BUBBLE COLUMN HYDRODYNAMIC STUDY WITH EXPERIMENTAL INVESTIGATION AND CFD COMPUTATIONS

A Thesis

Submitted to the College of Engineering
of Nahrain University in Partial
Fulfillment of the Requirements for
the Degree of Master of Science in
Chemical Engineering
by

Riham Abdul Majeed Saed

B.Sc. in Chemical Engineering 2003

Rabee Al-Awal 1428

March 2007

CERTIFICATION

I certify that this thesis entitled "**Bubble Column Hydrodynamic Study with experimental Investigation and CFD Computations**" was prepared by ***(Riham Abdul majeed Saeed)*** under my supervision at Nahrain– university college of Engineering in partial fulfillment of the requirements for the degree of Master of Science in chemical engineering.

Signature: Naseer Habobi

Name: Dr. Naseer A. Al-Habobi

Date: 29 / 4 /2007

Signature:

Name: Prof. Dr. Qasim J. Slaiman

(Head of Department)

Date: / /2007

CERTIFICATE

We certify, as an examining committee, that we have read this thesis entitled "**Bubble Column Hydrodynamic Study with experimental Investigation and CFD Computations**", examined the student ***(Riham Abdul Majeed Saeed)*** in its content and found it meets the standard of thesis for the degree of Master of Science in Chemical engineering.

Signature:
Name: Dr. Naseer A. Al-Habobi
(supervisor)
Date: 29/ 4/ 2007

Signature:
Name: Dr. Malek M. Mohammed
(Member)
Date: 2/ 5/2007

Signature:
Name: Dr. Basim O. Hasan
(Member)
Date: 29/ 4 / 2007

Signature:
Name: Dr. Balasim A. Abid
(Chairman)
Date: 2/ 5/ 2007

Approval of the College of Engineering

Signature: M. J. Jweeg
Name: Prof. Dr. Muhsin J. Jweeg
(Acting Dean)
Date: 22 /5 /2007

ABSTRACT

By achieving simulation of hydrodynamics in a bubble column reactor, have taken an important step towards designing an efficient slurry reactor for large-scale conversion of synthesis gas to liquid hydrocarbons. The simulation allows determining gas holdup, a key variable affecting the reaction rate of gas-to-liquid (GTL) conversion using Fischer-Tropsch (FT) chemistry. This information is important because there is the potential for the wide application of FT for conversion of refinery residues, an ever-increasing problem on a worldwide basis. A slurry reactor offers one such method, but detailed knowledge of the hydrodynamics of commercial-size reactors is required. Validated two-dimensional computational fluid dynamicó modeling allows simulating a commercial reactor.

This work have two parts, the first part experimental work with bubble column with specific conditions and the second deals with computational fluid dynamics *CFD* for bubble column have the same geometry and operating conditions.

The experiments of bubble column of (0.081m diameter and 1.03m height) were carried out with air-water, CO_2-*water*, and *Air-NaCl* six concentrations of *NaCl* solutions from 0.24 M to 0.8 M.

This study is to clarify experimentally the effect of superficial gas velocity on the gas holdup, and the mass transfer coefficient (k_La) where each of them increases as the superficial gas velocity increase. *NaCl* addition will increase the transition velocity, and the gas holdup for certain concentration and superficial gas velocity

Three-D simulation was applied in this work. Gas holdup computational results using Ansys *CFX* used on Eulerian-Eulerian multiphase model have been compared to experimental data. These results for air-water system were encouraging.

CONTENTS

Chapter Four: Design of Experiment

Chapter Five: Results and Discussions:

Chapter Six: Conclusion and Recommendation for Future Work

NOMENCLATURE

A	Pipe cross sectional area	m^2
a	Specific gas-liquid interfacial area based on aerated volume	m^{-1}
C_o	Initial concentration of oxygen	ppm
C_i	Saturated concentration of oxygen	ppm
C	Concentration of dissolved oxygen	ppm
c_o	Distribution parameter	-
C_D	Drag coefficient	-
d	Diameter of column	m
D_i	molecular diffusivity of solute in liquid phase	m^2/s
d_b	Individual bubble diameter	m
d_{vs}	Sauter mean bubble diameter	m
E_{LG}	Interphase momentum exchange coefficient	$Kg/m^2.N$
f	Drag function	-
F	External body force	N
F_{lift}	Lift force	N
F_{vm}	Virtual mass transfer force	N
g	Gravitational acceleration	$m^2.s^{-1}$
h_i	Dispersion height	m
h	Initial height	m
h/d	Normalized height of measurement location	-
$i\ d$	Inner diameter	m
I	Jump condition	-
J	Drift flux	$m.s^{-1}$
k_L	Individual mass transfer coefficient	$m.s^{-1}$
k_La	Volumetric mass transfer coefficient	s^{-1}

K	Thermal conductivity of liquid phase	J/m.s.K
m	Mass transfer from one phase to the other	K_g/s
Q	Phase volumetric flow rate	m^3/s
t	Time	s
T_{ij}	Time required for coalescence	s
u	Superficial fluid phase velocity	$m.s^{-1}$
U	Slip velocity	$m.s^{-1}$
u_G^*	Interstitial gas phase velocity	$m.s^{-1}$
u_{Gc}	Critical Superficial gas velocity	$m.s^{-1}$
$u_{b\infty}$	Terminal single bubble rise velocity	$m.s^{-1}$
V	Phase volume	m^3

Greek Letters

ρ	Phase density	k_g/m^3
σ	Interfacial tension	N/m
ν	Kinematic viscosity	m^2/s
ε	Phase holdup	-
τ	Phase stress-strain tensor	N/m
λ	Phase capillary coefficient	Pa.s
μ	Dynamic viscosity	Pa.s
μ_{eff}	Effective viscosity	Pa.s
μ_w	Viscosity of water	Pa.s

Subscripts

G	Gas phase
L	Liquid phase
o	Initial
i	Final

CHAPTER ONE

INTRODUCTION

1.1 Bubble columns

Bubble columns or airlifts are widely used in the chemical industry where heterogeneous gas-liquid or gas-solid reactions take place, particularly, in which the liquid film controls mass transfer processes due to the relative insolubility of gases *[Vázquez et al., 2000; Lye and Stuckey, 2001]*. Important applications of bubble columns include oxidation, hydrogenation, ozonolysis, alkylation, column flotation, wastewater treatment, etc. *[Yang et al., 2001; Wu et al., 2002]*.

The design parameters for bubble columns are: gas-liquid specific interfacial area, a, individual mass transfer coefficient, k_L, flow regime, bubble size distribution, and coalescence of bubbles. Most studies on bubble columns were devoted to the experimental determination of some of these parameters, and more specifically, of the volumetric mass transfer coefficient, k_{La}, which depends fundamentally on the superficial gas velocity and on the physical properties of the absorption phase. Numerous correlations have been proposed for the superficial velocity of the gas as well as for the viscosity of the liquid phase *[Joshi, 2001; Mitsuharu et al., 2001]*.

Facile construction and low costs make bubble columns highly attractive gas-liquid contactors. Fractional gas hold-up ε_G is an important parameter in the design and scale-up of bubble column reactors. It can be defined as the percentage by the volume of the gas in the two or three phase mixture in the column.

Bubble column has two fold applications. One hand, gas holdup in two phase systems gives the volume fraction of the phase present in the reactor and hence their residence time. On the other hand, the gas holdup in conjunction with the knowledge of mean bubble diameter allows the determination of interfacial area and thus leads to the mass transfer rates between gas and liquid phase.

Gas holdup depends mainly on the superficial gas velocity, and often is very sensitive to the physical properties of the liquid *[Shah, 1982]* .The spatial variation of ε_G, gives rise to pressure variation and eventually results in intense liquid phase motion. These secondary motions govern the rate of mixing, heat transfer and mass transfer.

Fortunately, the radial distribution of ε_G can be estimated through the knowledge of the ε_G-u_G relationship. It has several direct and indirect influences on the column performance. The direct and obvious effect is on the column volume. This is because the fraction of the volume is occupied by the gas and the respective phase volume becomes important depending upon the phase in which the rate controlling step takes place. The indirect influences are far reaching.

The spatial variations of ε_G, gives rise to pressure variation and eventually which can be conveniently established experimentally.

1.2 Advantages of Bubble Columns

The main advantages of bubble columns compared to other multiphase contactors are

1. Less maintenance due to absence of moving parts.
2. Higher values of overall mass transfer coefficient can be obtained.
3. Less floor space is occupied and bubble column reactors are less costly.
4. Slow reactions can carried out due to high liquid residence time.
5. Solids can be handled without any erosion or plugging problems.
6. Higher heat transfer rates per unit volume of reactors can be achieved *[Shah, 1982]*.

1.3 Types of Bubble Columns

There are many types of bubble column reactors as following (all types are shown in figure (1.1)).

a. Bubble column of single stage: the simplest type of bubble columns reactor with absence of moving parts. These types are particularly recommended for aerobic fermentation process. This type of bubble column was used in this study with no liquid flow. *[Schurgerl, 1980]*
b. Bubble column of multistage: this type of bubble column does not suffer from limitation, e.g. high energy dissipation.
c. Bubble column of multi channels: used to reduce back mixing and prevent unstable flow condition.
d. Bubble column with loop reactor: this type show a direct well defined circulation flow. Loop reactors can be classified, according to the design, into types with internal and external reactors.

e. Bubble column with jet reactor:
f. Three phase fluidized bed reactor: High pressure operations are common in industrial applications like resid hydrotreating, Fischer-Tropsch synthesis, coal methanation, methanol synthesis, polymerization and other reactions *[Fan, 1989]*
g. Slurry reactor: this type is used as chemical reactor for various processes in industrial practice.
h. Bubble column with static mixers: improved re-dispersion of the gas.
i. Down flow bubble column: this type is recommended when complete gas phase conversion and hence long bubble residence time is required *[Deckwer, 1986]*

1.4 The Aim of the Work

Construct of experimental apparatus and perform the following experiments:

1. Measuring Gas holdup for different conditions.
    a) (*Air-water*) system.
    b) (*Air- NaCl* solution) system with six different concentration of *NaCl.*
    c) (*CO2-water*) system.
2. Measuring the mass transfer coefficient with *air-water* system.
3. Using simulation of bubble column by *CFD* code for measuring gas holdup and hydrodynamic of bubble column.

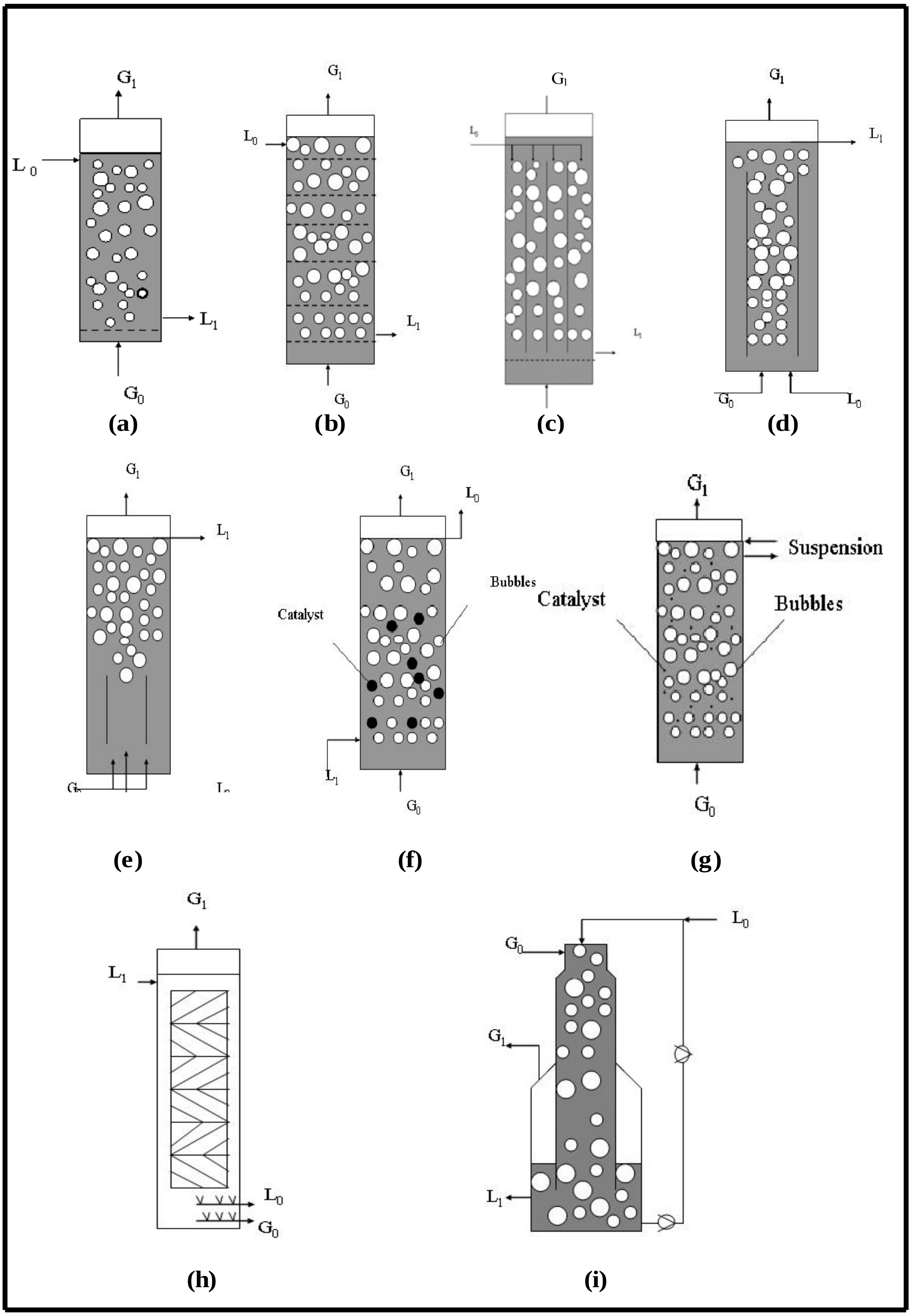


Figure (1.1) Types of bubble columns and modifications

CHAPTER TWO

LITERATURE SURVEY

2.1 Gas holdup

Gas holdup is one of the most important parameter that characterizing the hydrodynamics of bubble columns. It can be defined as the percentage by volume of the gas in the two or three phase mixture in the column.

Many authors studied the gas holdup and the parameters that have effect on it; one of them were Youshida and Akita (1965) who studied the effect of the following parameters on gas holdup:

1. Gas rate: gas holdup increased as the superficial gas velocity increased.
2. Nozzle Diameter: gas holdup is not affected by nozzle diameter.
3. Temperature: gas holdup is not affected by the temperature.
4. Liquid height: gas holdup is particularly independent of the liquid height.
5. Viscosity: gas holdup varies with liquid viscosity in an irregular manner.

Akita and Youshida (1973) also had presented data for column diameter range (0.152-0.6) m for various systems. It was found that the gas holdup varies with the density and viscosity of liquid, surface tension, and superficial gas velocity.

Kumar et al. (1976) found that the gas velocity in bubble column showed a remarkable effect on the gas holdup. Holdup data (ε_G <0.35) where in good agreement with those observed for columns of diameter 5.1×10^{-2} or larger when orifice sparger type distributor were used.

Batch and Pilhofer (1978) suggested that at high gas throughputs, the gas holdup mainly depends on the gas flow rate and on the physical properties of liquid and it is independent of column diameter for diameter of over 0.150 m.

Marruyam et al. (1981) found that at $u_G < u_{Gc}$ (where u_{GC} is the critical velocity and it was defined as the superficial gas velocity at the point incipient regular circulation and was obtained from gas velocity at maximum gas holdup), where the values of gas holdup calculated by static pressure difference agreed well with those from an increase in liquid level.

Shah et al. (1983) carried out experimental studies for two different diameter column using cocurrent and batch systems. Five aliphatic alcohols were investigated with different concentrations. The gas hold up was measured using hydrostatic head technique and it was found that the gas holdup increases with increasing in gas velocity, strongly influenced by the type of alcohol, and not affected by alcohol concentration significantly.

Uchida and Tsuyutani et al. (1989) presented gas holdup correlation as a function of operating variables such as liquid and gas superficial velocity, liquid slip velocity, and liquid properties by using (air-glycerolsolution) and (air-water) system. The gas holdup increased sharply with increasing liquid velocity. However when liquid velocity exceeds a certain value, gas holdup began to decrease.

Philipet al. (1990) determined gas holdup for highly viscous liquids, for highly viscous Newtonian liquids, the gas holdup in slug flow regime (at a given gas flow rate) increases with an increase in liquid viscosity, this is because the absolute velocity of slugs decrease as the viscosity increase due to (i) lower liquid circulation rate and (ii) Lower single slug rising velocity.

Thorat *et al.* (1998) studied the effect of sparger design and dispersion height on ε_G, gas holdup was investigated in a 0.385 m *id* bubble column. Perforated plates were used as spager. The height of the diameter ratio (h_i/d) was in the range of 1-8. In all cases in order to investigate the combined effect of sparger design and (h_i/d) ratio together with the coalescing nature of the liquid phase, three liquid systems were considered, namely, water, an aqueous solution of electrolyte and aqueous solution of carbixymethyl cellulose (CMC). With an increase in the height to diameter (h_i/d) ratio, ε_G was found to decrease when multipoint spargers (with hole diameter $<3\times10^{-3}$ m) were used. In contrast, for single point spargers, ε_G was found to increase. However, in both the cases, a limiting (h_i/d) ratio was observed beyond which the values of ε_G remain practically constant.

The above observations were found to hold for three gas-liquids systems: (i) air-water (ii) relatively less coalescing system air-aqueous solution of electrolyte and (iii) relatively more coalescing system air-aqueous solution of carboxymethyl cellulose (CMC). The limiting value of the (h_i/d) ratio (beyond which ε_G is independent of (*hi/d*) was found to be in the range of 4-5 for the air-water system, greater than 8 for the air-electrolyte system and 3 for the air-aqueous CMC system.

Veera and Joshi (2000) used 0.385 m I.D bubble column to investigate the variation of gas holdup using gamma ray tomography. Radial profiles were measured at three axial locations (h_i/d =0.259, 3, and 5). Two types of liquid phase were selected; one was coalescence inhibiting (aqueous solution of butanol) and the other coalescence promoting (aqueous solution of carboxyl methyl cellulose).perforated plates were used as spargers. The gas holdup profiles was found to dependent strongly on the sparger design for an air alcohol solution up to axial location of height to diameter ratio (h_i/d) of 3. Whereas for air- water system and CMC solutions, the dependence on (h_i/d) ratio was progressively less. Further the height of the dispersion was found to influence the holdup profiles near the sparger.

Bohn, 2000 used 3D simulation of bubble column, gas holdup was calculated by plotting the distribution of gas volume fraction vs. column height. The plot showed a point where the concentration of gas increased rapidly which is the gas-liquid interface. The next step was to compare the height of the liquid-gas interface to the original height of the liquid-gas interface. The expansion, or the amount of the liquid-gas interface rose, showed how much gas holdup in the reactor, from which the gas holdup is easily calculated.

Spica *et al.* (2001) used a commercial CFD code to simulate a two phase flow in bubble column both with a stagnant and flowing liquid phase.

Different sets of gas and liquid velocity were applied as inlet boundary conditions, obtaining the local gas volume fraction as well as liquid velocity profile for an air-water system was studied.

Mouza et al. (2004) presents (by a plot of u_G vs. ε_G) the effect of superficial velocity with respect to the column cross section on the overall gas holdup in bubble column reactor using CFD simulation. A comparison between the experimental data *[Mouza et al. (2004)]* with the simulation of CFD code that calculates satisfactorily the gas holdup at homogenous regime also was made.

Dhotr *et al.* (2004) studied the effect of sparger design and the column height to the diameter ratio on radial gas holdup profiles for three different gas-liquid systems: air-water, air-aqueous solution of butanol, and air aqueous solutions of carboxyl methyl cellulose.

