BEHAVIOUR OF BUBBLE COLUMNS WITH TWO AND THREE PHASES

by

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SUMMARY

Bubble columns are multi-phase contactors widely used in the process industries. For proper design, effective operation and analysis of such columns, knowledge of the fluid-dynamics, mixing and transport parameters is essential. Hitherto, most studies have been on simple two-phase systems in relatively small diameter columns at low phase throughputs. A study of the fluid dynamics and mixing characteristics of bubble columns was thus undertaken, following a comprehensive literature review that confirmed the need for further research. A two-dimensional column 15.3cm x 1.5cm in section and 143cm high, and a column 15.2cm in diameter were used to study the effects of operating conditions and phase properties on the gas hold-up, flow regimes, bubble dynamics and mixing in two- and three-phase systems. Complex two- and three-phase systems were also tested.

Air, water and a variety of solid particles were used as the gas, liquid and solid phases respectively. More complex systems involved the use of the following additives: alcohols, ketones, esters and electrolytes. Three methods were employed in measuring gas hold-up while flow regimes and bubble dynamics were characterised by visual observations, photographic techniques and qualitative analysis. The "stimulus-response" technique was used for mixing studies. In all cases, high flow rates corresponding to regimes of industrial importance were investigated.

Experimental results were qualitatively and quantitatively analysed. Column behaviour varied with the operating conditions, nature of the phases and column design. The gas hold-up increased with gas flow rate for all systems studied and was also sensitive to the properties of the phases. Flow regimes, bubble dynamics and dispersion coefficients all depended on the gas velocity and nature of the phases. Small bubbles and thus high interfacial areas were produced by the presence of organic solutes and electrolytes. These additives affect the coalescence behaviour of aqueous solutions drastically changing from quick coalescence as in water to coalescence restraining. However, the effect of additives was insignificant on the three-phase systems. The effect of solids was linked to the surface property of wettability. This led to a two-way classification of behaviour: while the wettable solids tended to suppress bubble coalescence, non-wettable solids enhanced coalescence. Consequently, the former resulted in high values of gas hold-up while the latter reduced gas hold-up and dispersion when compared to systems without solids.

A new concept aimed at a clearer understanding of column behaviour is presented. This is based on the analysis of molecular behaviour in bulk phases and at interfaces between phases. Using this approach, it should eventually be possible to predict the effect of various factors on the design and operation of bubble columns.

KEY WORDS:

Bubble columns. Gas hold-up. Flow regimes. Bubble dynamics. Liquid mixing.

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DEDICATION

This thesis is humbly dedicated to the glory of GOD,
to my wife, my family
and all who have contributed towards my progress.

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NOMENCLATURE

a_L	Gas-liquid interfacial area per unit volume, cm^{-1}
a_s	Liquid-solid interfacial area per unit volume, cm^{-1}
A	Concentration of reactant A, g/cm^3
C	Concentration, g/cm^3
C_N	Carbon number
C_o	Radial distribution parameter
d_B	Bubble diameter, cm
d_o	Orifice diameter, cm
d_p	Particle diameter, cm
d_{vs}	Sauter mean diameter, cm
D_c	Column diameter, cm
D_L	Dispersion coefficient, cm^2/s
E_i	Energy input rate, watt
F_B	Buoyancy force, dynes
F_s	Surface tension force, dynes
g	Gravitational acceleration, cm^2/s
G_v	Gas flow rate, litre/s
H	Liquid height in manometer, cm
j_{GL}	Drift flux, cm/s
K, K_o	Kinetic constants
K_G	Gas side mass transfer coefficient, cm/s
K_L	Liquid side mass transfer coefficient, cm/s
K_s	Liquid-solid mass transfer coefficient, cm/s

l	Length (Figure 4.3) cm
L	Distance between the two manometers, cm
L_i	Liquid height without aeration, cm
L_f	Aerated liquid height, cm
(P/V_L)	Power input per unit volume of liquid, watt/cm^3
Q	Gas volumetric flow rate, cm^3/s
r_c	Rate of chemical reaction, moles/s cm^3
r_E	Length of bubble neck (Figure 2.2), cm
r_F	Bubble radius, cm
R	Global reaction rate, mol/(mass catalyst) s
S	Cross-sectional area of column, cm^2
t	Time, s
U	Mean velocity in axial direction, cm/s
U_B	Velocity of bubbles, cm/s
$U_{b\infty}$	Single bubble rise velocity, cm/s
U_c	Circulation velocity, cm/s
U_o	Buoyancy velocity, cm/s
U_s	Slip velocity, cm/s
U_{SG}	Superficial gas velocity, cm/s
U_{SL}	Superficial liquid velocity, cm/s
V_c	Average liquid circulation velocity, cm/s
V_E	Volume generated in first stage (equation 2.6) cm^3
V_F	Final bubble volume, cm^3
V_Z	Vertical component of liquid velocity, cm/s
W	Work of adhesion, erg (J)

GREEK SYMBOLS

α, β	Constants defined by equation 7.1
γ	Generalised viscosity constant (equations 7.2 and 7.3)
δ	Film thickness, cm
ε_G	Gas hold-up (-)
$\varepsilon(o)$	Static gas hold-up (-)
$\varepsilon(t)$	Dynamic gas hold-up (-)
η	Catalyst effectiveness factor
θ	Contact angle of the liquid (on the given solid surface in the presence of gas)
μ	Dynamic viscosity g/cm s (= poises)
ν	Kinematic viscosity, cm^2/s (= μ/ρ)
ρ	Density, g/cm^3
σ	Surface tension, dynes/cm
τ_f	Detachment time duration, s

SUBSCRIPTS

G, g	Gas phase
L, l	Liquid phase
M	Gas and liquid mixture
S	Solid phase
i	Interface

DIMENSIONLESS PARAMETERS

Bo	Bond Number,	$\frac{\rho_L g d_B^2}{4\sigma}$
Da	Damkholer Number,	$\frac{r_c L_f}{U C_A}$
Fr	Froude Number,	$\frac{U_{SG}}{\sqrt{g D_c}}$
Ga	Galileo Number,	$\frac{g D_c^3}{\nu_L^2}$
Pe	Peclet Number,	$\frac{U_{SL} L_f}{D_L}$
Re	Reynold Number,	$\frac{\rho_L U_{SG} d_B}{\mu_L}$
Sh	Sherwood Number,	$\frac{K_L d_B}{D_L}$
St	Stanton Number,	$\frac{K_L a_L L_f}{U_{SL}}$
We	Weber Number,	$\frac{d_B U_{SG}^2 \rho_L}{\sigma}$

CHAPTER 1

GENERAL INTRODUCTION

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1.1 Process Applications

Chemical, biochemical, pharmaceutical and petroleum processes constitute a wide spectrum of operations where BUBBLE COLUMNS are used. These are cont-actors in which a gas is bubbled into a moving or stagnant pool of liquid in either co-current or counter-current flow. Figure 1.1 shows how the gas, G, is dispersed as a bubble swarm of high interfacial area and the liquid, L, may contain suspended solid, S, as either reactant or catalyst.

Bubble columns are very versatile systems for conducting a variety of gas-liquid and gas-liquid-solid operations, with or without chemical reaction. They are widely used as absorbers, strippers, reactors and fermenters in various processes where it is necessary to transfer matter and energy from one phase to another immiscible, or partially miscible phase. A review of process applications with examples is given in Figure 1.2.

Bubble columns are preferred to other multiphase cont-actors because of their simplicity of operation, relatively low investment and maintenance costs, and high values of effective interfacial area and of overall mass and heat transfer coefficients.

1.2 The Design Problem

The procedure for designing a contactor should start with an exact definition of the requirements, ie. the production level and yields. These quantities and the type of operation under consideration permit a first choice of the so-called adjustable operational conditions, which include phase velocities, temperature, pressure, direction of the flows,

Figure 1.1 Schematic representation of the Bubble column.

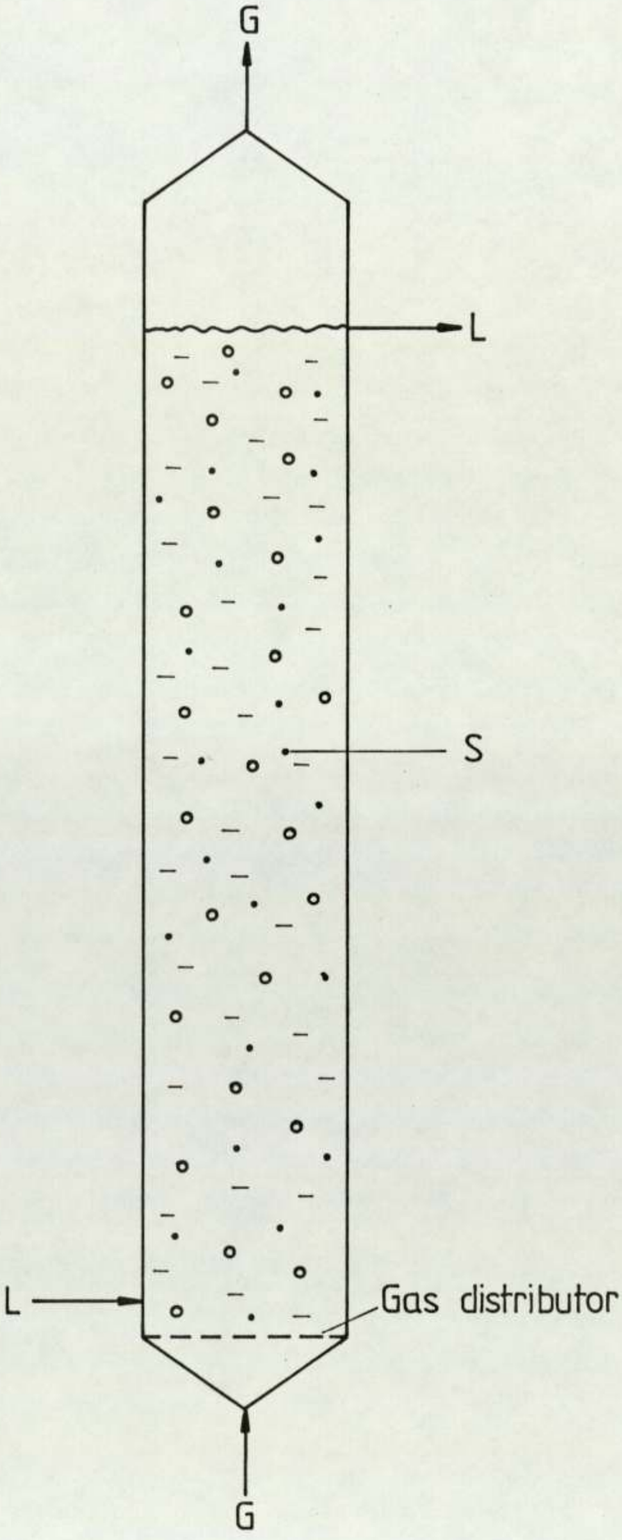
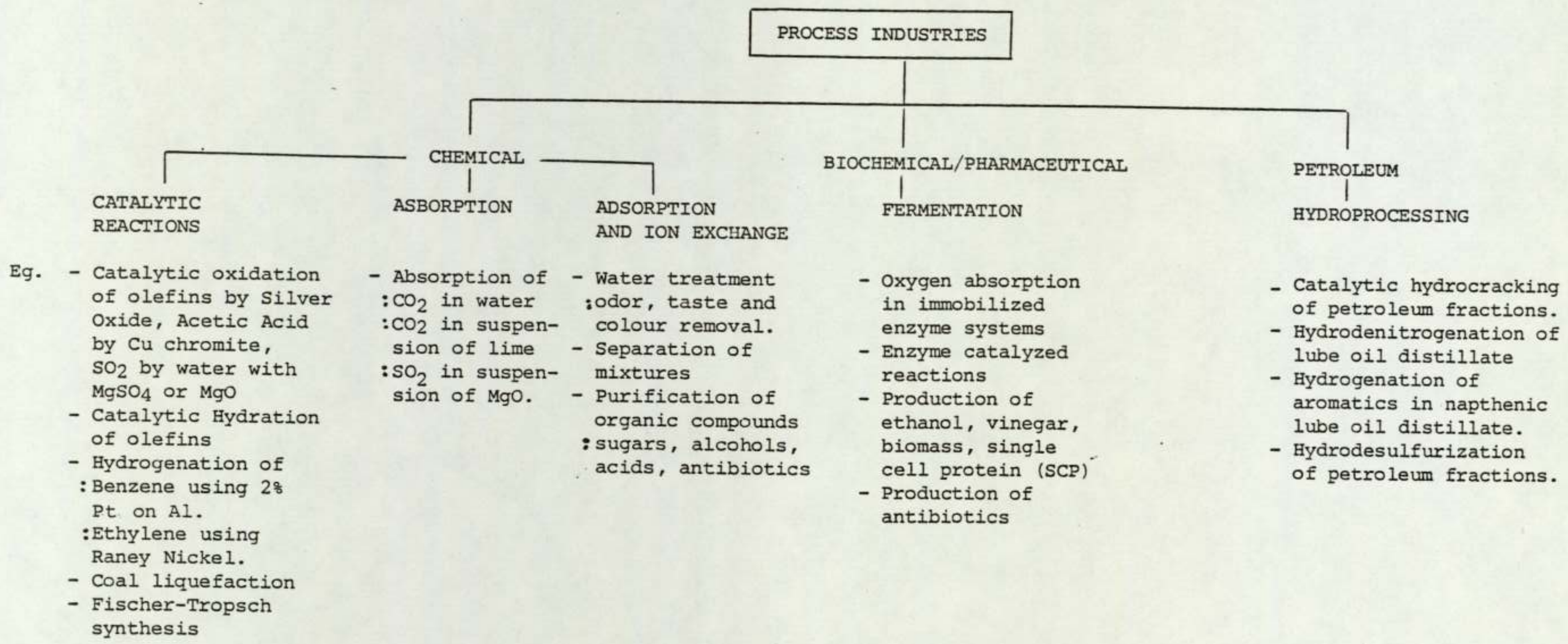


Figure 1.2 - Process Applications of Bubble Columns



- PROCESS INDUSTRIES
  - CHEMICAL
    - CATALYTIC REACTIONS
      - Eg.
        - Catalytic oxidation of olefins by Silver Oxide, Acetic Acid by Cu chromite, SO_2 by water with $MgSO_4$ or MgO
        - Catalytic Hydration of olefins
        - Hydrogenation of
          :Benzene using 2% Pt on Al.
          :Ethylene using Raney Nickel.
        - Coal liquefaction
        - Fischer-Tropsch synthesis
    - ASBORPTION
      - Absorption of
        :CO_2 in water
        :CO_2 in suspension of lime
        :SO_2 in suspension of MgO.
    - ADSORPTION AND ION EXCHANGE
      - Water treatment ;odor, taste and colour removal.
      - Separation of mixtures
      - Purification of organic compounds ;sugars, alcohols, acids, antibiotics
  - BIOCHEMICAL/PHARMACEUTICAL
    - FERMENTATION
      - Oxygen absorption in immobilized enzyme systems
      - Enzyme catalyzed reactions
      - Production of ethanol, vinegar, biomass, single cell protein (SCP)
      - Production of antibiotics
  - PETROLEUM
    - HYDROPROCESSING
      - Catalytic hydrocracking of petroleum fractions.
      - Hydrodenitrogenation of lube oil distillate
      - Hydrogenation of aromatics in napthenic lube oil distillate.
      - Hydrodesulfurization of petroleum fractions.

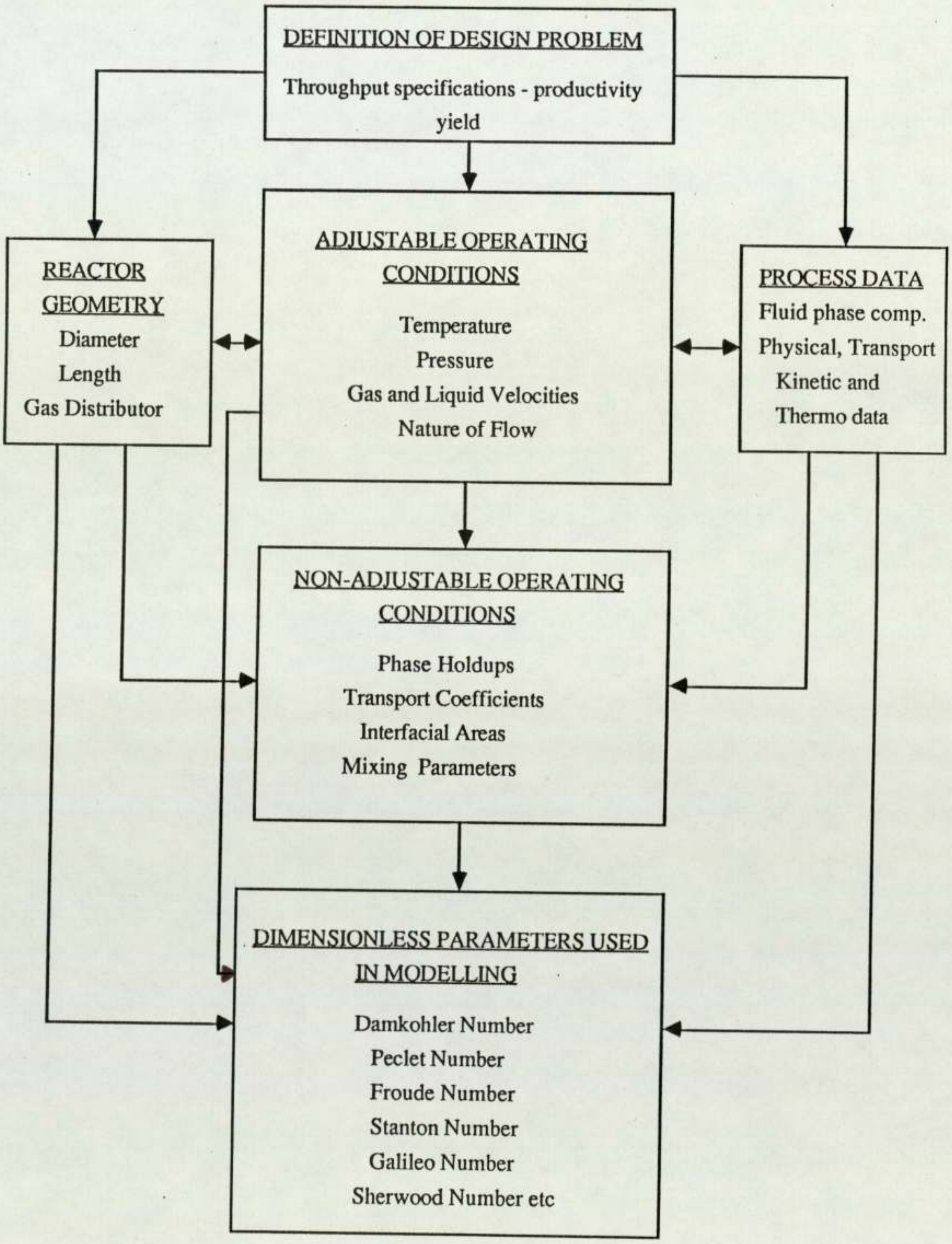
ie. co-current or counter-current. In addition, process data are needed: they comprise physical properties of the mixture and its components (densities, viscosities, heat and mass diffusivities, surface tension), phase equilibrium data and the chemical parameters. The latter are important as they include all the kinetic and thermodynamic information. These "first-level" quantities are interrelated in various ways as shown in Figure 1.3.

For the case of single phase columns the information given in this first level would be sufficient, provided additional knowledge on the residence time distribution is available. In multiphase contractor design, hydrodynamic properties constitute another group of important parameters. These are more or less "non adjustable" or "self adjusting" quantities dependent on the chosen reactor geometry, the adjustable operational conditions and the process data. Below this level are the phase hold-ups, the interfacial areas, the heat and mass transfer properties and the dispersion coefficients. All these quantities, ie. the geometry, process data, adjustable and non adjustable parameters, are then introduced in the column model equations derived on the basis of the physical and chemical phenomena which are suspected to take place within the column. Usually, the model equations have to be solved numerically as they contain strong non-linearities (eg. temperature dependency on reaction rates and solubilities and phase flow variations).

1.3 Scope of the Research

Though there is a considerable amount of knowledge on the various parameters affecting the design and operation of bubble columns, much of the research work has been on simple two-phase systems in relatively small diameter columns. Consequently, there are few or no design data for complex two-phase and multi-phase systems. Also, since most of the correlations apply to specific experimental set-ups and conditions, it is not surprising to find there are some discrepancies and inconsistencies in the published data.

Figure 1.3 - Design Procedure for Bubble Column (based on Deckwer, 1981)



A thorough experimental and theoretical analysis of bubble column behaviour is thus necessary. These points highlight the need for more research. The author's study is concerned with the fluid dynamics and mixing properties of two- and multi-phase bubble columns. Particular attention is paid to holdups, flow regimes, bubble dynamics and dispersion coefficients, being essential parameters in the design and operation of such columns. Complex two- and three-phase systems are also studied.

Background

There has been a considerable interest in bubble column research at the University of Aston. The initial group was composed of chemical engineers headed by Dr EL Smith and microbiologists supervised by Dr RN Greenshields. The microbiologists in the group were mostly concerned with the applications of bubble columns, in particular for beer and alcohol fermentations,and biomass and metabolite production using moulds and bacteria. The engineering aspects of the research - design, scale-up and operation of bubble column fermenters for both aerobic and anaerobic processes - have been carried out mainly in the Chemical Engineering Department.

The overall engineering research programme was divided into the following sub-projects:

1. properties of suspensions of micro-organisms;
2. behaviour of single bubbles in suspensions of micro-organisms;
3. behaviour of bubble swarms in bubble column fermenters;
4. properties of microbial aggregates and their behaviour in bubble column fermenters;
5. mass- and heat-transfer studies in gas-liquid systems in bubble columns;

6 development of mathematical models to aid in the design, scale-up and operation of bubble column fermenters.

Over the years, a number of studies has been started and successfully completed on early sections of the sub-projects (Smith et al, 1974, 1977, 1978). These have gone a long way in providing a solid background for the present author's work which includes sub-projects 3 and 4.

Methodology

Following a comprehensive review of the literature that highlights the need for more consistent and extensive design data, experiments were performed using two- and three-dimensional bubble columns. This work was followed by qualitative analysis of actual flow mechanisms, phase behaviour and column performance. A new approach is developed to complement the continuum method of analysis.

It is believed that the research will provide further insight into column design, operation and performance.

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CHAPTER 2

FUNDAMENTALS OF BUBBLE COLUMN OPERATION

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2.1 Mechanism of Bubble Formation

In a discussion of gas subdivision in liquids with or without solids, it is logical to begin with an examination of the bubbling process in its simplest form. The formation of bubbles at orifices therefore forms the subject matter of this section.

2.1.1 Discrete Bubble Formation

In the bubble column, gas is forced through an orifice or orifices submerged in a liquid (see Figure 1.1). Depending on the gas flow rate, the gas will exit from the orifice as individual bubbles (at low flow rates) or continuously as a jet (at high flow rates). The jet subsequently breaks up into bubbles of various sizes.

Figure 2.1 shows that as a bubble is formed, it is acted upon by three forces:

(i) buoyancy - against the downward force of the liquid;

(ii) surface tension force acting on that section of the bubble still in contact with the orifice;

(iii) convection currents in the fluid.

The effect of convection currents is usually assumed to be negligible in the development of simple theories of bubble formation. If the bubble is assumed spherical at the moment of release, the buoyancy force acting on it is given by:

$$F_B = 1/6\, \pi\, d_B{}^3 g\, (\rho_L - \rho_G) \qquad \text{........ 2.1}$$

Figure 2.1 Forces acting on bubble at the orifice.

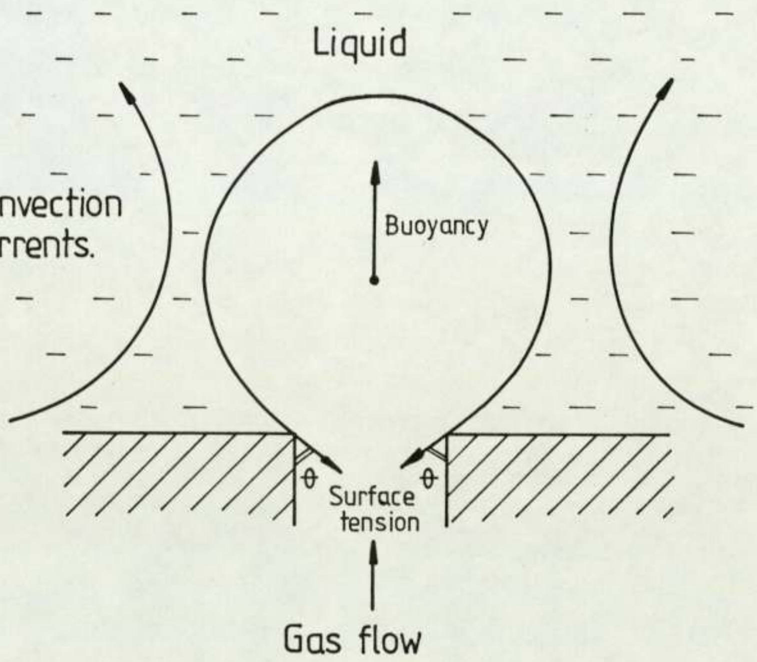
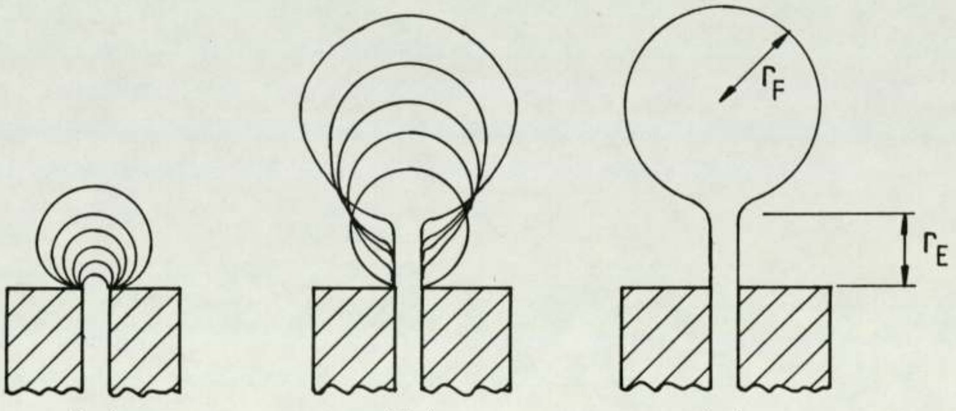


Figure 2.2 Stages of bubble formation at the orifice.



(a) Expansion stage.

(b) Detachment stage.

(c) Condition of detachment.

where:

d_B is the diameter of bubble at the instant of release

g - acceleration due to gravity

ρ_L - density of the liquid

ρ_G - density of the gas.

The surface tension force is given by:

$$Fs = \pi\, d_o\, \sigma\, (\text{Cos}\ \theta)\, f\, \frac{(d_o)}{a} \qquad \text{........ 2.2}$$

where:

d_o = the diameter of orifice

σ = surface tension of liquid

θ = angle between the perpendicular and the bubble surface at the orifice. (This is zero if the column is perfectly wetted by liquid).

$a = (2\gamma/\rho g)^{1/2}$ and

$f\, \frac{(d_o)}{a}$ is the shape factor (which is 1 for spheres).

For the case of low gas flow and when the viscosity of the liquid is low, the buoyancy and surface tension forces balance:

$$1/6\ \pi\ d_B^3\ (\rho_L - \rho_G)\ g = \pi\ d_o\ \sigma \qquad \text{........ 2.3}$$

$$d_B = \left(\frac{6\ \sigma\ d_o}{(\rho_L - \rho_G)g} \right)^{1/3} \qquad \text{........ 2.4}$$

Several workers have obtained results which indicate that bubble diameter, in the bubbly regime, is dependent on gas flow rate. Wraith (1971) derived the following equation from first principles, assuming a two stage growth period:

$$d_B = 1.275\ (G_v^2/g)^{0.2} \qquad \text{........ 2.5}$$

where G_v is the gas flow rate (litre/s)

g is in cm/s^2 and d_B is in cm.

For the Reynolds number range 200 < Re < 1000, the bubble size becomes considerably dependent on viscous and inertial forces. For this range, Ramakrishman et al (1969) proposed a model in which bubble formation takes place in two stages: expansion and detachment.

During the first stage, the bubble remains attached to the orifice and, during the second stage, the bubble moves away from the orifice keeping in contact with the orifice through a neck (see Figure 2.2). The final volume of the bubble is the sum of the volumes generated in each of the two stages -

$$V_F = V_E + Q\tau_f \qquad \text{........ 2.6}$$

where:

V_F = the final bubble volume

V_E = the volume generated in the first stage

Q = the gas volumetric flow rate

τ_f = the detachment time duration.

When the upward forces exceed the downward forces in the second stage the bubble moves away from the orifice but is connected by the bubble neck. The length of this neck is taken to be equal in magnitude to r_E, the bubble radius at detachment (ie. after the first stage), on the assumption that this condition avoids the possibility of the next bubble coalescing with the first. When the bubble base has travelled this distance, the bubble neck is broken and the bubble floats up.

2.1.2 Formation of Bubbles

When a gas is mixed with a liquid (with or without solids) through an array of orifices, a mass motion of gas bubbles is produced which gives rise to an intensive mixing of the liquid. Depending on the flow rates of the gas and liquid in the mixture, different hydrodynamic regimes arise and, consequently, changes in the structure of the mixture occur.

At low flow rates, the gas passes through the liquid at regular intervals in the form of bubbles generated from the orifices. Increasing the gas flow rate retards the circulatory downflow of liquid, and at a certain critical gas velocity the amount of liquid descending to the orifices is insufficient to generate new bubbles, at which point gas jets are produced. The gas jets are eventually broken up into bubbles due to the dynamic influence of the liquid.

2.1.3 Bubble Break up and Coalescence

In intensive mass-bubbling operations, bubbles of different sizes are formed in the gas-liquid/slurry mixture, this being a result of the breaking up and coalescing of bubbles. Bubble breakup and coalescence are part of the physical phenomena occurring during the journey of gas bubbles through the column. They are due mainly to the dynamic interaction between the liquid (or slurry) and gas. Bubble sizes depend on the physical properties of the liquid. Heat, agitation and vibrating are other factors. Bubble break up and coalescence have a considerable influence on column behaviour. In quantitative terms, the relationship between stable bubble size and some of the above factors is:

$$d_B \max \propto \frac{\sigma^{0.6}}{\left(\frac{P}{V_L}\right)^{0.4} \rho_L^{0.2}} \quad \text{........ 2.7}$$

coupled with the expression

$$\frac{P}{V_L} = \rho_L g U_{SG} \quad \text{........ 2.8}$$

If $d_B \neq d_B$ max, then either bubble breakup or coalescence must be expected down-stream from the distributor.

2.2 Gas-Liquid Dispersion and Three Phase Fluidisation

2.2.1 Gas-Liquid Dispersion

The expression gas-liquid dispersion is used to describe the mixing of a gas in a liquid phase. The contacting of a gas with liquid for the purpose of mass transfer of one or more components is a very important operation in chemical engineering; often the mass transfer is followed by chemical reaction. One or both of these phases are normally subdivided in such a contacting operation. This usually increases the interfacial area and produces relative motion between the phases. Both of these effects contribute to an increase in overall mass transfer rate.

2.2.2 Three-Phase Fluidisation

In three-phase fluidisation, contact is established between the three-phases - gas, liquid and solid particles. This may result in momentum, heat and mass transfer and chemical reaction. Such processes are becoming increasingly important in the process industries, examples of which are given in Figure 1.2.

Several different types of operation may be employed in order to obtain the desired contact between the three phases. They may be grouped into two main classes, depending upon the state of motion of the solid particles.

(i) Fixed-Bed Operations

In the first class, the particles form a fixed bed, and the fluid phases may be in either cocurrent or countercurrent flow. Two types of fixed-bed operations, characterized by distinctly different flow patterns, are in current industrial use.

Trickle-Flow Operation

In trickle-flow operation, the liquid phase flows downwards and may or may not cover the solid particles as a film. The gaseous phase moves in either co- or countercurrent, continuous flow.

Bubble-Flow Operation

In bubble-flow operation, the gaseous phase moves upwards as discrete bubbles. The liquid may be in either co- or countercurrent flow. The liquid holdup is thus relatively high.

(ii) Suspended-Bed Operations

In this group of operations, the solid particles are kept in a suspended state by momentum transfer from the liquid phase. Momentum may be transferred to the liquid phase by different means, and a distinction may be made on this basis between three different types of operation.

