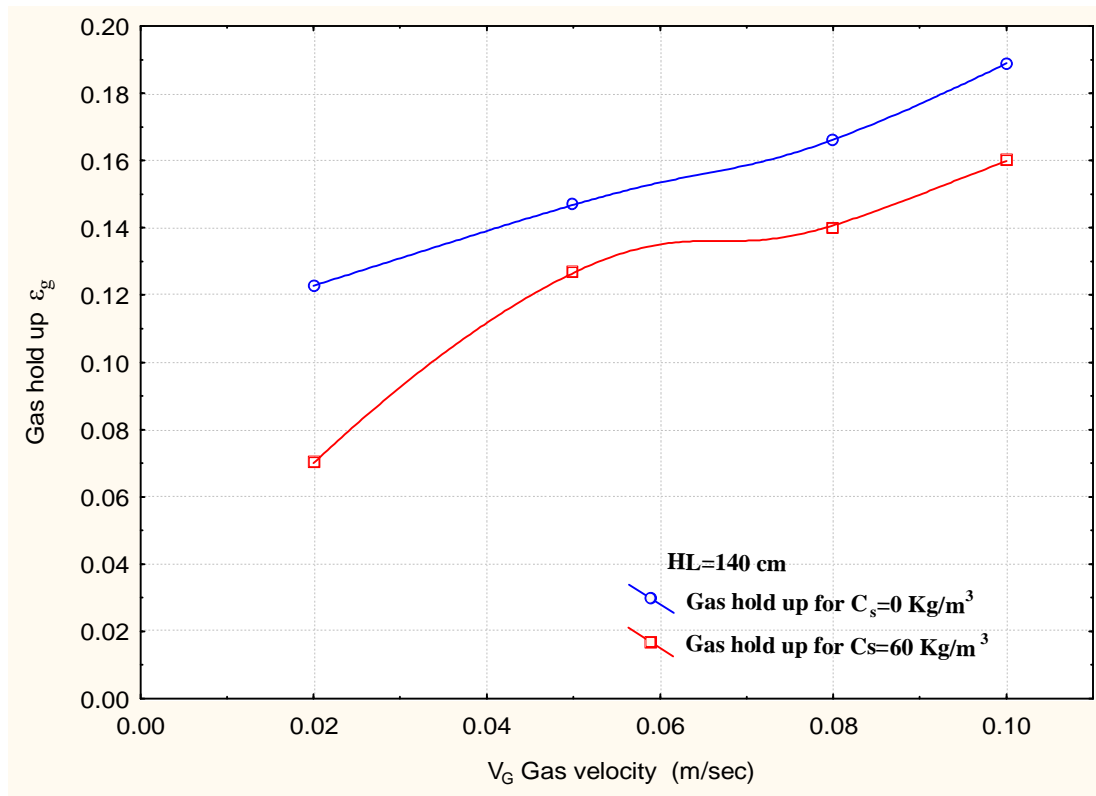


Figure (2) Gas hold up vs gas velocity for water system