Martis, (2004) used bubble column of a 4-inch diameter plexi-glass cylinder. Gas holdup and flow regime transition velocity measurements in a bubble column reactor were made. A transition velocity between homogeneous and heterogenous of about 3 cm/s was found. The transition velocity was found to be unaffected by standing water column height, provided that the height was larger than the diameter of the column

Table (2.1) Gas holdup correlation in literature

(All Dimensional Quantities Are in SI Units)

System	Range of Parameters	Correlation Proposed	Reference
Air-water Air-kerosene Air- Na_2SO_3 aq. soln. Air-Glycerol Air-light oil Air-ZnCl2 aq.soln. ρ_L, Kg/m^3: (780-1700) μ_L, Pa.s: (0.0009-.0152) σ, N/m: (0.25-0.076)	d, m > 0.1 u_G,m/s:(0.004-0.45)	$\varepsilon_G = \frac{1}{2+(0.35/u_G)(\rho_L\sigma/72)^{1/3}}$	**Hughmark (1967)**
Air-water Air-glycol aq.Soln. Air-methanol O_2-water	u_G,m/s: (0.003-0.4) u_L,m/s: (0-0.044) d, m: (0.125-0.6) $h_{I:}$, m: (1.26-3.5)	$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} =$ $C\left(\frac{gD^2\rho_L}{\sigma}\right)^{1/8}\left(\frac{gD^3}{\nu_L^2}\right)^{1/12}\left(\frac{u_G}{\sqrt{gD}}\right)$ C = 0.2 for pure liquid non electrolytes C = 0.25 for electrolytes	**Akita and Youshida (1973)**
Air-water Air-methanol aq. Soln. ρ_L, Kg/m^3: (910-1200) μ_L, Pa.s: (0.00043-0.02) σ, N/m: (0.0214-0.0728)	u_G,m/s: (0.042-0.38) d, m : (0.1-0.19) h_i, m: (0.6-1.35)	$\varepsilon_G = 0.505\, u_G^{0.47}(0.072/\sigma)^{2/3}(0.001/\mu_L)^{0.05}$	**Hikita and Kikukawa (1974)**
Air-Different liquids ρ_L, Kg/m^3: (800-1600) μ_L, Pa.s: (0.00043-0.02) σ, N/m: (0.0214-0.0728)	u_G,m/s,: (0.01-0.08) d, m : (0.0759-.61), h_i, m: (0.02-3.5)	$\varepsilon_G = 0.89\left(\frac{H_i}{D}\right)^{0.036(-15.7+\log K)}$ $\times\left(\frac{d_b}{D}\right)^{0.3}\left(\frac{u_G}{d_b g}\right)^{20.025(2.6+\log K)}$ $\times K^{0.047} - 0.05$ $K = \left(\frac{\rho_L\sigma^3}{\mu_L^4}\right)$, $d_b = 0.003m$	**Gestrich and Rashe (1975)**

Table (2.1): Continued

Air-water Air-Glycerol aq. Soln. Air-kerosene ρ_L, Kg/m^3: (800-1100) μ_L, Pa.s: (0.0009-0.0115) σ, N/m: (0.0312-0.072)	u_G,m/s:(0.0014-0.14) d, m: 0.05 and 0.1	$\varepsilon_G = 0.728U - 0.485U^2 + 0.0975U^3$ $U = u_G \left[\frac{\rho_L^2}{\sigma(\rho_L - \rho_G)g} \right]^{0.25}$	**Kumar *et al.* (1976)**
Air-Alcohols Air-Halogenated Hydrogens	u_G, m/s: 0-0.1 d,m : >0.1 h, m : >1.2	$\frac{\varepsilon_G}{(1-\varepsilon_G)} = 0.115 \left[\frac{u_G^3}{\nu_L g(\rho_L - \rho_G)/\rho_G} \right]^{0.23}$	**Batch and Pilhofer (1978)**
Gas-Liquid	Theoretical equation	$\frac{\varepsilon_G}{\left(1-\varepsilon_G^{1/3}\right)} = \frac{u_G}{u_{b\infty}}$	**Iordache and Nuntean (1981)**
Different Gases (Air, H_2, CO_2, CH_4, C_3H_8, N_2)-water Air-Organic liq. Air-Electrolyte ρ_L, Kg/m^3: (790-1170) μ_L, Pa.s: (0.0009-.0178) σ, N/m: (0.229-0.0796) ρ_G, Kg/m^3: (0.84-1.84)	u_G,m/s,: (0.042-0.38) d, m: 0.1 h_i, m: 0.65	$\varepsilon_G = 0.672 f \left(\frac{u_G \mu_L}{\sigma} \right)^{0.578} \left(\frac{\mu_L^4 g}{\rho_L \sigma^3} \right)^{-0.131}$ $\times \left(\frac{\rho_G}{\rho_L} \right)^{0.062} \left(\frac{\mu_G}{\mu_L} \right)^{0.107}$ f = 1.0 for electrolyte solutions, f = 1.1 $I >$ 1.0 kg ion/m^3 f = $10^{0.04141}$ $I <$ 1.0 kg ion/m^3, I =solution's Ionic strength	**Hikita *et al.*** **(1980)**

2.2 Volumetric Gas-Liquid Mass Transfer Coefficients

Several studies in literature dealt with mass transfer in bubble column, one of these studies was done by Braulick et al. (1965) studied the air oxidation of aqueous sodium sulfite solution in simple bubble contacting column of 3, 4, and 6 diameters. Mass transfer was found to increase with gas rate and with submergence/diameter ratio. The mass transfer characteristics of gas-sparged bubble column were found to be equivalent to those stirred vessels.

Youshida and Akita (1965) studied the effect of the following parameters on k_La:

1. Gas rate: data of sulfate oxidation in 0.077 m column diameter showed that the k_La values increased as superficial gas velocity increase.
2. Nozzle diameter: for many nozzles of various diameters it was found that the k_La is not affected by the nozzle diameter. This was explained by the fact that, in the range of his study, gas flows out of the nozzle as a continues jet stream and then was split into bubbles by turbulent motion of liquid in a zone several inches above the nozzle.
3. Liquid height: the data showed that the k_La values are particularly independent on liquid height.
4. Column diameter: k_La values increased as the column diameter increase.
6. Temperature: k_La values increased with increasing temperature.
7. Liquid viscosity: this parameter was studied by performing experiments on desorption of O_2 in 0.152 m. It was found that the k_La values decrease with increasing viscosity, probably owing to decrease in liquid diffusivity with increasing viscosity.

Akita, and Youshida (1973) proposed experimental data for k_La for various systems, they fond that, liquid phase diffusivity and the column diameter affect k_La, which is proportional to the gas holdup to the 1.1 power.

Deckwer et al. (1974) used two columns (0.15 and 0.20 m, 4.4 and 7.23 m high) with different gas distributors, measurements of k_La were carried out with tap water. Owing to the different gas sparger the k_La values of both columns differ by a factor about two.

Nakanoh and Youshida (1980) used 0.1455 m inside diameter and 1.9 m height to calculate k_La in both inelastic and viscolastic liquids and presented an equation for k_La. For inelastic liquids the relation between k_La and superficial gas velocity was found straight line of unity slope, for viscolastic liquids, k_La decrease as increase viscolastisity.

Vermeer and Krishna (1981) used 4 m tall, 0.19 m diameter column with N_2 and turpentine 5 as gas and liquid phases, respectively. They found that large bubbles have k_La values, (which agree with extrapolation of the value in literature) to be an order of magnitude higher than those calculated from correlations derived for small bubbles. Also values of k_La have been found to be unimpaired by the bubble coalescence.

Alvarez-Cuenca and Nerenberg (1981) calculated k_La values from both experimental and mathematical model; the result indicated that k_La increase, often linearly, with the superficial gas velocity. Further more, for liquid-phase range [0, 0.01] m/s, k_La had a minimum at u_L~ 0.075 m/s. for larger superficial liquid velocities, an increase in k_La was found.

Hikita *et al*. (1981) carried out an experiment with 0.1 and 0.19 m I.D., bubble column with single-nozzle gas sparger with several systems were used. For non- electrolyte solutions they found that k_La values is proportional to about 0.76 power u_G and it is not effected by either by nozzle diameter or column diameter. For electrolyte solutions the k_La values varies as 0.76 power of u_G in the same manner to non-electrolyte solutions, and these values are slightly larger than those for water, the extent depending on the nature of the electrolyte and its concentration.

Jeng et al. (1986) studied the effect of surface-active additive on bubble behaviors and mass transfer coefficient on CO_2 absorption in bubble column under fixed gas flow rate. Addition of small amount of surface-active in the liquid phase had the effect of retarding the coalescence of gas bubble, thus making the gas-liquid interfacial area, a, larger; it also had the effect of resisting the stretching or compression of the interface and reducing disturbance in the bulk fluids, therefore making mass transfer coefficient, k_La smaller. Because of these two opposing factors, the k_La had a maximum value at a very low surfactant concentration.

Uchida et al. (1989) proposed k_La correlation and compared their own experimental with the correlation values; the error on k_La is 40% maximum. Observed data were smaller than that correlated values for solutions containing surface active agent T-20 with 0.01vol.% Silicon oil which is used as foam, the foam inhibitor (i.e. silicon oil) is considered to affect mass transfer.

Wei-rong et al. (2004) carried out an experiment to determine the dependence of a, k_L, and k_La on the surface tension, the temperature of the absorption phase, and superficial velocity of gas. Mass transfer measurements were carried out using bubble column with 0.6 m; the internal and the external diameter of the column were 0.102 m and 0.108 m respectively. The absorption phase used was sodium carbonate-bicardonate buffer solutions with sodium arsenite as catalyst and DBS as surface tension modifier with CO_2 as inlet gas. The result showed that the k_La increased with increasing temperature, decreased with increasing surface tension modifier, or decreased with increasing the surface tension. The reduction of k_La as the surface tension modifier decreased may be attributed to the effect of the surface active-agent, which can reduce the interfacial movement when it occupied part of the surface of the bubble.

Our work contains two parts, the first part experimental work with bubble column with specific conditions and the second deals with computational fluid dynamics *CFD* for bubble column have the same geometry and operating conditions. The experiments of bubble column of (0.081m diameter and 1.03m height) were carried out with air-water, CO_2-*water*, and *Air-NaCl* six concentrations of *NaCl* solutions with (0.24-0.8) M *NaCl* solution range.

Table (2.2) Reported Studies for $k_L a$

System	Parameter's Range	Correlation proposed	Reference
Water-air Glycol-air Methanol-air Glycol aq. soln.-air Water-O_2 Water-He Water-CO_2	u_G,m/s: (0.003-0.4) u_L, m/s: (0-0.44) D, m: 0.152-0.6 Hi, m: (0.126-0.35) ρ_L, kg/m^3: (800-1600) μ_L, Pa.s: (0.00058-0.021) σ, N/m: (0.022-0.0742)	$\frac{k_L a D^2}{D_i} = 0.6\left(\frac{\upsilon_L}{D_i}\right)^{0.5}\left(\frac{gD^2\rho_L}{\sigma}\right)^{0.62}$ $\times\left(\frac{gD^3}{\nu_L^2}\right)^{0.31}\varepsilon_G^{1.1}$	**Akita and Youshida** **(1973)**
Gas-liquid	Theoretical Equation $*\left(\frac{d_b\rho_L u_G}{\mu_L\varepsilon_G}\right)^{0.5}$	$K_L a = 3.31\frac{D_i\varepsilon_G}{d_b^2}\left(\frac{\mu_L}{\rho_L D_i}\right)^{1/3}$	**Fair** **(1967)**
Air-water Sucrose aq. soln. –air CMC aq. soln. –air Soduim polyacrylate aq. soln. -air	u_G, m/s: < 0.1 ρ_L, kg/m^3: (995-1230) μ_L, Pa.s: (0.005-0.06) C = 0 for un elastic liquids C= 0.133 for elastic liquids m = 0.55 λ=characteristic relaxation time	$\frac{k_L a D^2}{D_i} = 0.09\left(\frac{\nu_{eff}}{D_i}\right)^{0.5}\left(\frac{gD^2\rho_L}{\sigma}\right)^{0.75}$ $\times\left(\frac{gD^3}{\nu_{eff}^2}\right)^{0.39}\left(\frac{u_G}{gD}\right)\left(1+C\left(\frac{u_{b\infty}\lambda}{d_{vs}}\right)\right)^{m-1}$	**Nakanoh and Yoshida** **(1980)**
Water-air Water -O_2 Water-H_2 Water-CH_4 Water-CO_2 Sucrose soln.–air n-butanol-air methanol soln.–air electrolyte soln.-air	u_G,m/s: (0.042-0.38) D, m : (0.10-0.19) Hi, m: (0.13-0.22) ρ_L, kg/m^3: (998-1230) μ_L, Pa.s: (0.0008-0.011) σ, N/m: (0.025-0.082) D_i, m^2/s: (4.6-26.0)	$k_L a = \frac{14.9 g f}{u_G}\left(\frac{u_G\mu_L}{\sigma_L}\right)^{1.76}\left(\frac{\mu_L^4 g}{\rho_L\sigma^3}\right)^{-0.248}$ $\times\left(\frac{\mu_G}{\mu_L}\right)^{0.243}\left(\frac{\mu_L}{\rho_L D_i}\right)^{-0.604}$ f = 1.0 for non electrolytes $f = 10^{0.068I}$ I< 1.0 Kg ion/m^3 $f = 1.114*10^{0.021}$ I > 1.0 Kg ion/m^3	**Hikita *et al.*** **(1981)**
Aqueous CMC soln. (1.0-2.0 wt%)	u_G,m/s: 0.08 D, m : 0.14 Hi, m: 2.6	$k_L a = 0.00315\ u_G^{0.59}\mu_{eff}^{-0.84}$	**Deckwer *et al.*** **(1981d)**
CO_2-carbinate-biocarbonate buffer soln.	D, m : 0.102 Hi, m: 0.6	$\ln(k_L a) = 11.43 + 0.796\ln(\sigma)$ $+1.045\ln(u_G) - \left[\frac{2.30\times10^3}{T}\right]$	**Wei-rong *et al.*** **(2004)**

CHAPTER THREE

THEORY

3.1 Hydrodynamics in Bubble Column

The most important process in a bubble column is the formation of gas bubbles at the gas sparger. The smaller bubbles, give larger area for mass transfer between gas liquid phase. Bubble swarm behavior in a bubble column is mainly determined by the gas superficial velocity *[Heijnen, 1984]*.

3.2 Bubble Dynamics

Bubble size, bubble rise velocity, bubble size distribution, and liquid and bubble velocity profiles have a direct bearing on the performance of bubble column. The bubble rise velocity is equivalent to the interstitial gas velocity, which follows from the superficial gas velocity *[Shah, 1982]*

$$u_G^{*} = u_G / \varepsilon_G \qquad \ldots(3\text{-}1)$$

3.3 Flow Regimes

In a bubble column, the gas rises through the liquid in three different ways, called flow regimes as shown in figure (3.1).

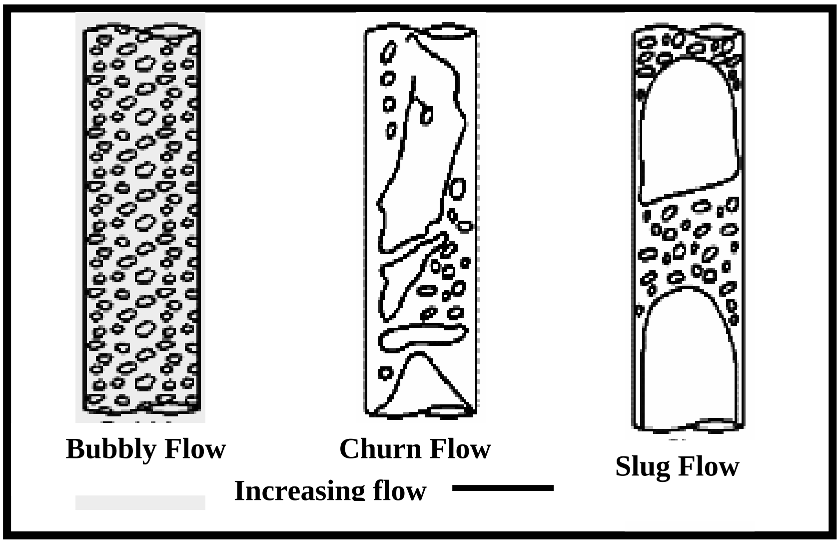


Fig. (3.1) schematic of flow patterns *[Shohman, 1982]*

3.3.1 Homogeneous (Bubbly) Flow

Homogenous flow is characterized by small, uniform bubbles with little to no interaction and slow velocities mostly straight up the column *[Martis, 2004]*.This regime occurs if the superficial gas velocity is less than 0.05 m/s *[Fair, 1967]* and the rise velocity of bubbles lies between 0.18 and 0.03 m/s *[Levich, 1962]* these data should be only be regarded as guideline which are valid for aeration of water *[Shah, 1982]*. The bubble generated at the sparger rise undisturbed virtually vertical axial oscillation.

Homogeneous or bubbly flow is observed at high liquid flow rate and low gas flow rate. The concentration of bubbles is uniform, particularly in the transverse direction. The process of coalescence and dispersion are practically absent in the homogeneous regime and hence the sizes of bubbles are entirely dictated by the sparger design and the physical properties of the gas and liquid phases *[Thorat, 1998]*.

Observed change from bubbly flow to transition flow is asymptotic, depending on various factors *[Koetsier et al. 1976; Lockett et al.; 1975; Wallis, 1969]* which affect the size of the gas bubble by altering the degree of coalescence.

3.3.2 Heterogeneous (Turbulent) Flow

Heterogeneous flow is characterized by larger bubbles formed when small bubbles coalesce and interact with each other and, as a result, the bubbles have a range of speeds in varied directions *[Martis, 2004]*.

In the heterogeneous regime the role of sparger design diminishes depending upon the column height. In fact, the total column height can be divided into two regions: the sparger region and the bulk region *[Thorat, 1998]*. The large bubbles take the form of spherical caps with very mobile and flexible interface. These large bubbles can grow up to diameter of about 0.15m.

Experimental evidence suggested that the liquid's axial (main stream) velocity in the plume is essentially uniform and that the thickness of boundary layer at the wall of the column is negligible. *[Youshida and Shiria, 1970]*

Since some basic bubble column properties such as mass transfer, gas holdup, and interfacial area are highly varied in the different flow regimes, finding the transition between the two regimes is critical to an accurate representation of reactor performance.

The limits of the transition region between homogeneous and the heterogeneous bubbling regimes are characterized by the onset and the development of liquid circulation patterns in bed *[Zahradinik et al., 1997]*

In this study we use a range of velocities (0.009657-2.37995) m/s that covers the homogeneous and heterogeneous flow regime.

3.3.3 Slug Flow

This regime occurs in columns with a small diameter. This slug flow is characterized by large bubbles that form slugs when stabilized by the column's walls *[Wild, 2003]*. Slug flow can be observed in columns of diameter up to 0.15 m *[Hills, 1976; Miller, 1980]*.

If the residence time within the column is "sufficiently" long, bubbles of a size comparable to the column diameter will be formed unless the foam is formed *[Anderson and Quinn, 1970]*. If the volume of these bubbles is greater than about 5×10^{-6} m^3, they will form the characteristic spherical-cap shape *[Batcher, 1967]*. These spherical-cap bubbles then agglomerate to form slugs *[Moissisr and Nicodeno, 1967]*

3.4 Bubble Formation and Growth

The formation of bubbles in liquid is a dynamic process in which the shape and the size of bubbles vary continuously. It is assumed that the liquid is displaced from the bubble which is forming as shown in figure below.

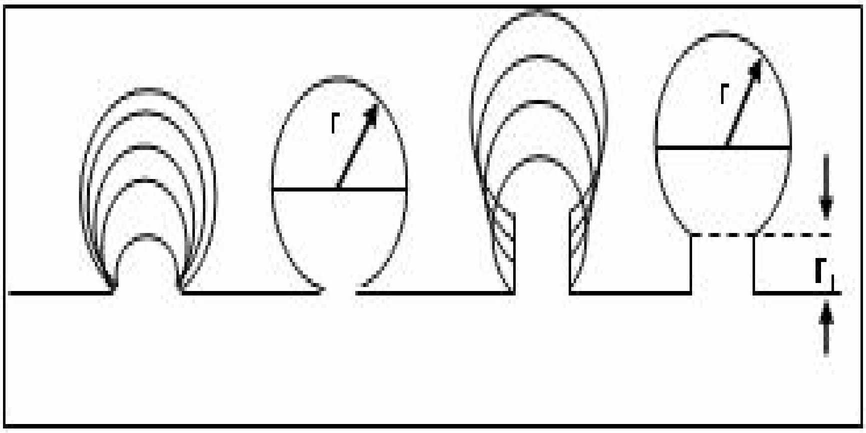


Fig (3.2) Bubble formation *[Kumar et al., 1967]*

During the expansion step, the bubble adheres to the capillary orifice, since the retaining forces are stronger than the detaching forces. As soon as the equilibrium of forces has been attained the bubble becomes detached but remains linked to the capillary through a neck. The bubble is severed when the neck attains a length which corresponds to the radius of the bubble at the conclusion of the expansion step *[Kumar, 1967]*. It is assumed that the bubble formation from orifice follows the following procedure in figure (3.3) and on the basis elementary force balance.

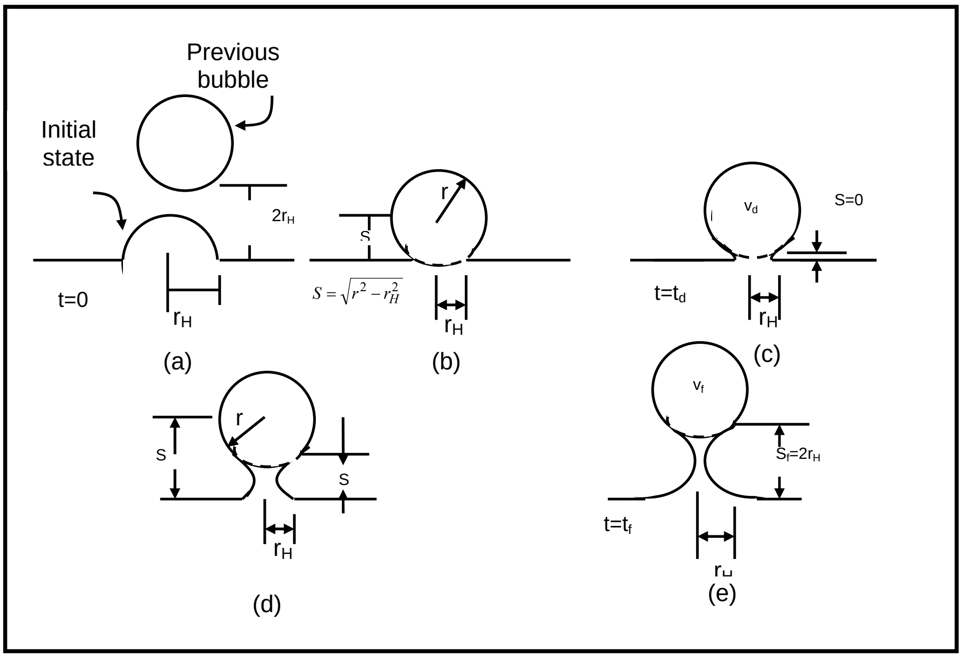


Fig. (3.3) Bubble Formation Stages.