Slurry Operation

In slurry operation, momentum is transferred to the liquid phase by the movement of gas bubbles; the liquid medium is stationary in most cases. The operation is usually carried out in columns with high height-to-diameter ratios. This mode of operation was used in the author's research work.

Stirred Slurry Operation

In stirred slurry operation, momentum is transferred to the liquid phase by mechanically stirring as well as by the movement of gas bubbles. Small particles are used in most cases, and the operation is usually carried out in tank reactors with low height-to-diameter ratios.

Fluidised-Bed Operation

In fluidised bed operation, the liquid flows upwards through a bed of solid particles which are fluidized by the combined action of the flowing liquid and gaseous phase. The gas moves as discrete bubbles through the liquid-fluidized bed. This type of operation was also used by the author.

2.3 Transport Phenomena

Having introduced the concept of two- and three-phase operation in bubble columns, it is necessary to give a brief discussion of the transfer mechanisms in such processes.

2.3.1 The Transfer Mechanism at the Gas-Liquid Interface

Several models have been proposed to describe the phenomena occurring when a gas phase is brought into contact with a liquid phase in the presence or absence of a solid phase. The model that have been most widely used so far is the two-film theory proposed by Whitman (1923) and by Lewis and Whitman (1953). "Penetration" and "surface renewal" models have also been used to describe the mass transfer processes.

2.3.2 Film Theory Analysis of Gas-Liquid-Solid Fluidisation

Solid particles are present in three-phase systems either as inerts, as catalysts, or as reactants. The analysis in the previous section could also be used for the case where the solid particles are inert. The film theory is used in this section for gas-liquid-solid reactions with solids present as a catalyst.

A schematic representation of the concentration distribution of a gaseous reactant is given in Figure 2.3. The following process steps are involved:

(1) mass transfer from the bulk gas phase to the gas-liquid interface
(2) mass transfer from the gas-liquid interface to the bulk-liquid phase
(3) mixing and diffusion in the bulk liquid
(4) mass transfer to the external surface at the catalyst particles
(5) reaction at the catalyst surface, including intraparticle diffusion effects.

Shah (1979) has given an extensive treatment for such processes. For a first-order, catalytic reaction involving a gaseous reactant, the following steady-state equations are valid under isothermal conditions.

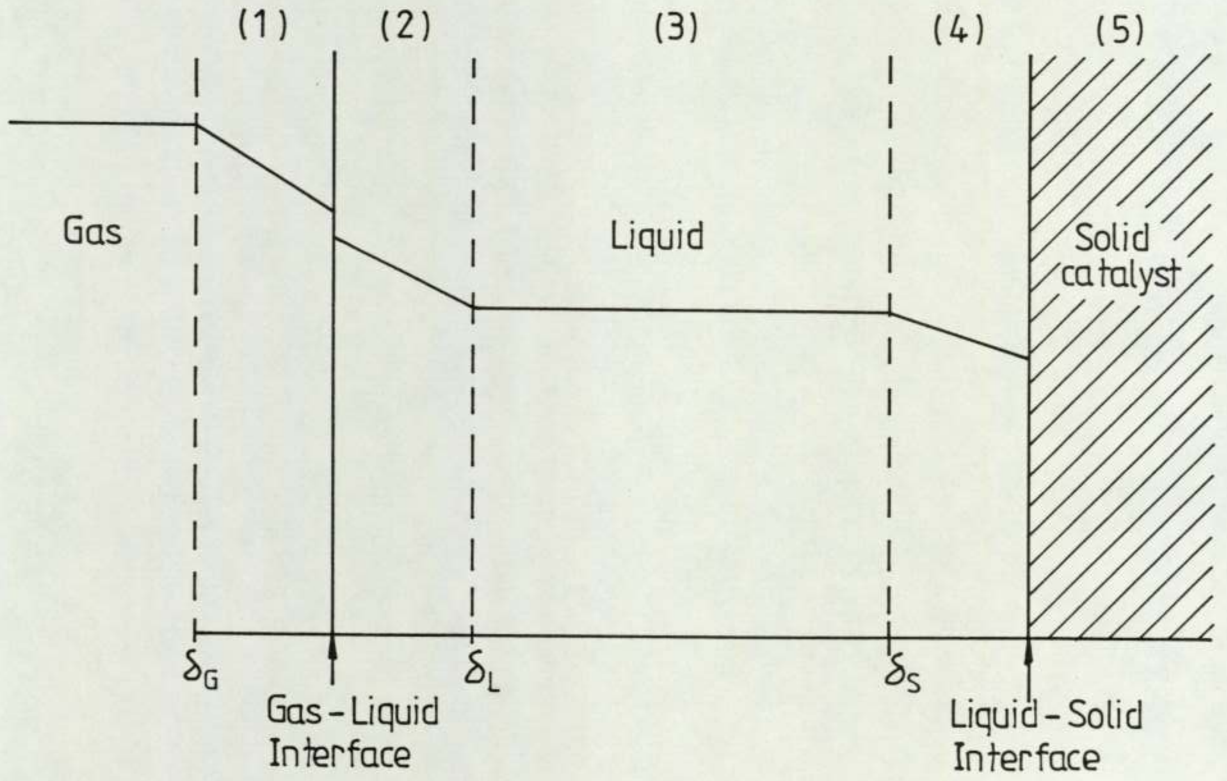
R = Rate of reaction = $K_G a_L (A_G - A_{iG})$: gas to gas-liquid interface 2.9

= $K_L a_L (A_{iL} - A_L)$: gas-liquid interface to bulk liquid

= $K_s a_s (A_L - A_S)$: bulk liquid to catalyst surface

= $K a_s A_s \eta$: reaction rate at the catalyst surface.

Figure 2.3 Concentration profile for a slurry catalysed reaction.



Henry's Law gives:

$$A_{iG} = HA_{iL} \qquad \text{........ 2.10}$$

If $R = K_o a_s A_G$

then

$$\frac{1}{K_o} = \frac{a_s}{a_L}\frac{1}{K_G} + \frac{a_s}{a_L}\frac{H}{K_L} + H\left(\frac{1}{K_s} + \frac{1}{K\eta}\right) \qquad \text{........ 2.11}$$

where

K_G = gas-side mass transfer coefficient

K_L = liquid-side mass transfer coefficient

K_s = liquid-solid mass transfer coefficient

K, K_o = kinetic constants

a_L, a_S = gas-liquid and liquid-solid interfacial areas per unit volume

A = concentration of reactant A

R = global reaction rate

η = catalyst effectiveness factors

δ_G, δ_L, δ_S = gas, liquid, solid film thickness (see Figure 2.3).

Subscripts

G, L, S = gas, liquid, solid phase

i = interface

Further discussion on this topic is given by Chaudhari et al (1980).

Heat Transfer

The gas-liquid, liquid-solid, intraparticle and wall heat transfer rates are also important. These could be derived analogously from the mass transfer rate except for the wall heat transfer rate which is characterised by a convective heat transfer coefficient between the slurry and column wall. Useful correlations have been developed for these mechanisms by authors such as Shah (1979).

2.4 Modelling Concepts

Three general types of models are used for the design and scale-up of reactors. These are:

1 TRANSPORT PHENOMENA MODELS using physiochemical principles
2 POPULATION BALANCE MODELS based on population balances
3 EMPIRICAL MODELS derived from empirical data fitting.

Examples of transport phenomena models are the phenomenological equations of change, that is to say continuum equations describing the conservation of mass, momentum and energy. Residence time distributions and other age distributions are examples of population balance models. Examples of empirical models are polynomials used to fit experimental data.

A number of models has been developed for various aspects of bubble column operation. Some of these are given in Tables 2.1 and 2.2 with their authors.

These three types of model have been used in providing a theoretical basis for this study.

Table 2.1 - Hydrodynamic Models for Bubble Columns

	MODEL	**AUTHOR**
1	Laminar Liquid Circulation and Bubble Street Model	Rietema & Ottengraph (1969, 1970)
2	Laminar Liquid Circulation and Bubble Chain Model	Crabtree & Bridgwater (1969)
3	"Gulf Stream" Model	Freedman & Davidson (1969)
4	"Circulation Flow Model"	Miyauchi et al (1970, 1979)
5	Liquid Circulation Model	Bhavraju et al (1978)
6	Energy Balance Method	Whalley & Davidson (1974)
7	Circulation Cell Model	Joshi & Sharma (1979)
8	Hydrodynamic Model for Three Phase Fluidized Beds and Slurry Reactors	Joshi (1980)

Table 2.2 - Mixing Models for Bubble Columns

	MODEL	AUTHOR
1	Mixing Model	Deckwer et al (1974)
2	Mixing Model	Baird and Rice (1975)
3	"The Recirculation Model"	Joshi (1980)
4	"Tanks in Series with Recirculation" Model	Joshi (1980)

2.5 Developments in Bubble Column Research

A comprehensive review of the literature has been carried out and a summary of this is given in subsequent chapters. The range of subjects, system parameters and operating conditions covered in recent studies (1980-86 March) of gas-liquid and gas-liquid-solid systems are given in Tables 2.3 and 2.4. This review has thus provided insight into developments in bubble column research. The keynote achievements are summarised below:

Keynotes on Recent Research

Two-Phase Systems

(a) Systems other than air-water have been given more attention; electrolytes and organics have been used due to interest in the use of bubble columns as fermenters.

(b) Studies have been made on larger columns.

(c) Exploration of higher values of superficial gas velocity has been undertaken.

(d) Loop reactors and draft tube systems have been given attention.

Three-Phase Systems

(a) Such systems have received more attention than before.

(b) A wide range of solids (other than ballotini) has been used.

(c) The effect of particle size, density and surface properties has been explored.

(d) The obvious need to study solids movement, mixing and concentration gradients (solid phase holdup and distribution) has been recognised.

TABLE 2.3 RANGES OF SYSTEM PARAMETERS FOR RECENTLY STUDIED GAS-LIQUID SYSTEMS.

REFERENCE	SUBJECT	SYSTEM COMPOUNDS	LIQUID PHASE Viscosity(cP) Density (g/cm^3) Surf Ten (dyne/cm)	COLUMN		PHASE VELOCITIES	
				Dia (cm)	Ht	Liquid (cm/s)	Gas
Hikita et al (1980)	Gas holdup Co current bubble column	Air, H_2, N_2 CO_2, CH_4, C_3H_8 Water Electrolyte Sucrose Methanol Buthanol Aniline	0.7-17.8 1.01-1.17 71.9-79.6	10	150	Batch	4.2-38
Hsu & Dudukoric (1980)	Gas Hold up Flow Regime Pressure drop Liq circulation Loop reactor	Air Water other liquids	0.8 -56 0.8-1.5 29-73	2-4.5	38-90 (H/D)	6-100	1.5-100
Kojima et al (1980)	Liq circulation Hold up	Air Water	- -	550	700	Batch	2.4-8.5
Miller (1980)	Gas hold ups Flow regimes Pressure drop Co current bubble column	Air Water (+triton) Corn Syrup (+triton)	9-49 1-1.28 35-73	22.9	914	5	30
Nakanoh & Yoshida (1980)	Gas hold ups Bubble dia	oxygen Water Sucrose CMC	1-64 1-1.23 64.4-71	14.55	190	Batch	2.78

Pal et al (1980)	Pressure drop Hold up	Air Water	- - -	5	200	0.02-0.065 (m^3/hr)	0.07-2.10 (m^3/hr)
Khang & Kothari (1980)	Axial dispersion Coeff.	Nitrogen Dist water	1 1 72.8	2.2	190	Batch	10-33.5 (Re)
Spedding & Ngriyen (1980)	Flow regime	Air Water	- - -	4.55		-5000kg/hr	-500kg/hr
Ueyama et al (1980)	Bubble behav	Air Water	- - -	60 550	300 1400	Batch	2-33
Ying et al (1980)	Gas holdups	Nitrogen Water Methanol (solid particles)	- - -	5.08 12.7	152	Batch	0-8.23
Field & Davidson (1980)	Mixing	Air Liquid	0.3 0.8 -	320	1900	3-4.5	4-5.5
Nishiwaki et al (1981)	Mixing	Air Water	- - -	12.2 29	300	Batch	1-20
Rice et al (1981)	Hold up Mixing	Air Water	- - -	4.4, 9.5 14.5, 29	18 (H/D)	0.36	0-8

Ulbrecht & Baykara (1981)	Mixing Hold up	Air water Carboxy-methylcullulin Polyethyleneoxide Polyacrylamide	- - -	15.2			0.6-1.41/min 0.5-1.3mm/s
Otake et al (1981)	Flow regime Hold up	Air Water Ethanol	0.88-1.0 - 4.45-7.2 (10^{-2}) pa.m	50	150	0-15	0.7-8.24
Ohsasa et al (1982)	Gas hold up Mixing	Air Water	- - -	10	82	0.2-3.3	0-10.5
Maruyama et al (1981)	Flow transition (2D column)	Air Water Glycerol Acetic Acid	- - -	30x1	130	Batch	0-25
Vermeer & Krishna (1981)	Hydrodynamics (Mass Transfer)	Nitrogen Turpentines	0.94 0.761 24.3	19	450	Batch	0-40
Gopal & Sharma (1982)	Flow pattern Liquid circ Gas hold up Pressure Drop	Air Water Ethanol CMC	0.8-9.5 - 45-72	11 i.d 20 o.d		0.1-0.8	1-9
Godbole et al (1982)	Gas hold up DGD	Air Water Glycerene CMC	1.3-246 996-1249kg/m^3	30.5	244	Batch	0-30

Schumpe & Deckwer (1982)	Hydrodynamics (mass transfer)	Air CMC	- 1.091-1.094 -	14 10.2	270 236	0.6	-18
Sachova (1982)	Mixing	Air Water	- - -	25	300	0.22-0.81	0-1.965
Miyahara et al (1982)	Gas hold-up	Air Water Ethanol	- - -	5 10	130	Batch	-50
Miyahara et al (1983)	Bubble formation	Air Water Ethanol Glycerine	1.1-147 999-1226 Kg/m^3 61-73.1	10 x 10	60	Batch	0.04-4
Kelkar et al (1983)	Gas hold-up Mixing	Air Alcohols	0.83-0.85 991.2-997.3 Kg/m^3 49-70	15.4 30	335 244	0-15 Batch	2-30 1-30
Kelkar et al (1983)	Hydrodynamics Mixing	Air NaCl $CaCl_2$ Na_2SO_4	0.976-1.61 998-1100 kg/m^3 70.5-74	15.4	325	0-15	3-30
Lewis & Davidson (1983)	Bubble Sizes Loop bubble col	Air Water	- -	45 i.d 190 o.d	450	17-68	1.7-23
Pandit & Joshi (1983)	Mixing	Air Water Glycerine Alcohols Inorganic Salts CMC & PEO	- - 47-66	15, 20 38.5, 100	2-8 H/D	Batch	1-25

Chen et al (1983)	Gas Hold up	Air Water	- -	15	100	Batch	0.6-6
Shah et al (1983)	Gas hold up (2 & 3 phases)	Air Water Alcohols NaCl (Coal/shale particles)	0.83-1.29 991-1074 kg/m^3 40-74.25	7.5	265	20-33	0.06-2.0
Walter & Blanch (1983)	Liquid Circulation Gas hold up Axial Mixing	Air Carbopol Soln Glycerol	- - -			Batch	0-4
Tsuge & Hibino (1983)	Bubble formation	Ar, He N_2, CO_2 Water Methanol Glycerine	0.874-48.5 0.997-1.22 65.1-71.8	12x12		Batch	0.1-10cm^3/s
Godbole et al (1984)	Hydrodynamics (mass transfer)	Air, O2 CMC	0.095-7.683$Pa.s^n$ 1000-1101 67.6-73.2	30.5	340	Batch	3-25
Godbole et al (1984)	Hydrodynamics (mass transfer)	Air, N2 Soltrol-130	1.3 751 kg/m3 34	30.5	340	Batch	3-24
Jones (1985)	Liquid circulation Draft tube bubble column	Air Water	1.0 1000kg/m^3 73	25		Batch	100-575ml/s
Molerus & Kurtin (1985)	Hydrodynamics	Air Water Ethyleneglycol	- - -	19	250	Batch	$10^{-3} \leq Re \leq 3000$
Kelkar & Shah (1985)	Gas Hold up Back mixing	Air CMC	0.002-0.165$Pa.s^n$ 67-71	15.4	335	3-10	3-30

Weiss et al (1985)	Gas hold up	N_2 Poly(dimethyl siloxane) (silicone fluids)	0.00092- 1.670 Ns/m^2 812-975 17.4-21.5	1.51	12.7	Batch	0-6
Devine et al (1985)	Axial mixing	Air CMC	30-115 987-999 678	15	335	1.5-30	7-41
Koide et al (1985)	Gas hold up (mass transfer) -effects of surface active substances	Air Distilled water n-hexanol(aq) n-heptanol " n-octanol " antifoam agent	$8.937x10^{-1}$ $997.0kg/m^{-3}$ 58.2-71.96	20x15x150		Batch	0.30-2.07
Yamashita F (1985)	Gas hold-up Effect of liq depth column Inclination & baffle plates	Air Tap water	- - -	16 8	270 350	Batch	1-21
Merchuk JC (1986)	Gas hold-up Liquid velocity	Air He, CO_2, Ar Freon 114 Tap water Ag Na_2So_4 Ag glycerine	- - -	15 x 1.5 x 250		Batch	0.2-5
Tinge & Drinkenburg (1986)	Liquid dispersion Gas hold-up Effect of non-verticality	Air Tap water	- - -	2.2, 5.8, 10.3	150	Batch	0-5
This study	Gas hold-up, Flow regimes Bubble dynamics Mixing	Air Water Aq Alcohols Ketones, Esters Electrolytes	1.2-1.46 998-1009 Kg/m^3 53-76	15.3 x 1.3x143 15.2	300	0-1 Batch	0-35

TABLE 2.4: RANGES OF SYSTEM PARAMETERS FOR RECENTLY STUDIED GAS-LIQUID-SOLID SYSTEMS

Reference	Subject	System Compounds	Liquid phase viscosity (cP) Density (g/cc) Surface tension (Dyne/cm)	Solid size (mm) Density (g/cc) Concentration	Column		Phase Velocities	
					Diameter (cm)	Height	Liquid/ slurry	Gas
Vasalos et al (1980)	Bed contraction Gas hold-up Bubble rise vel Ebullated bed	Hydrogen Kerosene Mineral oil Coal char CoMo H-coal catalyst	14.9 -	1.6 x 5.1 - 17.8 (vol % coal char in kerosene)	15.2	600	3	3-4.6
Ying et al (1980)	Phase hold-ups Mixing	Gas - N_2 Liquid - water methanol Solid - silica particles	- - -	<100 μm - 0-0.2 g/cm^3	12.7	152	0-1.52	1.1-11.2
Deckwer et al (1980)	Hydrodynamics	N_2 Molten parafin Powdered Al_2O_3	2-13 0.67-0.73 21.2-29.1	≤5 μm - 16 wt%	4.1, 10	-	Batch	<4
Dhanuka & Stepanek (1980)	Gas & liq Hold-up Pressure drop	$Co_2/N_2/O_2$ Water + HCO_3/CO_3 = Glass beads	- - -	2-5.9 2.96 -	5 -	750	6.1-13.8	0-8
Kusakabe et al (1980)	Hold-ups	N_2 Water (Cu SO_4) Cu beads	- - -	0.38-0.67 - -			1.1-3.1	1-5
Euzen et al (1981)	Gas hold-up	Air Water Alumina beads	- - -	1.4 950 -	15	900	0.5-2	0-14

Hassanien et al (1981)	Expansion - Contraction Bubble diameters	Air, N_2 Water (NaOH, Fe CN) Glass/plastic beads	- - -	5.1-10.4 1.34-2.51 -	9.4	-	0-20	0-16
Kato et al (1981)	Liquid Hold-up	Air Water(CMC) Glass beads	- - -	0.42-2.2 2.5 -	5.2 12.0	150	0.8-7	2-10
Linneweber & Blass (1981)	Gas & solid Hold-ups	Air Water Polystyrene	- - -	0.16-2.0 1.02-11.8 -	19	-	-	1-5
Sisak et al (1981)	Hold-ups	Air Water Sand	- - -	0.315-1.25 2.65 -	10.6	-	0.2-3.78	0.16-1.1
Alvarez-Cuenca et al (1981)	Mixing	Air Water Glass beads	- - -	1-5 - -	66 x 2.5	25	2.5-10	4-28
Kim & Kim (1981)	Mixing	Air Water (CMC, glycerol, methanol triton)	- - -	1.6 - 6 2.50 -	1.45	250	2-13	0-12
Kojima & Asano (1981)	Concentration distribution Bubble diameter Gas hold-up	Air Water-glycerol Glass Acryl	1-18.8 1-1.18 -	100, 159 μ 1.2-2.49 0.324 kg/m^2	5.5, 9.5	150	Batch	1-8
Kato et al (1982)	Concentration distribution	Air Water Glass	- - -	0.137 2.52 -	12.2 21.4	197 405	Batch 1.01-5.08	5.07-11.7 5.06-5.15

Kara et al (1982)	Hydrodynamics Axial mixing	Air Water Coal & dried Mineral ash	0.854-7.547 1.0-1.1 -	10-70µm 1.30 0-40% wt	15.2	335.3	0-10	3-30
Godbole et al (1983)	Hydrodynamics (mass transfer)	Air Water, CMC Polystyrene	- 1000-1009 Kg/m^3 68.2-74.5	33-650 µm 1.06-2.73	30.5	340	Batch	3-30
Kulkarni et al (1983)	Hydrodynamics (mass transfer)	Air Water Triton 114 Glass bead/ coal & shale particles	0.8-1.35m Pas 1000-1270 Kg/m^3 0.043-0.077	30-265µm 1.29-2.66 4-14.72% w/w	7.5	265	19.5-36	0.2-2.5
Kelkar et al (1984)	Hydrodynamics Mixing	Air Water (CMC, Triton) Polystyrene beads & oil shale particles	0.001-0.007 Pas 0.997 0.040-0.073	40-300µm 1-2.3 0-15% Wt	15.4	335	3-10	3-25
Sangnimnuan et al (1984)	Gas hold-up Backmixing	H_2, N_2 Tetralin Coal	- - -	- - 0-40% w/w	1.9	222	0.1-0.3	0.2-1.1
Koide et al (1984)	Gas hold-up	Air, N_2 Water (Glycerol, Glycol, BaCl, Na_2SO_4)	0.894-17.6 997-1178 59.5-73	47.5-192µm 2.5-8.77 0-200Kg/m^3	10, 14, 21.8, 30	300	-	3-15
Capuder & Koloini (1984)	Gas hold-up (Interfacial area)	CO_2, Air Water $CaCO_3$/ $Ca(OH)_2$	- - -	10-15µm - 0-18.8 Wt%	19	114	-	4-14

Van der Meer et al (1984)	Axial Solid dispersion	Air Water Ion-exchange particles	- - -	0.519-0.713 1196.8-1375.1 Kg/m^3 -	2-6	100	0.3-2	-
LS Fan et al (1984)	Solids mixing	Air Water Glass particles	- -	3-6 2.52 0.7-1.54 Kg	7.62	273	8.28-16.9	0-12.4
Kojima et al (1984)	Solid hold-up	Air Water Glass beads	- - -	0.105-0.177 - 16-274Kg/m^3	5.5 9.5	150	0.087-0.26	3-10.5 3-8.3
Smith & Ruether (1985)	Solid dynamics	N_2 Water Ethanol Glass spheres	- - -	44-210μm 2.42-3.99 120-150Kg/m^3	10.8	194	0-2	3-20
Sekizawa & Kubota (1985)	Solid concentration	Air Water Glass beads	- - -	1.06-1.53 2.55 4-400Kg/m^3	10 20	100, 200, 60, 80	0-2	1.7-10.8
Kato et al (1985)	Axial hold-up (gas and solids)	Air Tap water Water & porous alumina Glass sphere	- 1.8 Mg m^{-3} (slurry) -	0.52, 2.2, 5.2 2.52 Mg m^{-3} 0.4	12	216, 266	0-30	2-20
Kato et al (1985)	Longitudinal dispersion coefficient	Air Water Glass Porous alumina	- - -	0.52-5.2 1.7, 1.8, 2.5 Mg m^{-3} -	12, 19	250	0-25.3	2-20

This study	Gas hold-up	Air	1.19-1.48	0.40-1.23	15.2	350	0-1	0-35
	Flow regimes	Water	1000	1-1.23	15.3x1.3x143		Batch	
	Bubble dynamics	Aq Alcohols	1008 Kg/m^3	0-15% Vol				
	Mixing	Ketones, Esters, Electrolytes	48-58					
		Styrocell Ballotini Amberlite resin Diakon						

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CHAPTER 3

DESIGN OF EXPERIMENTS, EQUIPMENT AND ANALYSIS

DESIGN OF EXPERIMENTS, EQUIPMENT AND ANALYSIS

DESIGN OF EXPERIMENTS, EQUIPMENT AND ANALYSIS

3.1 Introduction and Design of Experiments

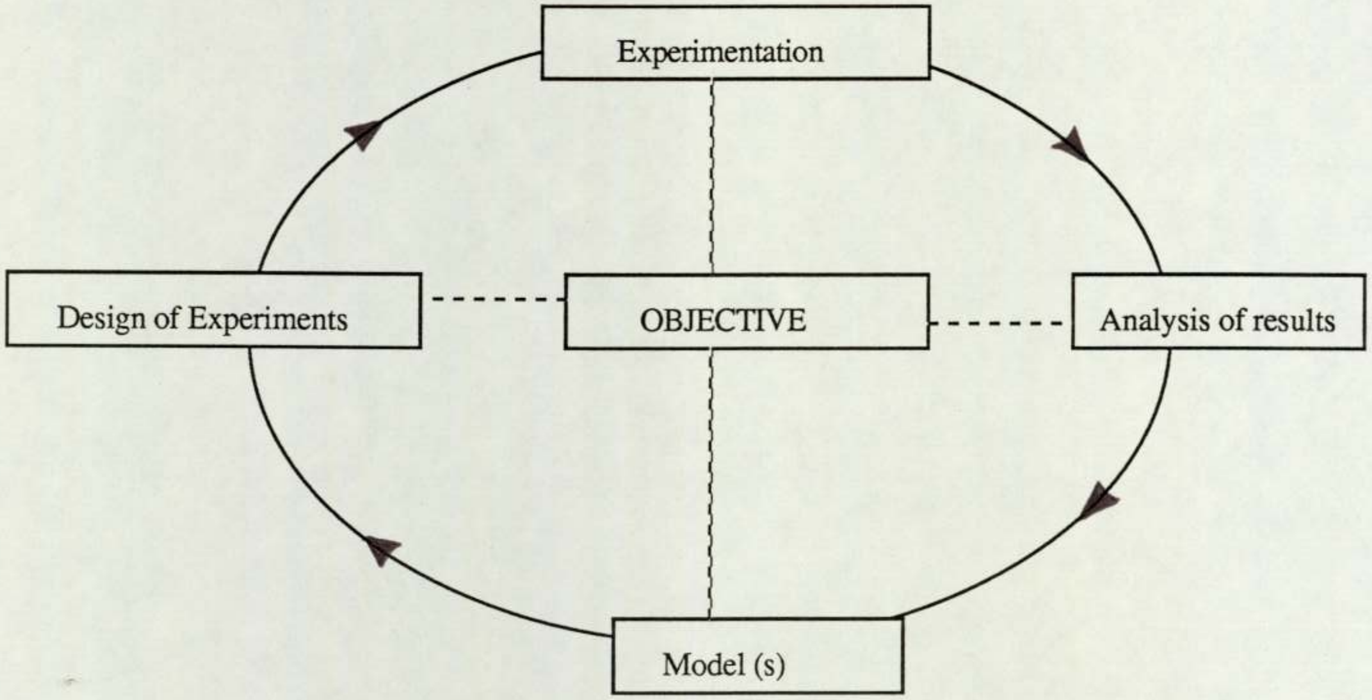
A general strategy for experimental work is shown in Figure 3.1. The design of experiments, experimentation and analysis of results are centered around the objectives of the research. The outer loop is followed until, in the judgement of the researcher, the objectives have been met.

When designing and analyzing experiments, it is necessary to have a clear idea in advance, of exactly what is to be studied, how the data are to be collected, and an understanding of how the data are to be analysed. Thus the procedure to be followed will involve:

1 Recognition and definition of the problem.
2 Choice of factors (independent variables).
3 Selection of the response (dependent variables).
4 Designing the experiments.
5 Performing the experiments.
6 Data analysis.
7 Preparation of conclusions and recommendations.

During the first stage of the experimental programme, work was carried out with the relatively simple air-water system. Experiments were undertaken to find the effects of operating parameters and system design on gas holdup, flow regime and bubble

Figure 3.1 - Strategy for Efficient Experimentation



dynamics. This was followed by a study of axial mixing in the liquid phase. The effects of soluble additives were then examined, the purpose being to check and extend the work of previous researchers.

In the second stage of the work, three-phase systems were investigated (see also Section 3.2). This involved the addition of solid particles to the system with a view to simulating the effects of microbial aggregates in fermentation processes. Other solid particles were used after consideration of other processes in which bubble columns can be employed. This was followed with studies of the effects of water-soluble additives and of axial mixing in the liquid phase.

All experiments were carried out using either a two-dimensional (2-D) or a three-dimensional (3-D) bubble column (see also Section 3.3). The terminology "2-D" is used to describe a thin rectangular column, while "3-D" refers to a vertical cylindrical column. Experiments were run both continuously and semi-continuously (using a batch of liquid). The mode of fluid flow was always co-current.

3.2 Choice of Materials

Solids

An important general classification of particles is:

(i) wettable and
(ii) non-wettable.

This stems from a consideration of the surface properties of particles. Other general properties considered are: shape, size, uniformity of size and density.

In the choice of solid particles, the main consideration was simulation of typical chemical engineering processes, bearing in mind the application of bubble columns in the chemical, biochemical, pharmaceutical and petroleum industries. The particles used were spherical, of uniform size and were: Styrocell, Ballotini, Diakon and Amberlite.

Liquid Additives

Both organic and inorganic compounds were chosen to study the effect of various constituents in process systems. The additives used were: alcohols, ketones, esters and electrolytes.

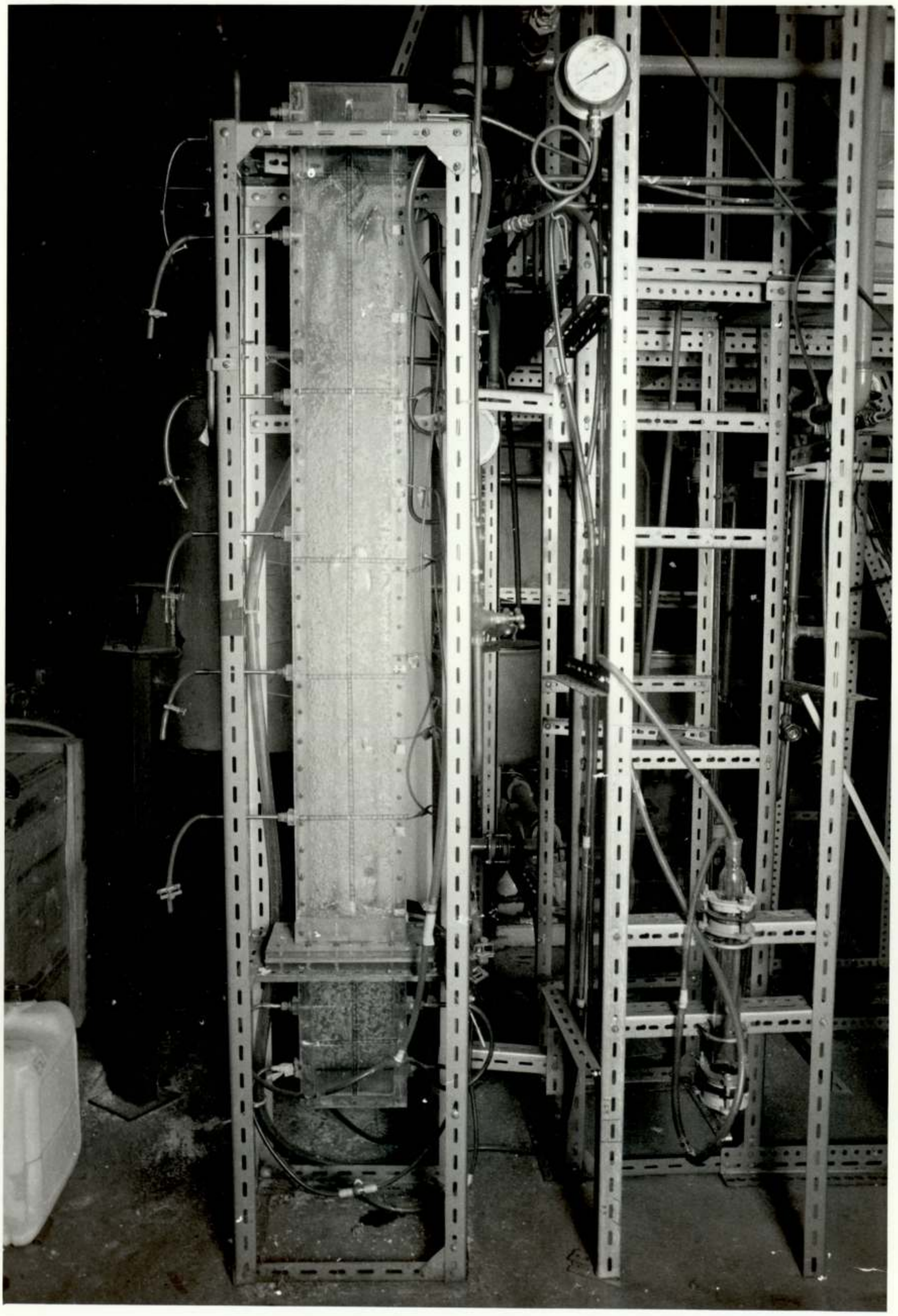
3.3 The Experimental Equipment

Experimental studies were conducted in a 2-D bubble column of dimensions 1.3cm x 15.3cm x 143cm and a 3-D bubble column 15.2cm in diameter and of variable height.