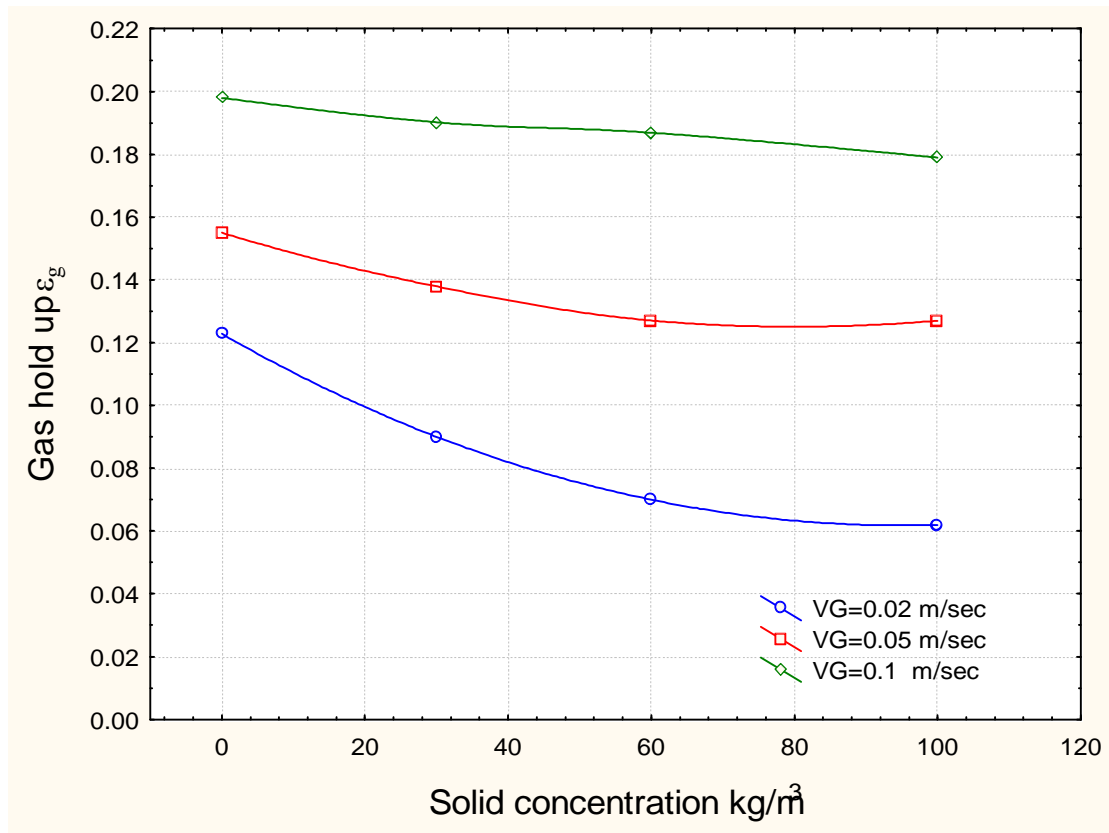


Figure (3) Gas hold up vs solid concentration for water system for various gas velocities