(a) bubble birth, (b) bubble expansion, (c) bubble lift-off, (d) bubble rise with continued expansion, (e) bubble detachment (final bubble size)

[Geary and Rice, 1991]

3.5 Bubble Coalescence

Bubble coalescence plays a significant role in determining bubble size distribution, gas holdup, interfacial area and bubble rise velocity, which govern the performance of bubble columns and distillation towers. When two bubbles come in contact, a thin liquid film forms between them, draining until an instability forms, then coalescence occurs. The bubble coalescence mechanism is considered to be a three-step process:

1) The approach of two bubbles to within a distance of 10^{-5} to 10^{-6} m.
2) Further thinning of liquid layer between bubbles to thickness about 10^{-8} m.

3) Rupture of the thin liquid layer via an instability mechanism. *[Kim and Lee 1987].* The bubble coalescence principally because of the relative motion of elements of dispersed phase of different sizes *[Jackson, 1964].*

The diameter of the gas bubble of the gas distributor is not necessarily the same as the bubble diameter in the bulk of the column. Bubble distributor can undergo coalescence and/or (re) dispersion process. The coalescence rate is independent on the liquid surface properties, varying from coalescence (e.g. pure liquids) to non coalescing (e.g. water-salt system), and depends on whether cations or anions are adsorbed at the surface of the bubbles. Therefore, the distinction between coalescing and non coalescing properties is very important in determining the performance of the bubble column *[Heijiner and Riet 1984] [Marucci and Nicodemo, 1967.* The effect of bubble formation frequency on bubble coalescence frequency showed opposite tendencies according to species of solute *[Kim, 1987].*

Coalescence process depends on the collision rate of the two bubbles and collision efficiency that is a function of time required for coalescence t_{ij} and contact time T_{ij}.

Collision is result of three mechanisms:

- Turbulence θ_{ij}
- Laminar shear θ_{ij}^{Ls}
- Buoyancy θ_{ij}^{B}, and the total coalescence rate is :

$$Q_{ij} = \left(\theta_{ij}^{T} + \theta_{ij}^{B} + \theta_{ij}^{Ls}\right)\ e^{\left(-t_{ij}/T_{ij}\right)} \qquad \ldots(3\text{-}2)$$

[Prince and Blanch]

It observed that there exist sufficiently strong attractive forces between bubbles in water to coalesce. In these cases the attractive vander waals forces overcome the repulsive forces and the total free energy diminishes, thus reducing the contact area as shown in figure (3.4) *[Aguilera, 2000]*

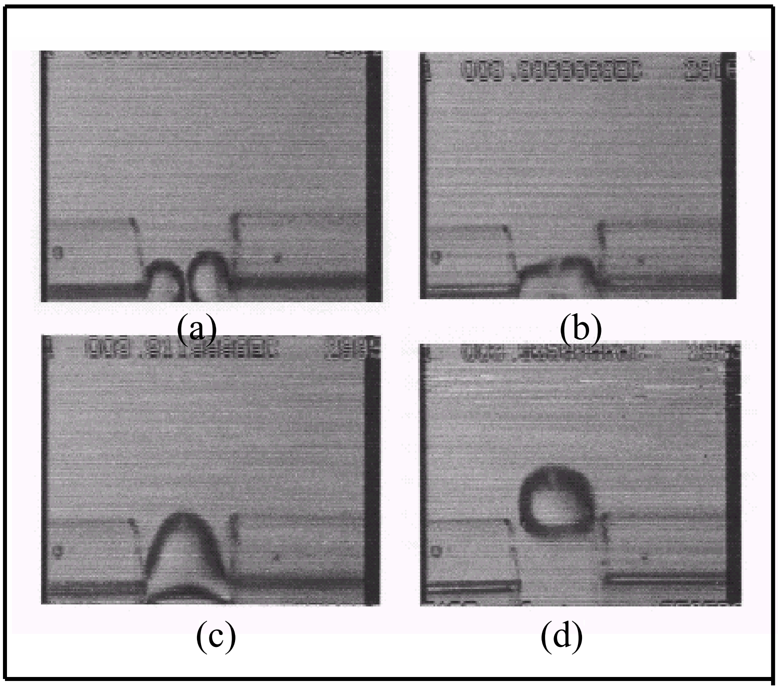


Fig. (3.4) (a) formation of bubbles, (b) Coalescence of bubbles, (c) Formation of new bubble, (d) a new elliptical bubble arises

[Aguilera, 2000]

Bubble coalescence was found to follow first-order rate dependence on bubble concentration *[Calderbank, 1962].*Collision arises from relative motion between bubbles, which specifically excluded by steady-state, uniform bubbling.

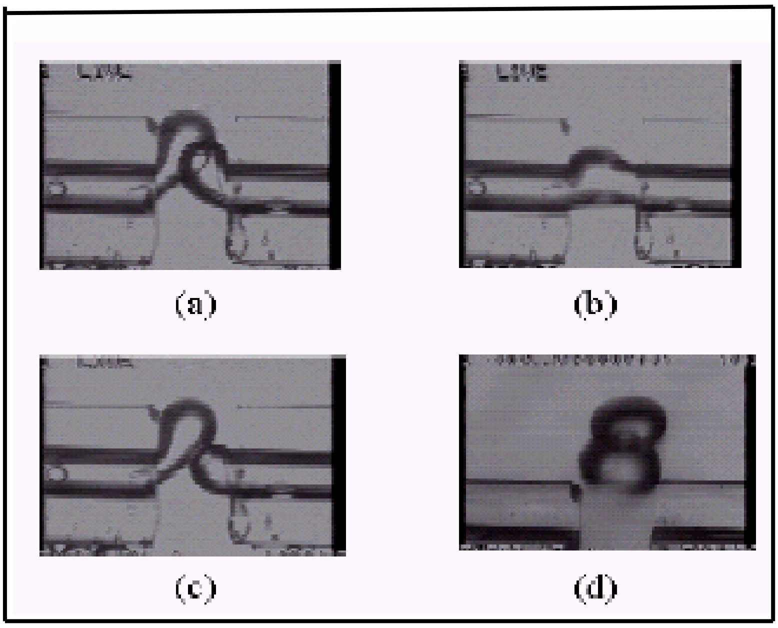
In reality, there will be random fluctuations, even in uniformly bubbling state; these will lead to initial collisions which promote coalescence. Once large bubbles have been formed, they will rise faster than the rest, overtaking and giving the possibility of further collision *[Hills, 1975].*

Under coalescence-suppressing condition the most evidence features observed are that the uniform bubble swarm begins to meander, indicating an occurrence of liquid circulation near the walls *[Maruyama, and Youshida 1981]*

3.6 Effect of Electrolyte Solutions

The aerated solutions of electrolytes behave differently from pure liquids. Although their physical properties differ only slightly from those of pure liquids, large deviations of gas holdup occur at superficial gas velocities above 0.04 m/s *[Batch and Pilhofer, 1978].*

It observed that the electrolytes act to prevent coalescence by substantial hydrodynamic repulsion force as shown in figure (3.5). For electrolyte solutions, the value of gas holdup increases with increasing electrolyte concentration. This means that non coalescing property of electrolyte reaches its limiting value at critical concentration. Therefore it is desirable to use an electrolyte concentration above the critical concentration value *[Thorat et al., 1998].*



(a) (b) (c) (d)

Fig. (3.5) (a) Formation of bubbles, (b) Superposition of bubbles, (c) Separation of bubbles, (d) Two spherical bubbles arise *[Aguilera, 2000]*

The gas holdup for lower electrolyte concentration is increased by ionic forces in the bulk of the solution rather than by reduce size bubbles from the gas distributor. The bubble coalescence rate is high and the gas holdup is low for pure water, presences of the electrolytes decrease the coalescence rate and, therefore, increase the gas holdup *[Jamal Ahmadi and Steinshagen, 1990]*

For all electrolytes studied, the appropriate dependence (ε_G vs. u_G) corresponding to respective bubbling regimes, were achieved at concentration close to the transition concentration, marking for individual electrolytes transition from total bubble coalescence to virtually complete coalescence suppression.

The effect of electrolytes presence on bubble voidage has been considerably more pronounced in the homogeneous and transition bubbling regimes than in heterogeneous regime and indicates magnifying effect of electrolyte addition on the differences between the (ε_G vs. u_G) dependence for the different bubbling modes *[Zaharadnik et al., 1997]*

For high electrolyte concentration and low superficial gas velocities, large bubbles form at the gas distributor plate. If the gas velocity is increased, these bubbles break up into smaller bubbles thus increasing the gas holdup and then the mass transfer *[Jamal Ahmadi and Stienhagen, 1990]*

The surface tension at the interface between bubbles and electrolyte solution is higher than that for pure water, due to the presence of dissolved ions. However because of the higher attractive forces between water molecules, these ions will be quickly removed from interface into the bulk of the solution. Therefore, the actual surface tension should be lower than the initial surface tension, approaching that of pure water for electrolyte concentration *[Jamal Ahmadi, 1990].*

The boundary between bubbly flow and churn turbulent flow for aqueous solution of non-electrolyte is more strongly affected by operating conditions and geometric of gas sparger and bubble column rather than by liquid properties such as viscosity, surface tension, etc. *[Uchida et al. 1989].*

The boundary (bubbly flow-churn turbulent flow) for aqueous solution of electrolytes or organic solutes is influenced by liquid properties (ionic strength or number of carbon atoms, etc.) because bubble coalescence is restrained by these solutes *[Keitel and Oken, 1982].*

3.7 Phenomena That Affect Bubble Column Performance

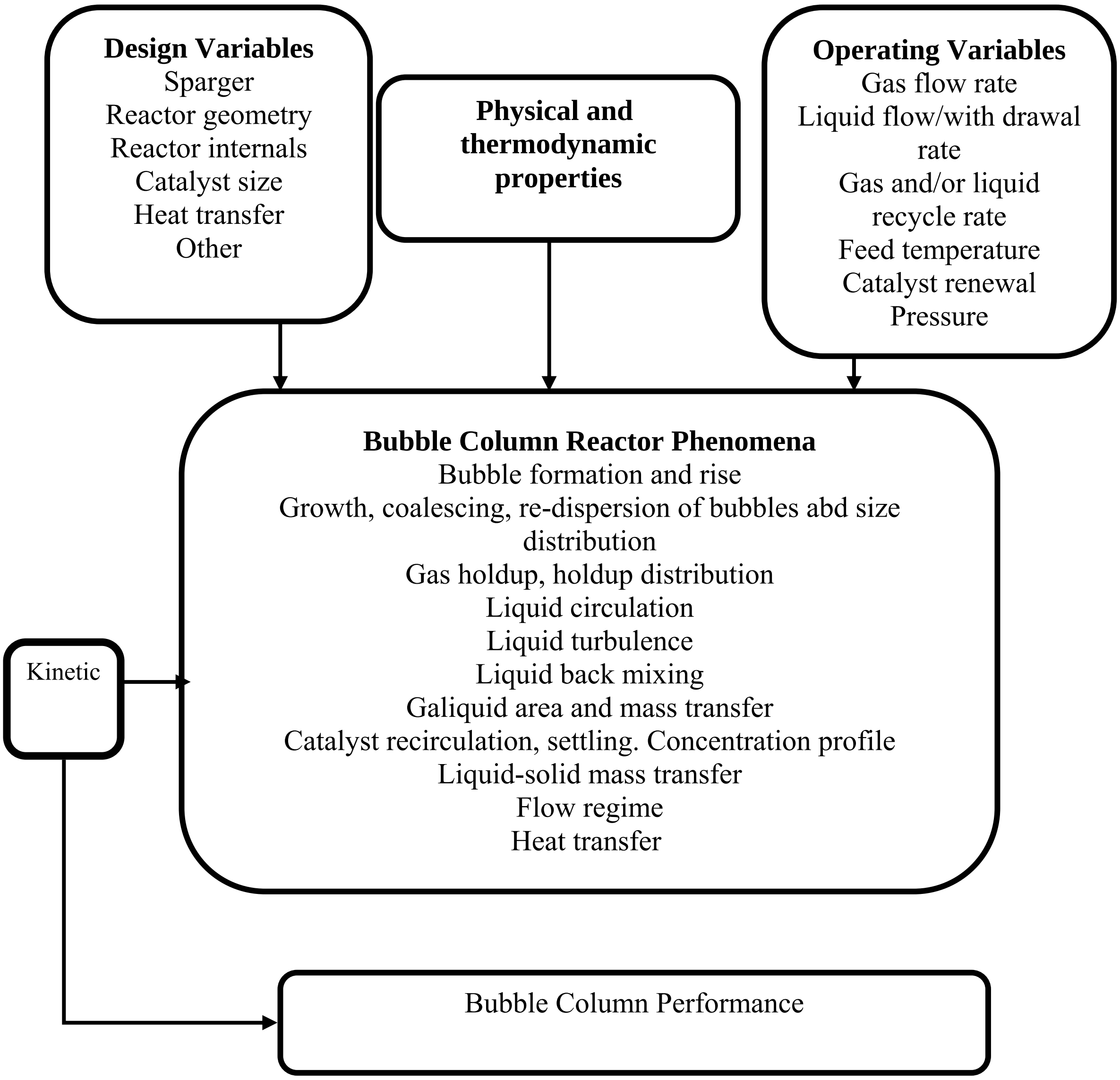


Fig. (3.6) Phenomena affecting bubble column performance

[Dudukov,1997]

3.8 Definition of Basic Parameter

The superficial velocities of the liquid and gas phases (u_L and u_G) are defined as the volumetric flow rate for the phase divided by the pipe cross sectional area

$$u_L = \frac{Q_L}{A} \text{ and } u_G = \frac{Q_G}{A} \qquad \ldots(3\text{-}3)$$

where Q_L and Q_G are the volumetric flow rate of liquid and gas respectively and A is the pipe cross sectional area.*[Chen, 2001]*

3.9 Gas Holdup Theory

Gas holdup, or voidage, is a dimensionless quantity that represents the percentage of total gas-liquid system is occupied by the gas. And is defined as

$$\varepsilon_G = \frac{V_G}{V_G + V_L} \qquad \ldots(3\text{-}4)$$

where $V_{G,}$ V_L are the gas and liquid volume respectively. When making calculation of gas holdup based on measurements of the bubble column in the laboratory, the above equation simplifies to

$$\varepsilon_G = \frac{h_i - h}{h_i} \qquad \ldots(3\text{-}5)$$

where h_i = total height of liquid and gas, and h= height of standing water. *[Martis, 2004]*.

Gas holdup can be measured by:

1. Directly by bed expansion
2. In directly by manometer method
3. Attenuation of γ or X- ray

Gas holdup is important property for design purpose, because of its indirect influence on column size and because it is indirectly related to the gas-liquid surface area and hence to the mass transfer *[Hills, 1976]*.

The gas holdup depends mainly on the superficial gas velocity of the form

$$\varepsilon_G \ \alpha \ u_G^n \qquad \ldots(3\text{-}6)$$

The values of n depend on the flow regime *[shah, 1982]*. For bubbly flow regime varies from 0.7 to 1.2 *[Rieth et al., 1968; Miyachi and Shyu, 1970; Zlokarnik, 1971; Hammer and Rashe, 1973; Schugerl et al., 1977; Batch and Pilhofer, 1978; Botton et al.1978; Deckwer et.al.1980.a]*

For churn turbulent or transition regime, the effect of the superficial gas velocity is less pronounced and the exponent n takes values from 0.4 to 0.7 *[Shah, 1982]*. Figure (3.7) shows that in homogeneous bubbly flow, where the gas holdup increases linearly when gas velocity increased.

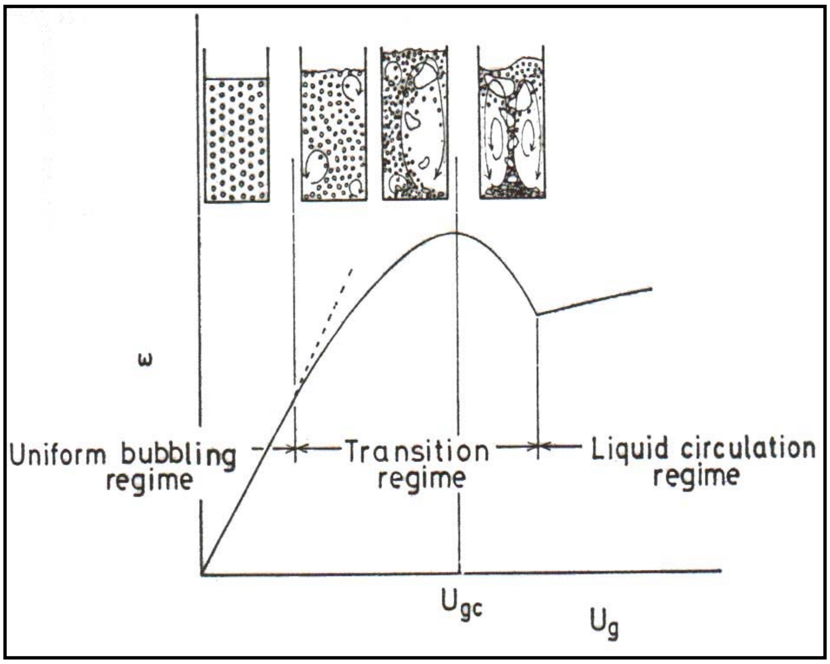


Figure (3.7) Qualitative sketch of observed flow pattern and gas holdup as a function of superficial gas velocity

During the flow transition the rate of increase gas holdup with gas flow velocity is much lower than in the bubbly flow region. The lowest superficial gas velocity of transition regime corresponds to the point where over of all gas holdup breaks away from ideal curve for uniform bubbling.

Near maximum the gas holdup, gross liquid circulation current sweeps the entire volume and the downward motion of gross liquid movement near the side wall affect the steady bubble formation at the base.

At maximum gas holdup, a symmetrical two-loops circulation upward in the middle and downward near the side walls appears for the first time among the violent and frequent interactions of large bubble and asymmetrical circulation *[Molina, 1999] [Maruyma, 1981].*

The numerous variables that can affect the gas holdup of a system including viscosity of both liquid and gas, the exact verticality of the column, types of sparger or scintered plate used to introduce the bubbles, and impurities in the liquid and gas *[Geary, 1991]*.

3.10 The Drift Flux Model

In vertical two-phase flow, the average gas void fraction in the pipe cannot be simply equated to the volume flowing gas fraction $u_G/(u_G+u_L)$, for two reasons. First, the bubbles rise in the aid due to the gravitational forces, so that the local gas velocity of the gas and liquid phase differs. Second if the gas void is distributed in the pipe such that a disproportionate amount of gas is in region of flow with a greater or lesser velocity than the average velocity in the pipe, then the average gas velocity will be further modified. Typically both gas void distributor and velocity distribution have coincident maximum at the pipe center so that, on average, the gas traveling faster than the liquid. Taking both of the factors into account, *Zuber and Findlay (1964, 1965)* present a drift-flux model so that has been accepted widely for the prediction of gas holdup and applied to various two-phase flow regimes in up flow regimes in up flow *[Nassos and Bankoff, 1967; Goveir and Aziz; Anderson and Hau, 1980]*.

$$u_G/\varepsilon_G = c_o(u_G+u_L)+(J\varepsilon_G)/\varepsilon_G \qquad \ldots(3\text{-}7)$$

If the value of $(J\varepsilon_G)/\varepsilon_G$ is constant or very small compared to the value of (u_G+u_L), by plotting a graph of (u_G/ε_G) against (u_G+u_L), the value of c_o

can be obtained. A change in the slope of the graph will indicate the bubbly flow regime has ended.

3.11 Transition Velocity

The flow regime transition velocity in a gas-liquid bubble column system can be found graphically by plotting the gas holdup against the drift flux velocity of the system. With the system in churn turbulent phase, the relationship is not linear, but is approximated that way to find a linear best fit line of data assumed to be heterogeneous. As the two regimes produce "linear" relationship with different slopes, the intercept of the two lines represents the transition velocity for the system *[Wild, 2003].*

The method of determining the transition velocity produces a good approximation; the actual nonlinearity of the churn-turbulent regime makes it difficult to produce a best-fit line close enough to the data. In these cases, the gas hold-up is graphed against the drift-flux of the system, pronounced linearity with sharper slop changes results.