3.3.1 The Two-Dimensional (2-D) Bubble Column System

The 2-D column was used because it enabled clear visual and photographic observations to be made. The column (see Figure 3.2) was constructed from perspex and has the vertical faces glued and screwed together; the distributor section and bed section are bolted together using flanges. The column has the following distributor arrangement:

Figure 3.2 General view of the two-dimensional bubble column



- a 0.2cm thick copper plate drilled with 0.1cm holes on a 0.6cm square pitch, and
- a 100 mesh wire gauze.

This combination provides reasonably uniform gas distribution under most operating conditions.

A flow diagram of the system is presented in Figure 3.3. Water directly from the mains supply (1) is introduced into the column through valve A using rotameter R1 (Type 7S). Mains compressed air (2) is available at 50 psig and is reduced, using a pressure regulator to 10 psig. Air is then passed into the column through rotameter R2 (Type 10A). Calibrated rotameter charts were used to determine flow rates.

An arrangement to measure pressure drop is also available. Eight tappings are located at the side of the column and these are connected to a common junction. Tappings can be isolated using clips such that only one is in use at any time. A tapping just above the support screen is used as the other arm of the manometer.

3.3.2 The Three-Dimensional (3-D) Bubble Column System

Experimental studies were also carried out in a vertical, cylindrical column of variable height and of 15.2cm diameter. The column (see Figure 3.4) is made of standard lengths of QVF 15.2cm bore pipe. The lowest section has a side-arm which is used for introducing the liquid. The top section of the column consists of a reducer with side outlet; any excess liquid (or solid) is usually washed out through this outlet.

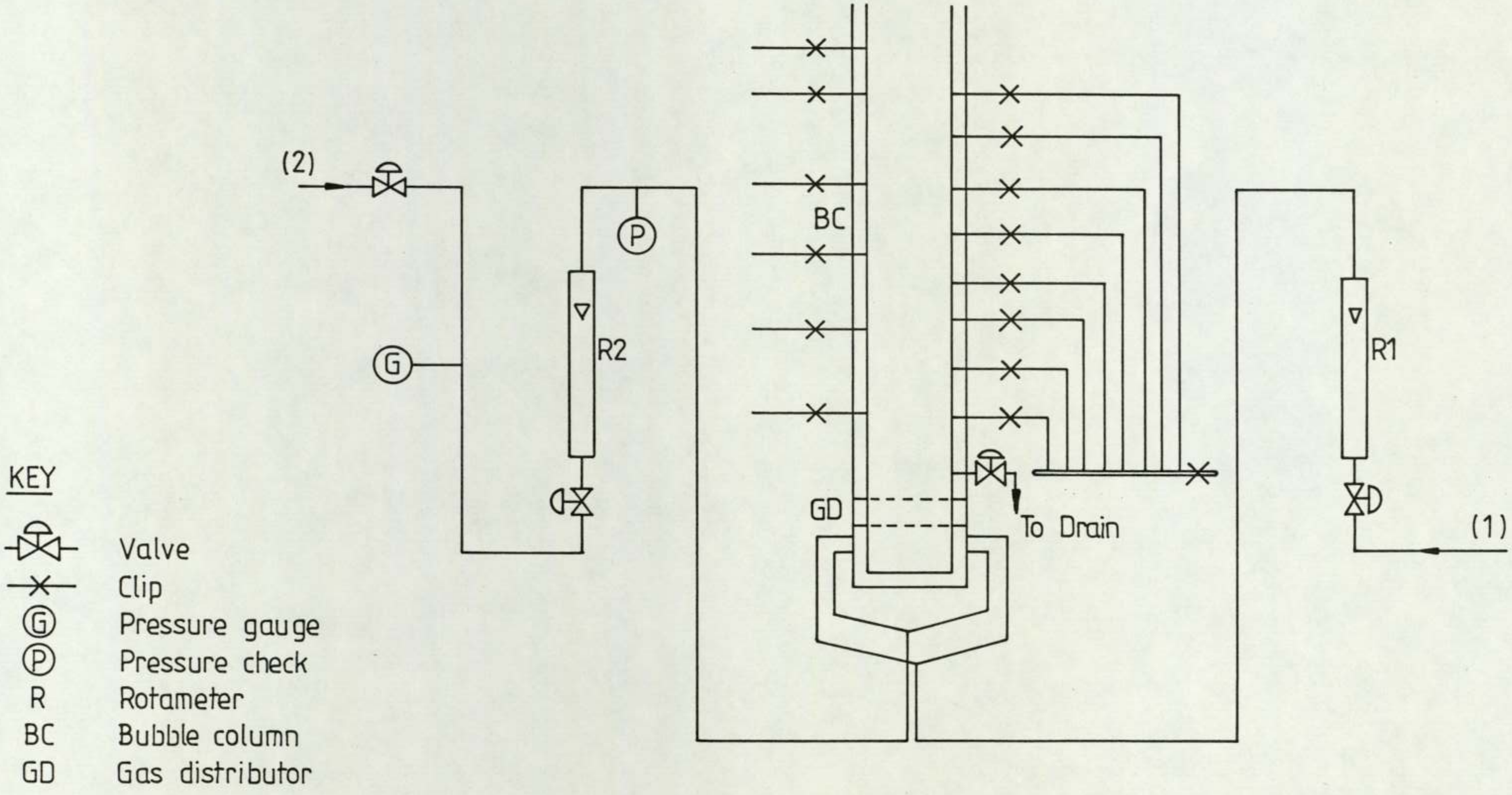
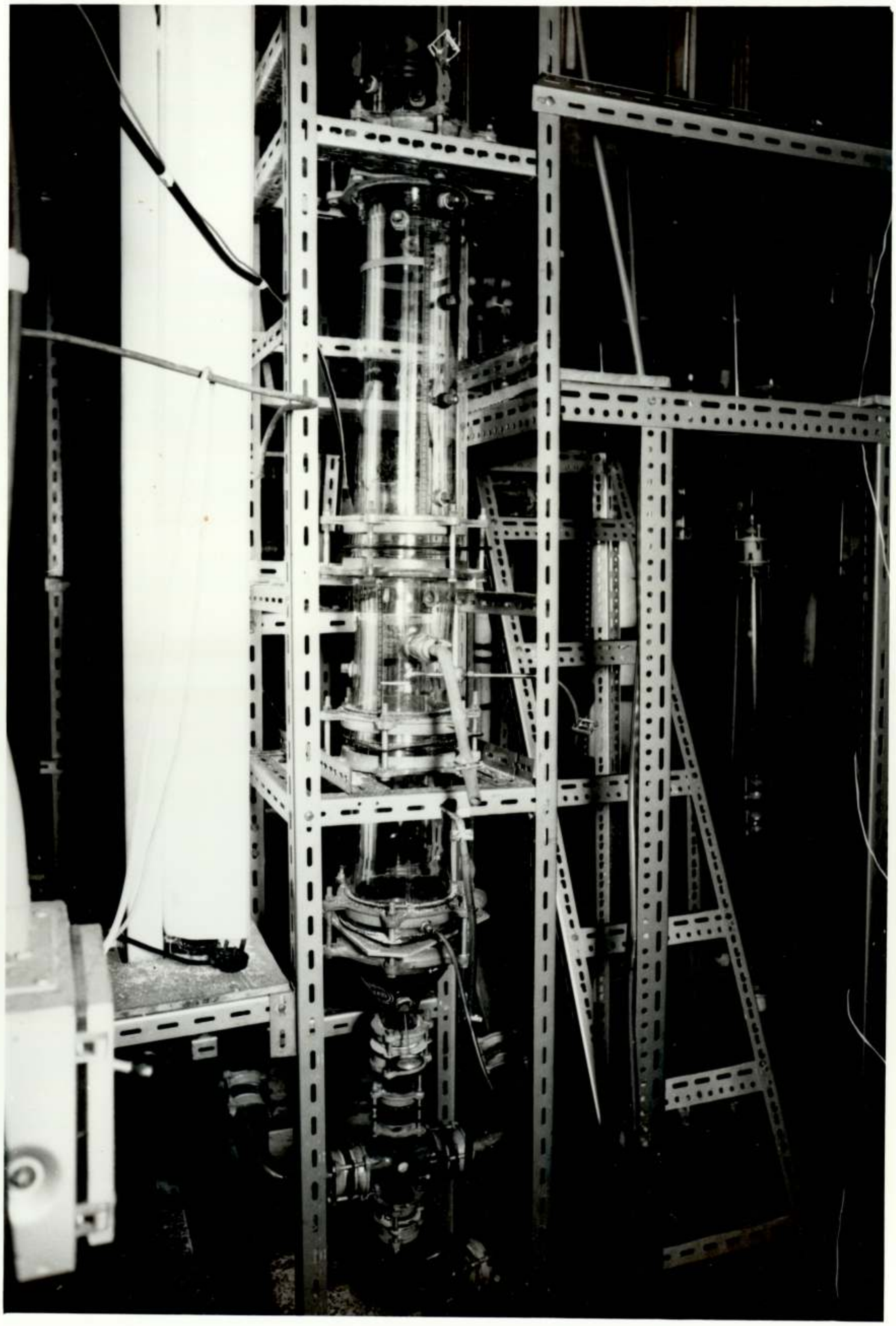


Figure 3.3 Flow diagram of the two-dimensional bubble column rig.

Figure 3.4 General view of the three-dimensional bubble column (Lower Section)



The general layout of the column and associated equipment is as shown in Figure 3.5. Water is introduced into the column from the mains supply (1) by opening valve D: rotameter R3 (Type 1OS) is used for metering the flow. The gas supply is obtained from a compressed air service main (2): this is fed through a pressure regulator to the metering section via a control valve E. The pressure is measured by means of a calibrated pressure gauge P. The metering section consists of two rotameters R4, R5 (Types 14A and 24A). Rotameter charts were used to convert experimental readings.

The gas distributor (see Figure 3.6) consisted of a circular plastic plate of the same diameter as that of the column and drilled with 61 holes of 0.75mm diameter on a 17.4mm triangular pitch.

To measure the hold-up (based on the manometric method) in any part of the column, 1.3cm diameter holes have been drilled along the length of the column. The holes are supplied with fittings so that 3.2mm O.D stainless steel tubes can easily be inserted into the column. These tubes are then connected by means of flexible PVC tubing to vertical glass tubes mounted at the top of the column. Besides each glass tube a self-adhesive downward scale has been fixed.

3.4 Measurement Methods

3.4.1 General Measurements

Extensive experiments have been performed following a very systematic procedure. Standard techniques of measurement were used and results cross-checked with other

Figure 3.5 Flow diagram of the three-dimensional bubble column rig.

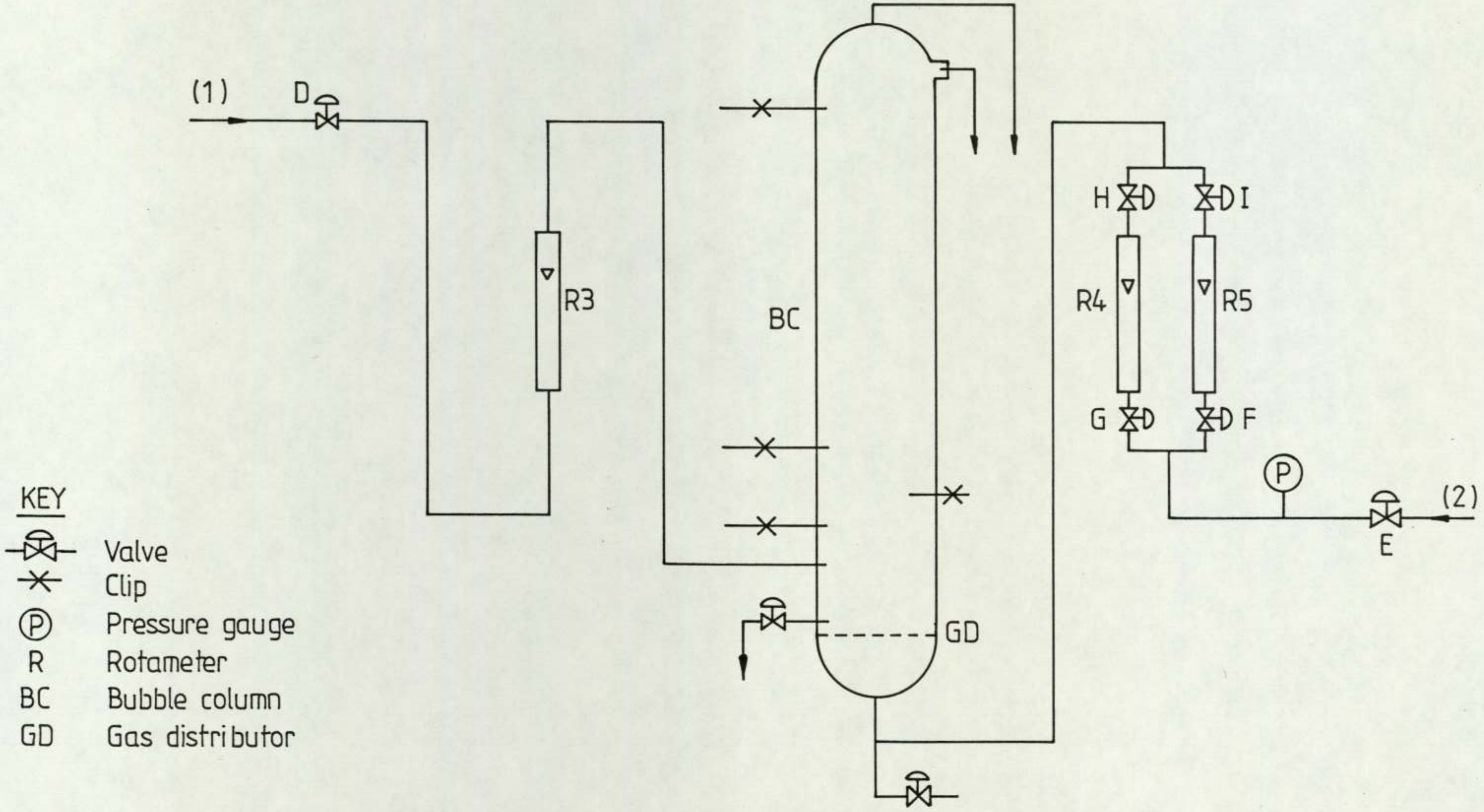
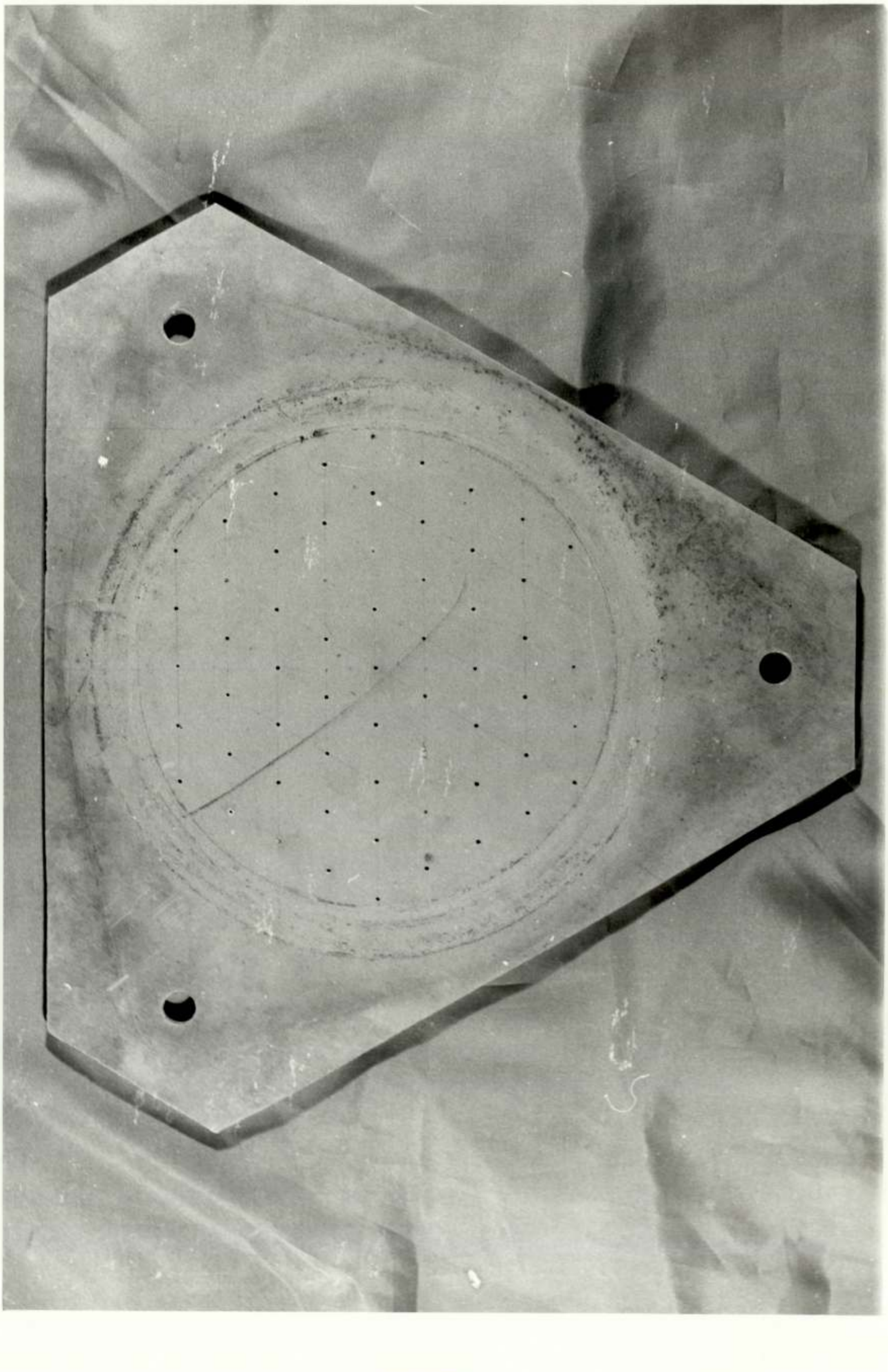


Figure 3.6 The three-dimensional column gas distributor



methods. Experiments were often repeated to check the consistency of results. Details of the measurement methods and experimental procedures are given in each relevant chapter.

3.4.2 Measurement of Physical Properties

The physical properties of all systems used were measured:

Density - a density bottle was employed in measuring density using the displacement technique.

Viscosity - viscosity was measured using the Haake RV Automatic Viscometer. This is shown in Figure 3.7.

Surface Tension - the Torsion Balance (see Figure 3.8) was used in measuring the surface tensions of the systems studied.

3.5 Analysis of Results

The methods of analysis are described in detail in each chapter. The general approach was first to make plots of experimental data and second to search for quantitative relationships between variables.

Graphical Method

This is the first method of analysis. For accuracy and neatness, a computer with graph plotting package (GINOGRAPH) was used for all the analysis. Graphs provide a qualitative view of relationships between the factors and variables studied.

Figure 3.7 The Haake RV12 Automatic Viscometer

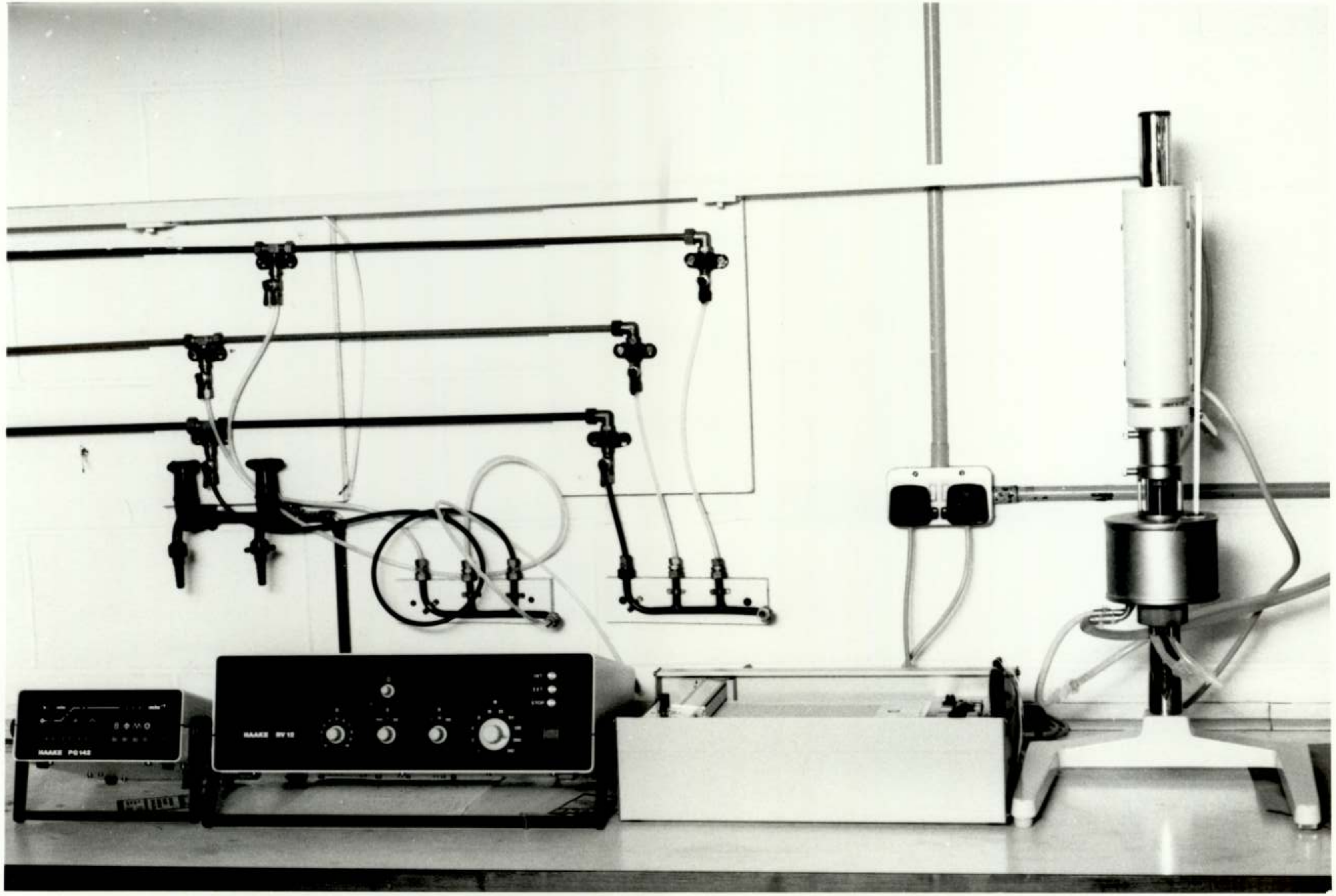
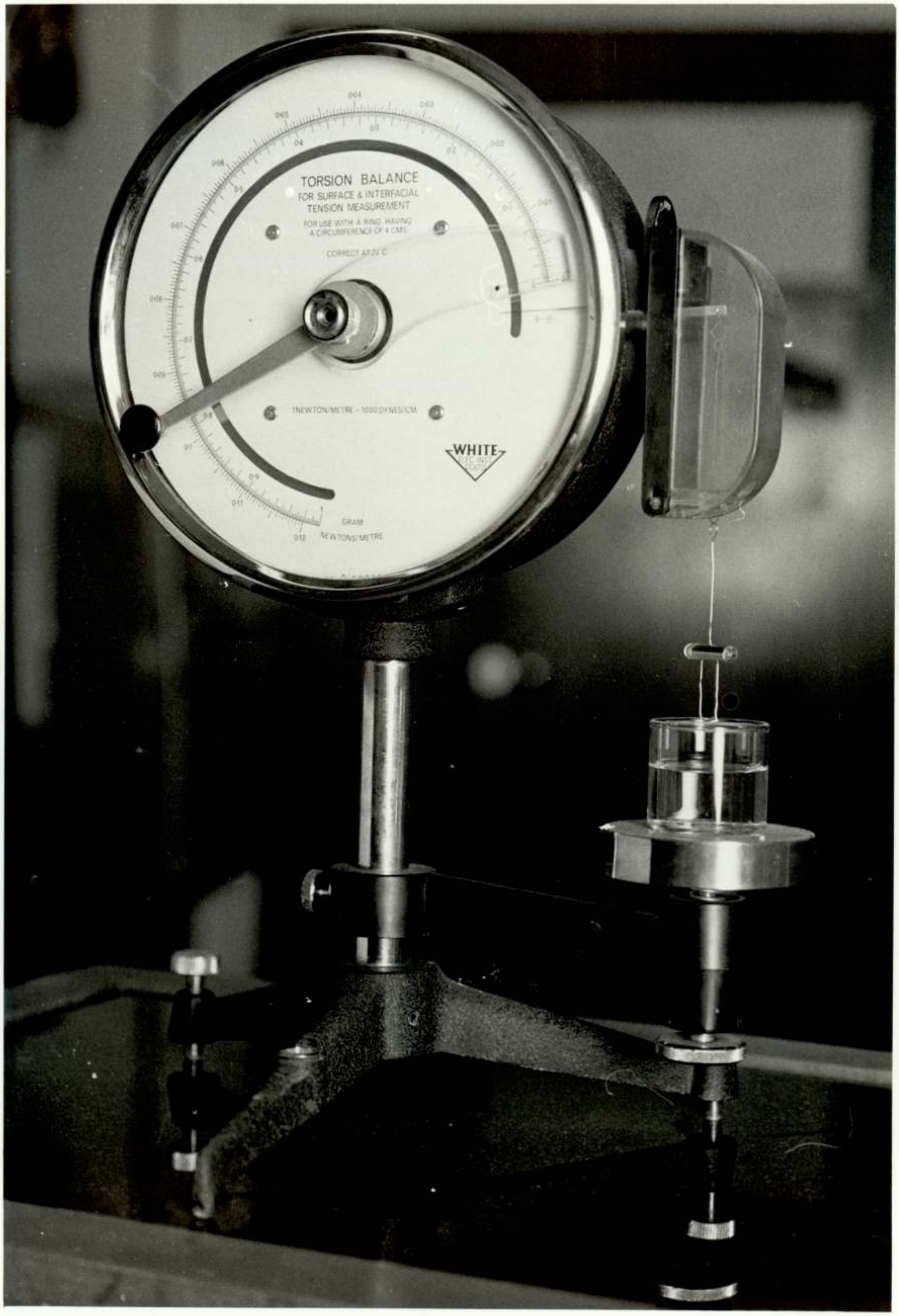


Figure 3.8 The Torsion Balance



Modelling and Correlation

These involve quantitative analysis of results. Existing models and correlations were tested and some new correlations are presented.

CHAPTER 4

GAS HOLD-UP IN TWO-PHASE SYSTEMS

GAS HOLD-UP IN TWO-PHASE SYSTEMS

GAS HOLD-UP IN TWO-PHASE SYSTEMS

4.1 Introduction

Gas hold-up is one of the most important parameters characterizing the performance and hydrodynamics of bubble columns. It is defined as the percentage by volume of gas in the two- or three-phase mixture in the column. Knowledge of gas hold-up makes it possible to estimate bubble residence time and, in conjunction with the mean bubble diameter, the interfacial area can be calculated. This is useful for calculating mass transfer coefficients. Gas hold-up is also used with other parameters for calculating mixing coefficients and chemical reaction rates.

Average gas hold-up, ε_G, is obtained as the volume fraction of gas within the total volume of the system.

$$\varepsilon_G = \frac{\text{Volume of gas}}{\text{Total volume}}$$

$$= \frac{(\text{Total volume}) - (\text{Liquid volume at rest})}{(\text{Total volume})} \qquad \text{........ 4.1}$$

4.2 Literature Review

From a survey of the literature, it is evident that there is an extensive amount of information on gas hold-up behaviour in bubble columns. Unfortunately, the results from many research programmes are only relevant to a particular situation since few attempts have been made to explain the observations from a fundamental viewpoint.

The divergence in published results (Figure 4.1) provides a point for concern and this has influenced the presentation of this literature review. Consideration of the factors affecting gas hold-up is followed by a description of the techniques of measurement; it is thought that such an approach will provide valuable information and a better appreciation of bubble column behaviour.

4.2.1 Factors Affecting Gas Hold-up

From previous studies, these can be classified into three groups, namely:

(a) operating parameters
(b) column design and
(c) the nature of the phases (physico-chemical properties).

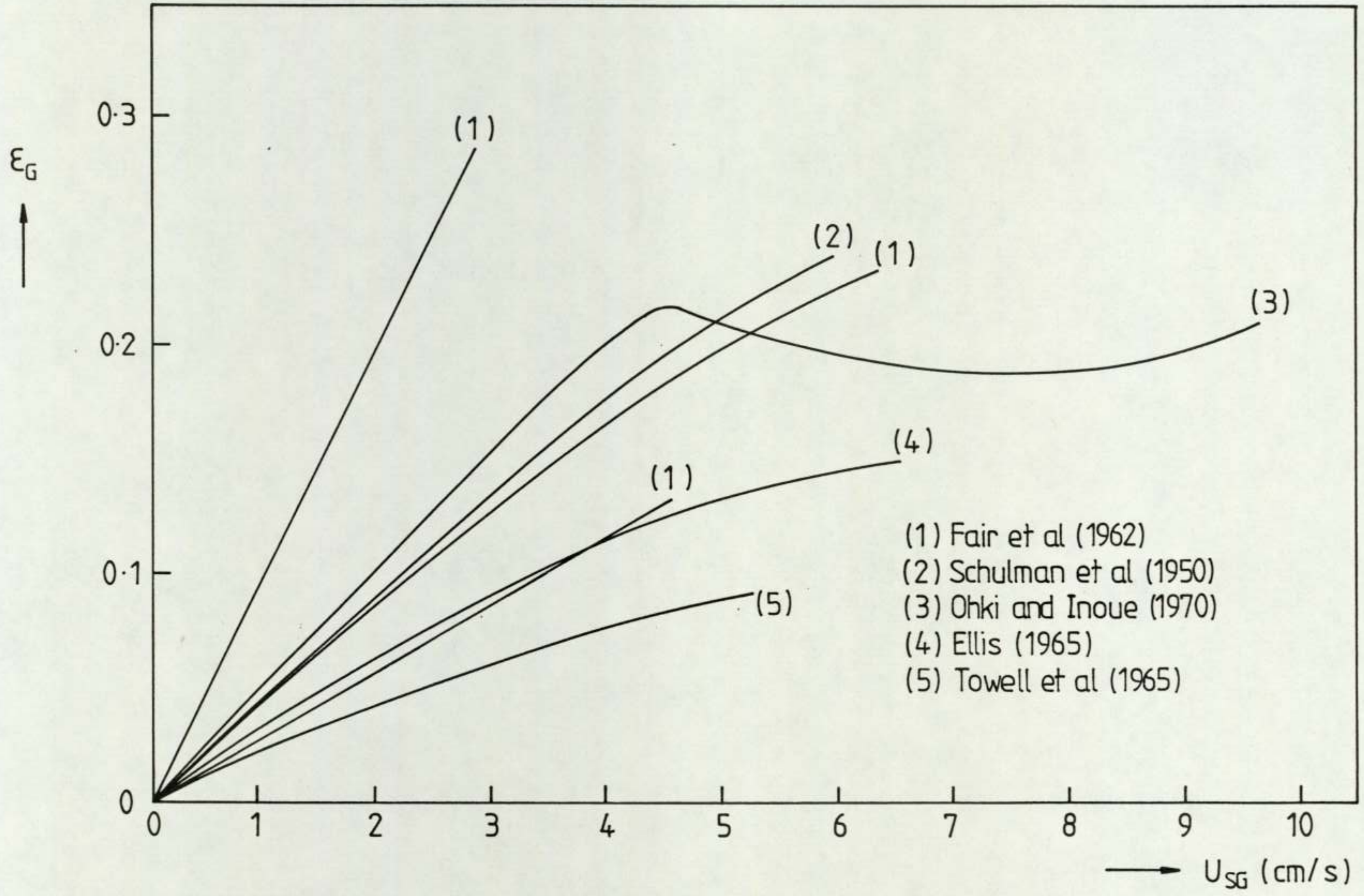
(a) Effect of Operating Parameters

The effect of operating conditions on gas hold-up has been found to be significant, the two most important parameters being the gas and liquid flow rates. These are usually presented as superficial gas and liquid velocities respectively, using the column cross-sectional area as the basis.