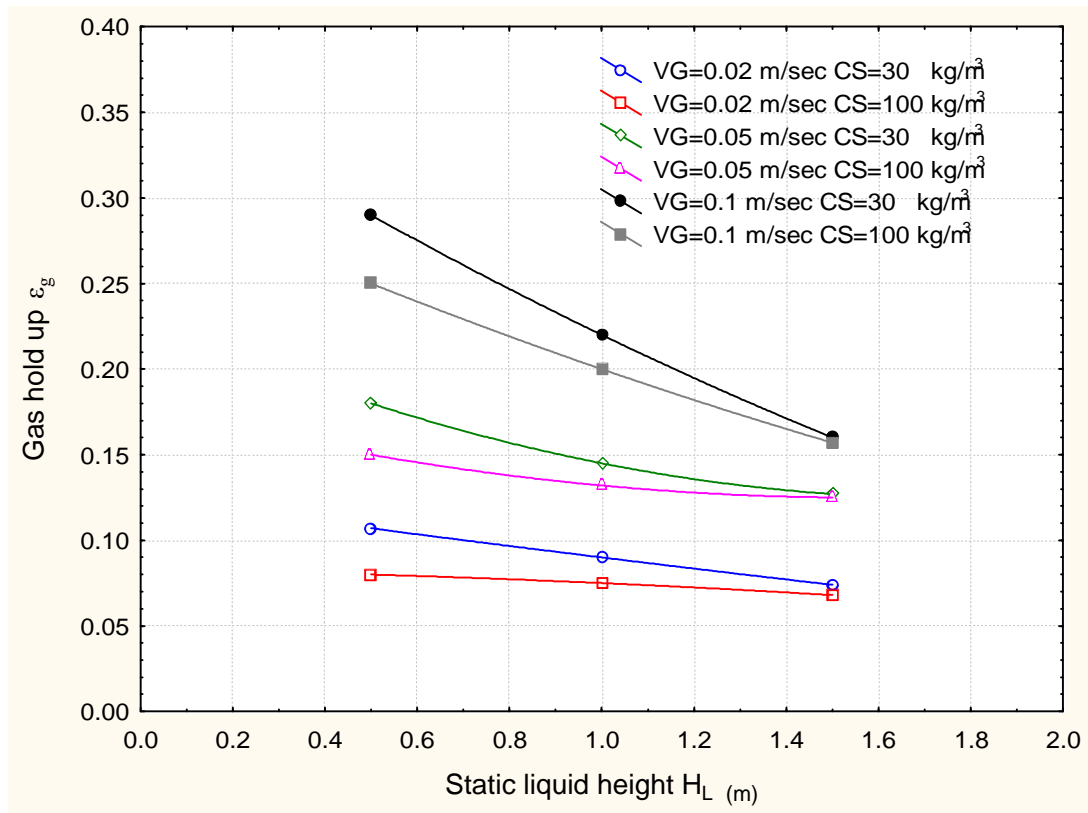


Figure (4) Gas hold up vs static liquid height for various gas velocities

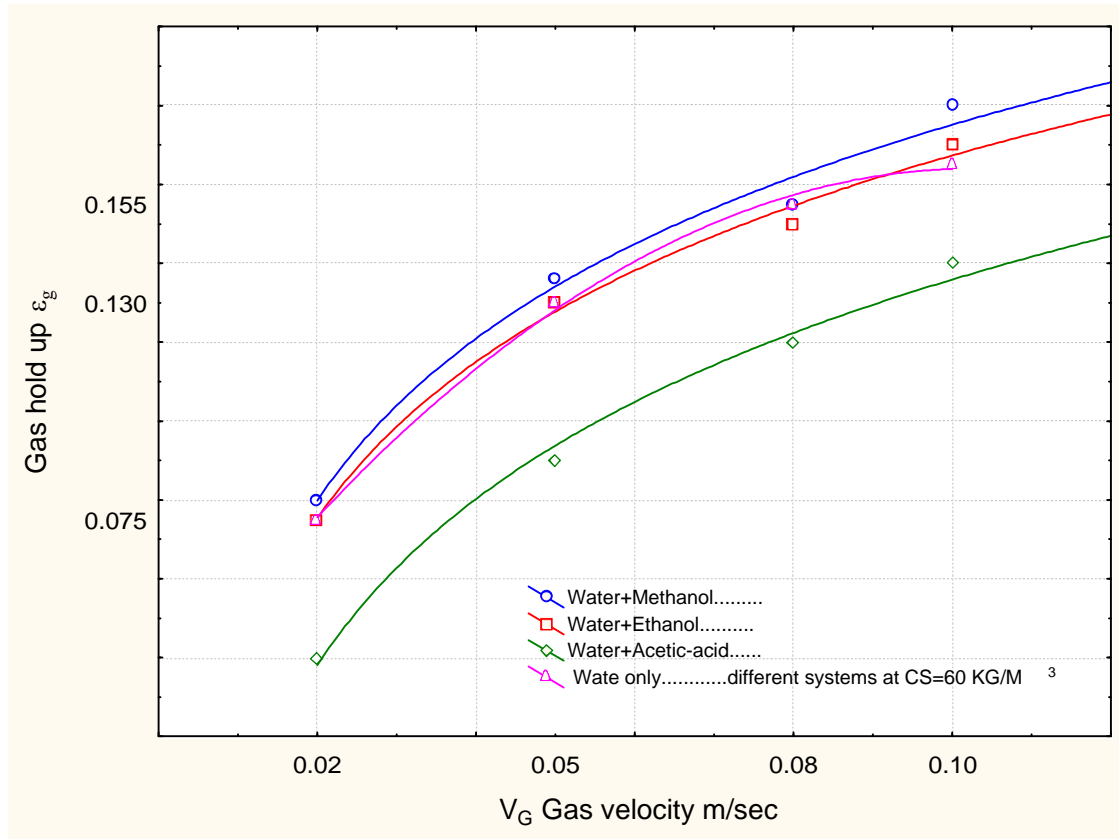