The drift flux is a velocity in (m/s) that relates the velocity of the gas to the velocity of the liquid components of the system.

$$j = \varepsilon_G (1-\varepsilon_G)\left(\frac{u_G}{\varepsilon_G} + \frac{u_L}{(1-\varepsilon_G)}\right) \qquad \ldots(3\text{-}8)$$

where j = drift flux and u_G, u_L = velocity of gas and liquid, respectively. In the case of stationary liquid phase (u_L= 0) this equation simplifies to

$$j = u_G(1 - \varepsilon_G) \qquad \ldots(3\text{-}9)$$

3.12 Mass Transfer

Mass transfer is the term used to denote the transfer of a component in a mixture from region in which its concentration is high to a region where the concentration is lower. The process can take place in gas or vapour or in liquid *[Coulson and Richardson, 1996].*

In the absence of chemical reaction, gas-liquid mass transfer to a low solubility gas bubble is controlled by molecular diffusion in the liquid phase. The mass transfer coefficient k_L depends on the bubble size. Firstly, because in other wise fixed condition k_L is found as a function of bubble diameter [Higbie, 1953; Frosslirg, 1983], and secondly, because bubble currature further decides the interface mobility. Large bubbles behave usually the fluid particles with mobile surface, while smaller bubbles tend to behave like solid spheres *[Clif et al., 1978].*

$$k_L a = \frac{-1.303(1 - \varepsilon_G)}{t} log \frac{C_i - C}{C_i - C_o} \qquad \ldots(3\text{-}10)$$

The Oxygen concentration measured by Oxygen-meter at 18 C^o

CHAPTER FOUR

DESIGN OF EXPERIMENT

4.1 Experimental

The experimental apparatus used in this work is shown in figure (4.1),

Fig. (4.1) The Complete Experimental Apparatus

The bubble column made of transparent plexiglass of 0.081m in diameter and 1.03 m height. Porous distributor was used as gas distributor. Tap water at room temperature was mainly used as liquid phase in the experiments, but results were also obtained for many concentrations of *NaCl* which are (0.24, 0.30, 0.36, 0.42, 0.6, 0.8) M *NaCl*.

Gas was taken from the air compressor; the volumetric flow rate was measure using two separate rotameters in order to reach wide enough range of gas velocities entering the column.

As shown in figure (4.1), the base that holds the column has a regulator screws to ensure exact verticality of the column with effective liquid height of 0.7 m. Runs were performed with different superficial gas velocities with the range from 0.009657 m/s to 0.131521 m/s. The systems that used in this work are air-water, air-*NaCl* solution of different concentrations, and CO_2-water.

4.2 Experimental Procedure

4.2.1 Measuring gas holdup

1. The column was filled with water; the liquid height was 0.7 m initially.
2. Gas was taken from compressor; the desired flow rate was measured by rotameter.
3. The overall gas liquid level measured to determine the gas holdup (foam, if present, was not included).
4. The above procedure was repeated with a new gas velocity for the three systems: air-water, CO_2-water, air-*NaCl* solutions of six different concentrations.

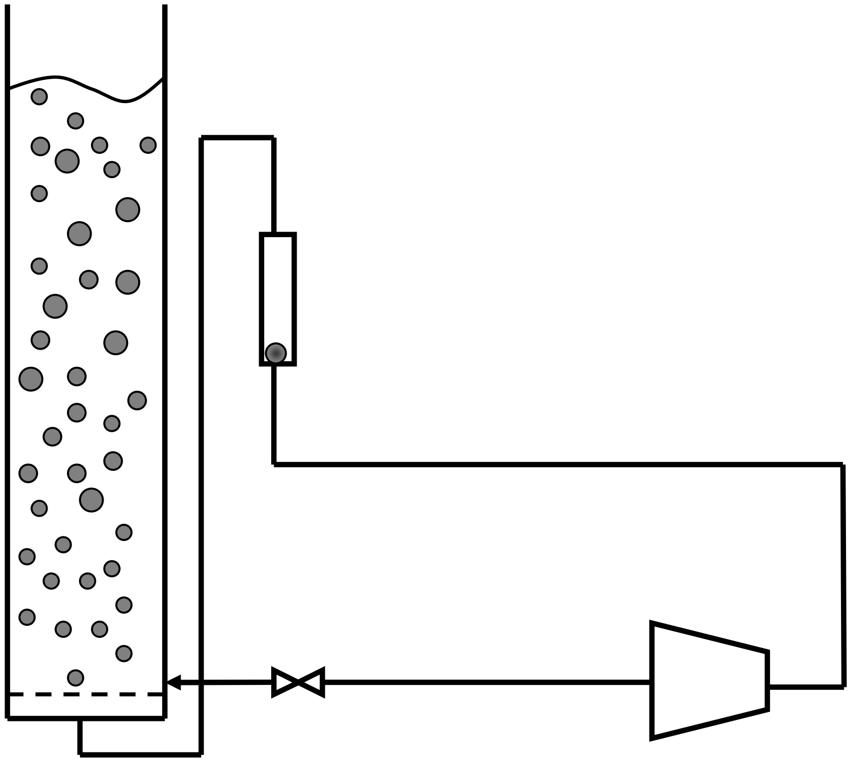


Fig. (4.2) Schematic Diagram of Experimental Apparatus.

(1) Bubble Column	(3) Rotameter
(2) Gas distributor	(4) Air Compressor

4.2.2 k_La Measurements

1. The column filled with 0.7 m height.
2. Gas was compressed from air compressor at a limited flow rate.
3. The concentration of the gassed water was measured every 7 min using dissolved oxygen meter by taking a sample of the gassed liquid using oxygen meter device.
4. The above procedure was repeated for another velocity.

The Experimental Devices

Dissolved Oxygen Meter

The dissolved oxygen concentration in water was measure using oxygen meter ORION 810 Aplus. This device calculate the concentration on the basis of known relationship between O_2 solubility, temperature, and acting atmospheric pressure.

This instrument measure and compensate automatically. The probe consists of an anode/cathode electrode system and non electrolyte separated from the environment by oxygen permeable membrane.

Air Compressor

Compressed air was supplied by means of an Angersoll-rand type compressor, which is a reciprocating compressor, containing a clearance unloaders consist of pockets or small resvoirs, which are opened when unloading is desired.

The desired air flow rate was setup using needle valve and the amount was measured with a gas meter. Also a gas rotameter was used to measure the air flow rate.

4.3.3 Gas Distributor

A porose distributor of cylindrical shape with 14.894% void fraction, 0.03 m diameter was used as gas distributor, this type of distributor prevents feed back .

4.3.4 Rotameters

Two calibrated rotameters were used in order to cover wide range of gas velocities which was from 0.009657 m/s to 0.131521 m/s.

Chapter Five

RESULTS AND DISCUSSION

5.1 Dynamic Simulation of Chemical Process as a Tool to design "The Real Problem" of Identification

Simulation is the construction and use of a computer-based representation, or model, of some part of the real world as a substitute vehicle for experiment and behavior prediction in an environment of change. This works examines the practical use of simulation, within experimental data for the bubble column application areas. The discussion draws on a study of the employment of simulation, which serves as an authentic pilot plant, to design a bubble column. Concluding the several gas velocities are carried out in order to identify the system.

5.2 Computer Simulation of Bubble Column

Rapid progress in three influencing technologies over the past two decades has brought computational fluid dynamic to the forefront of process engineering *[Haidari and Matthews, 2003]*.

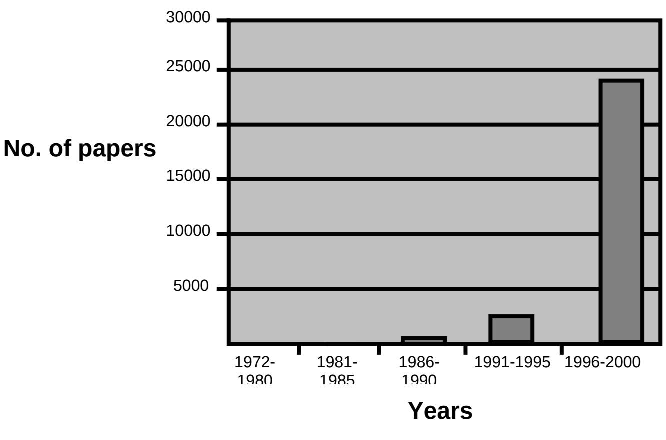


Fig. (5.1) CFD papers 1972-2000

ANSYS Inc., founded in 1970 as Swanson Analysis Systems Inc., develops and globally markets engi-neering simulation software and technologies widely used by engineers and designers across a broad spectrum of industries. The Company focuses on the development of open and flexible solutions that enable users to analyze designs directly on the desktop, providing a common platform for fast, efficient and cost-conscious product development, from design concept to final-stage testing and validation.

ANSYS provides unmatched technology, useablility and performance for engineering simulation. The complete ANSYS suite covers time-tested, industry-leading applications for structural, thermal, mechanical, computational fluid dynamics, and electromagnetic analyses, as well as solutions for transient impact analysis.

CFX is a leading supplier of computational fluid dynamics (CFD) software and services to the chemi-cal, manufacturing and power generation industries.

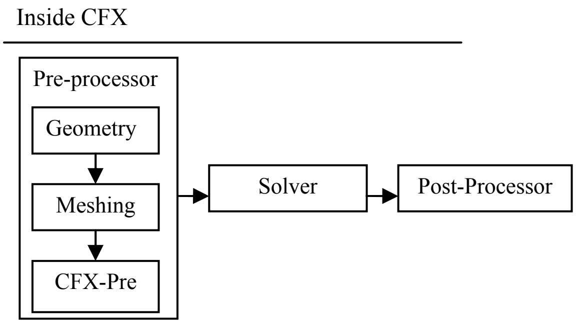


Fig. (5.2) Stages Inside CFX

In this chapter we used 3-D CFX demonstrates the Eulerian-Eulerian multiphase model to evaluate the gas holdup. The experiments were performed in 0.081 m diameter and with air- water at atmospheric pressure. The following parameters were used: air volume fraction of 0.25, air inlet velocity close to the terminal velocity, water inlet velocity zero, and symmetry planes for the cross sections.

The following figures show contours of air volume fraction at different superficial gas velocity:

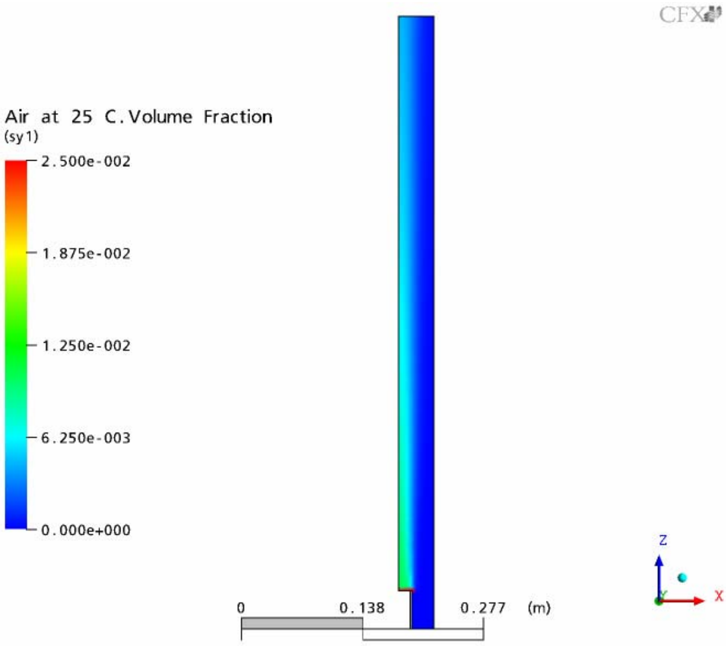


Fig. (5.3) Results showing contours of air volume fraction at 0.009657 m/s superficial gas velocity

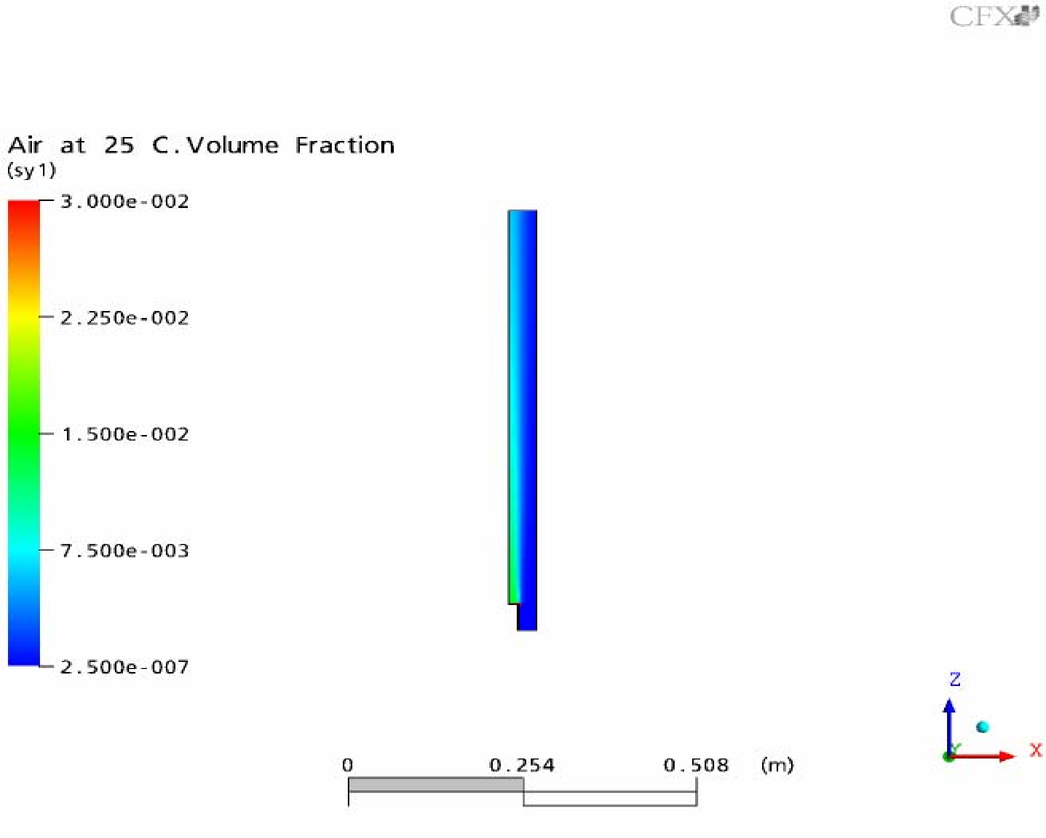


Fig. (5.4) Results showing contours of air volume fraction at 0.013114 m/s superficial gas velocity

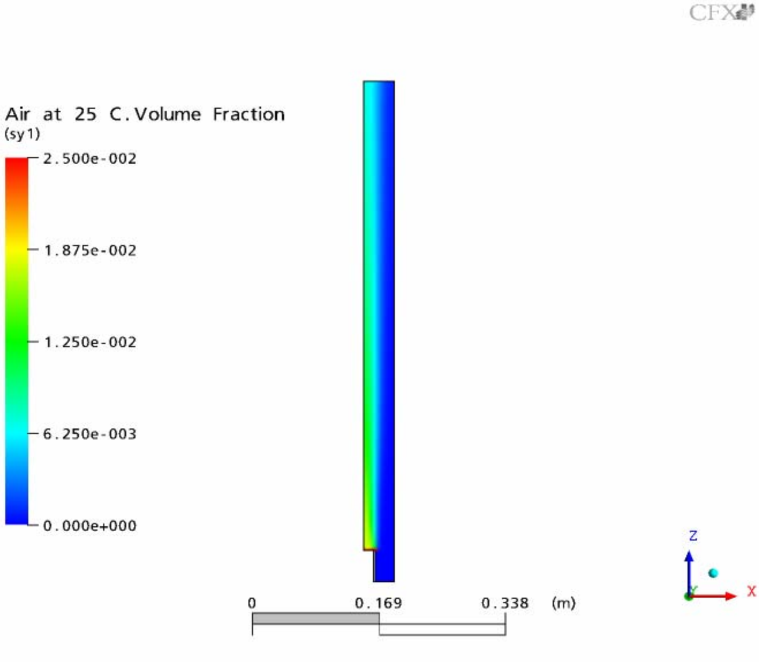


Fig. (5.5) Results showing contours of air volume fraction at 0.01657 m/s superficial gas velocity

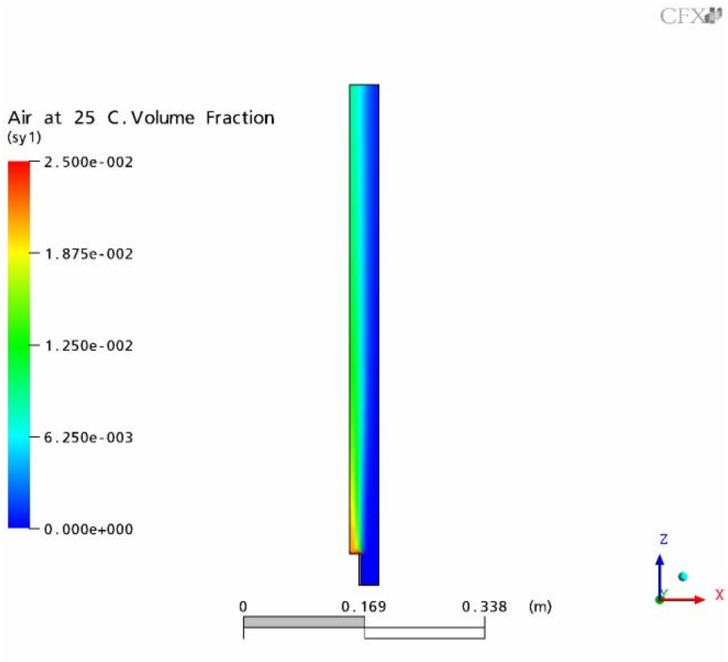


Fig. (5.6) Results showing contours of air volume fraction at 0.020026 m/s superficial gas velocity

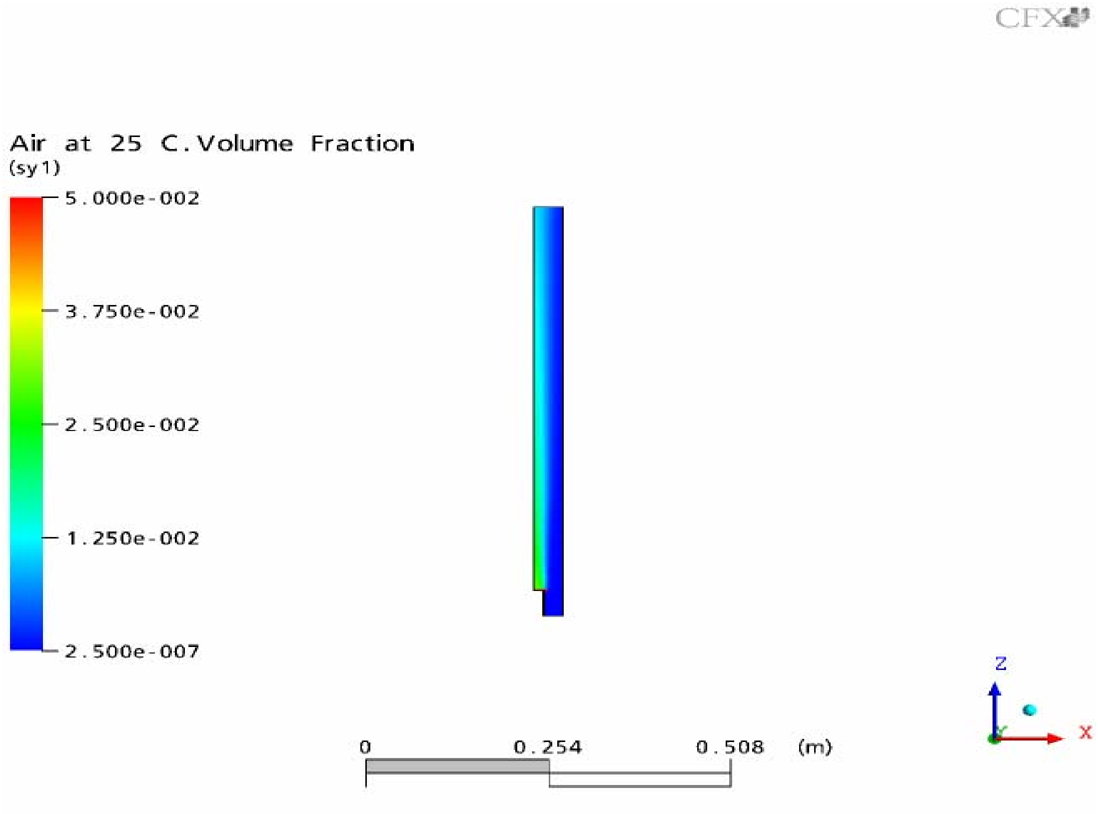


Fig. (5.7) Results showing contours of air volume fraction at 0.026939 m/s superficial gas velocity