Effect of Superficial Gas Velocity (U_{SG})

The hold-up in bubble columns is primarily a function of superficial gas velocity and published work in this area is plentiful. Figure 4.1 is a comprehensive picture of the

Figure 4.1 Gas hold-up values reported in the literature.



hold-up behaviour for air-water systems as observed by various workers. Reference to this Figure shows that the hold-up is nearly a linear function at low superficial gas velocities, but there is a falling off in the rate of increase of hold-up with gas velocity at the higher gas rates. A log-log plot of gas hold-up against gas velocity measured by various authors for air-water system is given by Shah et al (1982). The dependence of gas hold-up on gas velocity is generally of the form:

$$\varepsilon_G \propto U_{SG}^n$$

The value of n usually depends on the flow regime: for the homogeneous regime, the value of n varies from 0.7 to 1.2 (see Reith et al (1968), Schugerl et al (1977), Deckwer et al (1980)). In the heterogeneous regime, the effect of U_{SG} is less pronounced and n takes values from 0.4 to 0.7.

<u>Effect of Superficial Liquid Velocity</u> (U_{SL})

There are discrepancies surrounding the experimental results on the effect of liquid flow rate on gas hold-up. Akita and Yoshida (1973) found that for sodium sulphite-air systems, the effect is negligible for superficial liquid velocities up to 4.5 cm/s whether flow is countercurrent or cocurrent. In the same range of superficial liquid velocities, Argo & Cova (1965), Bischoff & Phillips (1966), Kato & Nishiwaki (1972), Reith et al (1968), Schugerl & Lucke (1977) found that there is no marked effect of liquid flow rate on gas hold-up.

Shayegan Salek (1974) suggested that increase in liquid phase velocity cause a quicker wash-out of the gas phase with a consequent reduction in gas hold-up. Ostergaard & Michelsen (1968), Hills (1976) and Downie (1972) also reported that an increase in liquid velocity caused a reduction in gas hold-up.

(b) Effect of Column Design

Under this sub-section, the effects of column geometry-height and diameter - and distributor design will be considered.

Column Height

Yoshida & Akita (1965) found that column height does not have a marked effect on gas hold-up. However, they suggested that for heights less than 100cm, end-effects might have an important influence on results. This was supported by Towell et al (1965). Fair et al (1962) concluded that, although the local value of gas hold-up can vary somewhat with height, the dependence of average gas hold-up on height is not marked; this was confirmed by Bhaga & Weber (1972).

Deckwer et al (1974), who carried out experimental studies on columns 440 and 723cm in height, concluded that liquid height does not affect the gas hold-up.

Column Diameter

The effect of column diameter on hold-up has also been a subject of considerable interest. Fair et al (1962) and Yoshida & Akita (1965) found no effect of column diameter when this exceeded 15cm but a slight decrease in hold-up was reported by Yoshida & Akita for the case of a 7.5cm column.

Shulman & Molstad (1950) found that changing from a column diameter of 5cm to one of 10cm had no effect on gas hold-up. However, a slight increase was observed when using a 2.5cm diameter column at fixed gas velocity. Reith et al (1968) observed much lower gas hold-ups for larger columns.

Ellis & Jones (1965) suggested that wall-effects increase the value of gas hold-ups in columns less than 7.5cm; this has been confirmed by Freedman & Davidson (1969).

Downie (1972) and Shayegan Salek (1974) also studied the effect of column diameter on gas hold-up. Four different columns of diameter 7.6, 15.2, 30.5 and 61cm were used. They concluded that as the column diameter increases, the gas hold-up decreases, at a fixed value of superficial gas velocity. More work on the effect of column diameter has been reported by Kato & Nishiwaki (1972), Argo & Cova (1965), Okhi & Inoue (1970) and Hikita et al (1980).

Distributor Design

Different investigators have reported varying views on the effect of gas distributor design on gas hold-up. Freedman & Davidson (1969) measured gas hold-up using different spargers. They examined the influence of distributor geometry. As compared to a distributor with large holes, spargers with small holes generated a higher hold-up for the same gas velocity. At higher gas rates, the trend in hold-up became similar although there was a small difference in the two hold-up values. Towell et al (1965) found that distributor design had little effect on gas hold-up. However, their "two-phase distributor" gave a higher hold-up at high liquid flow rates. This was because the increased input energy of pumping through the orifice produced smaller bubbles in the region close to the distributor.

Reith et al (1968) on examining several types of air distributors - single tubes, fine gauzes and perforated plates with different numbers of holes of various diameters - reached the conclusion that the gas hold-up is unaffected by changes in orifice geometry. Kato & Nishiwaki (1972) found that on decreasing the hole diameter of the gas distributor, the size of the bubbles generated decreased and gas hold-up increased.

Ellis & Jones (1965) found that perforated baffles placed inside a bubble column increased the gas hold-up significantly. This was confirmed by Fair (1967) who reported that when the baffles were vibrated, greater hold-ups could be obtained at any given superficial gas velocity.

Bischoff and Phillips (1966) also looked at gas hold-up in a column containing a number of perforated plates. They observed that the gas hold-up tended to level out at about 0.35 at a gas velocity of 18 cm/s and that the plate spacing had only a slight effect on gas hold-up.

(c) Effect of Physico-Chemical Properties of the Phases

The nature of the gas and liquid phases can also affect gas hold-up.

Gas Properties

Data showing the influence of gas properties are scarce and often lead to conflicting conclusions. Bhaga et al (1971) and Koetsier et al (1976) concluded from the results of their experiments that an increase in the gas density results in increasing gas hold-up. Akita & Yoshida (1973) and Shulman & Molstad (1950) reported that the nature of the gas has no effect. Hikita et al (1980) mentioned that the effects of gas density and gas viscosity are not highly significant.

Jamialahmadi (1982) using different gases found that the nonpolar ones (O_2, N_2) had a lower hold-up than the polar ones (CO_2, NH_3) with water as the liquid phase.

Liquid Properties

The effects of surface tension, viscosity and density are considered in this sub-section.

Surface Tension

The effect of surfactants is not clearly understood. Many investigators have tried different surfactants with varying conclusions. While Botton et al (1978) and Miller (1980) reported no effect of surface tension, Schugerl et al (1977) and Todt et al (1977) observed a significant increase in the gas hold-up with a decrease in the surface tension. Recently Oels et al (1978) carried out experiments with dilute alcohol solutions and found a significant increase in the values of gas hold-ups. Bach & Pilhofer (1978) observed no effect of surface tension on gas hold-up for pure liquids but they noted differences when liquid mixtures and electrolyte solutions were used. The addition of electrolytes increased the gas hold-up as noted by Akita & Yoshida (1973), Hikita et al (1980) and Freedman & Davidson (1969). Freedman & Davidson observed that the addition of electrolytes postponed the appearance of large bubbles in the column. Electrolytes and/or alcohol additives probably stabilize the bubble size by formation of an ionic and/or polar double layer at the interface which depresses the coalescence rate.

Viscosity

An increase in viscosity in the case of Newtonian liquids normally decreases the gas hold-up. Most of the work has been carried out with glycerol solutions or sugar solutions. Many investigators have reported a maximum with respect to viscosity near the vicinity of three centipoises (see Eissa & Schugerl (1975), Bach & Pilhofer (1978) and Buchholz et al (1978)). This has been explained on the basis of hindered gas bubble motion in the viscous fluids. At relatively low viscosities, drag forces are not large enough to cause bubble coalescence; also there is a more uniform distribution of bubbles and hence higher hold-ups.

Nishikawa et al (1977) and Nakanoh & Yoshida (1980) have reported hold-up values in non-Newtonian media. They observed higher values of gas hold-up than those predicted by Akita & Yoshida's correlation (1973).

Liquid Density

The effect of liquid density is also not clearly understood. Whereas Akita & Yoshida (1973), Hughmark (1967), Kumar et al (1976) observed a decrease in the gas hold-up with an increase in density, Hikita et al (1980) and Oshinowo & Charles (1974) observed an increase in the gas hold-up with an increase in the density.

4.2.2 Techniques for Measuring Gas Hold-up

Table 4.1 presents a summary of the different techniques for measuring the hold-ups of the different phases and their main characteristics.

The first three methods have been used in this research and compared. The results were cross-checked using photographic techniques involving still camera shots and video recordings. The principles involved in these three methods are now outlined.

Table 4.1 - Techniques for Measuring Hold-ups

TECHNIQUE	REMARKS	REFERENCES
Bed expansion	Poor precision when high rates of gas are reached.	Nicklin (1962), Hughmark et al (1961), Kim et al (1972)
Pressure Profile measurement (Manometric Method)	Pressure tappings all along the column may be needed.	Baker et al (1978), Bhatia et al (1972), Kato et al (1981), Kim et al (1975), Levsh et al (1968)
Simultaneous closure of the gas and liquid feed lines. (Disengagement Method)	A convenient method of measurement.	Dhanuka et al (1978), Kumar et al (1974), Kusakabe et al (1980) Razumov et al (1973), Vail et al (1970)
Saline tracers	Gives access to the axial dispersion of the liquid, poor precision, risk of changing the properties of the liquid.	El-Temtamy & Epstein (1979), Kim & Kim (1981)
Radioactive tracer	Problems of tracers in the gas phase; difficult technique.	Michelsen & Ostergaard (1970), Field & Davidson (1980)
Measurement of electrical conductivity	Calibration by another method necessary.	Begovich & Watson (1978), Dhanuka et al (1978), Bhatia & Epstein (1972)
Resistance or capacitance probes	Disturbance of the hydrodynamics.	Linneweber & Blass (1981), Ostergaard (1966), Razumov et al (1973)
γ-ray attenuation Light transmission Photography	Measurements are made in the two-phase part, which is in general not characteristic of the overall bed.	Sherrard (1966), Lee et al (1974a, 1978), Adlington & Thompson (1965), Lee et al (1974b) Hassanien et al (1981), Page & Harrison (1972, 1974)

(1) Bed Expansion Method

This is the most common and simple of the techniques. Gas hold-up is determined by directly measuring the height of the expanded, aerated liquid (L_f) and that of clear liquid without aeration (L_i) (see Figure 4.2).

From Equation 4.1

$$\varepsilon_G = \frac{SL_f - SL_i}{SL_f}$$

$$= \frac{L_f - L_i}{L_f} \qquad \text{........ 4.2}$$

(2) The Manometric Method

Gas hold-up can be determined by measuring the pressure at two or several points in the column using a manometric system. A and B in Figure 4.3 represent two manometers fixed at arbitrary positions C and D along the length of the column. The difference in the manometer levels, l, gives a direct indication of the hold-up in the section contained between the two tappings.

From equation 4.2, average gas hold-up

$$\varepsilon_G = 1 - \frac{L_i}{L_f} \qquad \text{........ 4.3}$$

Figure 4.2 The Bed Expansion method.

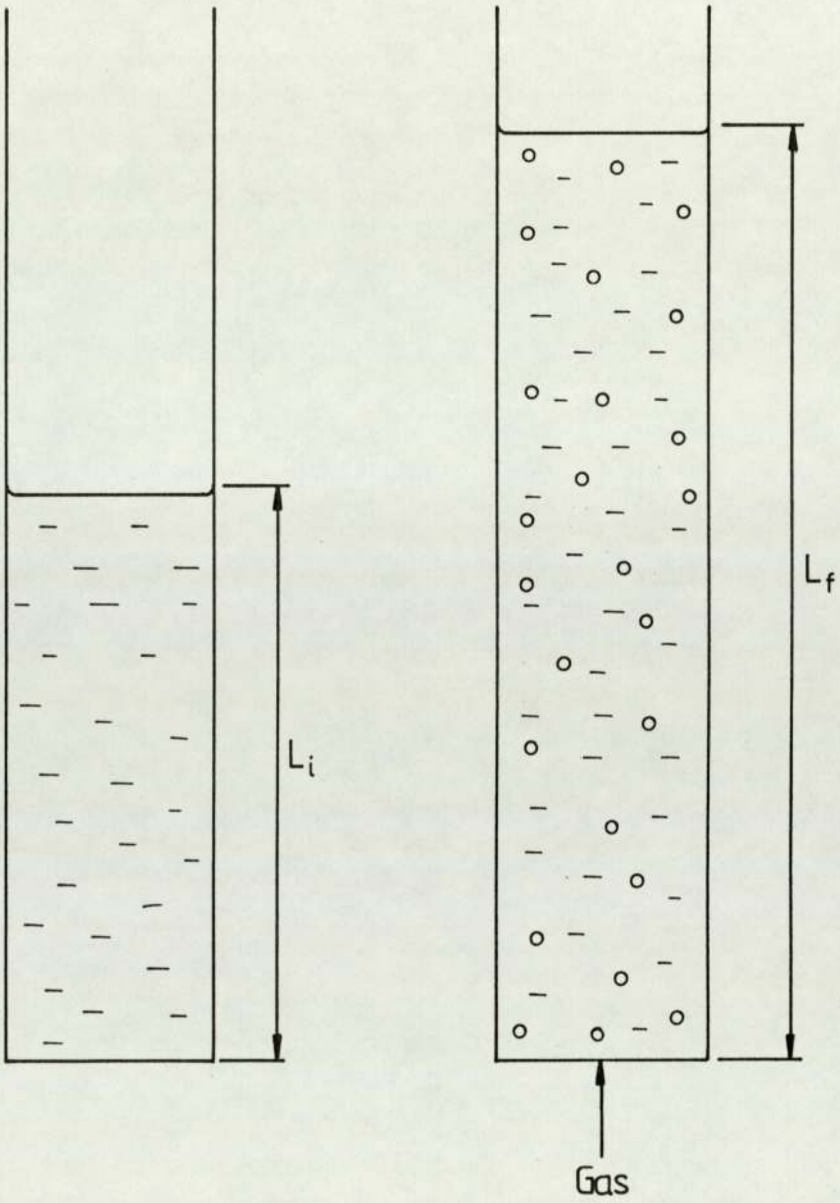
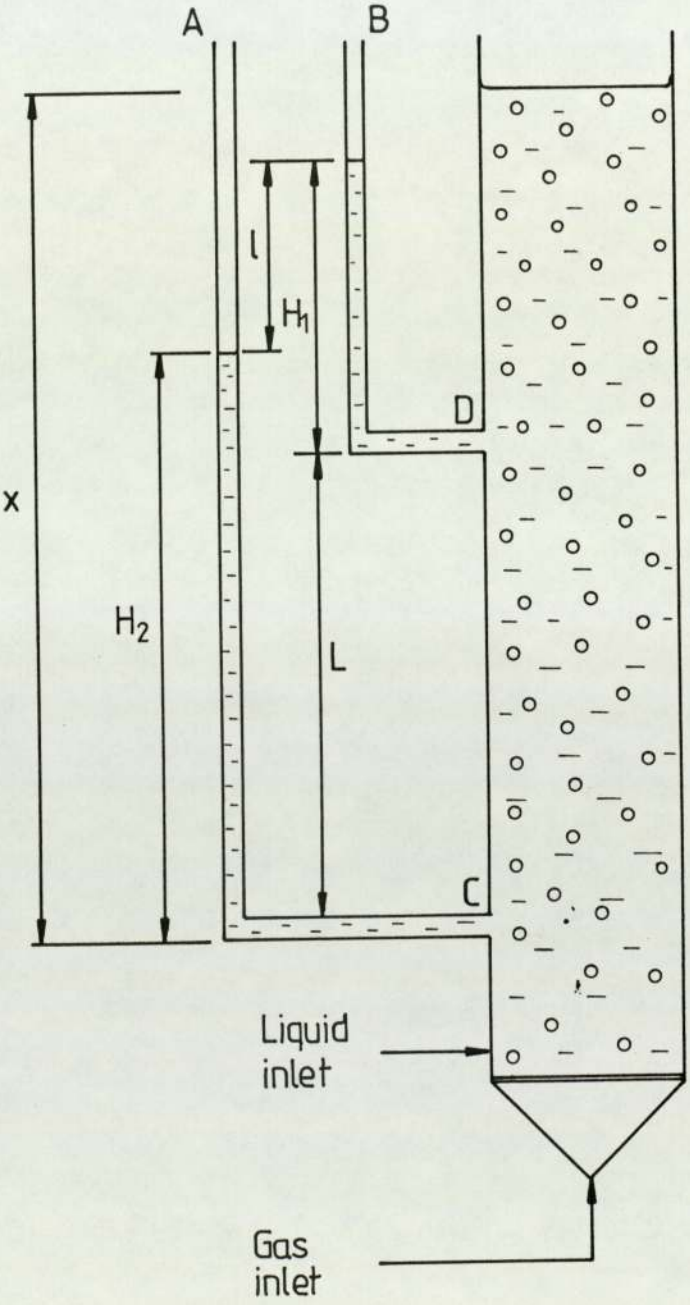


Figure 4.3 Representation of the Manometric method.



where

L_i is the height of liquid without aeration

L_f is the height of aerated liquid

S is the cross-sectional area of the column.

The density of the gas/liquid mixture may be defined as:

$$\rho_M = \frac{L_i S \rho_L + S (L_f - L_i) \rho_G}{L_f S}$$

$$= \rho_L \frac{L_i}{L_f} + \rho_G \left(1 - \frac{L_i}{L_f}\right)$$

$$= \rho_L - (\rho_L - \rho_G)(1 - L_i/L_f) \qquad \text{........ 4.4}$$

Combining equations 4.3 and 4.4 gives:

$$\varepsilon_G = \frac{\rho_L - \rho_M}{\rho_L - \rho_G} \qquad \text{........ 4.5}$$

Since $\rho_L >> \rho_G$, equation 4.3 becomes:

$$\varepsilon_G = \frac{\rho_L - \rho_M}{\rho_L} \qquad \text{........ 4.6}$$

Considering the pressures due to the hydrostatic head in the system, At C: $P_C = \rho_M x = \rho_L H_2$

At D: $P_D = \rho_M (x - L) = \rho_L H_1$

Therefore

$\Delta P = \rho_M L = \rho_L (H_2 - H_1) = \rho_L (L - l)$

$$\rho_M = \rho_L (1 - \frac{l}{L})$$

$$\rho_M / \rho_L = 1 - \frac{l}{L}$$

$$l/L = 1 - \rho_M/\rho_L$$

$$\frac{l}{L} = \frac{\rho_L - \rho_M}{\rho_L} \quad \text{........ 4.7}$$

Comparing equations 4.6 and 4.7 gives:

$$\varepsilon_G = \frac{l}{L} \quad \text{......... 4.8}$$

With this simple technique, gas hold-up can be measured for any section as well as the whole length of the column provided there are a sufficient number of manometers. Two manometers placed at different positions relative to the bottom of the column have been used to provide an indication of the overall gas hold-up.

(3) The Gas Disengagement Method

This technique has been developed for assessing bubble behaviour and bubble size distribution in bubble columns (Sriram and Mann, 1977). It is based on a theoretical approach which shows how the interaction of bubble size distribution and bubble rise velocity functions lead to predictions of the overall steady-state holdup. Since the approach is based on a physical understanding of how bubble flow at a given superficial velocity is related to the static hold-up, the theory is also extended to describe the disengagement of gas bubbles if the gas feed is cut off. The dynamic gas hold-up during gas disengagement is thus used to provide insight into the fundamentals of bubble column behaviour; bubble size distribution and liquid circulation intensity can thereby be assessed.

The theoretical background of this method has been given by Sriram and Mann (1977). For a batch system, the static gas hold-up is:

$$\varepsilon\,(o) = \frac{U_{SG}}{\int_o^\infty f\,(d_B)\,U\,(d_B)\,dd_B} \qquad \text{........ 4.9}$$

Where $f\,(d_B)\,dd_B$ is the volume fraction of bubbles having sizes between d_B and $d_B + dd_B$ and $U\,(d_B)$ is the rise velocity of a bubble of size d_B.

The overall or static gas hold-up $\varepsilon\,(o)$ is related to the dynamic gas hold-up $\varepsilon\,(t)$, after a time t, by:

$$\varepsilon\,(t) = \varepsilon\,(o) \int_o^\infty f\,(d_B)\;(1 - t\,U\,(d_B)/L\,)\quad dd_B \qquad \text{........ 4.10}$$

Equations 4.9 and 4.10 show that the static and dynamic disengagement hold-ups are functions of the size distribution and rise velocity of the bubbles.

Because this technique also provides, in principle, information about bubble dynamics, the experimental method is described in Chapter 5.

4.3 Experimental Programme

4.3.1 Choice of Factors

A review of the literature has revealed lack of consistent and coherent information. In some cases, investigators have reached contradictory conclusions; possible reasons for this have been treated in Section 4.2.1. The choice of factors in this research are outlined below:

Operating Parameters

The behaviour of columns was investigated primarily by varying the superficial gas and liquid velocities. The gas velocity was in the range: $0 \leq U_{SG} \leq 35$ cm/s, while the liquid velocity was within the range: $0 \leq U_{SL} \leq 1$ cm/s. In previous studies, Shayegan Salek (1976), Jamialahmadi (1982), relatively low values had been used, eg. $U_{SG} \leq 15$ cm/s. The author decided to investigate higher values, corresponding to different flow regimes, in view of their importance to industrial applications.

Column Design

Experiments were performed using the two different types of bubble column. These are designated the 2-D and 3-D columns and have been described previously in Chapter 3. The 2-D column allows clear and visual observations of column behaviour to be made. The 3-D cylindrical column bears close resemblance to industrial columns.

Different sizes of column have been used in previous studies of the effect of column diameter (Downie (1972) and Shayegan Salek (1974)). The effect of using different types of gas distributor was also explored.

Nature of the Phases

Because of the interest of the author in fermentation processes, air was used as the gas phase in all experiments.

Water and aqueous solutions were employed as the liquid phases. The additives used are discussed briefly in Section 4.3.2.

Three-phase systems were studied in the later stages of the programme and these are described in detail in Chapter 6.

4.3.2 Choice of Systems

The literature survey has indicated the need for further studies on the effects of the nature of dissolved salts (which are normally part of a fermentation media), organic compounds and the physical properties of the phases on the performance of bubble columns. A systematic method of study was developed as outlined below:

(a) Air-water systems

(b) The effect of additives in the liquid phase - alcohols, ketones, esters and electrolytes were used.

The physical properties of each system are presented in Table 4.2.

Table 4.2 - Physical Properties of Aqueous Organic and Inorganic Solutions

Data were taken at 22^{o}C. Solutions were made up in distilled water.

SYSTEM (VOL %)	DENSITY (Kg m^{-3})	SURFACE TENSION (Nm^{-1})	VISCOSITY (Ns m^{-2} x 10^{3})
<u>Alcohols</u>			
0.5% Methanol	999	0.070	1.37
0.5% Ethanol	999	0.064	1.35
0.5% i-Propanol	998	0.057	1.39
0.5% n-Butanol	998	0.055	1.41
<u>Ketones</u>			
0.5% Propanone	999	0.060	1.2
0.5% Butanone	999	0.063	1.35
<u>Esters</u>			
0.5% Ethyl Acetate	998	0.060	1.41
0.5% n-Proply Acetate	999	0.056	1.38
0.5% n-Butyl Acetate	999	0.053	1.22
<u>Electrolytes</u>			
1.0% NaCl	1007	0.073*	1.38
1.0% KCl	1006	0.072*	1.24
1.0% $CaCl_2$	1005	0.076*	1.46
1.0% Na_2SO_4	1009	0.074*	1.41

* By Hikita et al (1980)

4.4 **Experimental Results**

The results from experiments have all been tabulated and are presented in Appendix A. The graphical output was made using a computer graph plotter for the purpose of accuracy, neatness and clear presentation.

For convenience, experimental results obtained by other members of the Aston team are presented in this chapter.

4.4.1 Effects of Operating Parameters

Effect of Superficial Gas Velocity

Figures 4.4 and 4.5 show the results of gas hold-up measurements as a function of superficial gas velocity, with superficial liquid velocity as a parameter. These are for air-water systems in the two- and three-dimensional bubble columns respectively. The data used to plot these graphs are given in Tables A1 and A2 of Appendix A.

As shown in the figures, gas hold-up increased with an increase in the gas velocity. The increase was very rapid at low gas velocities and tended to level off at higher values.

Effect of Superficial Liquid Velocity

Figures 4.4 and 4.5 also portray the effect of liquid velocity on gas hold-up. At low gas velocities (< 10 cm/s), gas hold-up was unaffected by a change in the liquid velocity. However, as gas velocity was increased further, there was a slight decrease in gas hold-up.

Figure 4.4 Effect of superficial gas and liquid velocity on gas hold-up in the two dimensional bubble column.

Two phase system (Air - Water)

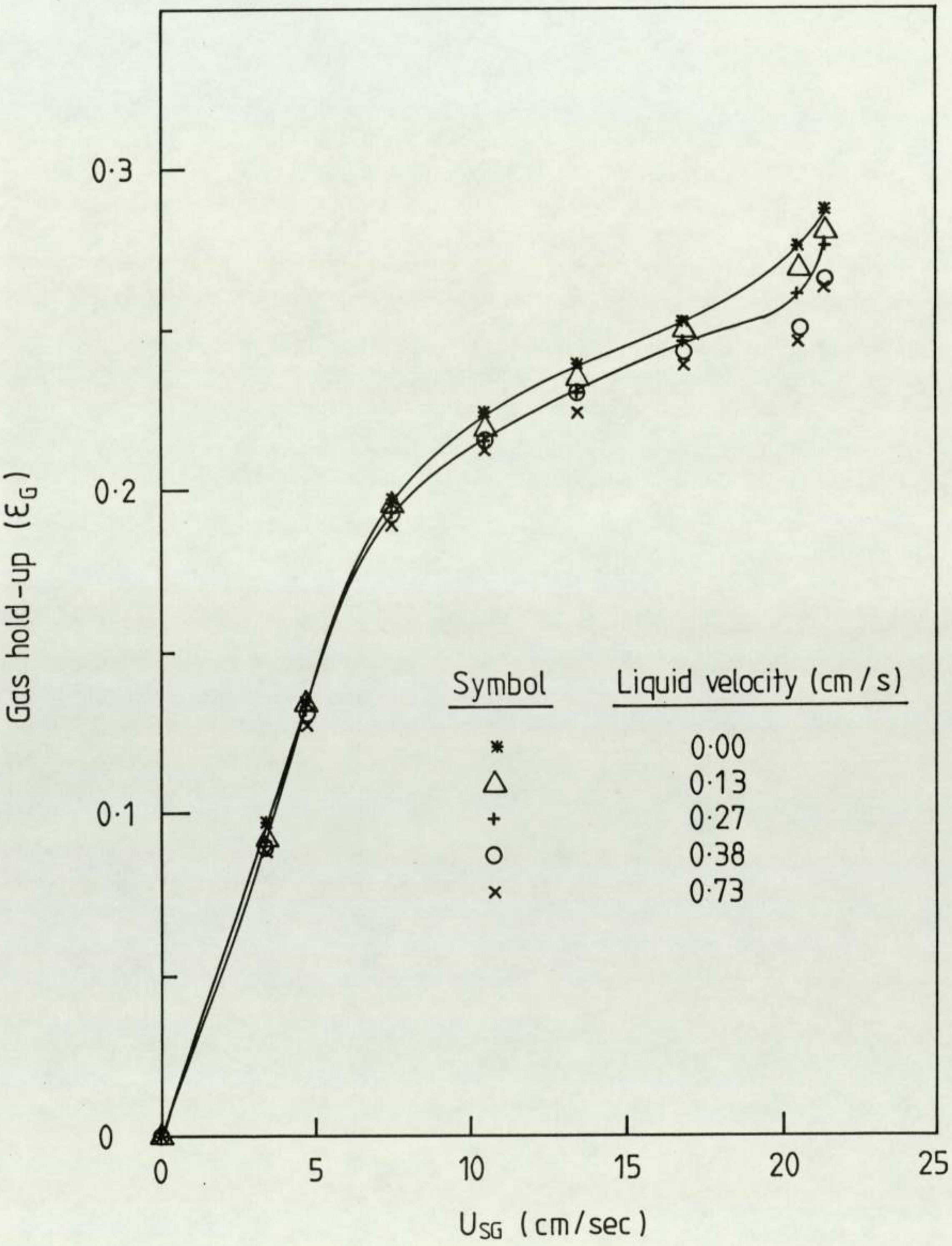
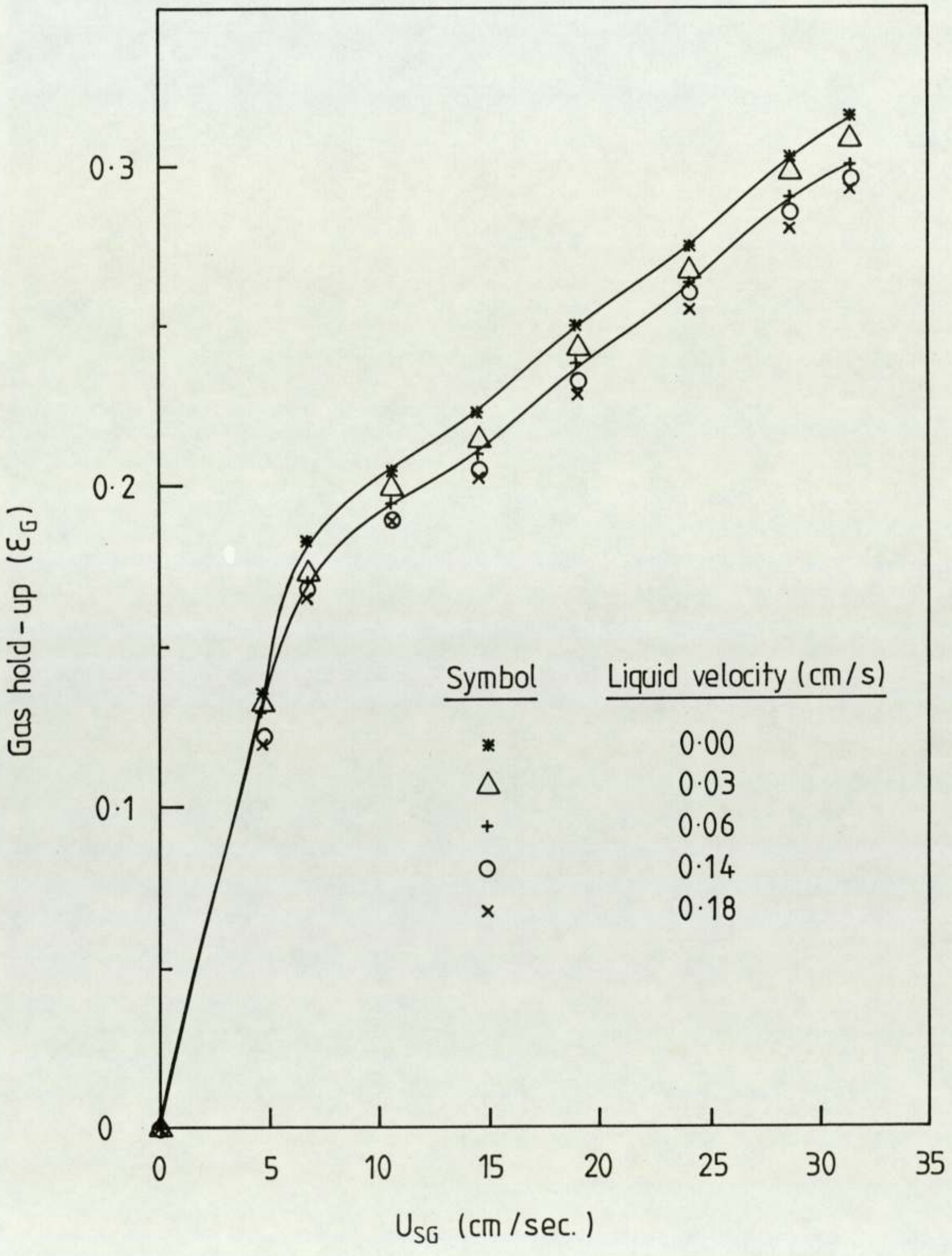


Figure 4.5 Effect of superficial gas and liquid velocity on gas hold-up in the three dimensional bubble column.
Two-phase system (Air - Water)



Effect of Operating Temperature

The effect of liquid-phase temperature has been studied using tap water as the liquid phase. A range of temperatures from 20°C to 70°C was chosen because this covers temperatures frequently used in fermentation processes. The results showing how the temperature affected gas hold-up in the three-dimensional column are presented in Figure 4.6; see also Smith et al (1986). Increase in temperature of the liquid phase causes a decrease in hold-up in the mid-range of superficial gas velocity.