Figure (5) Gas hold up vs gas velocity for different liquid phase systems

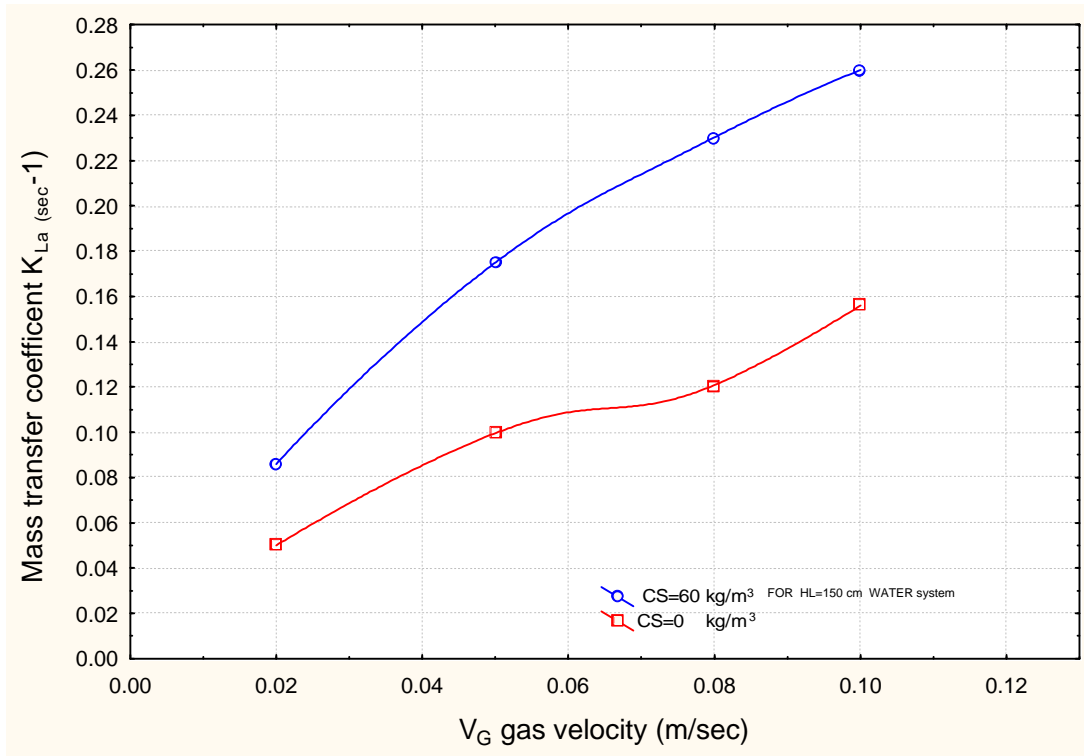


Figure (6) Mass transfer coefficient vs gas velocity for water systems