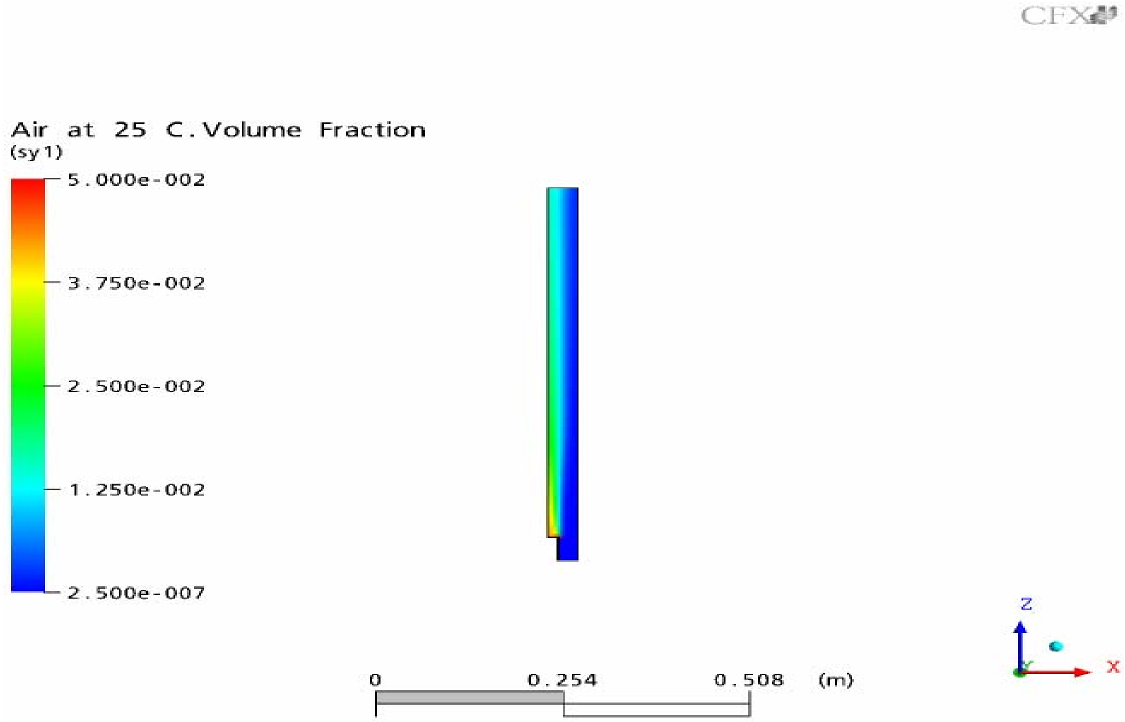


Fig. (5.8) Results showing contours of air volume fraction at 0.039355 m/s superficial gas velocity

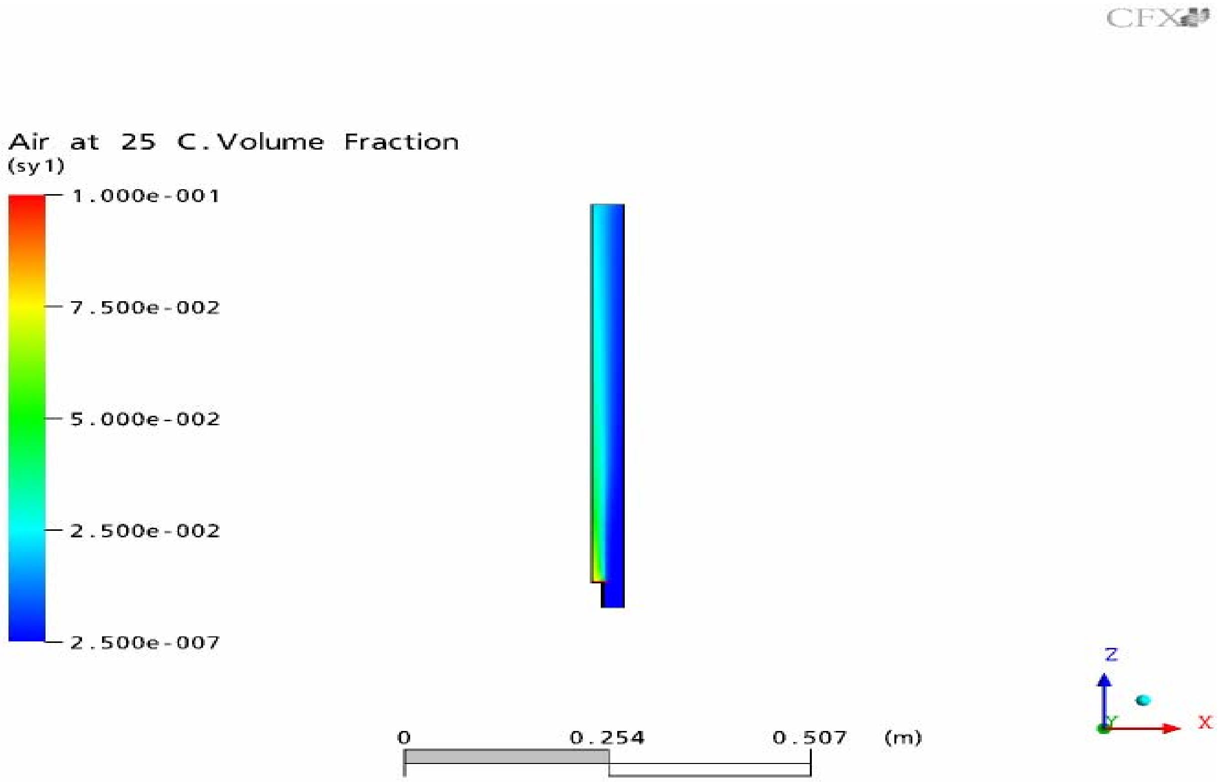


Fig. (5.9) Results showing contours of air volume fraction at 0.070077 m/s superficial gas velocity

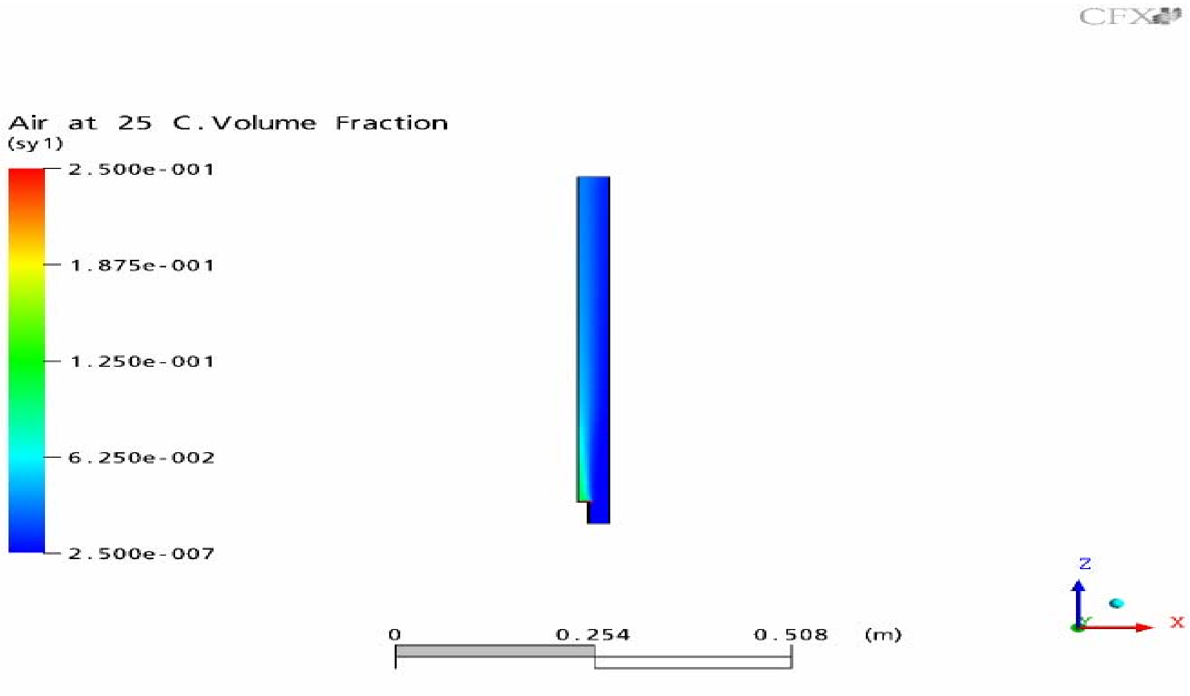


Fig. (5.10) Results showing contours of air volume fraction at 0.100799 m/s superficial gas velocity

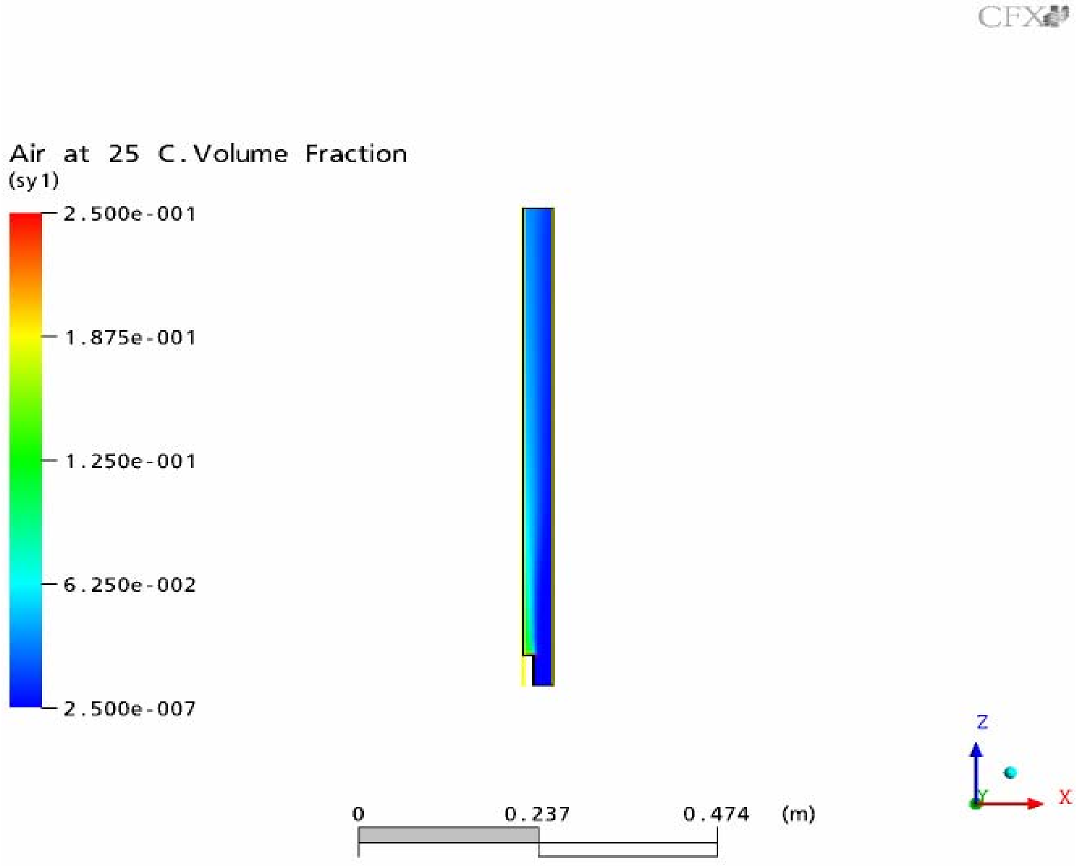


Fig. (5.11) Results showing contours of air volume fraction at 0.131521 m/s superficial gas velocity

5.3. Experimental Results

5.3.1 Gas Holdup

Three variables were used in semi-batch bubble column, which included superficial gas velocities, two types of gases, and the concentration of *NaCl*.

The effect of superficial gas velocity on gas holdup can be shown in figure (5.12) to (5.19). These figures indicated that the gas holdup increases with increasing superficial gas velocity. This is due to the fact that higher superficial gas velocity gives bubbles with lower rising velocity. This will lead to form large residence time and hence higher gas holdup.

Tap water with aqueous solution with different commercial *NaCl* concentration was used. As shown in figures (5.12) and (5.13) the addition of *NaCl* with (0.24, 0.3) M will increase the gas holdup as compared with that of water for superficial gas velocity below 0.03 m/s.

This is because of *NaCl* acts that prevent coalescence by the repulsive forces that acts and the bubbles unable to coalescence.

As shown in figures (5.12) to (5.14) the gas holdup curve of *NaCl* solution fall below that of water for superficial gas velocity above 0.03 m/s. At this velocity the sudden formation of spherical cap bubbles that could be observed caused this falling.

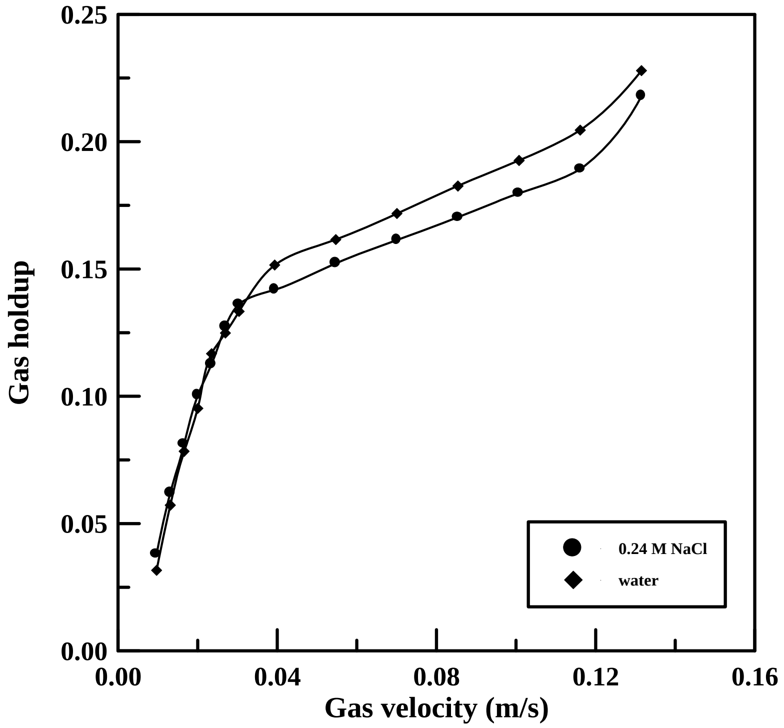


Fig. (5.12) Gas holdup vs. superficial gas velocity for water and 0.24 M *NaCl*

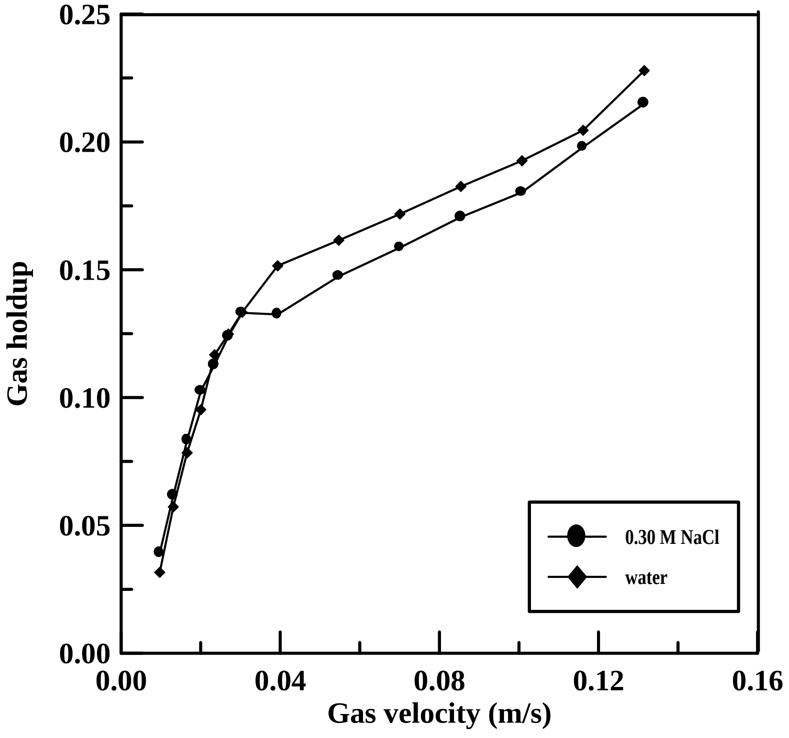


Fig.(5.13) Gas holdup vs. superficial gas velocity for water and 0.3 M *NaCl*

As shown in figures (5.14) and (5.15), for concentration of 0.36 and 0.42 M *NaCl* and superficial velocity below 0.03 m/s, the gas holdup decrease until the gas holdup curve of 0.42 M take the same behavior of water curve.

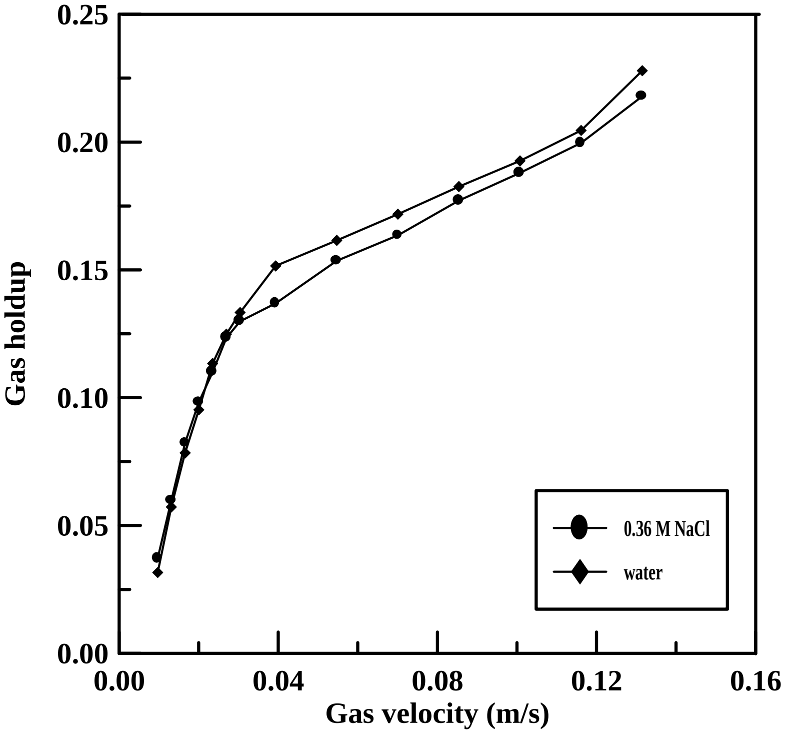


Fig. (5.14) Gas holdup vs. superficial gas velocity of water and 0.36 M *NaCl*

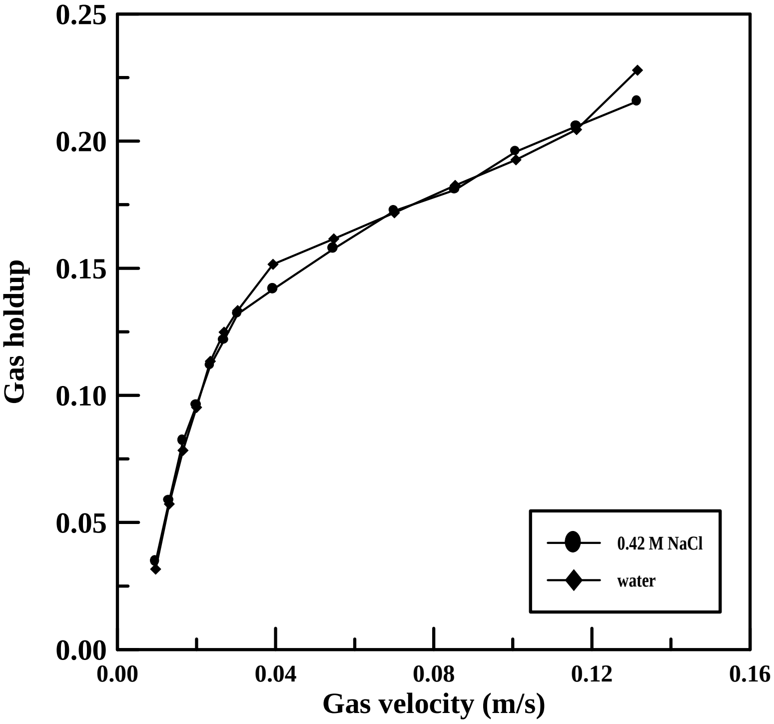


Fig. (5.15) Gas holdup vs. superficial gas velocity for water and 0.42 M *NaCl*

As show in figures (5.16) and (5.17) for high *NaCl* concentration and for superficial gas velocity below 0.06 m/s, the gas holdup curve of *NaCl* solution fall below that of water, this is due to increasing *NaCl* ions which are (Na^+, Cl^{-1}) in the solution that will overcome the water ions so that the coalescence property will stop causing the formation of large bubbles and then decreasing the gas holdup.