4.4.2 Effect of Column Design

Working with four different columns (7.6cm, 15.2cm, 30.5cm and 61.0cm in diameter), Shayegan Salek (1974) showed the effect of column diameter on gas hold-up for air-water systems. His data also included results for different gas distributors.

Gas hold-up decreased with increasing column diameter. Porous plates gave higher hold-ups compared to values obtained with perforated plates at the same gas velocity.

4.4.3 Effects of Phase Physico-Chemical Properties

Effect of Alcohols

Results with four different alcohols, are plotted in Figure 4.7; the basic data will be found in Table A3 in Appendix A.

Gas hold-up increased with an increase in gas velocity in all cases, although it was strongly influenced by the type of alcohol present. The gas hold-up decreased in the following order:

Figure 4.6 Effect of temperature and superficial gas velocity on gas hold-up in the three dimensional bubble column. Two-phase system (Air-Water)

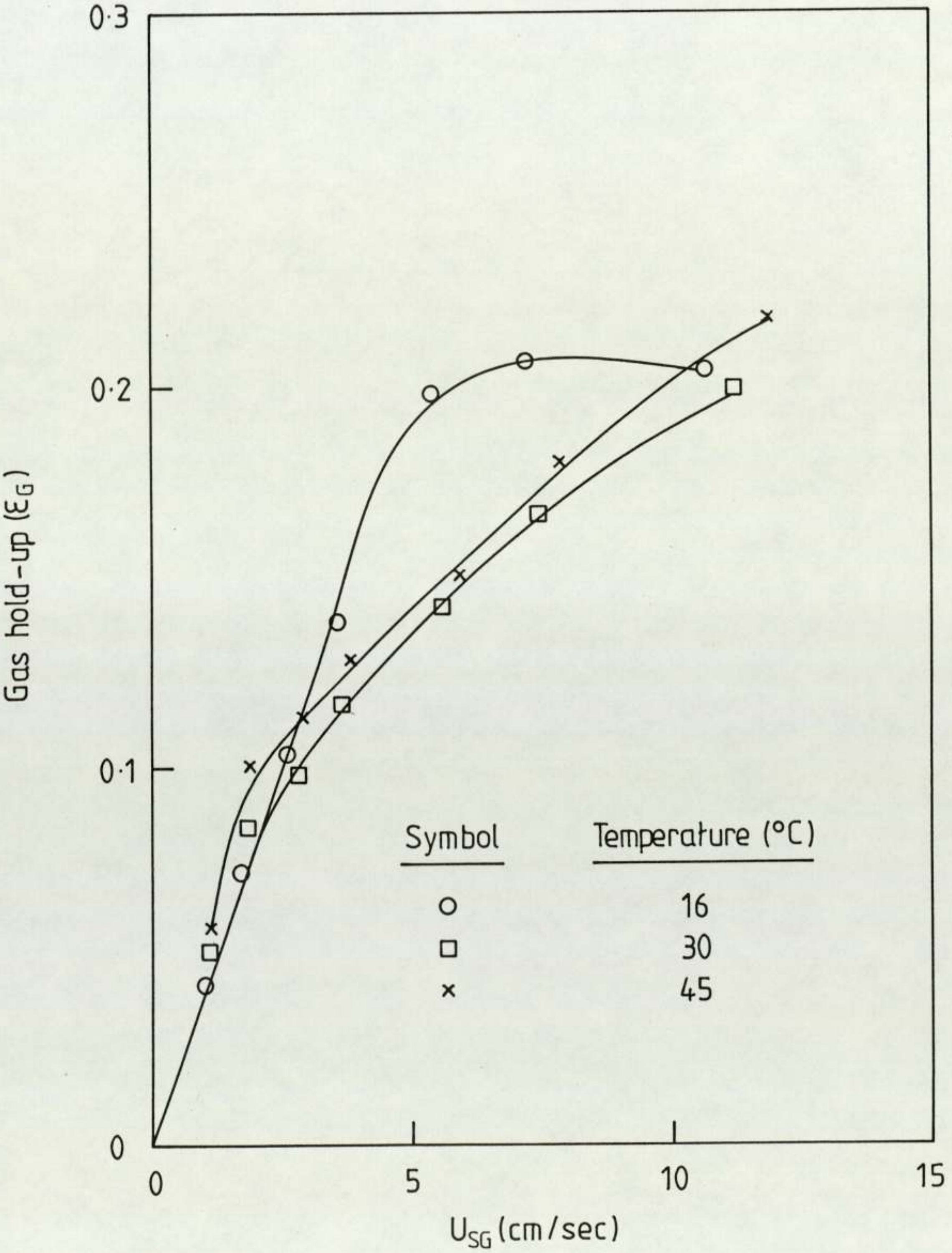
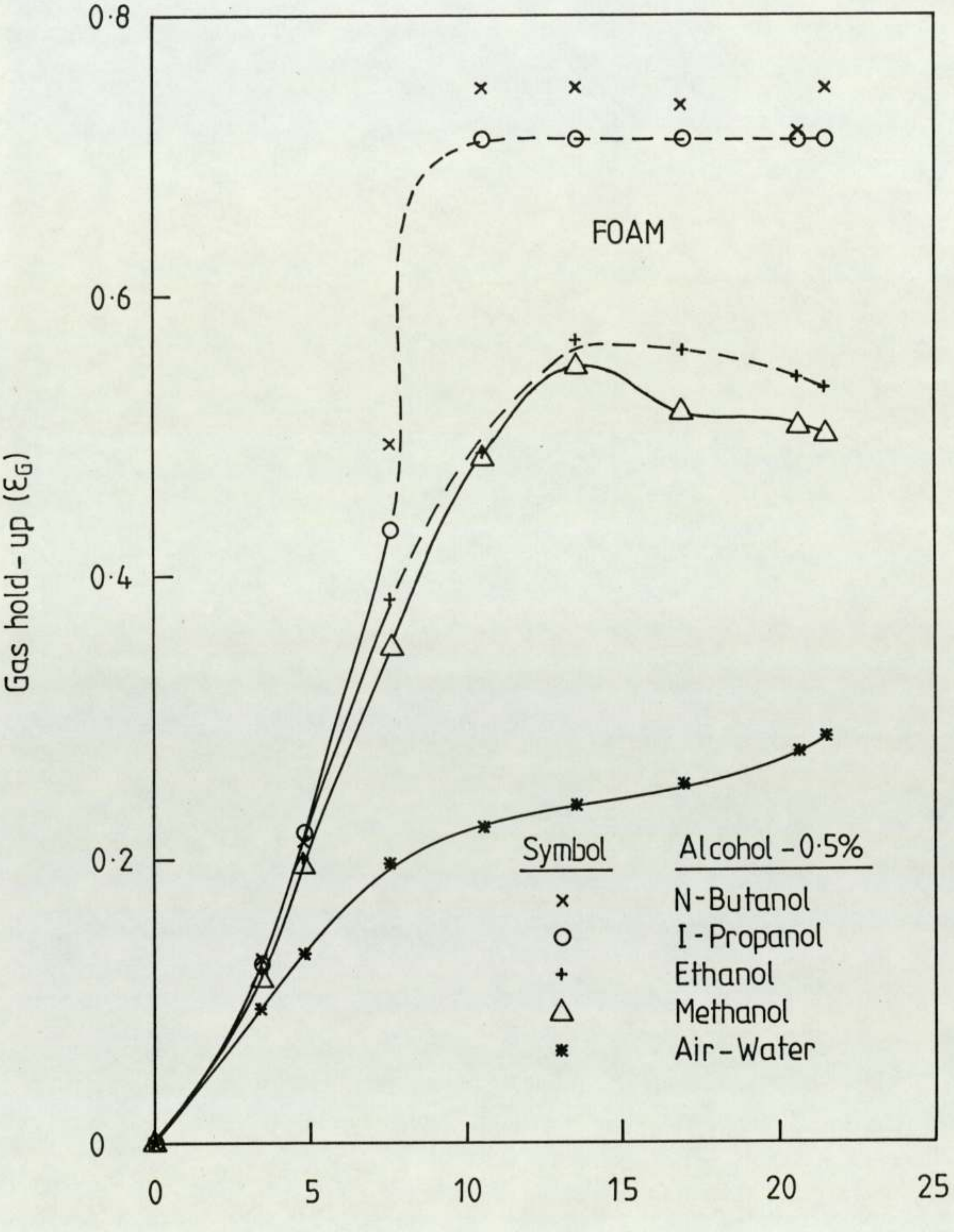


Figure 4.7 Effect of alcohols on gas hold-up in the two dimensional bubble column.

Two-phase system (Gas-Liquid)



n-butanol > i-propanol > ethanol > methanol > water

Foaming was observed at high gas velocities and the data points at U_{SG} values > 5 cm/s are merely indicative of hold-up.

Effect of Ketones

Two different ketones were used. The results are shown in Figure 4.8 with the data listed in Table A4 in Appendix A.

The addition of ketones results in higher hold-up values than in air-water systems. Hold-up is also influenced by the chain length of the ketone. The decrease is of the order

butanone > propanone (acetone) > water

Foaming was observed at the higher gas velocities as was the case when alcohols were the additives.

Effect of Esters

The effect of using different esters on gas hold-up is presented in Figure 4.9. The data are given in Table A5 in Appendix A.

Gas hold-up shows an increase with gas velocity in all cases. Again the effect is influenced by the type of ester present. Whereas ethyl acetate increases the gas hold-up relative to that for the air-water system, the longer chain esters have the opposite effect.

Figure 4.8 Effect of Ketones on gas hold-up in the two dimensional bubble column.

Two-phase system (Gas-Liquid)

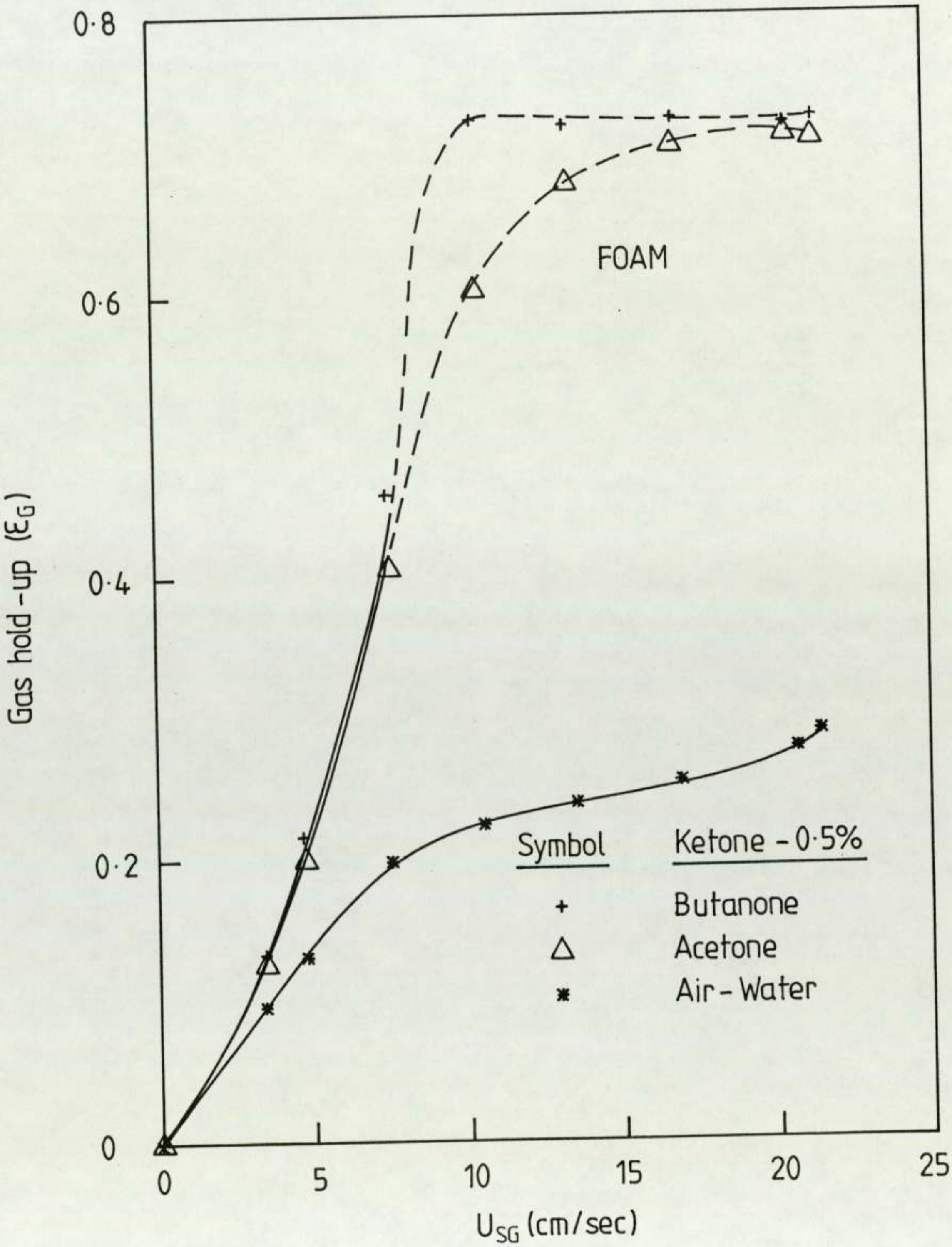
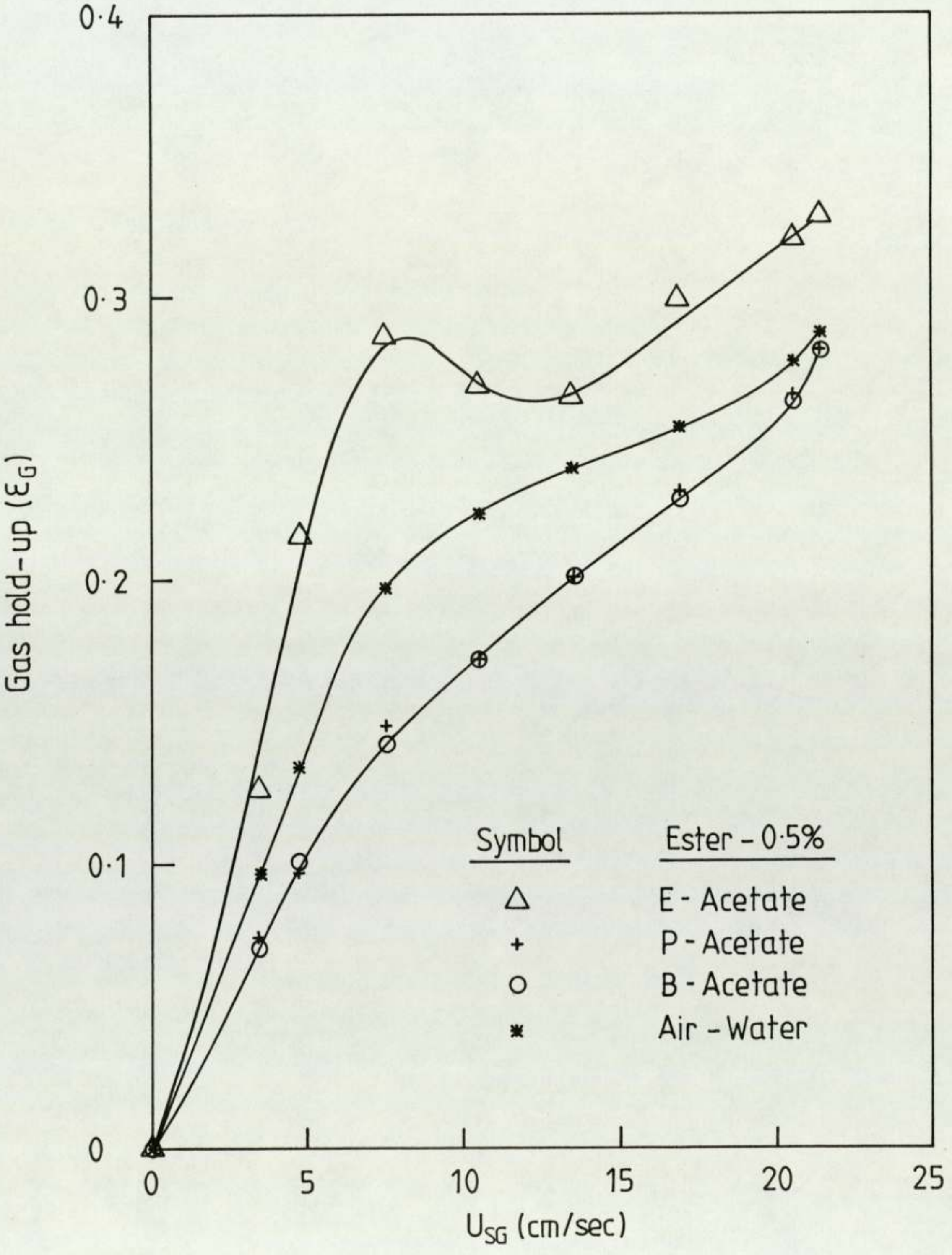


Figure 4.9 Effect of Esters on gas hold-up in the two dimensional bubble column.

Two-phase system (Gas-Liquid)



Effect of Electrolytes

Four different electrolytes were used in this part of the programme. The results are shown in Figure 4.10 with the full data given in Table A6 in Appendix A. All the electrolyte solutions gave higher values of gas hold-up than air-water systems.

4.5 Discussion

4.5.1 Effect of Operating Parameters

Effect of Superficial Gas Velocity (U_{SG})

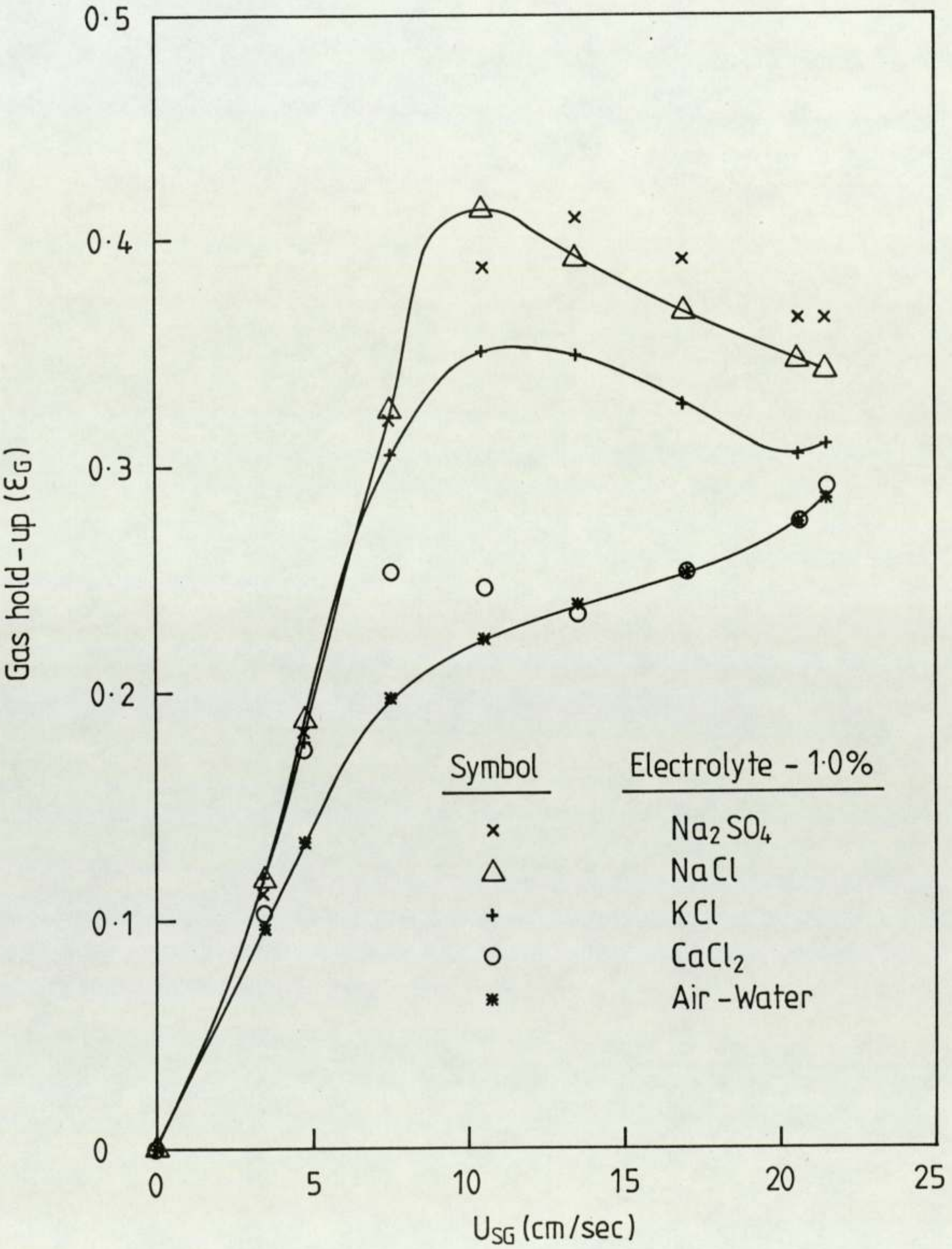
A swarm of bubbles rises uniformly within the bubble column when the superficial gas velocity is low (< 4 cm/s) and when bubbles of uniform size are generated at the gas distributor. This is the so called "bubbly-flow regime". When the gas velocity is increased, bubbly flow ceases to be uniform and the "churn-turbulent flow regime" commences. In the bubbly flow regime, gas hold-up increases almost linearly with superficial gas velocity and then tends to level off at higher velocities as turbulent flow sets in. Similar trends have been observed by Yamashita & Inoue (1975), Aoyama et al (1968) and Deckwer et al (1974).

Agitation by the Gas Phase

When a gas is bubbled into a column of liquid through a number of orifices, such as a perforated plate, the pressure energy contained in the gas serves two purposes: it is used in creating gas bubbles and in agitating the liquid as the bubbles force their way to the top of the column. The amount of agitation of the liquid phase caused by bubbles depends on the superficial gas velocity.

Figure 4.10 Effect of Electrolytes on gas hold-up in the two dimensional bubble column.

Two-phase system (Gas-Liquid)



In the bubbly flow regime, the bubbles are able to clear a way in an axial direction without any collision or coalescence occurring. The column is, therefore, uniform in appearance and the extent of liquid phase agitation is of little significance. This is sometimes termed homogeneous flow.

On increasing the gas velocity, eventually a point is reached when the bubbles are able to overcome the surface tension forces and coalesce. Increasing the gas velocity above this transition point leads to the formation of many large bubbles which ascend at the centre of column following a wavy pattern: this is often referred to as the churn-turbulent regime, and in small diameter columns, slug flow can develop (see also Chapter 5). At higher gas velocities, the slugs become unstable and break up resulting in even higher gas hold-ups. Such flow has been termed heterogeneous.

Generally, an increase in superficial gas velocity causes a considerable increase in gas hold-up.

Other means of aiding coalescence in the column includes heating, mechanical agitation (Reith and Beek (1970), Falch and Gaden (1969)) and vibration (Buchanan et al (1962), Biard (1962), Boyle (1927)). These methods have been discussed by Jamialahmadi (1982).

Effect of Superficial Liquid Velocity (U_{SL})

Superficial liquid velocity has little effect on gas hold-up for the range studied. As liquid velocity increases, there is a decrease in gas hold-up particularly outside the bubbly flow regime. The liquid velocity cause a quicker wash-out of the gas phase and with more liquid in the system, the gas hold-up reduces, though slightly.

Effect of Operating Temperature (see also Section 4.7)

Using a moderate energy input, the intermolecular forces between the water molecules become weaker. This results in the bubbles coalescing easily leading to a reduction in gas hold-up over the range of U_{SG} from 3 - 10cm/s.

4.5.2 Effect of Column Design

As column diameter increases, the intensity of liquid agitation increases. This leads to a lower hold-up. Similar results have also been reported by Siemes et al (1957).

The dependence of gas hold-up on gas distributor design is slight provided gas distribution is uniform and the height to diameter ratio of the column is large (Shayegan Salek (1974)).

4.5.3 Effect of Alcohols

Gas hold-up increases with an increase in gas velocity. It is strongly influenced by the type of alcohol present as illustrated in Figure 4.7. From measurement of physical properties, the only property which differs significantly for water and aqueous solutions of short chain alcohols is surface tension. Hence, alcohols are regarded as surfactants. It is observed that the presence of alcohols makes bubbles become more rigid and reduces the bubble-rise velocity; it also induces a noncoalescing tendency in two-phase flow. Levich (1962) suggested that surface-active substances tend to accumulate to the interface between the bubble and the liquid. When a bubble moves through the liquid, adsorbed surface active material is swept to the rear; this creates a concentration gradient and hence surface tension gradient which opposes the tangential shear stress. This phenomenon increases the drag on the bubble and hence reduces the rise velocity. The unusual increase in the gas hold-up with an increase in the superficial gas velocity is thus the result of the retardation in bubble-rise velocity and the noncoalescing tendency of the medium.

The bubble rise velocity decreases with an increase in the chain length of alcohols. Gal-or and Waslo (1968), Lindland and Terjesan (1965), Raymond and Zieminski (1971) found that bubble rise velocity decreases with an increase in surfactant concentration. This was supported by Davis and Acrivos (1966) who observed a critical point at which the surface film collapses. Any further addition after that point does not affect the surface gradient. Keitel and Onken (1982) reported that the coalescence-hindering property increases with an increase in the chain length and that the critical concentration of alcohol can be directly related to the number of C-atoms in the chain length.

The lower the bubble rise velocity and the greater the coalescence hindering quality of the surfactant the higher the gas hold-up. Thus the observed trend in gas hold-up is:

butanol > propanol > ethanol > methanol > water

4.5.4 Effect of Ketones

The presence of ketones, like alcohols, results in higher values of gas hold-up (see Figure 4.8). This trend is also influenced by the type of ketone present. Ketones behave like alcohols and are associated with very high hold-up values (> 70%) especially at high gas velocities. The very low bubble-rise velocity coupled with the non-coalescing quality of these additives are responsible for such high values. The observed trend in gas hold-up is:

butanone > acetone > water

As was the case with alcohols, foaming was observed with ketones in solution.

4.5.5 Effect of Esters

Esters behave quite differently from the other compounds tested. Whilst the presence of ethyl acetate results in higher gas hold-up values, both propyl- and butyl-acetate result in a reduction compared to values with water. With ethyl-acetate, the bubbles were small in size with low rise velocity whereas, in propyl- and butyl-acetate solutions, there was much coalescence and faster bubble rise. These facts explain the trend in gas hold-up values:

ethyl-acetate > water > n-propyl-acetate > n-butyl acetate

4.5.6 Effect of Electrolytes

The presence of electrolytes, like surfactants, results in higher values of gas hold-up (see Figure 4.10). Addition of electrolytes to water does not alter any measurable physical property significantly (except for a slightly higher surface tension). However, the presence of electrolyte does change the coalescence behaviour of the liquid media and, hence, indirectly affects the hydrodynamic characteristics. Smith et al (1986) pointed out that the electrolyte makes the liquid more cohesive and an electrical double layer formed at the interface suppresses bubble coalescence. The effectiveness of salts in inhibiting coalescence is related to their ionic strength and position in the lyotropic series, which provides a measure of the "salting out" effect.

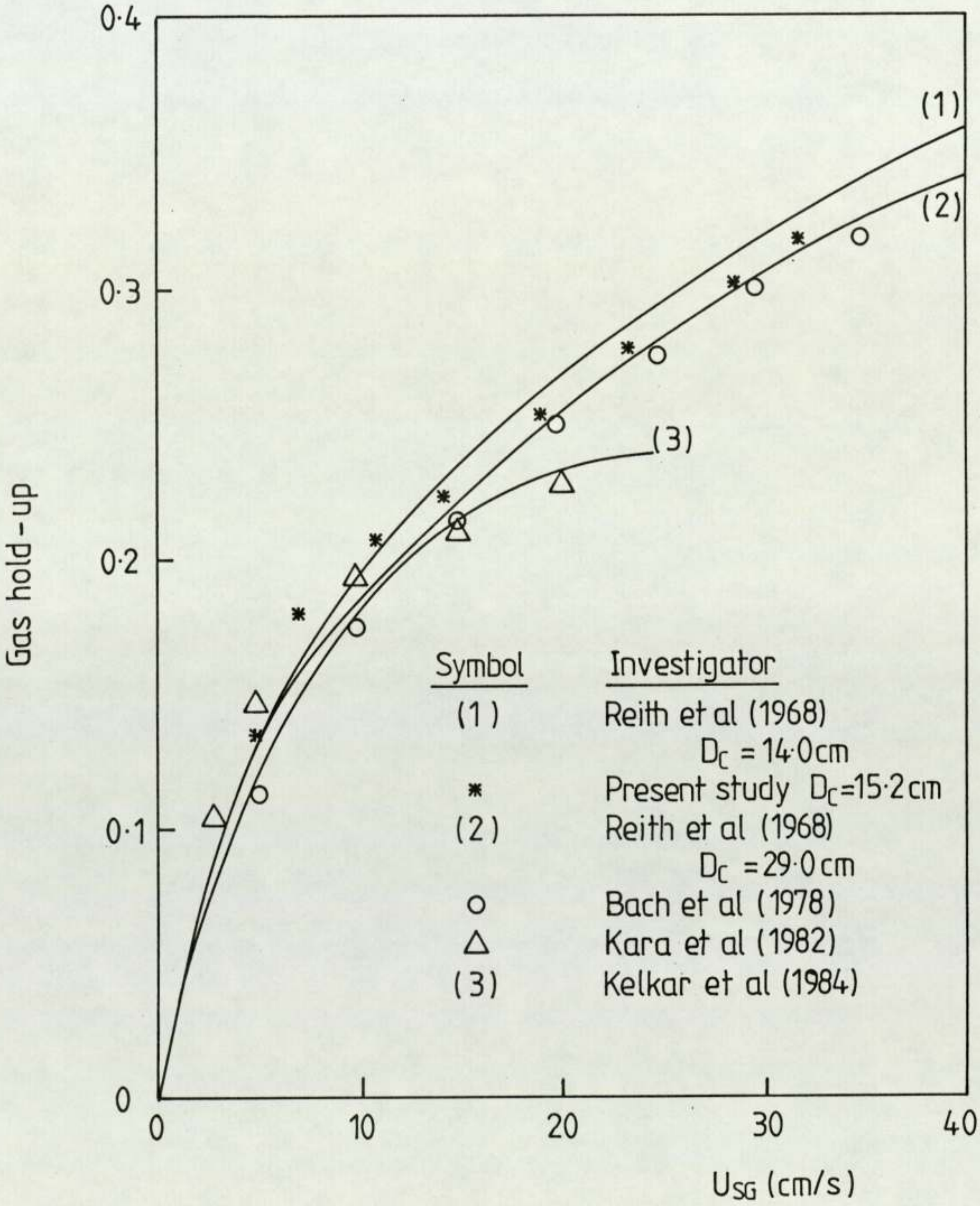
Freedman and Davidson (1969) reported that the addition of an electrolyte postpones the appearance of large bubbles and results in increased gas hold-up values. It seems that an electrolyte solution can maintain the small bubble size by virtue of its non-coalescing tendency. However, at high gas velocities (≥ 6.5 cm/s) this non-coalescing tendency diminishes and results in a decrease in gas hold-up. In a dilute alcohol solution, the non-coalescing tendency is much stronger; in addition the bubble rise velocities are lower than those in water. This is responsible for the higher gas hold-up values with alcohols compared to the values with electrolytes. In the electrolyte solutions, a cloud of fine bubbles was formed. This is thus a clear indication that the high gas hold-up values result from the relatively small size of the bubbles coupled with the ability of electrolytes to maintain the bubble sizes (at low gas velocities).