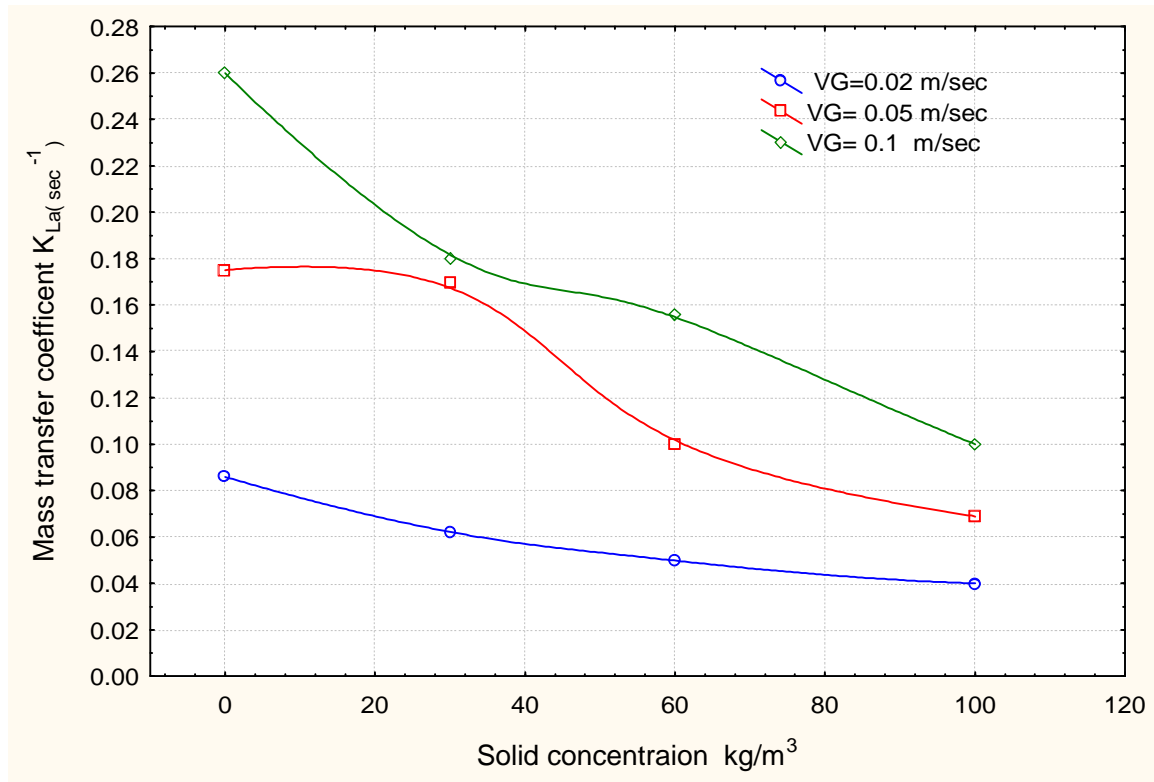


Figure (7) Mass transfer coefficient vs solid concentration for water system for various gas velocities