When the gas velocity increases above 0.06 m/s these large bubbles will break-up into small bubbles this caused increasing the gas holdup. These results are in good agreement with *Jamal Ahmadi* and *Stienhagen, 1990.*

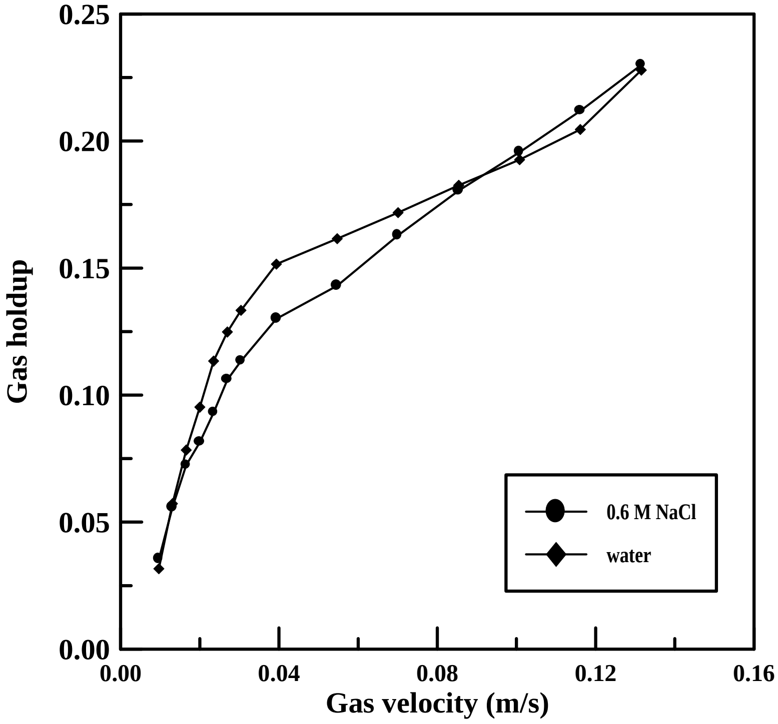


Fig. (5.16) Gas holdup vs. superficial gas velocity for water and 0.6 M *NaCl*

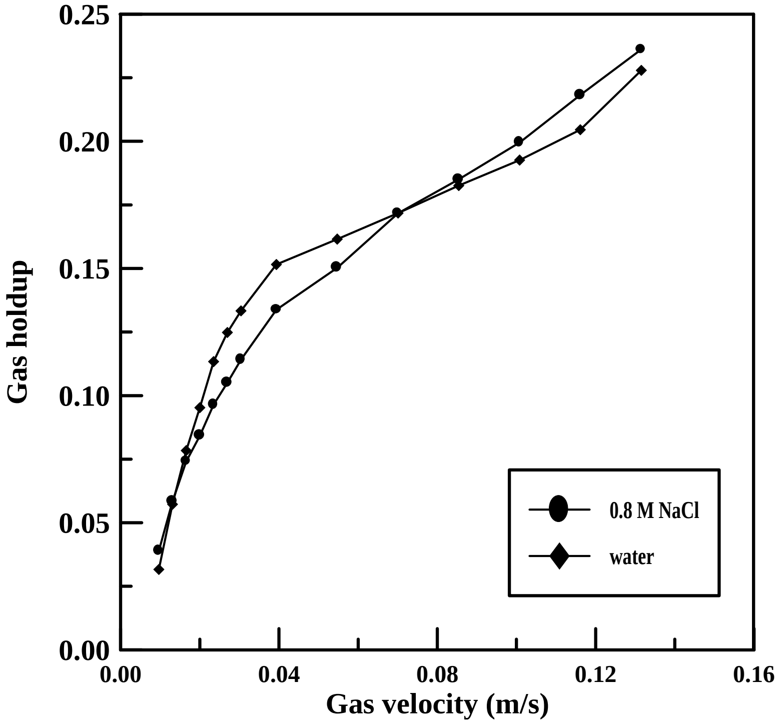


Fig. (5.17) Gas holdup vs. superficial gas velocity for water and 0.8 M *NaCl*

As shown in figure (5.18) computational results have been compared to the experimental data; the theoretical results were multiplied by factor 10 because of the difference in the distributor porosity between the simulation and the experimental work.

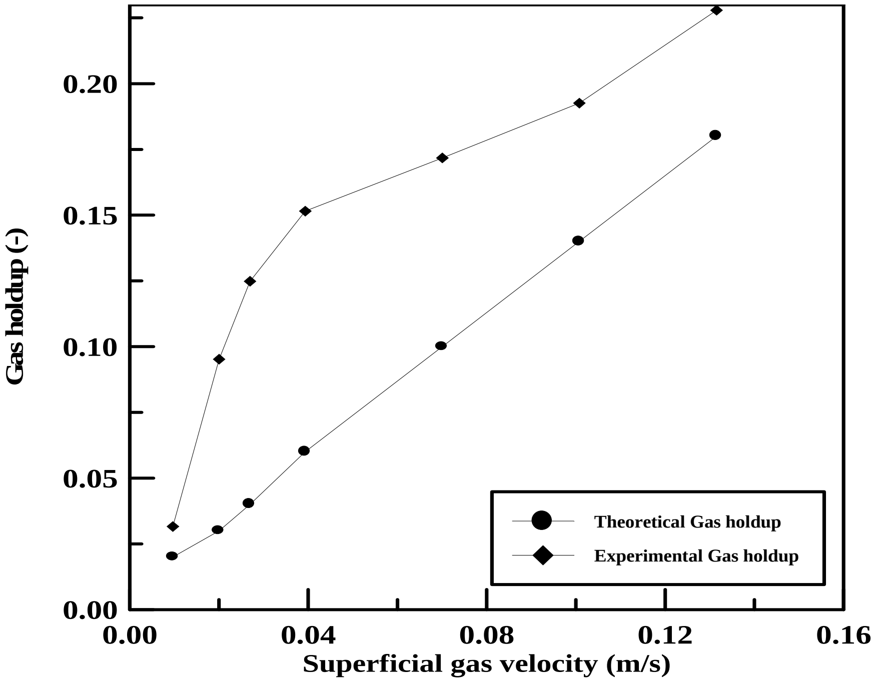


Fig.(5.18) Comparison between theoretical and experimental work

As shown in figure (5.19), the CO_2 was used as gas phase with water instead of air; the results were slightly differing from that of water. This is due the fact that the effects of gas density have little effect on gas holdup.

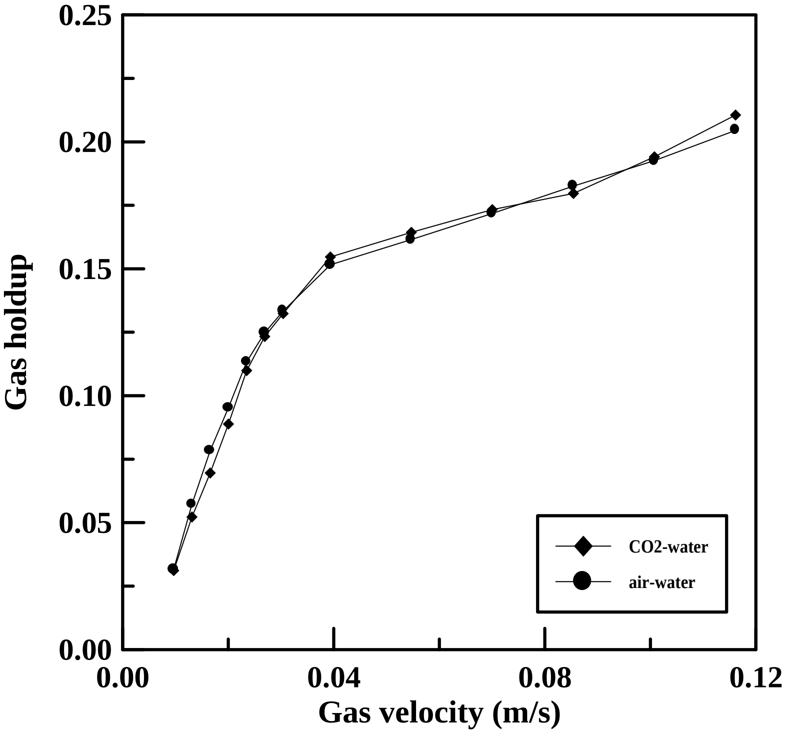


Fig. (5.19) Gas holdup vs. superficial gas velocity for (*CO_2*-water) system compared with (air-water) system.

5.3.2 Transition Velocity

As shown in figure (5.20) the intersection of the two lines represents transition velocity from bubbly to churn turbulent flow for (air-water) system, it was about 0.027 m/s. these results are in good agreement with the observation of *Martis, 2004*.

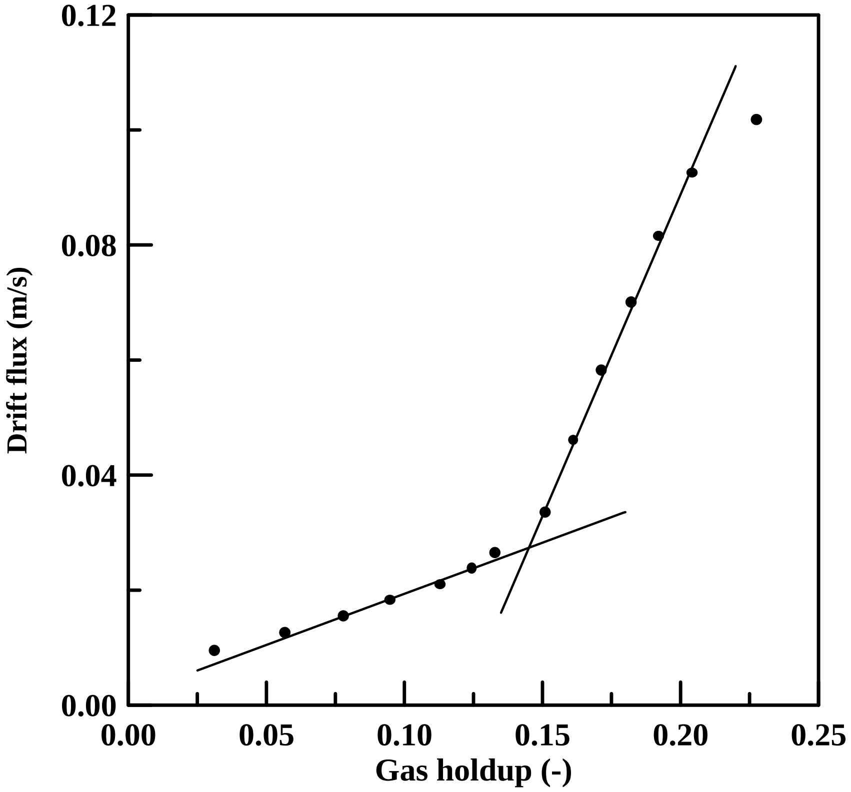


Fig. (5.20) Drift flux vs. gas holdup for (air-water) system

As shown in figures (5.21) to (5.26), the transition velocity for air with the different concentrations of *NaCl*, the transition velocity values was about, 0.022, 0.021, 0.023, 0.025, 0.028, and 0.029 m/s respectively.

These results show that the transition velocity from bubbly region to turbulent region increases as the *NaCl* concentration increase. This is because of the effect of salts that preventing coalescence between the bubbles. This will lead to increase the laminar region range and hence increase the transition velocity to the turbulent region.

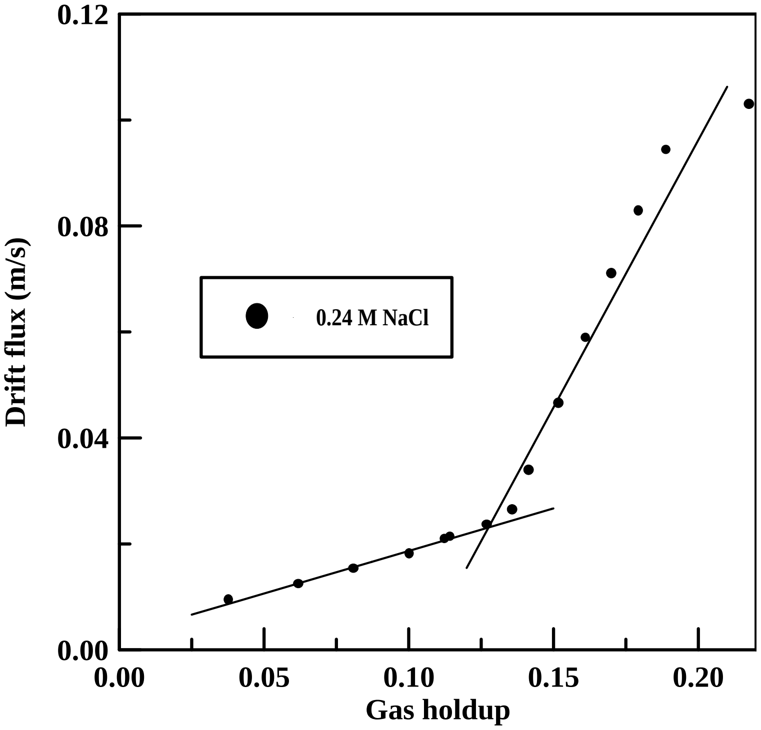


Fig. (5.21) Gas holdup vs. Drift flux for (air- 0.24 *NaCl*)

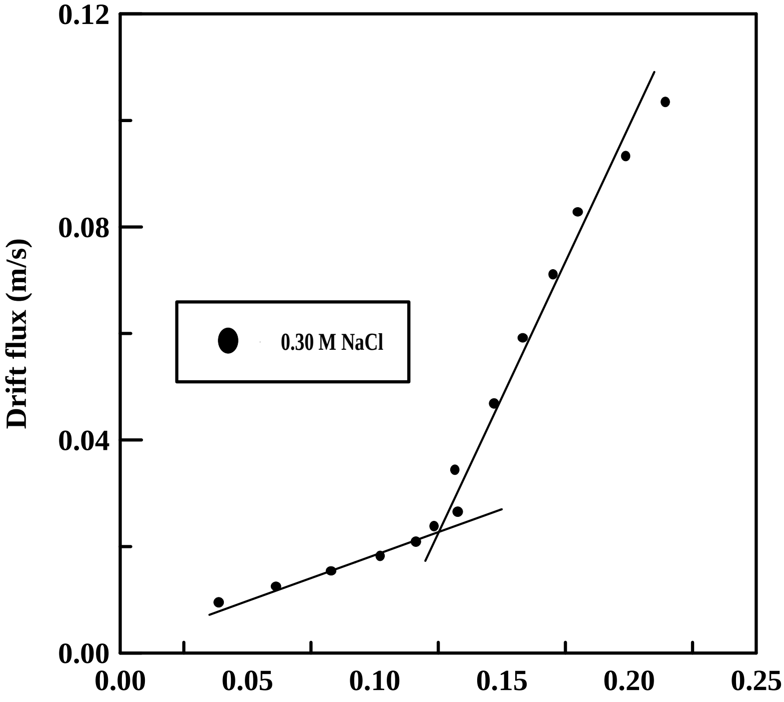


Fig. (5.22) Gas holdup vs. Drift flux for (air- 0.30*NaCl*)

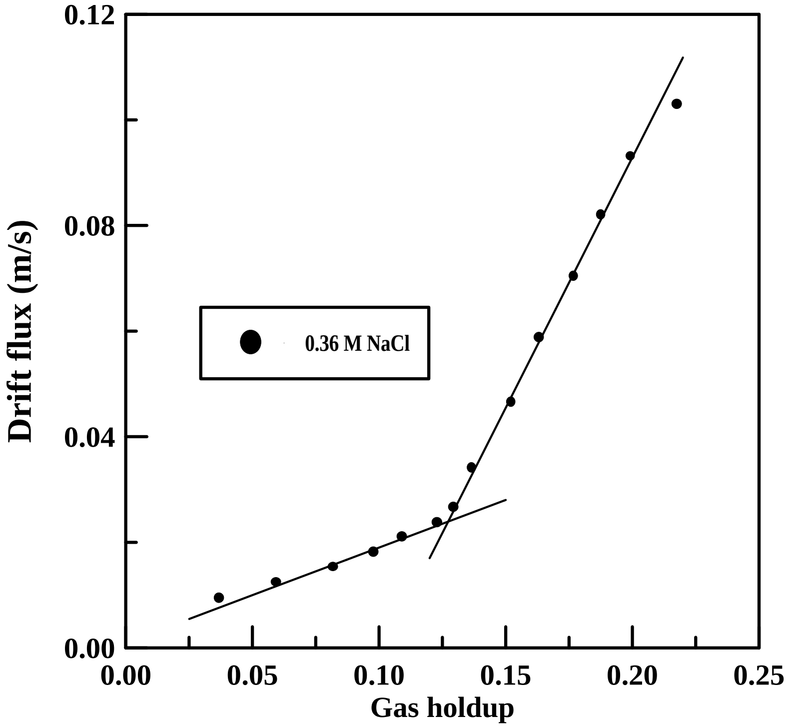


Fig. (5.23) Gas holdup vs. Drift flux for (air- 0.36 *NaCl*)

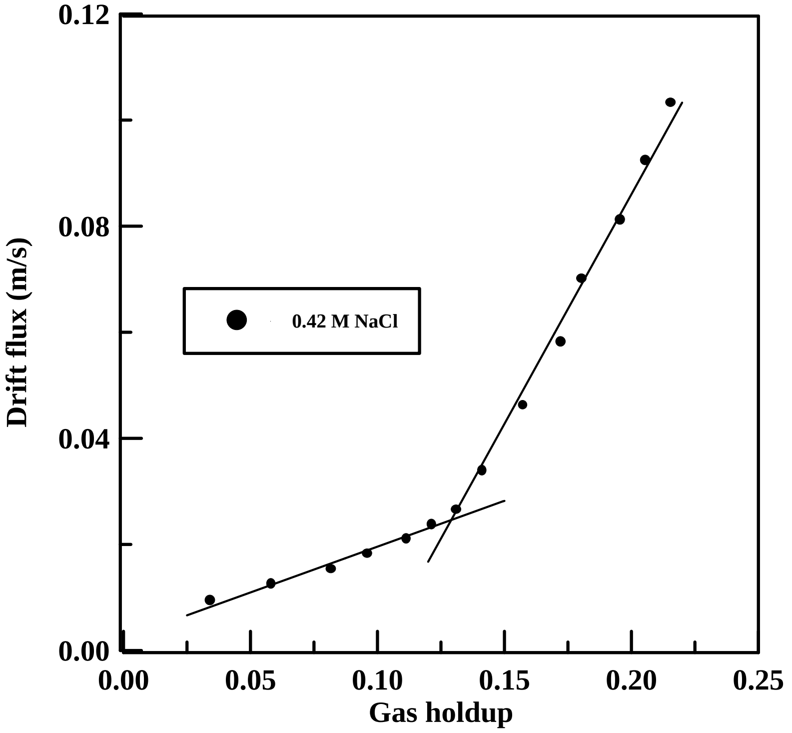


Fig. (5.24) Gas holdup vs. Drift flux for (air- 0.42 *NaCl*)

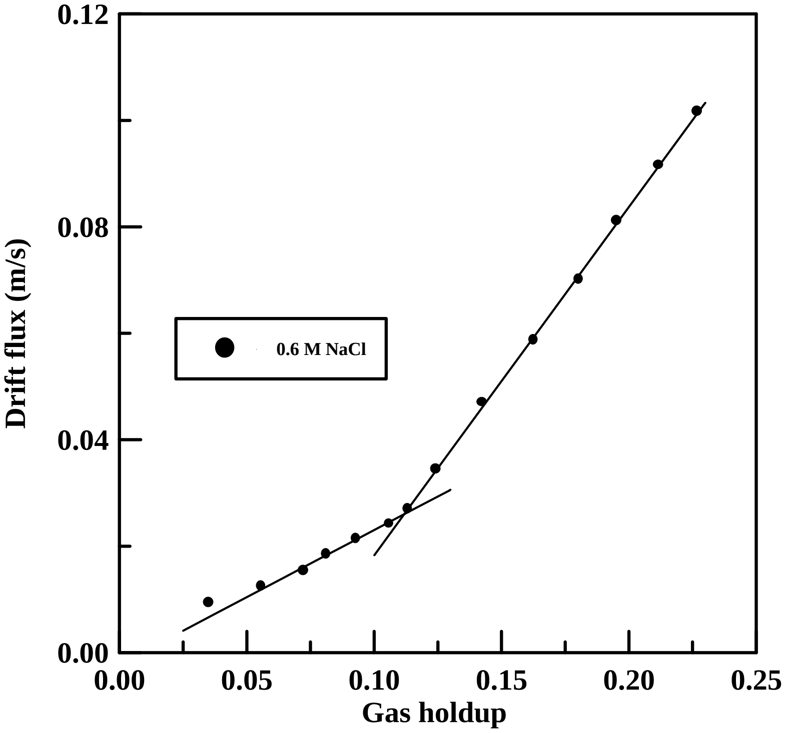


Fig. (5.25) Gas holdup vs. Drift flux for (air- 0.6 *NaCl*)

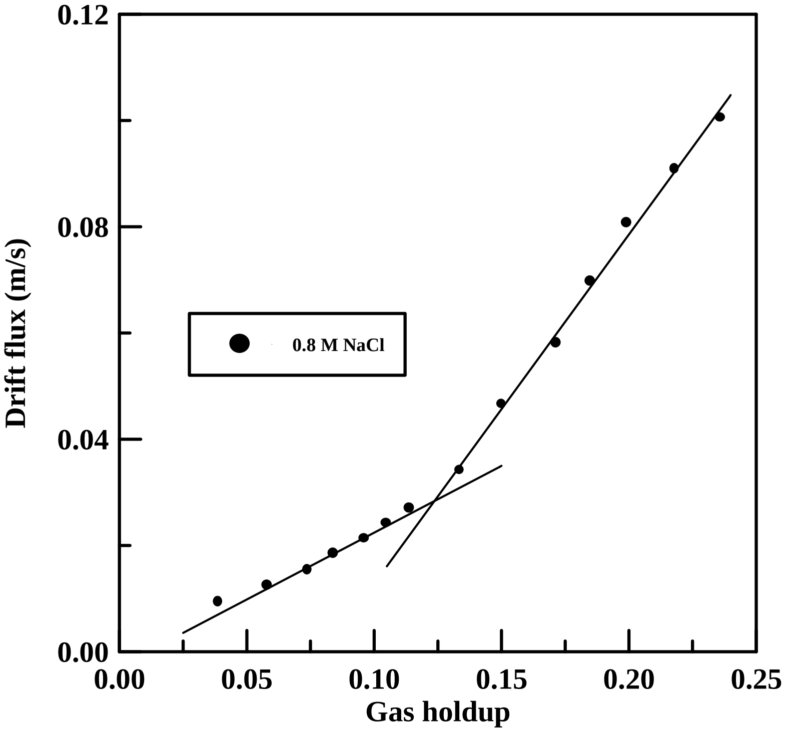


Fig. (5.26) Gas holdup vs. Drift flux for (air- 0.8 *NaCl*)

As shown in figure (5.27), the transition velocity for (*CO2*- water) **s**ystem was about 0.029 m/s.