4.6 Modelling and Correlation

4.6.1 Comparison with Published Results

The experimental gas hold-up data for the air-water system are compared with the results reported by Reith et al (1968), Bach et al (1978), Kara et al (1982) and Kelkar et al (1984) in Figure 4.11. The agreement is satisfactory. Comparison with the results of Reith et al, also reveal the effect of column diameter on gas hold-up; there is a decrease in hold-up on going from a column diameter of 14.0cm to one of 29.0cm.

Figure 4.11 Comparison of air-water gas hold-up data with literature results.



Data on the effect of liquid additives especially at high superficial gas velocities are very scarce in the literature. Kelkar et al (1983) obtained a similar trend for alcohols in solution. Electrolytes were also found to cause an increase in gas hold-up when introduced into gas-liquid systems.

4.6.2 Empirical Correlations

A computer program (see Appendix I) has been developed which calculates gas hold-up from published correlations and compares them with the experimental values. Four correlations were chosen for this purpose and are listed below:

(a) Akita and Yoshida (1973)

$$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = A \quad B_o^{1/8} \; G_a^{1/12} \; F_r \qquad \text{....... 4.11}$$

A = 0.2 for pure liquids and non-electrolytes

= 0.25 for electrolytes

where

$B_o = g D_c^2 \rho_L / \sigma$ - Bond Number

$G_a = g D_c^3 / \nu_L^2$ - Galileo Number

$F_r = U_{SG} / \sqrt{g D_c}$ - Froude Number

(b) Kumar et al (1976)

$$\varepsilon_G = 0.728\, U - 0.485\, U^2 + 0.0975\, U^3 \qquad \text{........ 4.12}$$

where

$$U = U_{SG} \left(\frac{\rho_L^2}{\sigma (\rho_L - \rho_G)\, g} \right)^{0.25} \qquad \text{........ 4.13}$$

(c) Mersmann (1978)

$$\frac{\varepsilon_G}{(1 - \varepsilon_G)^4} = 0.14\, U_{SG} \left(\frac{\rho_L^2}{\sigma (\rho_L - \rho_G) g} \right)^{1/4} \left(\frac{\rho_L^2 \sigma^3}{\mu_L^4 (\rho_L - \rho_G) g} \right)^{1/24} \left(\frac{\rho_L}{\rho_G} \right)^{5/72}$$

$$\left(\frac{\rho_L}{\rho_L - \rho_G} \right)^{1/3} \qquad \text{........ 4.14}$$

(d) Hikita et al (1980)

$$\varepsilon_G = 0.672 \left(\frac{U_{SG}\, \mu_L}{\sigma} \right)^{0.578} \left(\frac{\mu_L^4 g}{\rho_L \sigma^3} \right)^{-0.131} \left(\frac{\rho_G}{\rho_L} \right)^{0.062} \left(\frac{\mu_G}{\mu_L} \right)^{0.107} \qquad \text{........ 4.15}$$

The results are presented in Tables B1 and B14 in Appendix B.

Experimental values of gas hold-up compare reasonably with those based on the correlations for water, aqueous esters and electrolytes solutions as liquid phases. The correlations of Akita and Yoshida (1973) and Hikita et al (1980) gave similar values, whereas the correlations of Mersmann (1978) predicted high figures. The correlation of Kumar et al (1976) fits the data well. Due to foaming of aqueous alcohols and ketones, higher hold-up values were observed compared to correlated values.

Correlation of Experimental Data

In the case of the air-water system the hold-up data were empirically correlated with gas velocity. It is believed that this correlation can be suitably applied to perforated plates at high gas and liquid throughputs. The correlation is as follows:

$$\varepsilon_G = 0.079\ U_{SG}^{\ 0.4} \qquad \text{........ 4.16}$$

U_{SG} is in cm/s.

The correlation can be applied to the 3-D column as well as the 2-D column. The average error is ± 5%.

4.6.3 Slip Velocity Model

The model considers gas bubbles as rising relative to the liquid surrounding them at a velocity called the slip velocity. This is defined as the mean relative velocity:

$$U_S = \frac{U_{SG}}{\varepsilon_G} \pm \frac{U_{SL}}{(1-\varepsilon_G)} \qquad \text{........ 4.17}$$

where the + ve sign corresponds to countercurrent flow
and the - ve sign corresponds to cocurrent flow.

The theory was originally proposed by Lapidus & Elgin (1957). U_S was assumed to be independent of the flow rates of either phase and only a function of single bubble rise velocity in an infinite medium, $U_{b\infty}$, and the gas hold-up.

Mathematically,

$$U_S = U_{b\infty} f(\varepsilon_G) \qquad \text{....... 4.18}$$

Various expressions have been proposed for f (ε_G) and are summarized by Lockett & Kirkpatrick (1975). The results are barely distinguishable at low values of gas hold-up, but they differ at higher values.

The Turner (1966) assumption: For batch systems, Turner assumed that the slip velocity is independent of ε_G, so that:

$$U_S = U_{b\infty} \qquad \text{........ 4.19}$$

Thus

$$\varepsilon_G = \frac{U_{SG}}{U_{b\infty}} \qquad \text{........ 4.20}$$

This was applied to the measured hold-up data and gave a good fit at low U_{SG}. This agrees with the findings of Freedman & Davidson (1969) who showed that for batch systems, Turner's expression described the measured data for air-water and air-electrolyte systems fairly well up to gas velocities at about 0.05 m/s. From the gas hold-up, superficial gas and liquid velocity, slip velocity was calculated according to equation 4.17. This was then plotted against gas hold-up (Figure 4.12) and against gas velocity (Figure 4.13). The correlation:

Figure 4.12 The slip velocity as a function of gas hold-up.

Two-phase system (Air - Water)

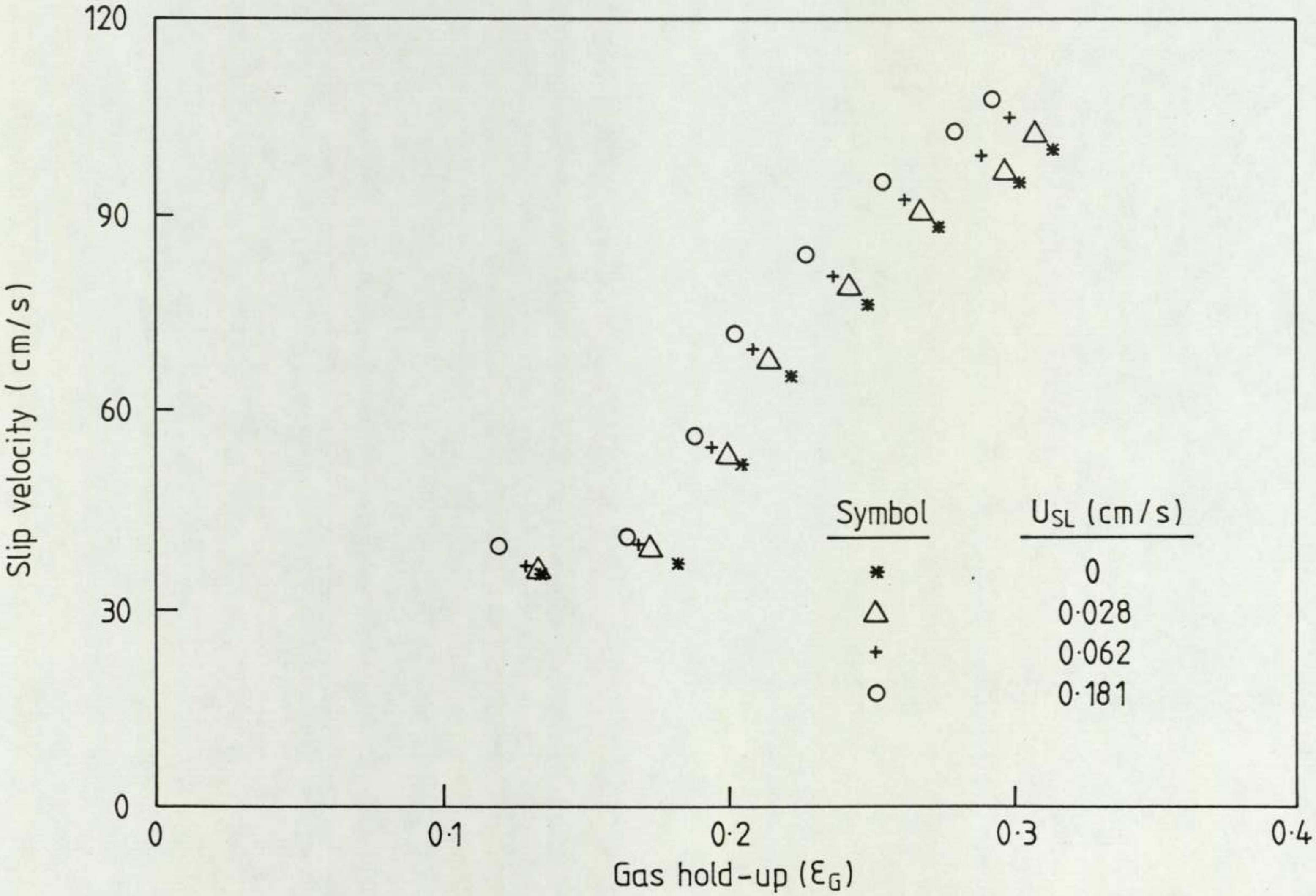
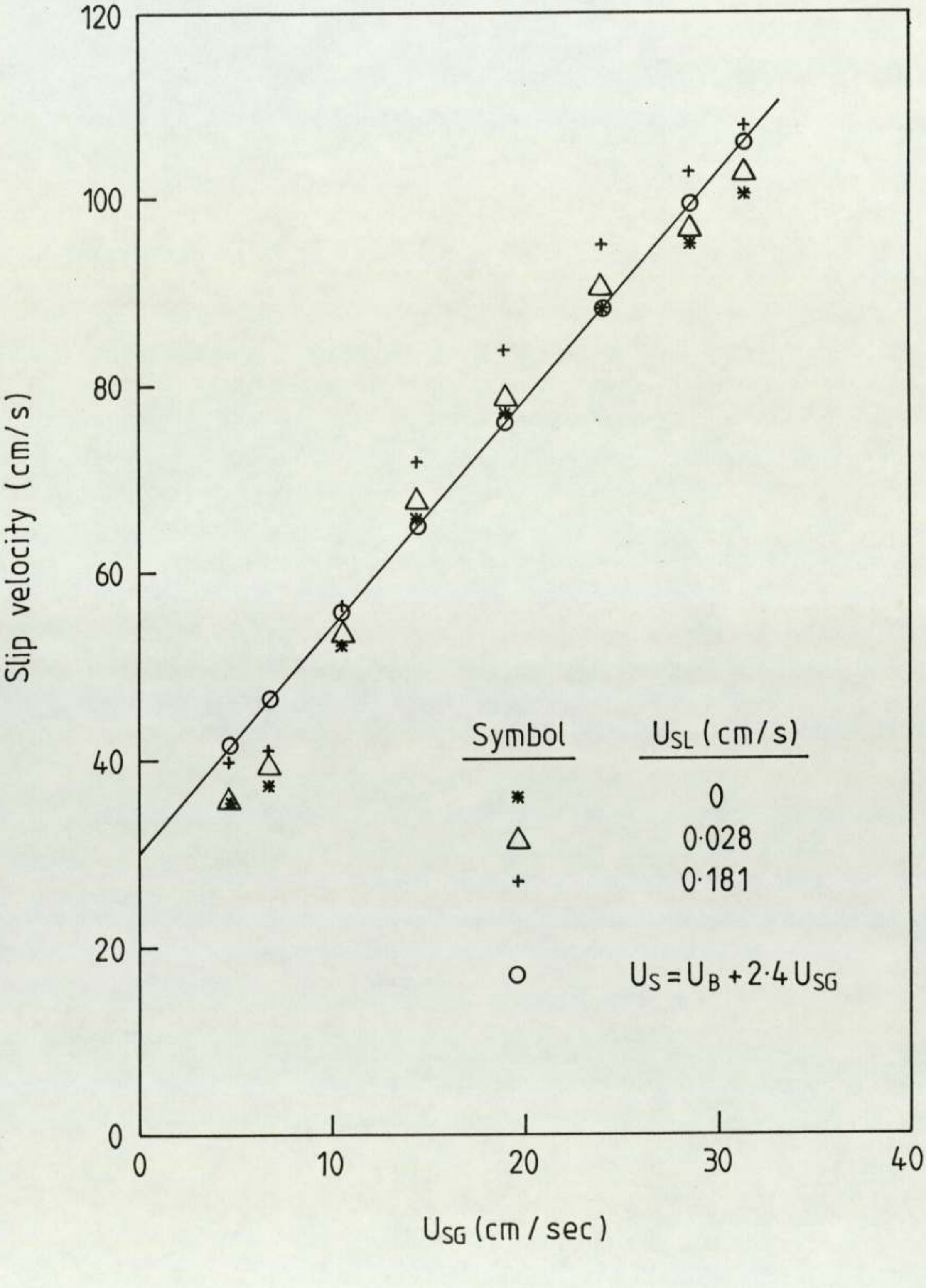


Figure 4.13 The slip velocity as a function of superficial gas velocity.

Two-phase system (Air-Water)



$$U_S = U_{b\infty} + 2.4\, U_{SG} \qquad \text{........ 4.21}$$

satisfactorily represents the data for air-water systems.

Drift Flux Model

Wallis (1962, 1969), proposed the method of characterizing the relative motion using the DRIFT FLUX of the gas, j_{GL}.

The drift flux is defined as the volumetric flux of gas relative to the surface moving at an average velocity. If the total velocity U_T is defined as:

$$U_T = U_{SG} + U_{SL}, \qquad \text{........ 4.22}$$

the drift flux is defined as $j_{GL} = U_{SG} \pm \varepsilon_G U_T$ 4.23

$$\text{or } j_{GL} = U_{SG}(1 - \varepsilon_G) \pm U_{SL}\, \varepsilon_G \qquad \text{........ 4.24}$$

where the + ve sign corresponds to countercurrent flow and the - ve corresponds to cocurrent flow.

The drift flux is also defined as:

$$j_{GL} = U_S\, \varepsilon_G\, (1 - \varepsilon_G) \qquad \text{........ 4.25}$$

Lapidus & Elgin (1957) suggested that the slip velocity U_S is a function of the single bubble rise velocity and gas hold-up. Thus:

$$j_{GL} = U_{b\infty}\, \varepsilon_G\, (1 - \varepsilon_G)\, f_1(\varepsilon_G)$$

or

$$j_{GL} = U_{b\infty}\, f_2(\varepsilon_G) \qquad \text{........ 4.26}$$

If the drift flux is plotted against the gas hold-up, the change in the slope of the curve will indicate the transition from one flow regime to another. Various expressions for drift flux are also reported by Lockett and Kirkpatrick. Among them are:

The Richardson & Zaki (1954) Expression

For batch systems, this is :

$$U_{SG}\,(1 - \varepsilon_G) = U_{b\infty}\, \varepsilon_G\, (1 - \varepsilon_G)^{2.39} \qquad \text{........ 4.27}$$

$$\frac{(1 - \varepsilon_G)}{\varepsilon_G\,(1 - \varepsilon_G)^{2.39}} = \frac{U_{b\infty}}{U_{SG}} \qquad \text{........ 4.28}$$

The Turner (1966) Expression

For batch systems, this is :

$$U_{SG}\,(1 - \varepsilon_G) = U_{b\infty}\, \varepsilon_G\, (1 - \varepsilon_G) \qquad \text{........ 4.29}$$

$$\varepsilon_G = \frac{U_{SG}}{U_{b\infty}}$$ (see also equation 4.20)

Plots of drift flux against gas hold-up are presented in Chapter 5. These are used to describe flow regime characteristics.

4.6.4 The Zuber-Findlay Model

Zuber and Findlay (1965) modified the drift flux model for the churn-turbulent regime to account for the radial distribution of hold-up which can lead to breakdown of bubbly flow.

For cocurrent flow, equation 4.23 is modified as:

$$\langle j_{GL} \rangle = \langle U_{SG} \rangle - \langle \varepsilon_G \; U_T \rangle \qquad \text{........ 4.30}$$

where < > corresponds to radially averaged value.

Dividing though by $\langle \varepsilon_G \rangle$ and rearranging gives:

$$\frac{\langle U_{SG} \rangle}{\langle \varepsilon_G \rangle} = \frac{\langle j_{GL} \rangle}{\langle \varepsilon_G \rangle} + \frac{\langle \varepsilon_G \; U_T \rangle}{\langle \varepsilon_G \rangle} \qquad \text{........ 4.31}$$

Zuber and Findlay defined a radial distribution parameter

$$C_o = \frac{\langle \varepsilon_G \; U_T \rangle}{\langle \varepsilon_G \rangle \langle U_T \rangle} \qquad \text{........ 4.32}$$

which is a rough indication of the non-uniform radial distribution.

Equation 4.31 then becomes:

$$\frac{\langle U_{SG} \rangle}{\langle \varepsilon_G \rangle} = \frac{\langle j_{GL} \rangle}{\langle \varepsilon_G \rangle} + C_o \langle U_T \rangle \qquad \text{........ 4.33}$$

If the value of $\langle j_{GL} \rangle / \langle \varepsilon_G \rangle$ is constant or very small compared to the value of $C_o \langle U_T \rangle$, by plotting a graph of U_{SG} / ε_G against U_T, the value of C_o can be obtained (see Figure 4.14).

The intercept on the y-axis gives an indication of bubble rise velocity because as

$$U_{SG} \longrightarrow 0, \quad \frac{j_{GL}}{\varepsilon_G} \longrightarrow U_{b\infty}$$

The data obtained in this study agree well with those of Hill (1976). The empirical correlation:

$$\frac{U_{SG}}{\varepsilon_G} = 30 + 2.4 \ (U_{SG} + U_{SL}) \qquad \text{........ 4.34}$$

adequately describes the results.

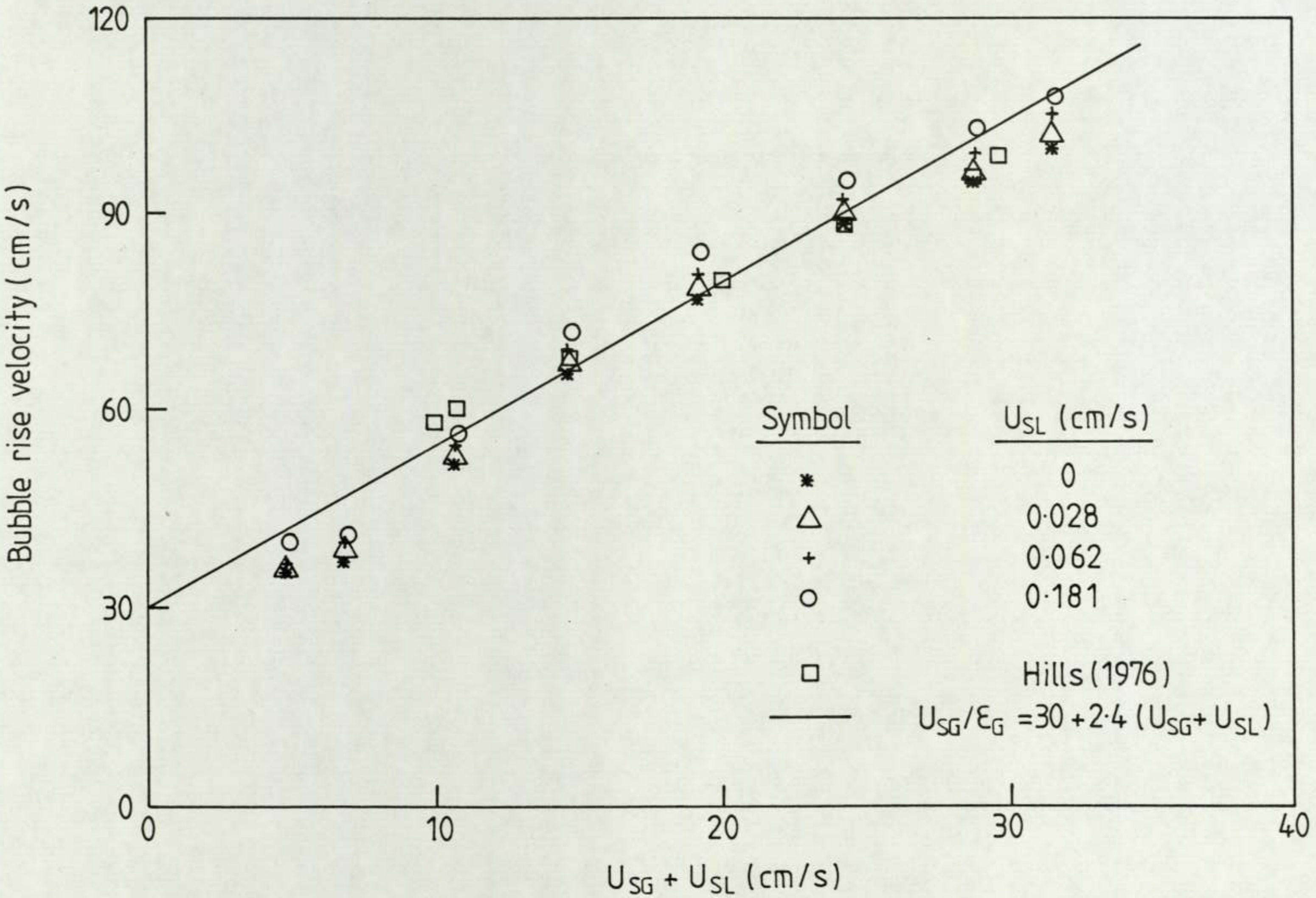
4.7 Corpuscular Analysis of Phase behaviour

4.7.1 Introduction

Two broad approaches to the description of fluid systems exist: these are the continuum approach and the corpuscular approach. In a corpuscular model, it is recognized that, for all real systems, typical behaviour is caused by the concerted actions of components.

Figure 4.14 Bubble rise velocity as a function of total phase velocity.

Two-phase system (Air-Water)



However, in the engineering sciences most of the problems commonly encountered are treated in terms of the continuum approach. In a continuum description, the corpuscular nature of reality is ignored and the system is considered to be continuous in space. The continuum approach is preferred because these models are conceptually simpler and can more readily be treated mathematically. Nevertheless, in view of the present-day image of matter, the corpuscular approach must be considered a more realistic description.

In some instances, continuum models can be formally derived from a corpuscular description by the use of suitable averaging techniques. In these cases, the treatments become equivalent. In other situations, direct equivalence cannot be shown and the continuum treatment must be handled with care.

The corpuscular approach, introduced by Jamialahmadi (1982) and Smith et al (1986), is developed further here and used to explain phase behaviour in gas-liquid systems. The approach considers what happens on the molecular level in systems by analysing the molecular behaviour in bulk phases and at interfaces between phases. The methodology involves analysing each individual phase on a molecular scale, and then considering what happens when these phases come into close contact. Emphasis is given to the physical properties and chemical structure. It is believed that this approach will provide further insight into the behaviour of phases in two- and multiphase-systems.

Structure and Intermolecular Forces of Water

Water, an inorganic compound, has an important characteristic - its unexpectedly high boiling point. This is due to hydrogen-bonding, in which a hydrogen atom serves as a bridge between two electronegative atoms, one being held by a covalent bond and the other by electrostatic forces (as shown below):

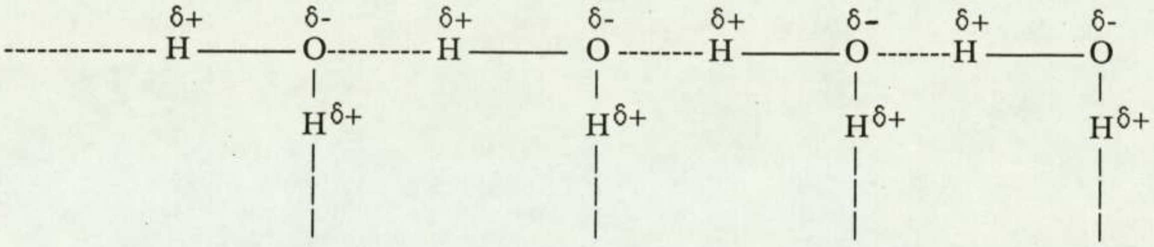


Figure 4.15 - Structure and Intermolecular Forces of Water

Liquids whose molecules are held together by hydrogen bonds are called associated liquids; their abnormally high boiling points are due to the energy required to break the hydrogen bond.

Molecules at the surface of water, or any liquid, are subject to attractive forces on one side only, ie. by the molecules present in the interior, since there are obviously no liquid molecules above the liquid surface. Surface tension is the result of the unbalanced intermolecular forces trying to pull the molecules at the interface towards the interior. The magnitude of surface tension depends upon the nature of such forces.

Structure of the Gas Phase

Air is a combination of 21% oxygen and 79% nitrogen. The polarity of these gas molecules is zero because of zero dipole moment; they are thus regarded as non-polar. Therefore, they cannot form any kind of physical bond with polar liquids, such as water. As a result, at the interface between water and air, the molecules of water at the surface can only form bonds with the interior water molecules. Their exterior sides are thus free and unbonded. It is to be noted that there are gases which behave differently. Examples are carbon dioxide and ammonia, which are polar gases, and have considerable dipole moments.

4.7.2 Dispersion of Gas into Liquid

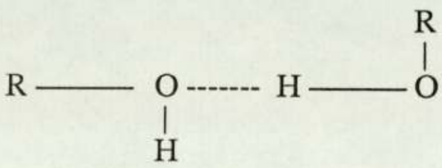
A polar gas will form bonds with a polar liquid causing a reduction in surface tension, bubble size and rise velocity; this leads to an increase in hold-up. Typical examples are: carbon dioxide or ammonia bubbled into water. Non-polar gases and liquids behave likewise, eg. air bubbles in kerosene.

Bubbling a non-polar gas into a polar liquid or a polar gas into a non-polar liquid causes the opposite effect. There is much coalescence due to non-compatibility, leading to lower gas hold-up values. This explains the situation when air is bubbled into water and ammonia into kerosene.

Bubble coalescence has a detrimental effect on column performance. It is reduced or avoided by all possible means. Amongst effective methods are the use of multistage columns and baffles. Another means is the use of additives in the liquid phase, as is discussed below.

4.7.3 Addition of Alcohols

Alcohols are compounds of the general formula ROH, where R is any alkyl or substituted alkyl group. All alcohols contain the hydroxyl (-OH) group, which, as the functional group, determines the properties characteristic of this family. The very polar -OH group contains hydrogen attached to the very electronegative element, oxygen, and therefore permits hydrogen bonding. The physical properties reflect the effects of this hydrogen bonding.



$$\begin{array}{ccccccc} & & & & & & R \\ & & & & & & | \\ R & \text{———} & O & \text{------} & H & \text{———} & O \\ & & | & & & & \\ & & H & & & & \end{array}$$

Like water, alcohols are associated liquids with abnormally high boiling points due to the greater energy needed to break the hydrogen bonds that hold the molecules together.

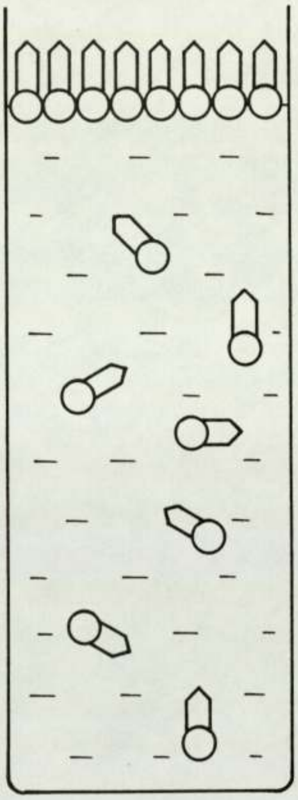
The solubility behaviour of alcohols also reflects their ability to form hydrogen bonds. Since alcohol molecules are held together by the same sort of intermolecular forces as water molecules, there can be mixing of the two kinds of molecules: the energy required to break a hydrogen bond between two water molecules or two alcohol molecules is provided by formation of a hydrogen bond between a water molecule and an alcohol molecule. This is true, only for the lower alcohols, where the -OH group constitutes a large portion of the molecule. A long aliphatic chain with a small -OH group at one end is mostly alkane and its physical properties show this. The change in solubility with carbon number is a gradual one: the first three primary alcohols are miscible with water; n-butyl alcohol is soluble to the extent of 8%; n-pentyl and the higher alcohols still less.

The presence of a small quantity of alcohol reduces the surface energy. Alcohol molecules have a tendency to accumulate at the water surface; when this happens the surface tension goes down because the surface tension of alcohols is low compared to that of water. The nature of the alkyl group (hydrocarbon chain) is such that it tends to be repelled by water while the hydroxyl group mixes with water. This repulsion of the hydrocarbon group away from water tends to concentrate the alcohol molecules at the air-water interface (see Figure 4.16).

Bubbling Air into Aqueous Alcohol Solutions

The overall effect is that the alcohol molecules tend to "anchor" the bubbles to the bulk of the liquid. The strong intermolecular forces existing results in the bubble rise velocity being lowered. The bubbles are of small sizes as coalescence is very much inhibited.

Figure 4.16 Alcohol molecules in water.



The "circles" represent the polar OH group.
The "tails" represent the non-polar hydrocarbon group.

There is much difficulty in bubbles breaking the strong force holding the molecules of solution together. The residence time of bubbles is thus increased giving rise to higher hold-up values than for water alone.

A comparison of dipole moments of the alcohols shows that there is no apparent difference between their hydrophilic groups (OH). However, when comparing the length of their hydrophobic groups, there is an increase from methanol to butanol. There is therefore, a greater tendency for butanol molecules to build up at the air-liquid interface compared with methanol molecules.

It is clear that alcohols cause a reduction in surface tension (see Table 4.2). The degree by which surface tension is reduced depends on the chain length, concentration of alcohol and the efficiency of packing of its molecules at the gas-liquid interface. Due to these effects, the lowering of surface tension of water is in the order:

butanol < propanol < ethanol < methanol < water

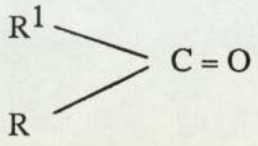
Consequently, gas hold-up is decreased in the order

butanol > propanol > ethanol > methanol > water

The higher alcohols behave differently and become increasingly insoluble. A comparison of the polarity with the polarity of water shows that the hydrogen bonds between water-water molecules are much stronger than those between water and long chain alcohols.

4.7.4 Addition of Ketones

Ketones are compounds of the general formula RR^1CO. The groups R and R^1 may be aliphatic or aromatic.



Ketones contain the carbonyl group, C = O, and it is this that largely determines their properties. The polar carbonyl group makes ketones polar compounds. Hence, they have higher boiling points than non-polar compounds of comparable molecular weight.

By themselves, they are not capable of intermolecular hydrogen bonding since they contain hydrogen bonded only to carbon; as a result they have lower boiling points than comparable alcohols.

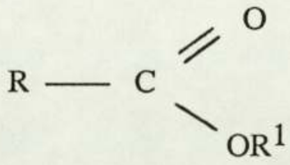
The lower ketones are appreciably soluble in water, presumably because of hydrogen bonding between solute and solvent molecules. Thus the intermolecular forces will increase significantly. Measurement of surface tension also shows a reduction. When air is bubbled into water-ketone solution, there is a resistance to the rising bubbles and coalescence; the effect is similar to that for alcohol solutions. The strong inter-molecular bonds in the bulk causes a reduction in bubble velocity and coalescence; therefore higher hold-up values arise.

Butanone molecules have a higher polarity and a greater tendency for forming bonds. They should thus suppress the coalescence of bubbles more than acetone resulting in higher gas hold-ups in the order:

butanone > acetone (propanone) > water

4.7.5 Addition of Esters

Esters are functional derivatives of carboxylic acids in which the -OH of the carboxyl group has been replaced by $-OR^1$. They are of the general formula $RCOOR^1$; R and R^1 may be alkyl or aryl.



The presence of the C = O group makes esters polar compounds having boiling points about the same as those of ketones of comparable molecular weight.