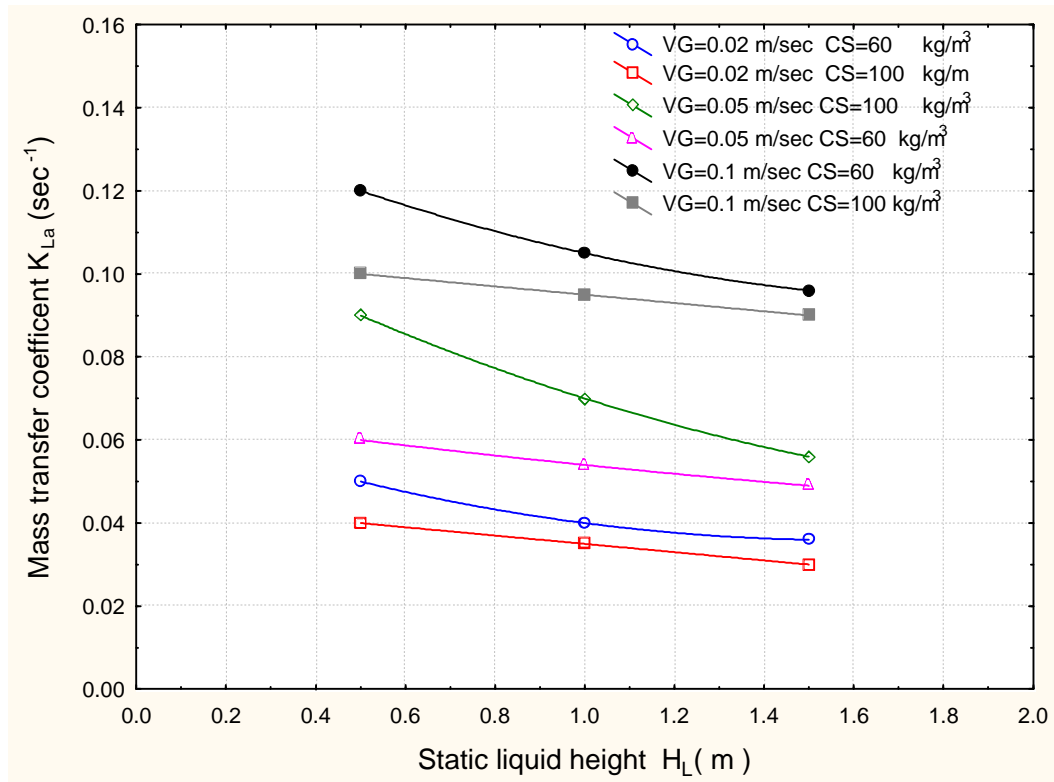


Figure (8) Mass transfer coefficient vs static liquid height for various gas velocities

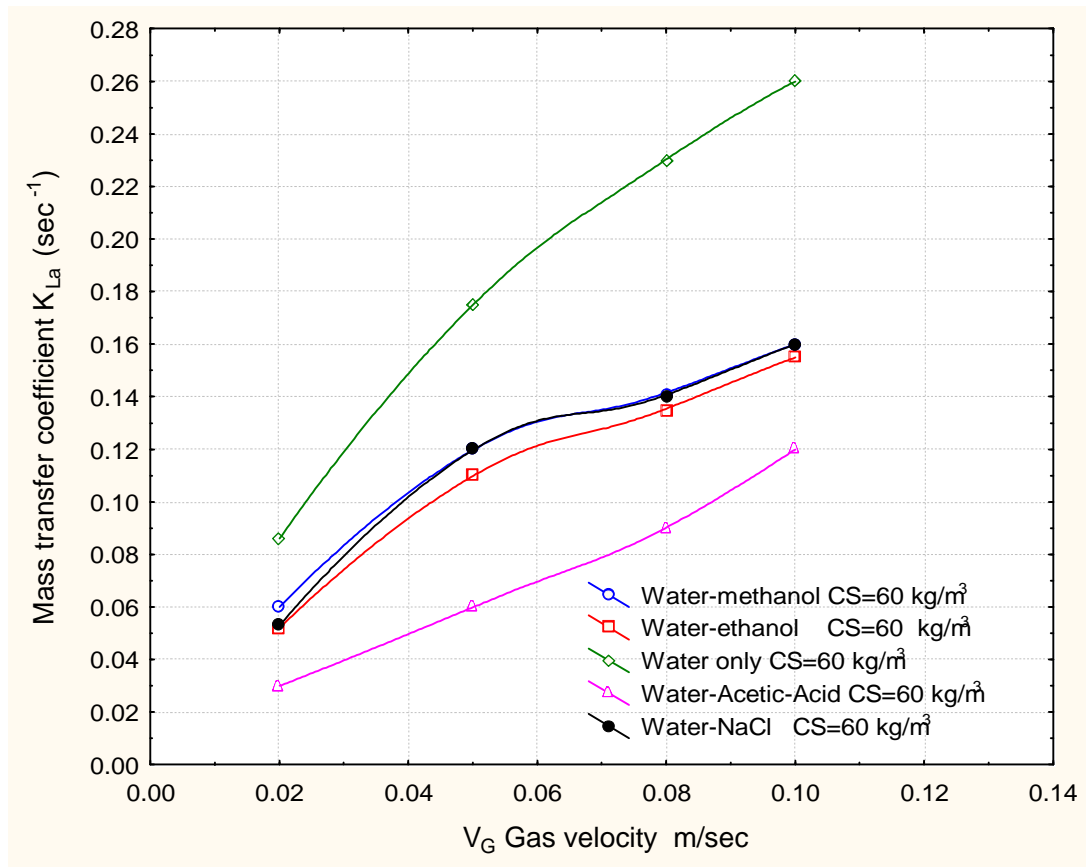


Figure (9) Mass transfer coefficient vs gas velocity for different liquid phase system