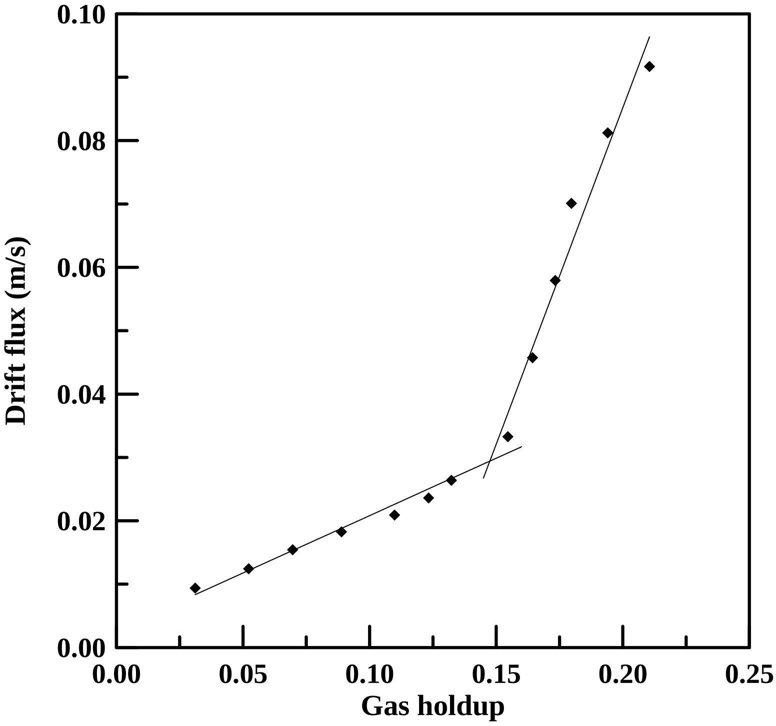


Fig. (5.27) Gas holdup vs. Drift flux for (*CO2*- water) system

5.3.3 Flow Regime Identification Using the Drift Flux Plot

The flow regime identified using the drift flux plot, as shown in figures (5.28) to (5.31) when the (superficial gas velocity/gas holdup) m/s is plotted against the superficial gas velocity m/s, the change in the slope of the curve indicate the transition from homogeneous region to the heterogeneous region. These results were in good agreement with *Dr.Ted Heindel*

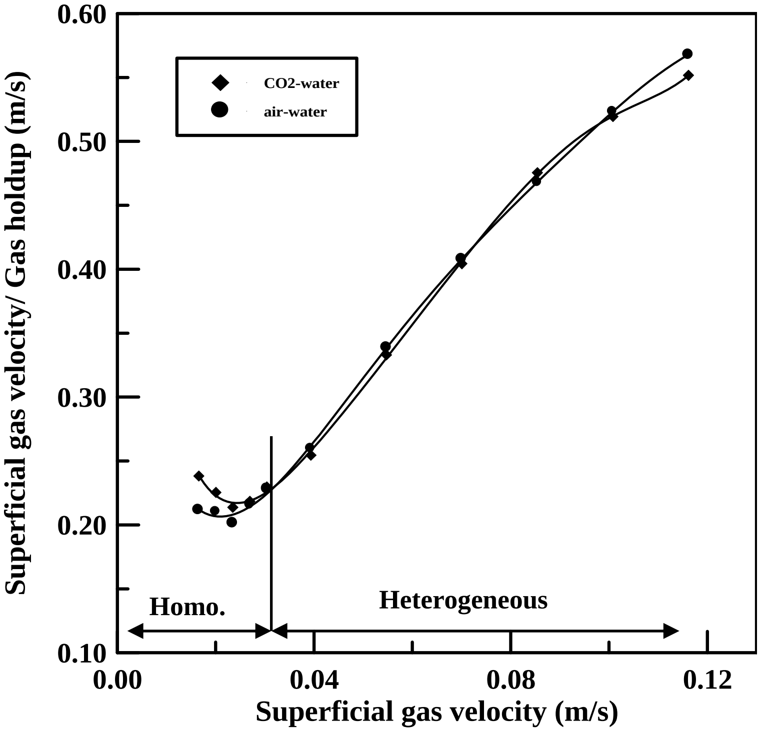


Fig. (5.28) Flow Regime Identification for Air-water, CO_2-water Systems

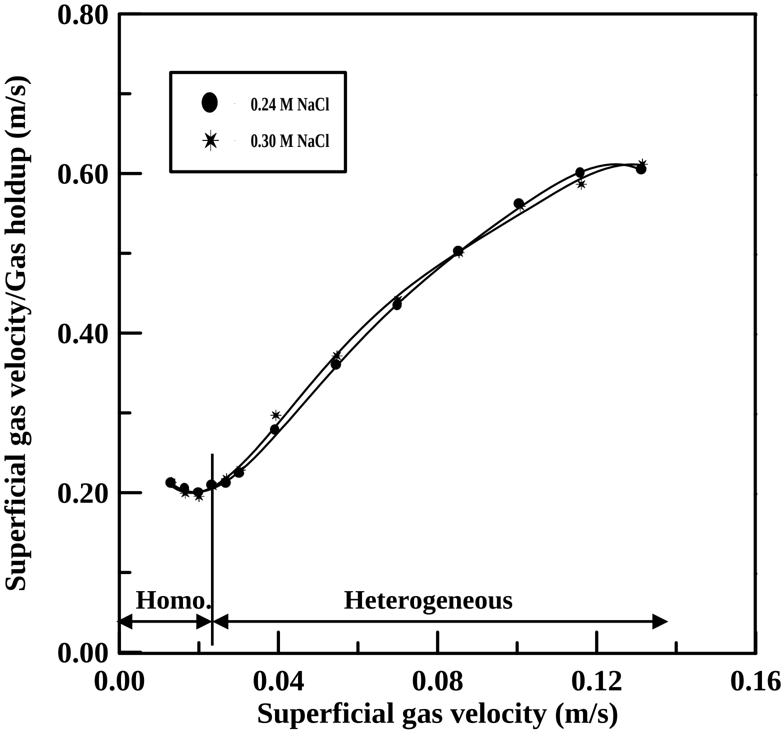


Fig. (5.29) Flow Regime Identification for 0.24, 0.30 M *NaCl*

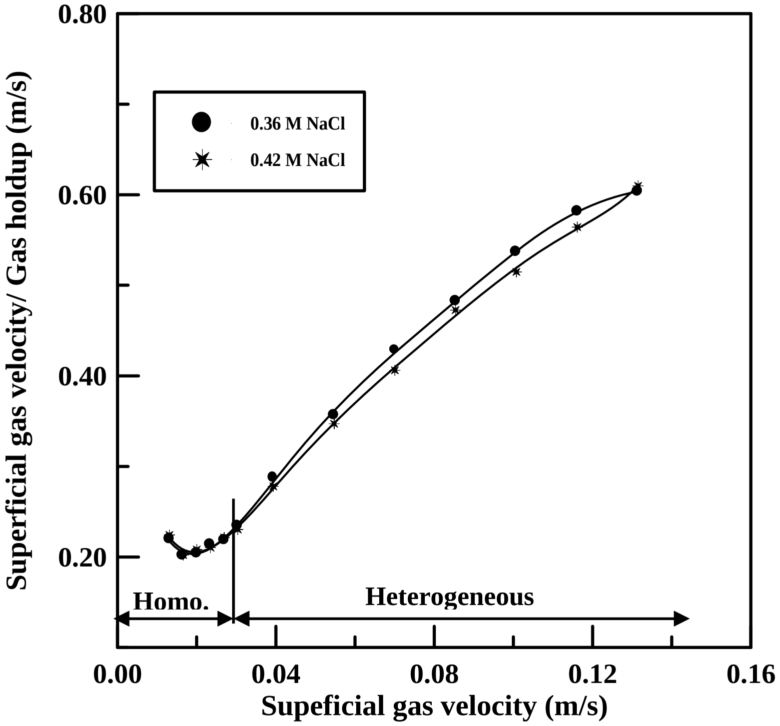


Fig. (5.30) Flow Regime Identification for 0.36, 0.42 M *NaCl*

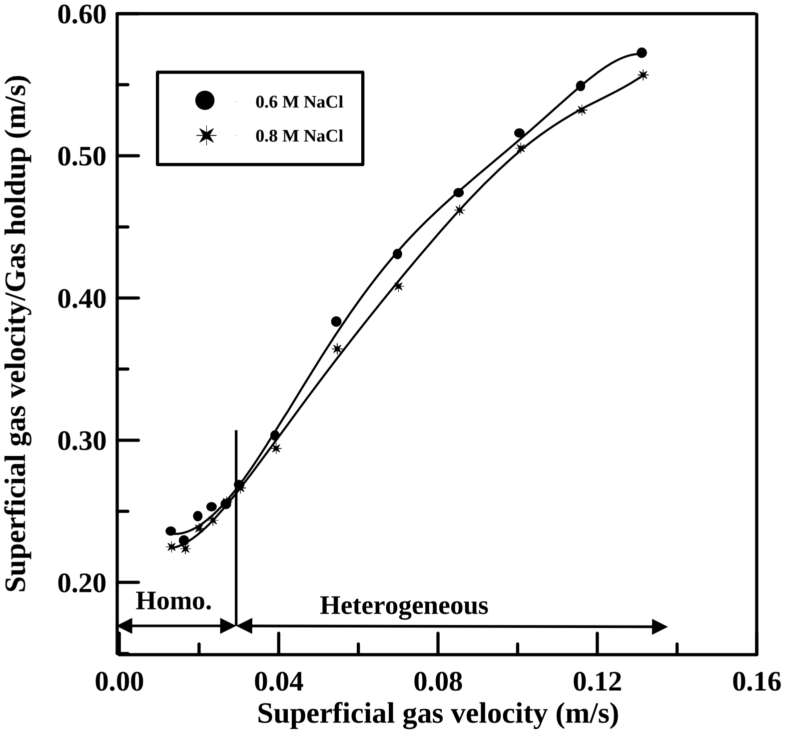


Fig. (5.31) Flow Regime Identification for 0.6, 0.8 M *NaCl*

5.4 Mass Transfer Coefficient

As shown in figure (5.32), the mass transfer coefficient increases as the superficial gas velocity increase. This is because of increasing the superficial gas velocity will increase the turbulence at the interface and hence increasing the mass transfer coefficient. Increasing the superficial gas velocity will also lead to decrease the thickness of the diffusion layer at the liquid side, so that resistance to mass transfer will decrease thus increasing the mass transfer coefficient.

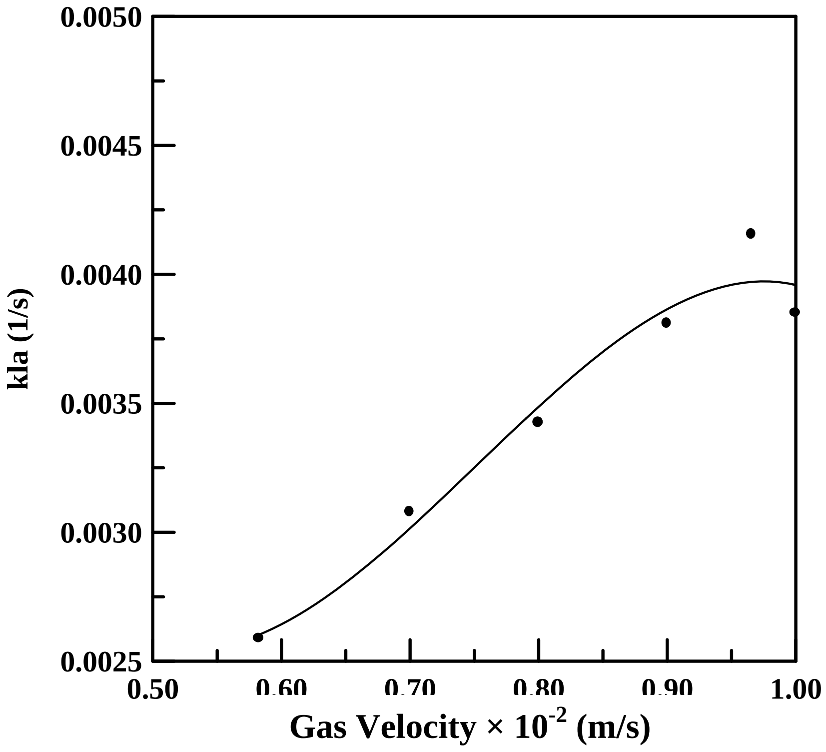


Fig. (5.32) Mass transfer coefficient vs. Gas velocity for (air-water) system

As figure (5.33) the relation between the gas holdup and the mass transfer coefficient is linear, where the mass transfer coefficient increase as the gas holdup increases. This is due to increasing the amount of air (which contains oxygen) in the two-phase system, this will lead to oxygen transfer from gas phase to liquid phase because of the difference in the oxygen concentration between the two phases.

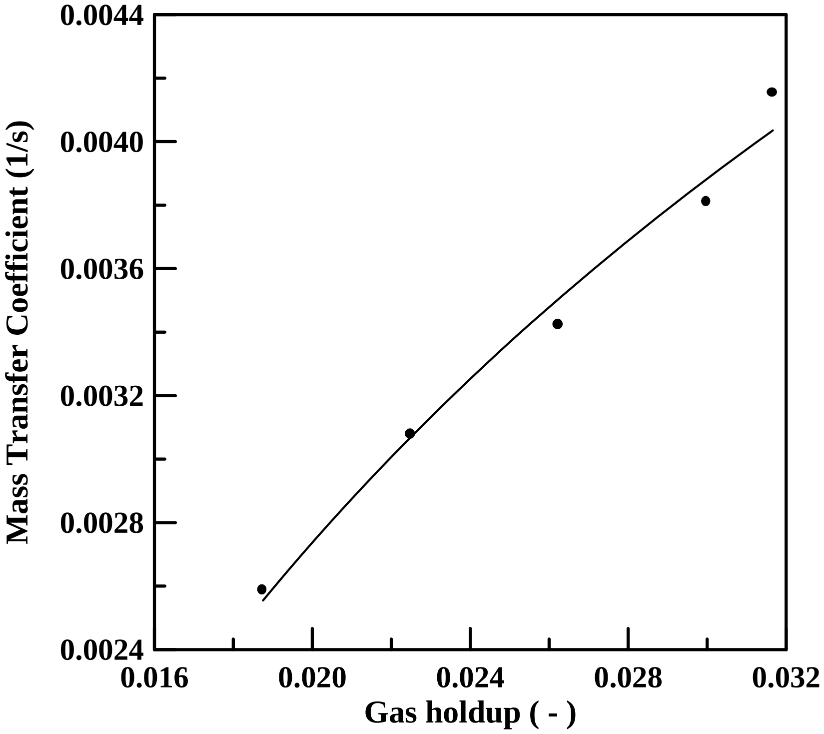


Fig. (5.33) Mass Transfer Coefficient vs. Gas holdup for (Air- water) system

CHAPTER SIX

CONCLUSION AND RECOMMENDATION FOR FUTURE WORK

6.1 Conclusions

1. Increasing the superficial gas velocity leads to increase in gas holdup for the three systems and the mass transfer coefficient.
2. At the critical concentration of *NaCl* solution, which is *0.42 M*, the gas holdup curve takes the same behavior of water gas holdup curve.
3. The addition of *NaCl* concentration below the critical concentration will increase the gas holdup for superficial gas velocity below 0.03 m/s compared to that of water while for higher velocity the gas holdup curve falls below of water with slightly increasing in the gas holdup as the *NaCl* concentration increase.
4. The addition of *NaCl* with concentration above the critical concentration and for superficial velocity below 0.06 m/s will decrease the gas holdup compared to that of water. When the superficial gas velocity increased above 0.06 m/s the gas holdup will increase.
5. The transition velocity increase with increasing the concentration of *NaCl.*
6. The flow regime identified using the drift flux plot.
7. Computational results have been compared to experimental data. The theoretical results were multiplied by factor 10 because of the difference in the distributor porosity between the simulation and the experimental work.

6.2 Recommendations for Future Work

From the present study, it is noticed that further studies in the following areas would be desired.

1. Using 3D fluid dynamic using other liquids such as alcohol.
2. The effect of adding solids on gas holdup, transition velocity, and mass transfer coefficient.
3. The effect of other type of electrolytes to make a comparison between these types and their influence on gas holdup and flow transition.
4. This study concerned only with measuring the mass transfer coefficient only with air-water system. Therefore studies to cover the mass transfer using the electrolytes system.
5. A modified method for measuring gas holdup is necessary such as gamma-ray absorbtion, pressure measurements, gas holdup measurements by optical probe, and electrical resitance tomography.

REFRENCES

1. Alvarez- Cuewca, M., and Nerenberg, M. A.,"O_2 Mass Transfer in Bubble Column Working at Large Gas and Liquid Flow Rates", *AIChE J.*, vol. 27, No.1, pp. (66), 1981.
2. Akita, K., Youshida, F., "Gas Holdup and Volumetric Mass Transfer Coefficient in Bubble Column", *Ind. Eng. Chem. Process Des. Dev.* ,vol. 12, No. 1, 1973.
3. Akita, K. and Youshida, F., "Bubble Size, Interfacial Area, and Liquid-Mass Transfer Coefficient in Bubble Column", *Ind. Eng. Process. Des. Dev.,* vol.13, No. 13, pp. (84), 1974.
4. Anderson, J. L., and Quinn, J. A., *Chem. Eng Sci*. to published 1970.
5. Aguilera, M. E., and Lo'pez de Ramos, A. L., “Effect of Electrolytes on Bubble Coalescence in Columns Observed Through Visualization Techniques”. University Simo'n Bolivar. Apartado Postal 89000. Caracas 1080-A Vene Zuela. http://www. Ichmt.org/abstract/vim-01/abstract
6. Batch, H.F., and Pilhofer, T., "Variation of Gas holdup in Bubble Column with Physical Properties of Liquid and Operating Parameters of Columns" *Ger. Chem. Eng.*, I, pp.(270), 1978.
7. Batshelor, G. K., "Introduction to Fluid Mechanics", pp. (474), Cambridge 1967.
8. Backurst, J. R., Harker, J. H., Coulson, J. M., and Richardson, J. F., “ Chemical Engineering”, vol. 1, fluid flow, Heat Flow and MassTransfer Flow.
9. Bruwlick, W. J., Fair, J. R., and Lerne, B. J., “Mass Transfer in a Sparged Contactor: Part1. Physical Mechanisms and Controlling Parameters”, *A. I. Ch. E. Journal*, vol. 11, No. 1,pp. 73, 1965.

10. Botton, R. D., Casseraty and Charpentier, J. C., "Influence of Column Diameter and High Gas Throughputs on the Operation of a bubble Column", *Chem. Eng. J.*, vol. 16, pp. (107), 1978.
11. Chen, Y., "Modeling of Gas-Liquid Flow in Pipes: Flow Pattern Transition and Drift Flux Modeling", June, 2001. http://pangea.stanford.edu/~ygchen/
12. Calderbank, P. H.; Young, M. B. ; Bibby, R.; "Coalescence In Bubble Reactors and Absorber", Symp. On drops and bubbles, Cambridge University, England, pp. (91), 1962.
13. Deckwer, W.D., Burchart, R.; and Zoll, G.,"Mixing and Mass Transfer in Tall Bubble Column", *Chemical Eng. Sci.*,1980.
14. Dudukovic, M.P., and Al-Dahhan, M., "Bubble Column Flow: Experimental and Modeling", Chem. Eng. Lab. (CREL) Washington University, St. Louis, Mo. 63130.
15. Dhotre, M.T.; Ekam bam, k., and Joshi, J.B. (2004),"CFD Simulation of Sprger Design and Height to Diameter Ratio on Gas holdup Profiles in Bubble Column Reactor", *Experimental Thermal and Fluid Science*, vol. 28, pp.(407-421).
16. Dechwer, w. D., " Design Simulation of Bubble Column Reactors', in *Chemical Reactor Design and Technology*, Edited by Huge, 1. , de Lasa, Nato ASI series, Series E, No. 110, pp. (411), 1985. (cited in Suker 2004).
17. Fair, J. R., "Designing Gas-Sparged Reactors", *Chem. Eng*. vol. 74, 67, July, 3, 1976.
18. Geary, N., "On Bubble Column", PhD. Thesis, Louisiana State University, 1991.