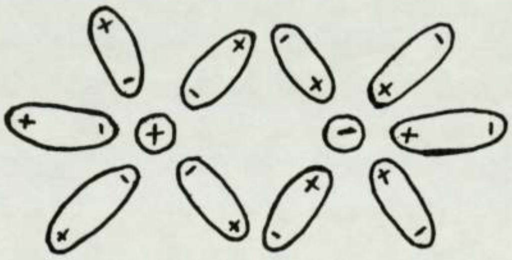
Ethyl acetate is soluble in water to an appreciable extent due to hydrogen bonding between the solute and solvent molecules. A reduction in surface tension may be predicted and is observed from measurement. On bubbling air into ester solutions, the inter-molecular bonds in the bulk cause a reduction in bubble velocity and coalescence leading to higher hold-up values. There is, however, a critical point at which a decrease is observed before further increase. This is due to the bubbles being able to overcome intermolecular forces; at that point little coalescence results.

The higher esters studied in this programme like alcohols, behave differently. This may be due to a decrease in polarity and quality of hydrogen bonding. The strength of anchorage of the films formed by these esters at the bubble surface with the bulk liquid is also weak. The bubbles become more mobile and coalescence results. This leads to a reduction in gas hold-up in the order:

methyl acetate > ethyl acetate > water > propyl acetate > butyl acetate

4.7.6 Addition of Electrolytes

Electrolytes are ionic compounds in which the structural units are positive and negative ions. They also form crystals of extremely strong rigid structure since the electrostatic forces holding each ion in its position are powerful. These powerful inter-ionic forces can be overcome only at very high tempeatures leading to very high melting points. When an ionic compound dissolves, the structural units become separated from each other by solvent molecules. The energy required to break the bonds between solute particles is supplied by the formation of bonds between the solute particles and the solvent molecules. Energy is necessary to overcome the powerful electrostatic forces holding an ionic lattice. In general, only water and a few other highly polar solvents are able to dissolve ionic compounds appreciably. In solutions each ion is surrounded in the bulk by a cluster of water molecules.



Electrolytes in water cause an increase in surface tension and an electrical double layer is formed at the gas-liquid interface. As a result, the bubble rise velocity and the tendency for bubbles to coalesce decrease.

The overall result is an increase in gas hold-up compared to that in air-water systems. However, there is a critical point up to which the non-coalescing tendency increases significantly and after which the effect becomes negligible. This occured in the author's experiments at a gas velocity greater than 8 cm/s, leading to a decrease in gas hold-up. The value of gas hold-up also depends on the quality of ionic bond of the electrolyte.

4.7.7 Formation of Foam

Solutions of alcohols and ketones lead to the formation of foam at high gas velocities (> 10 cm/s). This is a problem of considerable technical importance.

It is to be noted that pure liquids do not foam. Production of foams, thus, necessarily involves mixing a surface-active compound in water or less commonly another liquid. The compound having dissolved in water can be viewed as being made up of two parts. One part is hydrophilic suggesting thereby that it can mix with water; such parts are the polar groups. On the other hand, the other (usually hydrocarbon) part is non-polar and will not mix with water. Such hydrophobic parts tend to migrate away from the bulk and project towards air.

When a molecule consists of two segments, one hydrophobic and the other hydrophilic, that molecule tends to orient partly towards the water phase and partly towards gas or air phase. Thus it serves as a "link" between water and air. This causes frothing or foaming, whereby a large number of air bubbles persist at the water surface. This was observed experimentally. Increasing molecular weight results in a loss of solubility and so the properties of the interface will become like those associated with slightly soluble surfactants.

4.8 Conclusions

(1) Gas hold-up varies with operating parameters, column design and nature of phases constituting the system.

(2) Increase in gas velocity increases gas hold-up while increase in liquid velocity, height, temperature, column diameter cause a decrease.

(3) Both organic solutes and electrolyte affect coalescence behaviour of aqueous solutions drastically within a narrow concentration range, changing from quick coalescence as in water to coalescence restraining; bubble-rise velocity is also reduced.

(4) Whilst aqueous solutions of alcohols, ketones, lower esters and electrolytes enhance gas hold-up, the higher esters cause a decrease.

(5) For the organics, considerable increase in gas hold-up with chain length was observed with foaming at very high gas velocities.

(6) The Slip Velocity Model and Zuber-Findlay Model are useful in presenting correlations (eg. for air-water system).

(7) The corpuscular approach is systematically developed and used to give insight into the behaviour of phases in process systems. The effects of organic and inorganic constituents on bubble size, rise velocity, coalescence and hold-up can thus be explained. With this approach, it is now possible to predict the effect of any compound on column behaviour and performance. This complements the continuum approach.

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CHAPTER 5

FLOW REGIMES AND BUBBLE DYNAMICS IN VERTICAL TWO-PHASE FLOW

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FLOW REGIMES AND BUBBLE DYNAMICS IN VERTICAL TWO-PHASE FLOW

5.1 Introduction

Bubbling a gas into a column of liquid through a distributor usually results in streams of bubbles, the characteristics of which depend on various parameters. Of these characteristics, one of the most important is the size of the bubbles themselves; this parameter along with gas hold-up determines the surface area available for mass- and heat-transfer in gas-liquid operations. Of equal importance are the bubble rise velocity and bubble size distribution; they have a direct bearing on the performance of columns and influence both design and scale-up.

The flow regime in bubble columns is also an important factor affecting the hydrodynamics, transport and mixing properties. Pressure drop, hold-up of various phases, interfacial areas, mixing coefficients and interphase mass- and heat-transfer coefficients depend strongly on the prevailing flow regime.

5.2 Literature Review

5.2.1 Flow Regimes

Many investigators (Govier and Aziz, 1972; Lockett and Kirkpatrick, 1975; Wallis, 1969; Kawagoe et al, 1976; Hills, 1976; Miller, 1980; etc) have proposed different criteria to differentiate flow regimes. Wallis characterized the upward movement of bubble swarms into three separate flow regimes: these regimes occur in order of increasing flow rate.

Bubbly Flow or Quiescent Bubbling

This regime is characterized by almost uniformly sized bubbles with uniform radial distribution. The regime occurs if the superficial gas velocity is less than about 5 cm/s (Fair, 1967) and the rise velocity of the bubbles lies between 18 and 30 cm/s (Levich, 1962). The theory of bubble flow has been developed by Lapidus and Elgin (1957), Richardson and Zaki (1954) and Wallis (1962).

Churn Turbulent or Heterogeneous Regime

At higher gas velocities, the homogeneous gas-in-liquid dispersion cannot be maintained and an unsteady flow pattern occurs. This heterogeneous flow regime is characterized by large bubbles moving with high rise velocities in the presence of small bubbles (Hills and Darton, 1976). The large bubbles take the form of spherical caps with a mobile and flexible interface.

Slug Flow Regime

In small diameter columns, at high gas flow rates, large bubbles are stabilized by the column wall leading to the formation of bubble slugs. Bubble slugs can be observed in columns of diameters up to 15cm.

The various flow regimes are schematically presented in Figure 5.1. The dependence of flow regime on column diameter and gas velocity can be roughly estimated from Figure 5.2 (after Shah et al, 1982).

Kawagoe et al (1976) studied gas hold-up as a function of different flow regimes and analysed the data in each of the three regimes in different ways. They found that gas hold-up is linearly proportional to the gas velocity in the bubbly flow regime, while it is almost independent of gas velocity in the churn turbulent regime.

Figure 5.1 Flow regimes in bubble column.

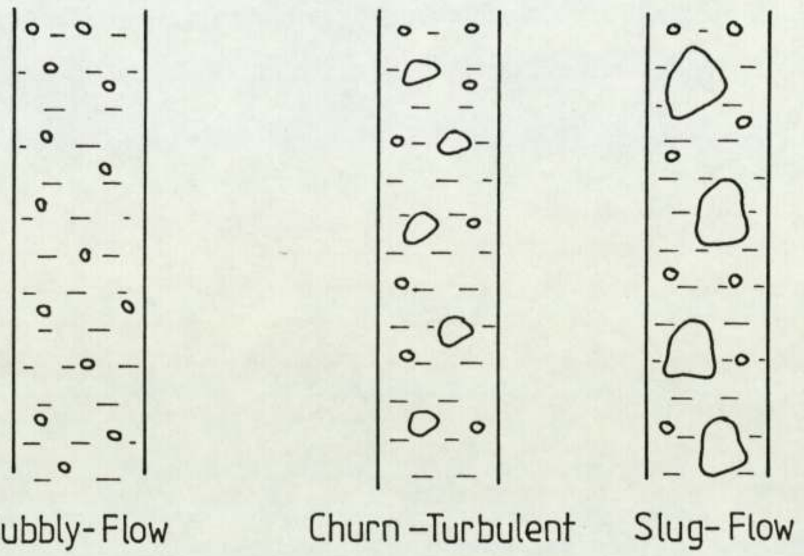
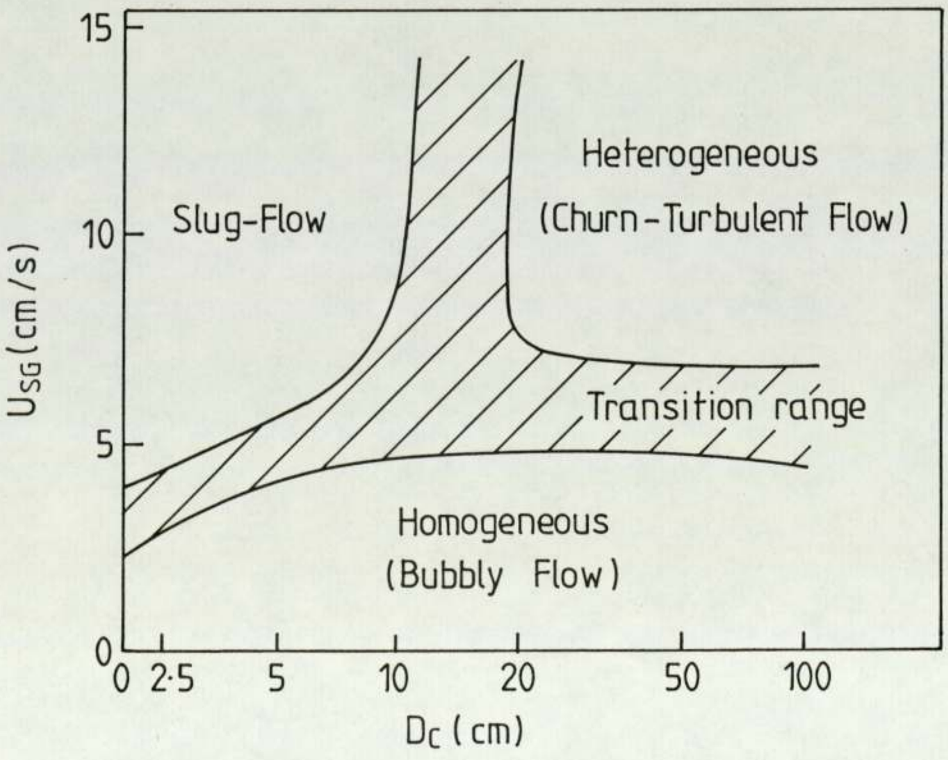


Figure 5.2 Dependence of flow regime on gas velocity and column diameter



Flow regimes have been characterized by some investigators either using conductance probes or by visual observation (Oshinowo and Charles, 1974; Barnea et al, 1980). Comprehensive flow regime charts have been presented by Govier and Aziz (1972) and a more detailed discussion is given by Miller (1980).

Bach and Pihofer (1978) have suggested that the churn-turbulent regime is most commonly encountered in industrial bubble columns. Knowledge of the transition from bubbly flow to churn-turbulent and slug flow regimes is important because the system performance depends strongly on flow regime.

5.2.2 Bubble Dynamics

Measurement of Bubble Size

Many methods are available to determine bubble size. Photographic techniques are widely used because of their simplicity (Akita and Yoshida (1974), Burckhart and Deckwer (1975), Schugerl et al (1977)). Other methods are based on light scattering and light reflection (Calderbank et al (1960)). In the past decade, various optical and electrical probes have been developed to determine bubble sizes (Burgess and Calderbank (1975), Calderbank and Pareira (1977), Yamashita et al (1979), Koide et al (1979)). A dynamic bubble disengagement technique has been described for estimating bubble size distribution (Sriram and Mann (1977), Vermeer and Krishna (1981)). Though the original bubble size distributions obtained from the various techniques differ, the volume-to-surface mean bubble diameters, d_{VS}, (sauter mean diameter) tend to be similar.

Correlations for Bubble Size

On the basis of Kolmogoroff's isotropic turbulence theory, Calderbank (1967) proposed a number of equations for d_{vs} in stirred vessels. All the equations have the general structure.

$$d_{vs} = C \frac{\sigma^{0.6}}{(P/V_D)^{0.4}\, \rho_L^{0.2}} \, \varepsilon_G^{\,n} \left(\frac{\mu_G}{\mu_L}\right)^{0.25} \qquad \text{........ 5.1}$$

where the constant C and exponent n depend on the stirrer type and the kind of liquid phase. (P/V_D) represents the energy dissipation rate per unit volume of dispersion. If the gas is sparged by two-phase nozzles, the correlation also applies to bubble columns (Unno and Inoue (1980), Joshi (1980), Kubie (1981)).

When the gas is sparged by single orifices, perforated or sintered plates, the diameter of the generated bubbles is only slightly dependent on the gas velocity (Akita and Yoshida (1974), Schugerl et al (1977), Deckwer et al (1978)). Akita and Yoshida determined the bubble size distribution in bubble columns using a photographic technique. The gas was sparged through perforated plates and single orifices, and various liquids (water, aqueous and pure glycol, methanol, carbon tetrachloride) were used. On the basis of their data, they proposed the following correlation:

$$\frac{d_{vs}}{D_c} = 26 \left(\frac{D_c^2\, g\, \rho_L}{\sigma}\right)^{-0.5} \left(\frac{g D_c^3}{\nu_L^2}\right)^{-0.12} \left(\frac{U_{SG}}{\sqrt{g D_c}}\right)^{-0.12} \qquad \text{........ 5.2}$$

for columns up to 30cm in diameter and gas velocities up to 7 cm/s. For $D_c > 30$cm, bubble diameter becomes independent of column diameter and they found no effect of orifice diameter in the bulk region away from the sparger.

Bubble sizes and distributions in various hydrocarbons, organic liquids, highly viscous Newtonian and non-Newtonian solutions have been reported in the literature. (Quicker and Deckwer (1981), Buchholz et al (1978), Franz et al (1980), Nakanoh and Yoshida (1980)).

Kumar et al (1976) and Bhavaraju et al (1978) have proposed correlations for bubble sizes which involve the orifice Reynolds Number.

Coalescence and Break-up

Additionally, bubble diameter depends on the specific gas-liquid system and its properties with respect to coalescence. Coalescence is significantly influenced by the physical properties of the liquid (Chapter 4). Calderbank (1967) and Vasalos et al (1980) reported high rates of bubble coalescence in highly viscous liquids. Low viscosity liquids are observed to show significant break-up. As the author has found in the case of aqueous solutions, the coalescence properties are mainly dependent on the added salts and the organic substances present. The bubble diameter d_{VS} is the actual stable bubble size. If d_B (diameter of bubbles as formed) is greater than or less than d_{VS}, then break-up or coalescence will occur.

Rise Velocity of Bubble Swarms

The mean rise velocity of the bubbles in a bubble swarm is equivalent to the interstitial gas velocity. In the bubbly flow regime, the rise velocity may vary from 3 to 20 cm/s (Schugerl et al 1977, Oels et al 1978). If the flow is churn-turbulent, rise velocities are considerably higher (> 80 cm/s). The transition from bubbly to churn-turbulent flow is usually accompanied by a sharp increase in rise velocity.

The velocity of bubbles when gas is bubbled steadily through a liquid is given by:

$$U_B = \frac{Q}{\varepsilon_G A} \qquad \text{........ 5.3}$$

Also the velocity of bubble relative to stagnant liquid above them (buoyancy velocity) is given by:

$$U_o = (1 - \varepsilon_G)\ U_B$$

ie.

$$U_B = \frac{U_o}{(1 - \varepsilon_G)} \qquad \text{........ 5.4}$$

Thus from equations 5.3 and 5.4,

$$U_B = U_o + \frac{Q}{A} \qquad \text{........ 5.5}$$

The detailed theoretical development of equations 5.3 to 5.5 is given by Nicklin (1962).

5.3 Experimental Programme and Procedure

Flow regimes were studied in the two- and three-dimensional bubble columns descriptions of which are given in Chapter 3. Air was used as the gas phase. This was bubbled at different flow rates into a static column of liquid. Flow regimes were characterized by:

- visual observations
- photographic techniques
- qualitative analysis.

Visual observations of the prevailing flow patterns were made and changes noted. This was aided by using the photographic techniques. A JVC (GX-N7OE) Colour Video Camera was used to record various flows; these were analysed on a TV screen by frame holding where necessary. Still photographs of operations were taken at selected gas velocities using a Canon AE1 programmable camera. From hold-up measurements, flow regimes were qualitatively deduced using the drift flux model.

Using superficial gas velocity and buoyancy velocity measurements, bubble rise velocity was determined (see equations 5.3 - 5.5). The photographic technique was employed in measuring bubble sizes. Typical results are presented in Chapter 7.

The Gas Disengagement Method

The theory has been described in Chapter 4. For the 2-D column, a cine camera was used to follow changes in dispersion height. For easy measurement of height, an adhesive measuring tape was fixed on the column to indicate the liquid level. The column was covered at the back with black material to provide a black-background, and front lighting was provided by two flood-lights at different angles to the column. Liquid was fed in to a known height and the gas (air) was then supplied at fixed flow rates. After some minutes, the gas flow was suddenly cut off and the 8mm cine-camera was used to film the gas-liquid level during gas-disengagement at 8 frames per second. The gas-liquid section was filmed at different flow rates and the dynamic gas hold-up was calculated as a function of time. This was achieved by projecting the film and "freezing" the picture every 4 frames, corresponding to a time interval of 0.5 seconds. The dynamic gas hold-up was calculated by the displacement method for each time interval until gas disengagement was complete.

The whole process was repeated with the 3-D column using a video-camera and making direct analysis on a TV screen.

Experimental Programme

The systems studied were:

- Air-water
- Air-water with additives in the liquid phase. The additives were alcohols, ketones, esters and electrolytes.

The operating conditions were similar to those used in Chapter 4. The experimental programme entailed a careful study of the effects of operating conditions and the nature of the liquid phase on flow regimes and bubble dynamics.

5.4 **Experimental Results**

The observed flow patterns in different systems at different operating conditions are presented in Figures 5.3 and 5.11 to 5.15. From these Figures, the effects of superficial gas velocity, liquid velocity and organic additives on flow regimes and bubble dynamics are clearly portrayed.

Effect of Superficial Gas Velocity

At low flow rates (< 3 cm/s), bubbles detaching from the gas distributor are about 5mm in diameter and ascend through the liquid phase without colliding or coalescing. The degree of backmixing at the sides of the column is also relatively low (see Figure

5.3). At higher superficial gas velocities (> 5 cm/s), the bubbles coalesce within a few centimeters from the gas distributor and large bubbles form; these ascend at the centre of the column along a wave-like path.

As the gas velocity is further increased, bubbles get bigger resulting in the formation of slugs. The slugs transport a considerable amount of liquid in their wakes; also a large number of small bubbles rise with the slugs. When the slugs leave the system, the small bubbles in the wake cannot overcome the downwards liquid flow and so may get dragged downwards with the liquid at the sides of the column. Increasing the gas velocity further leads to very fast movement of the bubbles. The flow is unsteady and takes on a wavy form. Depending on the initial liquid height in the column, turbulence may result (see Figure 5.4).

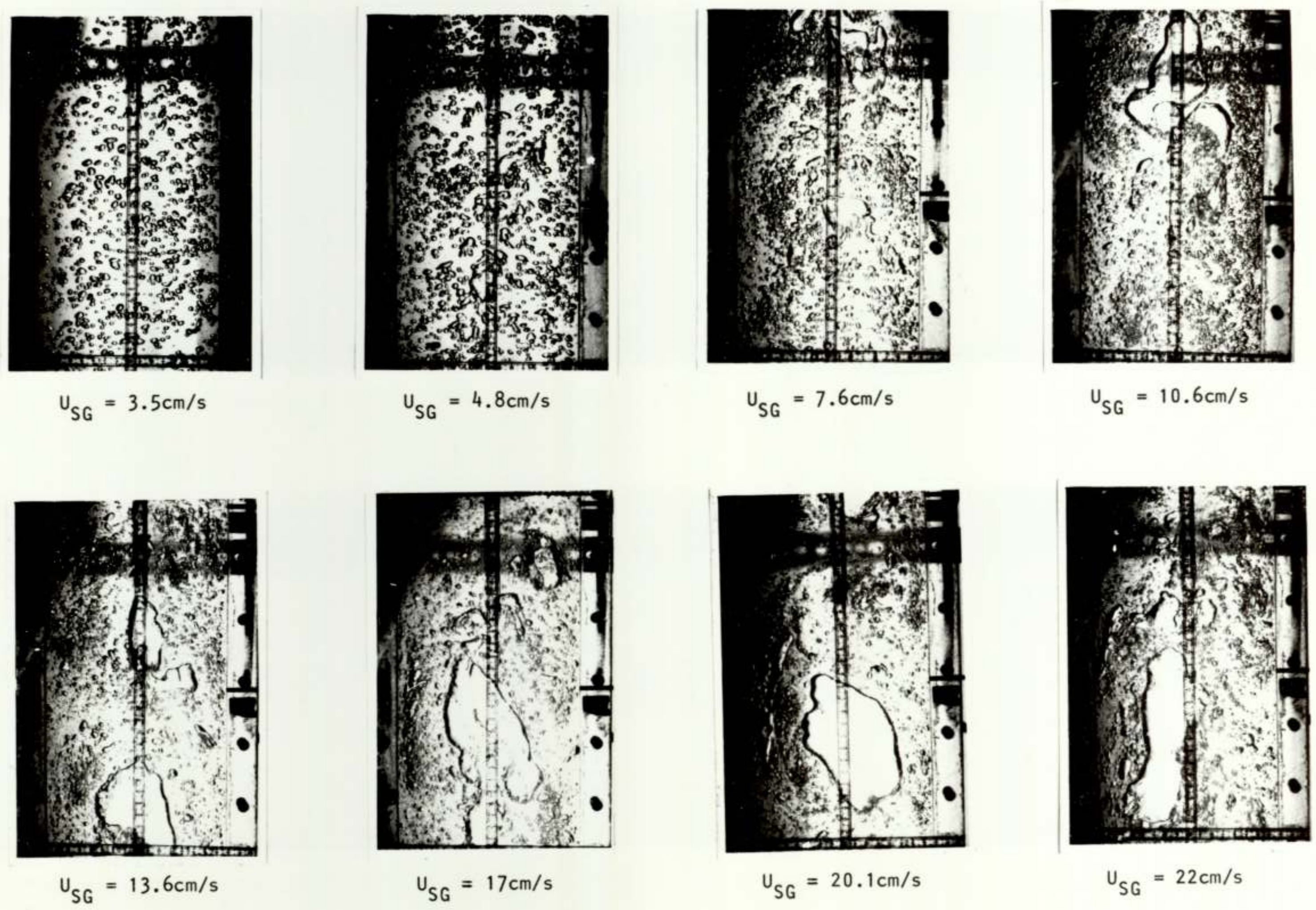
Bubble Rise Velocity, Buoyancy Velocity, Bubble Size Distribution

Variations of bubble rise velocity and buoyancy velocity with superficial gas velocity are shown in Figures 5.5 and 5.6. Bubble rise velocity increases with increasing gas velocity whilst buoyancy velocity decreases. These follow the expected trend (see equations 5.3 - 5.5). Figure 5.5 also shows the effect of liquid additives on bubble rise velocity. The rise velocity passes through a minimum before increasing at higher gas velocities.

Bubble size distribution for air-water system is quantified using the dynamic gas disengagement method. The gas hold-up during gas disengagement is plotted as a function of time for three different liquid heights as shown in Figures 5.7 to 5.9. Figure 5.10 shows a typical fall of liquid level with time during gas disengagement.

All the data used in plotting Figures 5.5 to 5.10 are given in Tables C1-C6 of Appendix C.

Figure 5.3 Effect of superficial gas velocity on flow regimes and bubble dynamics (two phase flow)



U_{SG} = 3.5cm/s

U_{SG} = 4.8cm/s

U_{SG} = 7.6cm/s

U_{SG} = 10.6cm/s

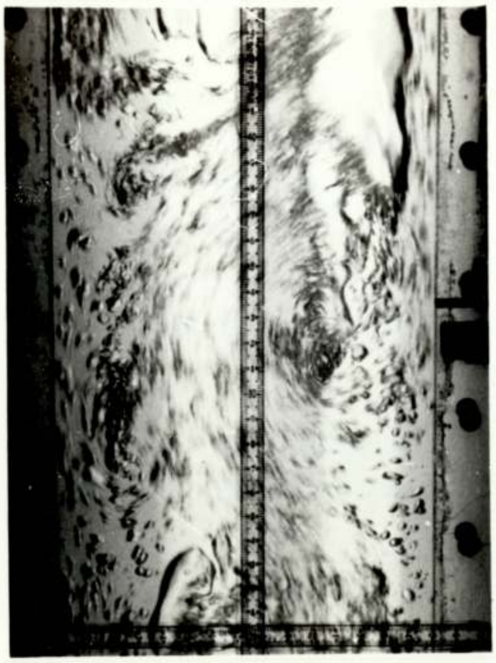
U_{SG} = 13.6cm/s

U_{SG} = 17cm/s

U_{SG} = 20.1cm/s

U_{SG} = 22cm/s

Figure 5.4 "Turbulence" Flow Pattern



(a) $U_{SG} \geq 18$cm/s (low initial liq. height)

(b) $U_{SG} > 20$cm/s (low initial liq. height)

Figure 5.5 Variation of bubble velocity with superficial gas velocity.

Two-phase system (Gas - Liquid)

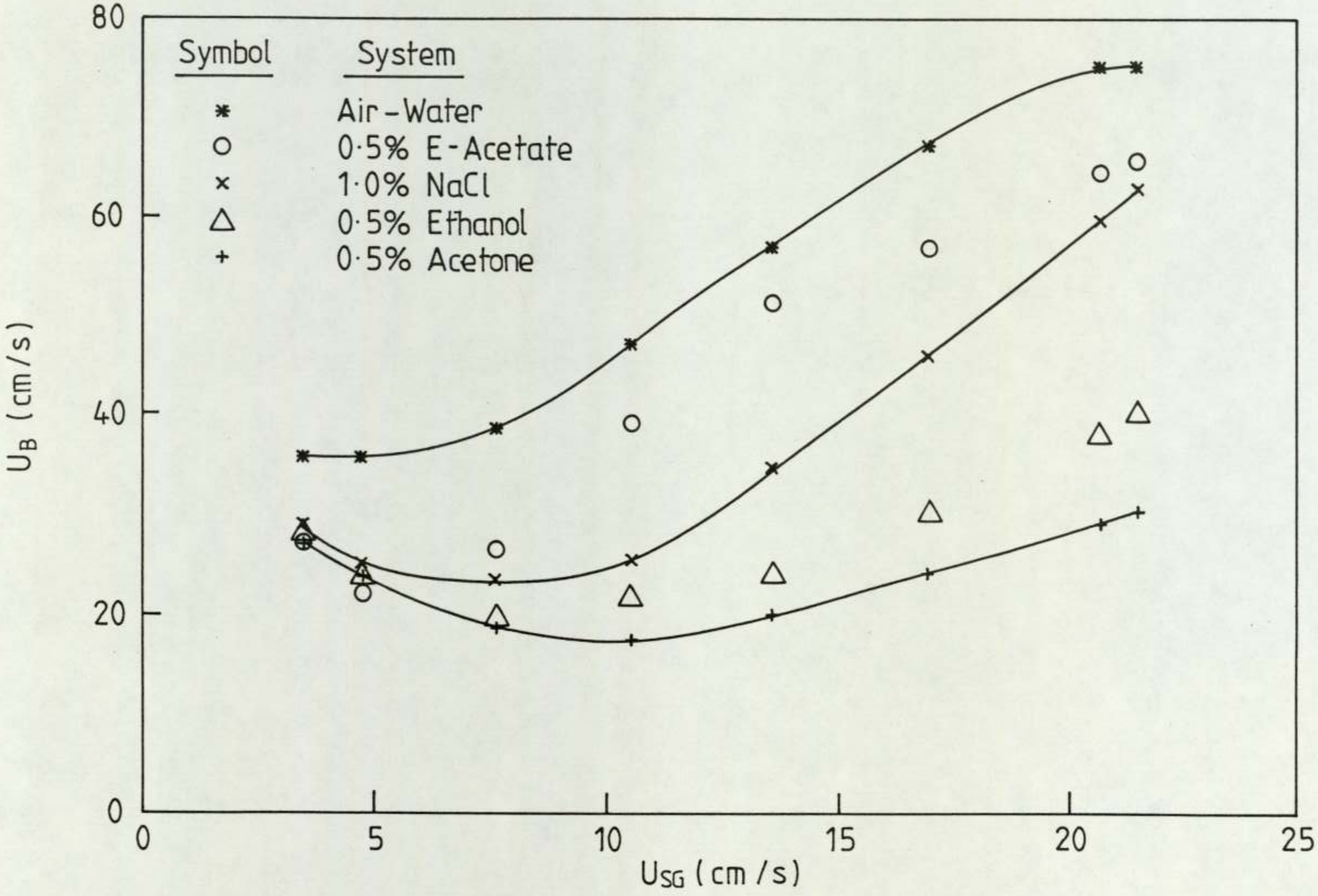


Figure 5.6 Typical variation of buoyancy velocity with superficial gas velocity.