19. Hughmark, G. A., "Holdup and Mass Transfer in Bubble Column", *I. and EC. Process Design and Development*, vol. 6, No. 2, pp. (218), 1967.
20. Hikita, H.; Asai, S.; Taniga, K.; Segawak, K.; and Kitao, M., "The Volumetric Liquid Phase Mass Transfer Coefficient in Bubble Column", *The Chemical. Eng. J.*, vol. 12, pp. (89), 1976.
21. Heijnen, J. J., and Vantriet, K., *Chem. Eng. J.*, vol. 28, pp. (B21-B22), 1984.
22. Hammer, H. and Rashr, W., "Blasengro Ben- Haafigkeits-Verteilungen and Phasengrenrezflachen in Blasensaulen", *Chem Eng. Tech.*, vol. 45, pp. (968), 1973.
23. Hills, J. H., "The Rise of Large Bubble Through a Swarm of Smaller Ones", *Trans. Instn. Chem. Eng.*, vol. 35, 1975.
24. Hills, J. H., "The Operation of Bubble Column at High Throughputs", *Chem. Eng. J.*, vol. 12, pp. (89-99), 1976.
25. Haidari, A. H., and Matthews[2], B., "Future Trends for Computational Fluid Dynamics in the Process Industry", Third International Conference on CFD in the Minerals and Process Industries, pp. (331), 2003.
26. Jeng, J. J., Maa, J. R., and Yang, Y. M., "Surface Effect and Mass Transfer in Bubble Column", *Ind. Eng. Chem. Process Des. Dev.*, vol. 25, pp. (974), 1986.
27. Jamal Ahmadi, M., and Muller-Stienhagen, H. "Effect of Electrolytes Concentration on Bubble Size and Gas holdup in Bubble Column", *Trans. I Chem. E*, vol. 68, part A, 1990.
28. Jackson, R. *Trans. Instn Chem. Engrs*, 1964, CE 107.
29. Joshi, J.B., "Computational Flow Modeling and Design of Bubble Column Reactors", *Chem. Eng. Sci.*, vol. 56,pp. 5893-5933, 2001.

30. Kumar, A.; Degalieesan, T.E.; Laddha, G.S.; and Holelsher, H.E." Bubble Swarm Characteristic in Bubble Column", *Can. J.* vol. 54, pp. (503), 1967.
31. Kelkar, B.G.; Godbole, S. P. ; Honath, M. F.; and Shah, Y. T.; Carr, N. L. ;Deckwer, W. D., "Effect of Addition of Alcohols on Gas holdup and Back mixing in Bubble Column", *AIChE. J.*, vol. 29, No. 3, pp. (361), 1983.
32. Koetsier, W. T., Van Swaaij, W. P. M., and Van Der Most, M., J. *Chem. Eng. Japan*, 9, 332, 1976.
33. Kim, J. W., and Lee, W. K., "Coalescence Behavior of Two Bubbles in Stagnant Liquids", vol. 20, no. 5, pp. (448), 1987.
34. Keitel, G. and oken, U., "Inhibition of Bubble Coalescence by Solutes in Air/Water Dispersion", Chem. Eng. Sci., vol. 37, pp. (1635-1638), 1982.
35. Kumar, R., and Kuloor, N. R., "Blasenbildung in Nichtviskosen Flussigkeiten Under Konstanten Flie B beding Ungen", *Chem Eng. Tech.*, vol. 19, pp. (657-660), 1967.
36. Levich, V. G., "Physicochemical Hydrodynamics", Prentice-Hall, Englewood Cliff, N. J. (1962). (cited in Shah et. al., 1982).
37. Lye, G.J., Stuckey, D.C., "Extraction of Erythromycin- A using Colloidal Liquid Aphrons: Part II. Mass Transfer kinetics", *Chem. Eng. Sci.*, vol. 56, pp. 97-108, 2001.
38. Lockett, M. J., and Kirkpatric, R. D., *Trans. Inst. Chem. Engrs.*, 53, 267, 1975.
39. Mouza, K. A.; Kazakis, N. A.; and Paras, S. V.," Bubble Column Reactor Design Using a CFD Code", International conference "From

scientific Computing and Computational Engineering", www:http//Philon. Cheng. Auth. gr.

40.Maruyama, T.; Youshida, S.; and Mizushina, T.; "The Flow Transition in Bubble Column", *J. of Chem. Eng. of Japan*, vol. 14, No. 5, pp. (352), 1981.

41.Martis, E. L., "Flow Regime Transition in Bubble Column Reactor", Augest, 2004.

http://www.env.wustl.edu/RE4/2004/students/martis/final

42.Miller, D. N., "Gas holdup and Pressure Drop in Bubble Column Reactor", *Ind. Eng. Chem. Process Des. Dev.*, vol. 19, pp. (371), 1980.

43.Marrucci. G., Nicodeno, L., *Chem. Eng. Sci.*, vol. 22, pp. (1257-1265), 1967.

44.Moissisr and Griffith, P. J., *Heat Transfer. Trans. ASME.* pp. (29) February, 1962.

45.Miyaychi, and Shyu, C. N., "Flow of Fluid in Gas-Liquid Columns", *Kagaku Kogaku*, vol. 34, pp. (958), 1970.

46.Mitsuharu, I., Hiroki, U., Toshifumi, I., "Mass Transfer Characteristics in Gas Bubble Dispersed Phase Generated by Plunging Jet Containing Small Solute Bubbles". *Chem. Eng. Sci.*, vol. 56, pp. 6225-6231,2001.

47.Nakanoh, M., and Youshida, F., "Gas Absorption by Newtonian and Non Newtonian Liquid in Bubble Column", *Ind. Eng. Chem. Process, Des., Dev.*, vol. 19, pp. (190), 1980.

48.Philip, J.; Proctor, J. M.; Nirajan, K.; and Davidson, J. F., "Gas holdup and Liquid Circulation in Internal Loop Reactor Containing Highly Viscous Newtonian and Non Newtonian Liquids", *Chem. Eng. Sci.*, vol. 45, No. 3, pp. (651), 1990.

49. Prince, J. and Blanch, H.W., "Bubble Coalescence and Breakup in Air-Sparged Bubble Column", *AIChE, J.*, vol. 36, No. 10, pp. (1485-1499), 1990.
50. Paras, U. V., and Joshi, J. B., "Measuring of Gas Holdup Profiles in Bubble Column by Gamma-ray Tomography", *Trans. I. Chem. Eng.*, vol. 78, part A, April 2000.
51. Reith, T. "Physical Aspects of Bubble Dispersion in Liquids", Doctoral Thesis, TH Delft., 1986. (Cited in Shah et al., 1982)
52. Shah, Y. T.; Kelkar, B. G.; Godbole, S. P.; and, Deckwer, W. D., "Design Parameters Estimation for Bubble Column Reactor", *AIChE. J.*, vol. 28, No. 3, 1982.
53. Shohman, O., "Flow Pattern Transition and Charateristic in Gas- Liquid Two Phase Flow in Inclined Pipes", PhD. Thesis, 1982.
54. Shügerl, K., Luck, J., and Oles, U., "Bubble Column Bioreactors*", Adv. Biochem. Eng. Ed.*, T. K. Ghose, A. Fiecher, and Blakeborogh, B., vol. 7, no. 1, 1977. (cited in Shah et al., 1982).
55. Shügerl, k., *Chem. Ing. Tech*. vol. 52, pp. (951), 1980.
56. Thorat, B. N.; Shevad, A. V.; Bhilegaovkar, K. V.; Aglawe, R. H.; Parasu, U.; Thakre, S. S.; Pandit, A. B.; Sawant, S. B.; and Joshi, J. B., "Effect of Sparger Design and Height to Diameter Ratio on Fractional Gas holdup in Bubble Column", *Trans. I. Chem.*, vol. 76, part A, October, 1998.
57. Uchida, S.; Tsuyutani, S.; and Seno, T.; "Flow Regimes and Mass Transfer in Counter- Current Bubble Column", *Can. J.*, of Chem. Eng., vol 67, October, 1989.
58. Vermeer, D. J., and Krishna, R., "Hydrodynamics and Mass Transfer in Bubble Column Operating in Churn Turbulent Regime", *Ind. Eng. Chem. Process Des. Dev.*, vol. 20, No. 3, 1981.

59. Wei-romg, Z.; Hui-Xiang, S.; and Duhui, W., " Modeling of Mass Transfer Characteristic of Bubble Column Reactor with Surfactant Present", *J. of Zhejiang university Science*, April 2004.

60. Vázquez, G., Cancela, M.A., Riverol, C., Alvarez, E.,Navaza, J.M., "Application of the Danckwerts Method in a Bubble Column: Effects of Surfactants on Mass Transfer Coefficient and Interfacial area", *Chem. Eng. J.*, vol. 78, pp. 13-19, 2000.

61. Wild, G.; Poncin, S.; Li, U.; and Olmose, E.; "Some Aspects of The Hydrodynamics of Bubble Column", International, *J. of Chemical Reactor Eng.*, 1(review, R7), 2003.

62. Walter, J. F., and Blanch, H.W., "Bubble Break-up in Gas Liquid Bioreactors", *Chem. Eng. J.*, vol.032, pp. (87), 1986.

63. Wallis, G. B., "One Dimensional Two Phase Flow", Mc Graw-Hill, New York, 1969.

64. Wu, Z.C., Zhou, M.H., Huang, Z.W., Wang, D.H., "Electrocatalysis Method for Wastewater Treatment using a Novel *beta*-lead Dioxide Anode". *Journal of Zhejiang University SCIENCE*, vol. 3(2), pp. 194-198, 2002.

65. Yang, Y.P., Wu, X.Y., Xu, X.H., Wang, D.H. "Treatment of Dyeing Wastewater by Photo-assisted Fenton System" *Journal of Chemical Engineering of Chinese Universities*, vol. 15(3), pp. 242-247 (in Chinese), 2001.

66. Youshida, F., and Akita, K., 'Performance of Gas Bubble Column: Volumetric Liquid-Phase Mass Transfer Coefficient and Gas holdup", *AIChE. J.*, vol. 11, No. 9, (1965).

67. Zahradnick, J.; Fialova, M.; Kastanck, F.; Green, K.; and Thomas, N. H., "The Effect of Electrolytes on Bubble Coalescence and Gas holdup in

Bubble Column Reactor", *Trans. Inst. Chem. Eng.* vol. 37, part A, pp. (341), 1995.

68. Zahradnick, J. G.; Kuncov'a, G.; and Fialov'a, M.;" The Effect of Surfactants on Bubble Coalescence and Gas holdup In Bubble Column Reactors Containing Viscous Liquid", Proceeding rnicka, Slovakia, 15-19 June (1997), (Ed. S. Schlosser), pp. (505-508), 1997.

69. Zlokarnik, M., "Eignung Von Einlochboden als Gasverteiler in Blasensaulen", *Chem. Ing. Tech.*, vol. 43, pp. (329), 1971.

APPENDIX "A"

Table (A-1) Results for 0.24 M *NaCl*

Gas velocity (m/s)	**Gas holdup (-)**	**Drift Flux (m/s)**	**Gas velocity/ Gas holdup (m/s)**
0.009657	0.038021	0.00929	0.253991
0.013114	0.062081	0.0123	0.21124
0.01657	0.081164	0.015225	0.204155
0.020026	0.10045	0.018014	0.199363
0.023482	0.112614	0.020838	0.208518
0.026939	0.127363	0.023508	0.211514
0.030395	0.136158	0.026256	0.223233
0.039355	0.141806	0.033774	0.277527
0.054716	0.1522	0.046388	0.359501
0.070077	0.161342	0.058771	0.434338
0.085438	0.170288	0.070889	0.501726
0.100799	0.179688	0.082687	0.560967
0.11616	0.189189	0.094184	0.613989
0.131521	0.217877	0.102866	0.603648

Table (A-2) Results for 0.3 M *NaCl*

Gas Velocity (m/s)	Gas holdup (-)	Drift Flux (m/s)	Gas velocity/ Gas holdup (m/s)
0.009657	0.039121	0.009279	0.24685
0.013114	0.06166	0.012305	0.212682
0.01657	0.08317	0.015192	0.19923
0.020026	0.102564	0.017972	0.195254
0.023482	0.112801	0.020741	0.208172
0.026939	0.123905	0.023601	0.217417
0.030395	0.133127	0.026349	0.228316
0.039355	0.13252	0.034158	0.296974
0.054716	0.147381	0.046652	0.371255
0.070077	0.158654	0.05896	0.441697
0.085438	0.170616	0.070861	0.500762
0.100799	0.180328	0.082622	0.558976
0.11616	0.198	0.093034	0.586667
0.131521	0.215	0.103269	0.611726

Table (A-3) Results for 0.36 M *NaCl*

Gas velocity (m/s)	**Gas holdup (-)**	**Drift Flux (m/s)**	**Gas velocity/ Gas holdup (m/s)**
0.009657	0.037139	0.009298	0.260023
0.013114	0.059772	0.01233	0.2194
0.01657	0.082268	0.015207	0.201415
0.020026	0.098229	0.018059	0.203871
0.023482	0.11	0.020913	0.213473
0.026939	0.123356	0.123356	0.218384
0.030395	0.129894	0.023616	0.233998
0.039355	0.136868	0.026447	0.28754
0.054716	0.153567	0.033969	0.356301
0.070077	0.163513	0.046368	0.428571
0.085438	0.177116	0.058619	0.482384
0.100799	0.187935	0.070306	0.53635
0.11616	0.199695	0.081855	0.581687
0.131521	0.217877	0.092963	0.603648

Table (A-4) Results for 0.42 M *NaCl*

Gas velocity (m/s)	Gas holdup (-)	Drift Flux (m/s)	Gas velocity/ Gas holdup (m/s)
0.009657	0.034483	0.009324	0.280051
0.013114	0.058507	0.012347	0.224144
0.01657	0.081967	0.015212	0.202155
0.020026	0.096191	0.0181	0.20819
0.023482	0.111675	0.02086	0.210271
0.026939	0.121706	0.02366	0.221345
0.030395	0.132	0.026398	0.230265
0.039355	0.141631	0.033781	0.27787
0.054716	0.157641	0.046091	0.347092
0.070077	0.172577	0.057984	0.406062
0.085438	0.180807	0.06999	0.472537
0.100799	0.195864	0.081056	0.514638
0.11616	0.205899	0.092243	0.56416
0.131521	0.215686	0.103154	0.60978

Table (A-5) Results for 0.6 M *NaCl*

Gas velocity (m/s)	**Gas holdup (-)**	**Drift Flux (m/s)**	**Gas velocity/ Gas holdup (m/s)**
0.009657	0.03537	0.009315	0.273028
0.013114	0.055755	0.012383	0.235208
0.01657	0.072438	0.01537	0.228747
0.020026	0.0814	0.018396	0.24602
0.023482	0.093068	0.021297	0.25231
0.026939	0.106003	0.024083	0.254134
0.030395	0.113363	0.026949	0.268121
0.039355	0.13	0.034457	0.302731
0.054716	0.143	0.046909	0.382629
0.070077	0.162846	0.058666	0.430327
0.085438	0.180488	0.070017	0.473372
0.100799	0.195556	0.081087	0.515448
0.11616	0.21186	0.09155	0.548287
0.131521	0.23	0.101673	0.57183

Table (A-6) Results for 0.8 M *NaCl*

Gas velocity (m/s)	Gas holdup (-)	Drift Flux (m/s)	Gas velocity/ Gas holdup (m/s)
0.009657	0.038902	0.009281	0.248239
0.013114	0.058296	0.01235	0.224955
0.01657	0.074074	0.015343	0.223695
0.020026	0.084169	0.01834	0.237926
0.023482	0.096386	0.021219	0.243625
0.026939	0.10505	0.024109	0.25644
0.030395	0.114111	0.02693	0.266363
0.039355	0.133753	0.034091	0.294236
0.054716	0.150228	0.046496	0.36422
0.070077	0.171679	0.058047	0.408186
0.085438	0.18502	0.06963	0.461777
0.100799	0.199466	0.080693	0.505344
0.11616	0.218241	0.090809	0.532256
0.131521	0.236155	0.100462	0.556927

Table (A-7) Experimental and Theoretical Values of Gas holdup

Superficial Gas Velocity (m/s)	Experimental Gas Holdup (-)	Theoretical Gas Holdup(-)
0.009657	0.013589	0.02
0.020026	0.095218	0.03
0.026939	0.124818	0.04
0.039355	0.151515	0.06
0.070077	0.171761	0.1
0.100799	0.192618	0.14
0.131521	0.227941	0.18

Table (A-8) Effect of Gas Velocity on Mass transfer coefficient

Gas velocity (m/s) × 10^{-2}	Mass Transfer Coefficient (s^{-1})
0.5825	2.587221E-3
0.7	0.0030769
0.8	0.0034231
0.9	3.80769E-3
0.9657	4.153226E-3
1	0.00385

APPENDIX "B"

General Equations Used in Bubble Column Simulation

The Euler-Euler approach

The Eulerian model solves a set of n energy, momentum and continuity equations for each phase. Coupling is achieved through the pressure and interphase exchange coefficients. Particle tracking (using the Lagrangian dispersed phase model) interacts only with the primary phase. The equations being solved in the Eulerian model are as follows:

Continuity equation (void fraction equation) for the gas phase from total of n phases

$$\frac{\partial}{\partial t}\left(\varepsilon_G\right)+\nabla.\left(\varepsilon_G u_G\right)=\frac{1}{\rho_G}\left(\sum_{\rho G}^{n} m_{LG}-\varepsilon_G\frac{d\rho_G}{dt}\right) \quad \ldots(1)$$

A similar term appears also in the momentum equations. The term is zero by default, but can be specified either as a constant, or by a user-defined function.

Momentum equation for the gas phase:

$$\frac{\partial}{\partial t}\left(\varepsilon_G\rho_G u_G\right)+\nabla.\left(\varepsilon_G\rho_G u_G u_G\right)=-\varepsilon_G\nabla p+\nabla.\tau_G+\varepsilon_G\rho_G g+\sum_{p=1}^{n}\left(E_{LG}\left(u_L-u_G\right)+m_{LG}u_{LG}\right)$$
$$+\varepsilon_G\rho_G\left(F_G+F_{lift,G}+F_{vm,G}\right)$$
…(2)

Where

$$\tau = \varepsilon_G \mu_G \left(\nabla u_G + \nabla u_G^T\right) + \varepsilon_G \left(\lambda_G - \frac{2}{3}\mu_G\right) \nabla . u_G I \qquad \ldots(3)$$

And the interphase momentum exchange coefficient E_{LG} can be expressed in a symmetric form as:

$$E_{LG} = \frac{\varepsilon_L \left(\varepsilon_L \rho_L + \varepsilon_G \rho_G\right) f}{\tau_{LG}} \qquad \ldots(4)$$

Where

$$\tau_{LG} = \frac{\left(\varepsilon_L \rho_L + \varepsilon_G \rho_G\right)}{18\left(\varepsilon_L \mu_L + \varepsilon_G \mu_G\right)} \left(\frac{d_{bL} + d_{bG}}{2}\right)^2 \qquad \ldots(5)$$

And

$$f = C_D Re/24 \qquad \ldots(6)$$

Where

$C_D = 24(1 + 0.15Re0.687)/Re$ for Re $\leq$ 1000,

$C_D = 0.44$ for Re> 1000.

Relative Reynolds number is computed from

$$Re = \frac{\left|u_L - u_G\right| d_{bp}}{\mu_G} \qquad \ldots(7)$$

A general form of the K_{LG} used is

$$E_{LG} = \frac{\varepsilon_G \varepsilon_G \varepsilon_L f}{\tau_L} \qquad \ldots(8)$$

$$\tau_L = \frac{\rho_L d_{bL}^2}{18 \mu_G} \qquad \ldots(9)$$

الخلاصة

عند عمل محاكاة لوصف العمودالفقاعي هو خطوة نحو تصميم هذه الأعمدة بكفاءة و بأحجام كبيرة.

المحاكاة تسمح لنا بالتنبؤ بكمية احتجاز الغاز, وهو العمل الرئيسي المؤشر لمدى الأتصال بين السائل والغاز.

هذا العمل يتكون من جزئين الأول عملي بعمود فقاعي بظروف تشغيليه مختارة بابعاد 0،081م قطرًا, 1،03م إرتفاع بأستخدام هوا- ماء, ثاني أوكسيد الكاربون, هواء- ماء مع ملح كلوريد الصوديوم بتراكيز مختلفة من 0،24 مولاري إلى 8، 0مولاري حيث تم دراسة تأثيرسرعة الغاز مع إضافة الملح على معامل احتجاز الغاز و السرعة الأنتقالية وتم ايضًا دراسة تأثير سرعة الغاز على معامل انتقال الكتلة k_{La}.

والجزء الثاني بأستخدام CFX (ديناميك المائع الحسابي) للعمود الفقاعي بذات الظروف. تم تطبيق المحاكاة ثلاثية الأبعاد بأستخدام Ansys مع نموذج Eulerian-Eulerian ومقارنتها مع مثيلاتها العملية وكانت النتيجة جيدة.

شكر و تقدير

اود ان اعبر عن خالص شكري و تقديري للمشرف الدكتور نصير الحبوبي لما قدمه لي من توجيهات قيمه و نصائح سديده طوال فترة انجاز البحث.

اود ايضا ان اشكر جميع اساتذة وموضفي قسم الهندسة الكيمياوية لابدائهم المساعدة اللازمه اثناء هدا العمل.

ولا انسى ايضا ان اشكر من رباني على طريق الخير و المعرفه اعزمن في الوجود امي وابي والى زوجي واخواتي.

م. رهام عبد المجيد

دراسة العوامل الهيدروديناميكية للاعمدة الفقاعية عمليًا وحسابيًا باستخدام ديناميك المائع العددي (CFD)

رسالة

مقدمه الى كلية الهندسة في جامعة النهرين

و هي جزء من متطلبات نيل درجة ماجستير علوم

في الهندسة الكيمياوية

من قبل

رهام عبد المجيد سعيد

بكلوريوس علوم في الهندسة الكيمياوية 2003

ربيع الاول 1428

آذار 2007