Two-phase system (Air - Water)

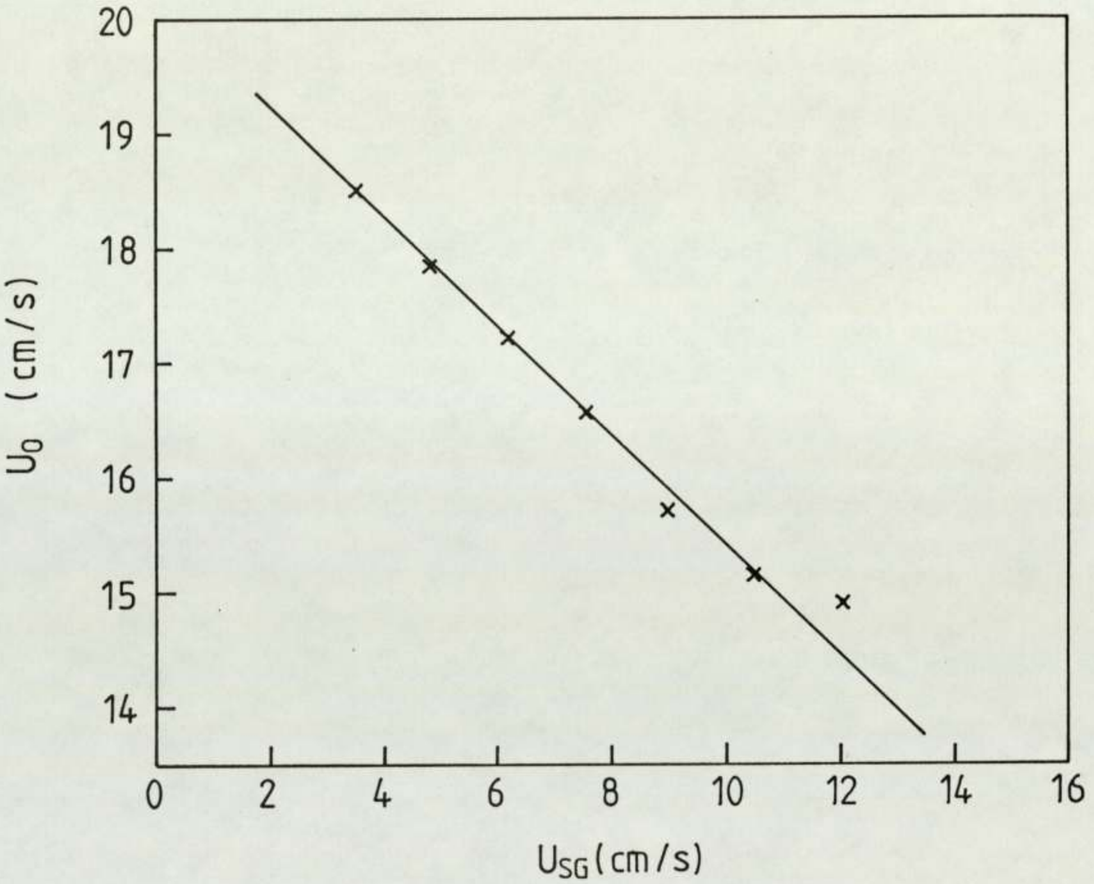


Figure 5.7 The variation of dynamic gas hold-up with time in the three dimensional bubble column (Li = 90cm)

Two-phase system (Air - Water)

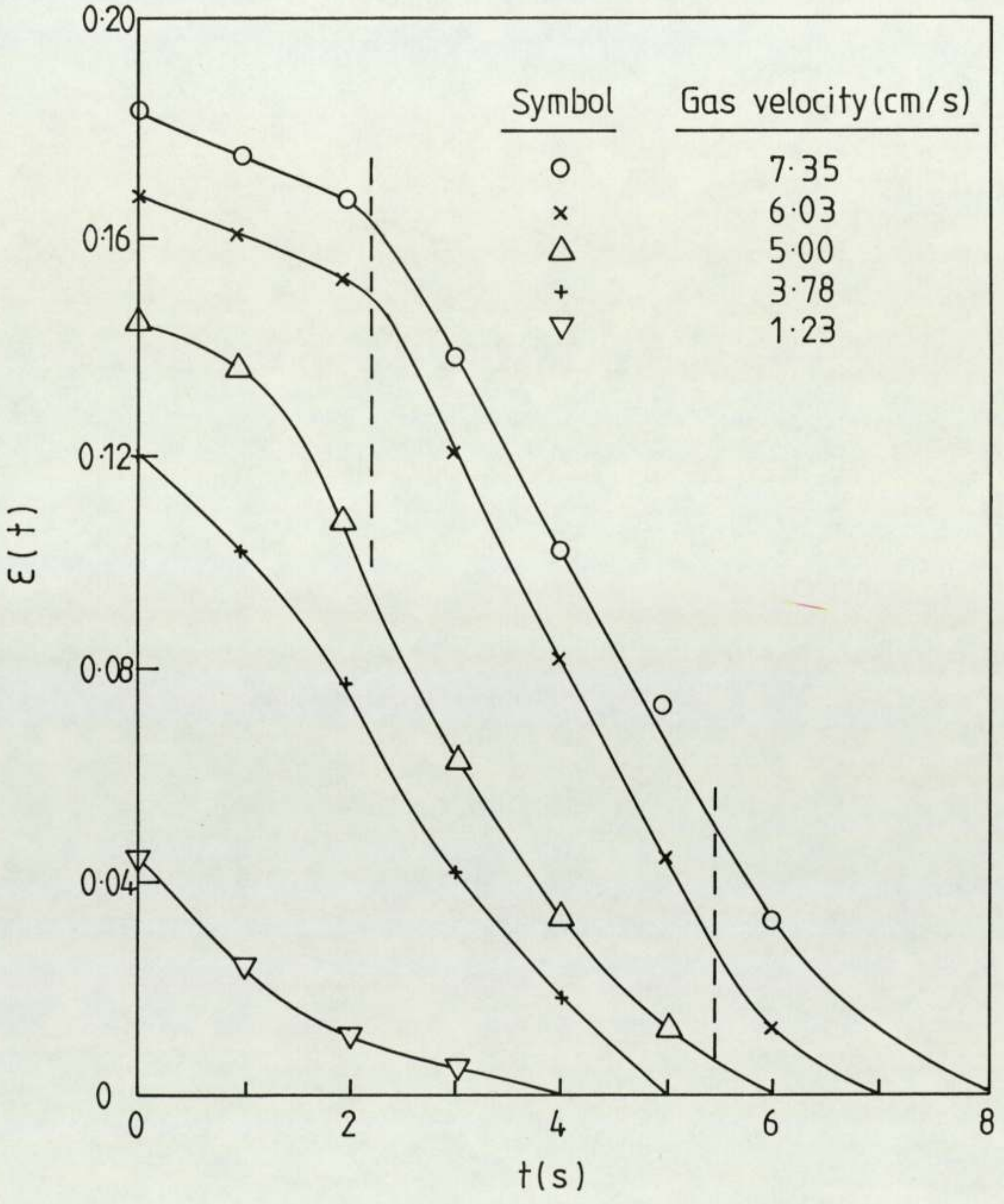


Figure 5.8 The variation of dynamic gas hold-up with time in the three dimensional bubble column (Li = 100cm)

Two-phase system (Air-Water)

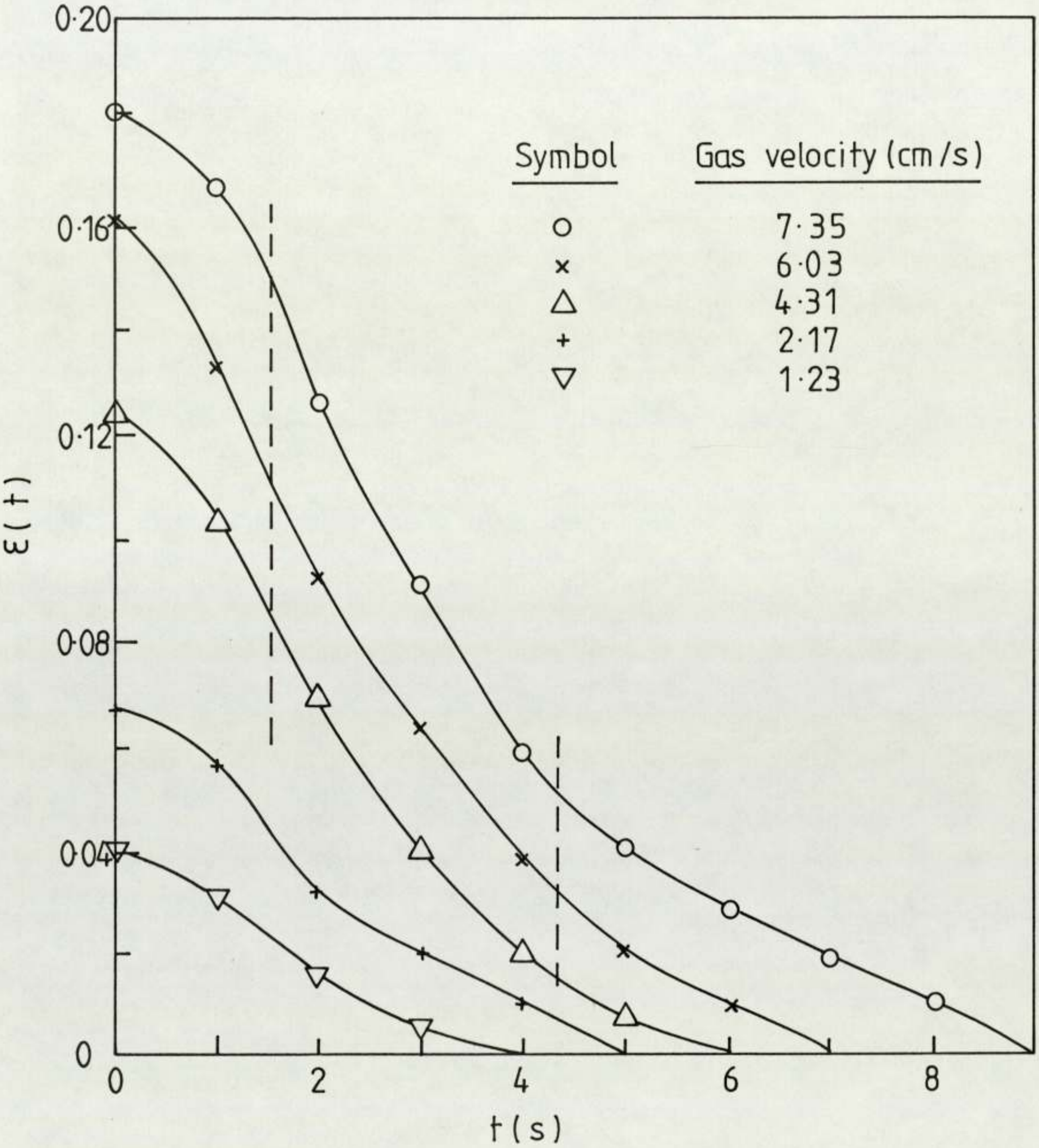


Figure 5.9 The variation of dynamic gas hold-up with time in the three dimensional bubble column. (Li = 140 cm)

Two-phase system (Air - Water)

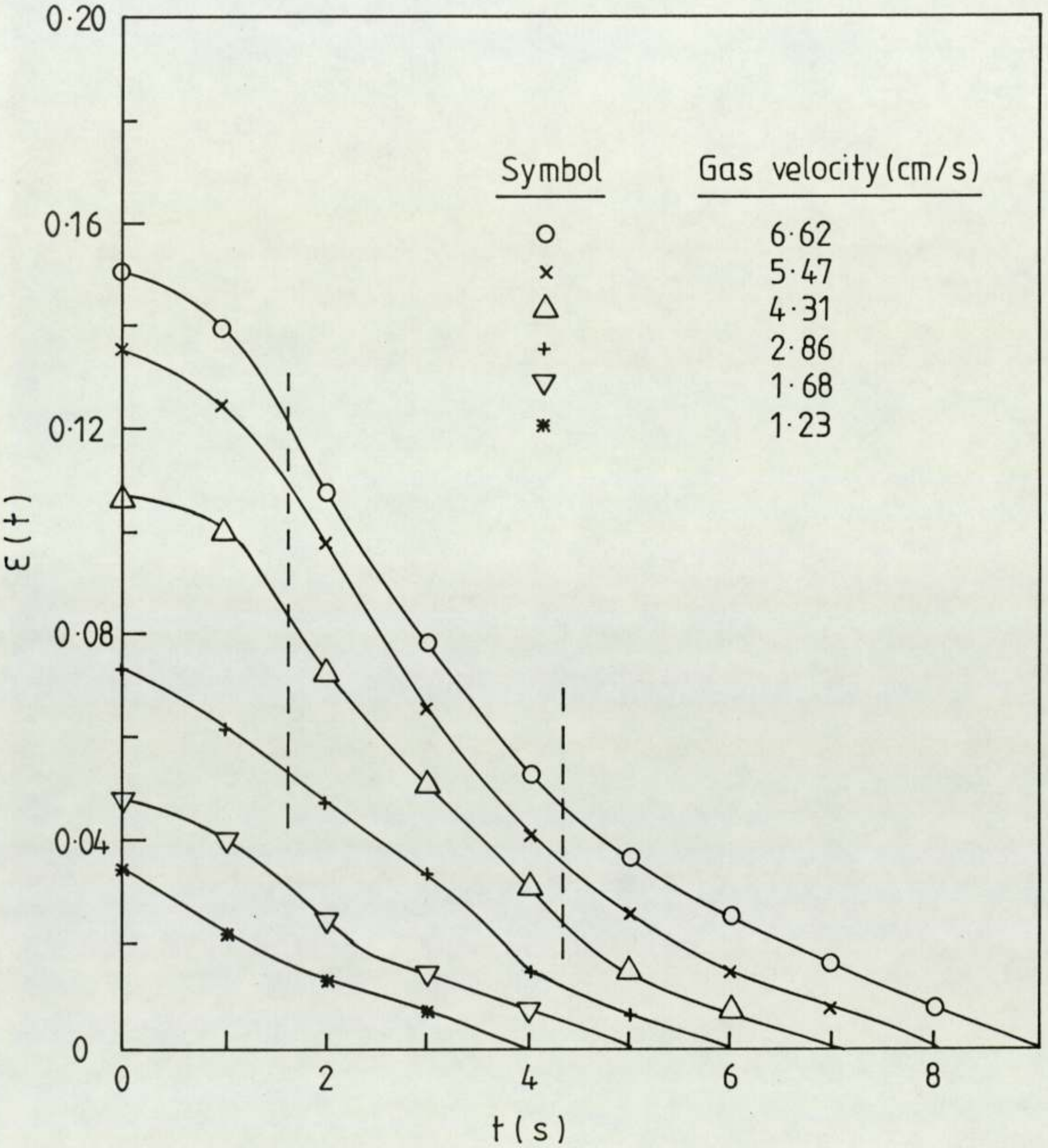
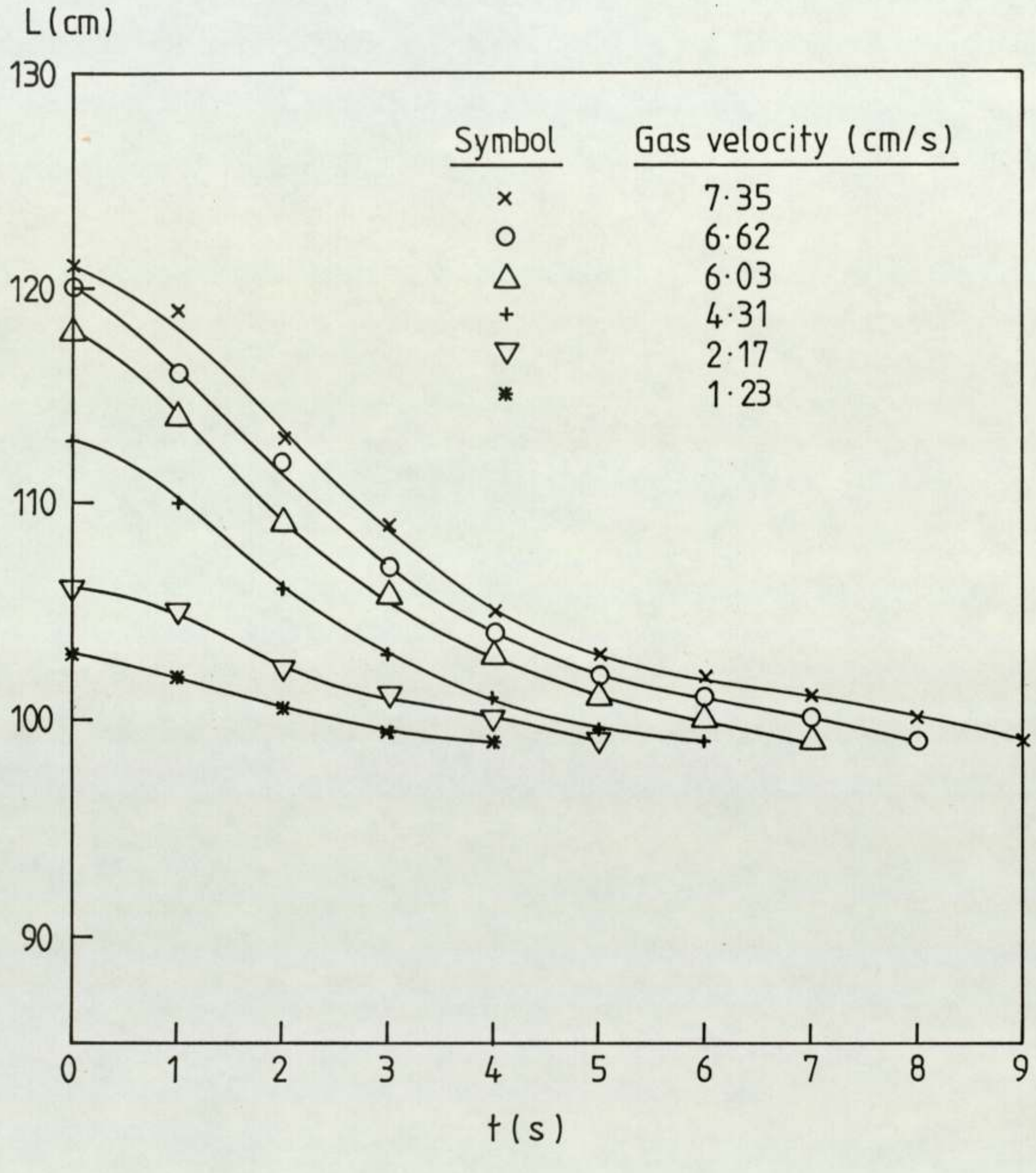


Figure 5:10 Fall of liquid level with time
Initial liquid height - 100 cm



Effect of Superficial Liquid Velocity

The effect of superficial liquid velocity on flow regime and bubble dynamics is shown in Figures 5.11. Increasing liquid velocity seems to enhance bubble coalescence; this effect is not very significant.

Effect of Liquid Additives

Figures 5.12 to 5.15 show the effect of organic additives on flow regimes and bubble dynamics. The bubbles are small in size and uniformly shaped. Aqueous solutions of most of the organic compounds studied by the author resulted in non-coalescence of the bubbles, lowering bubble rise velocity (Figure 5.5) and increasing the bubble residence time (see Chapter 4). These account for the relatively high gas hold-up data compared to those for air-water systems. Even at very high gas flow rates the bubbly flow regime was still maintained. However, foaming was observed in some cases. The effect on flow regime and bubble dynamics increases with the chain length of the additive.

5.5 **The Drift Flux Model**

As mentioned in Chapter 4, the drift-flux model is essentially a separated flow model in which attention is focussed on the relative motion rather than on the motion of the individual phases. The drift flux of the gas is defined as the volumetric flux of gas relative to the surface moving at an average velocity (Wallis 1969). It is given as:

$$j_{GL} = U_S \, \varepsilon_G \, (1 - \varepsilon_G) \qquad \text{........ 5.6}$$

Figure 5.11 Effects of superficial gas and liquid velocities on flow regimes and bubble dynamics (two phase flow)

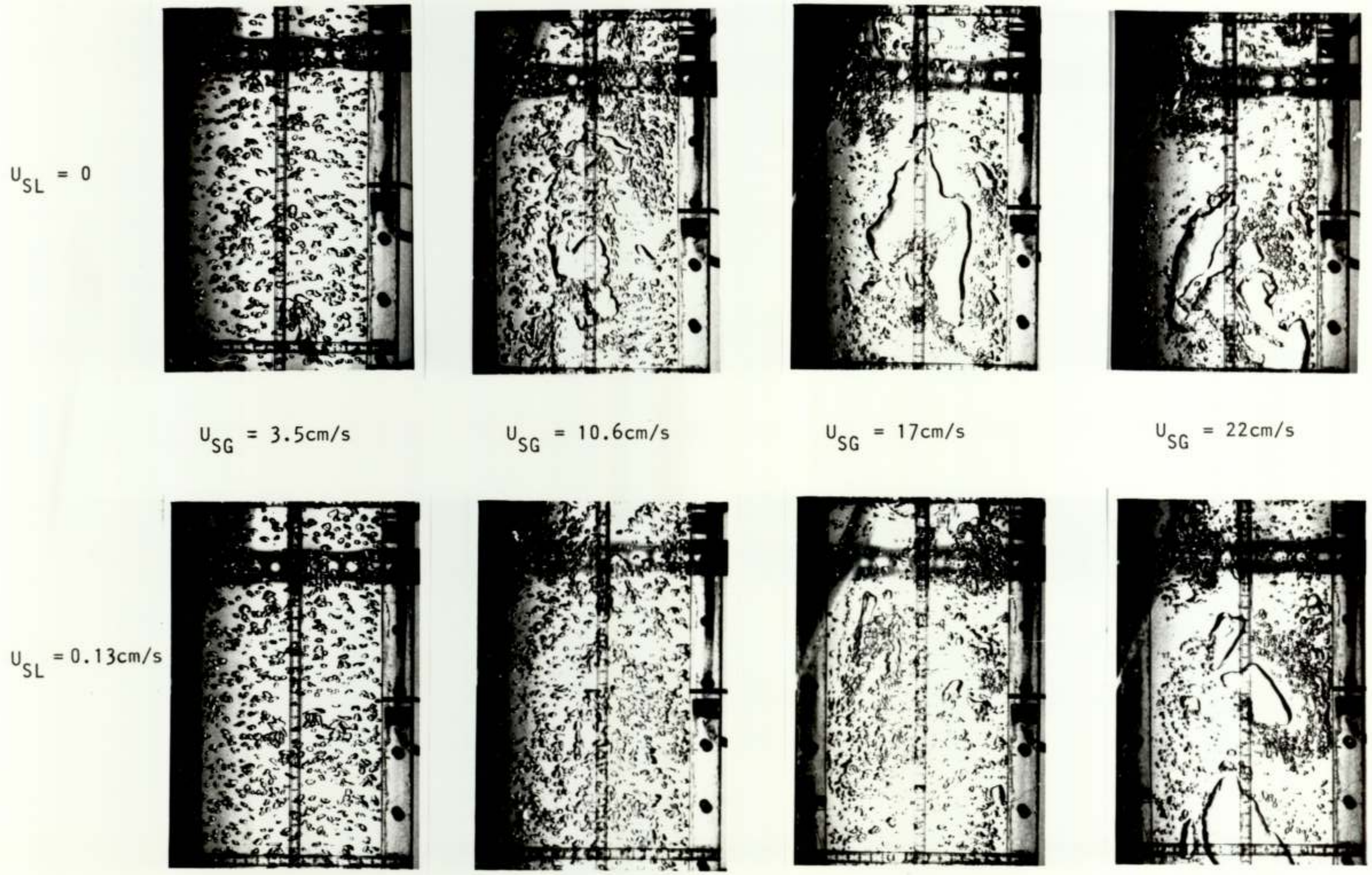


Figure 5.11 (continued)

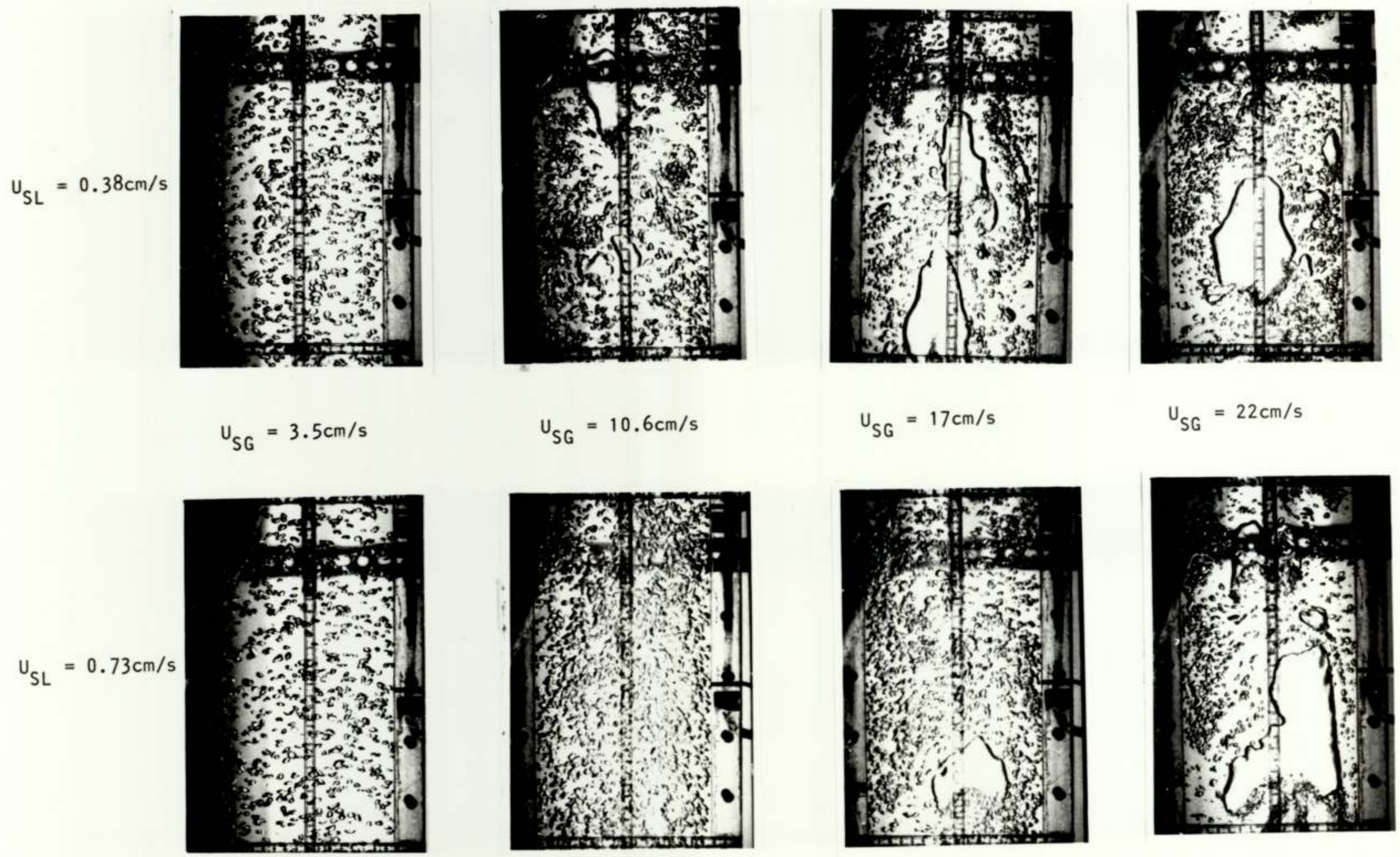


Figure 5.12 Effect of methanol and ethanol on flow regimes and bubble dynamics (two-phase flow)

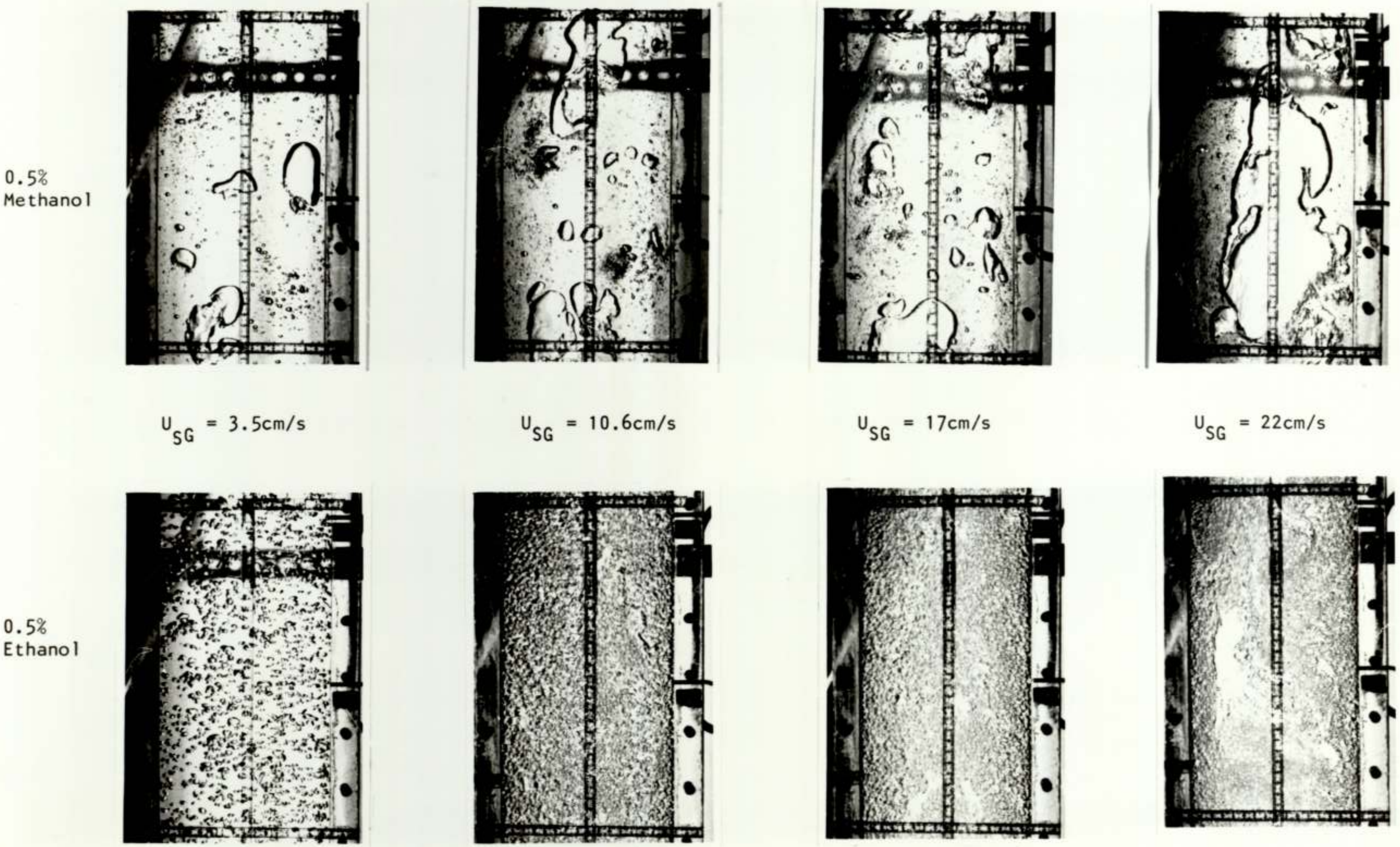


Figure 5.13 Effects of propanol and butanol on flow regimes and bubble dynamics (two-phase flow)

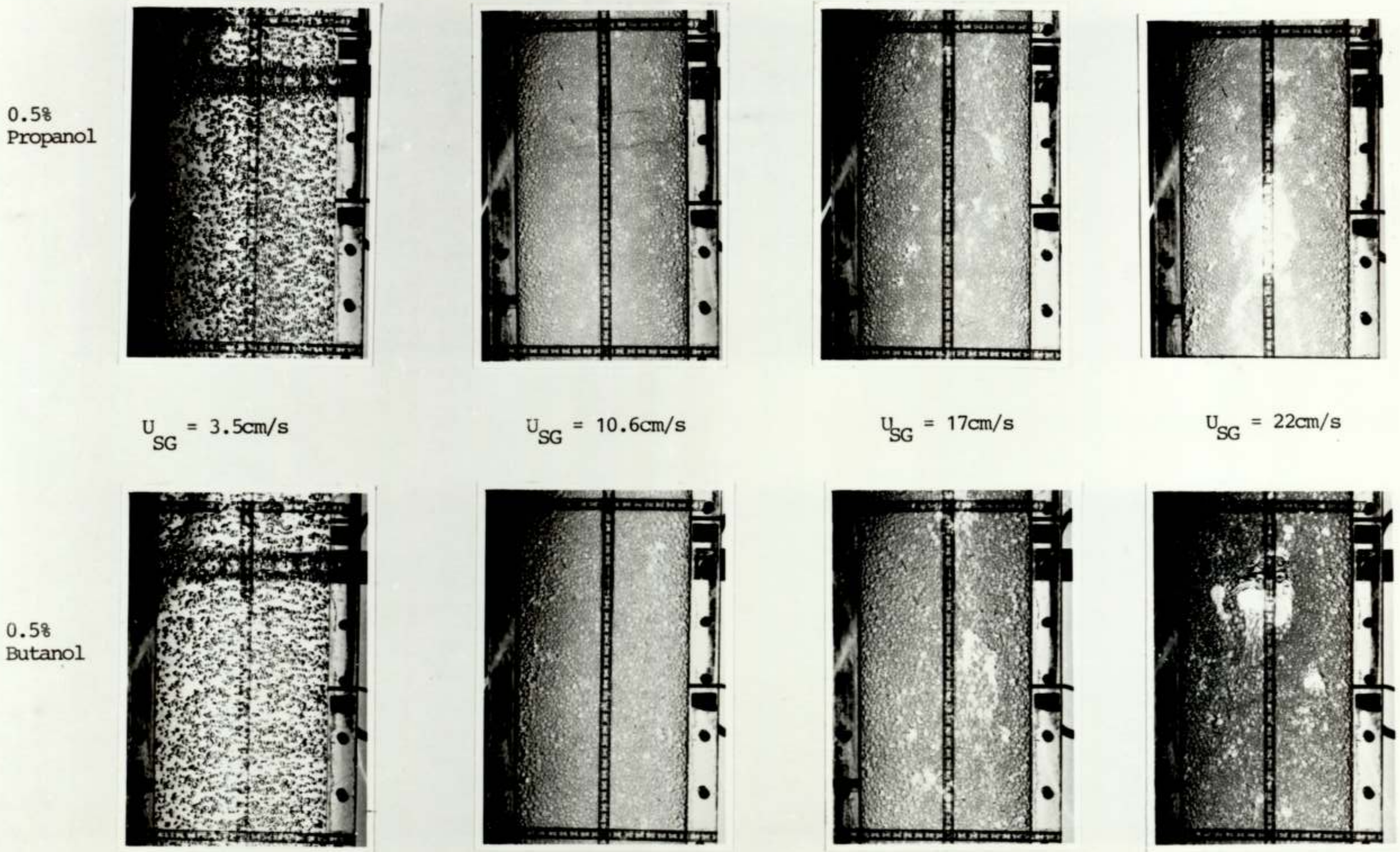


Figure 5.14 Effects of propanone and butanone on flow regimes and bubble dynamics (two-phase flow)

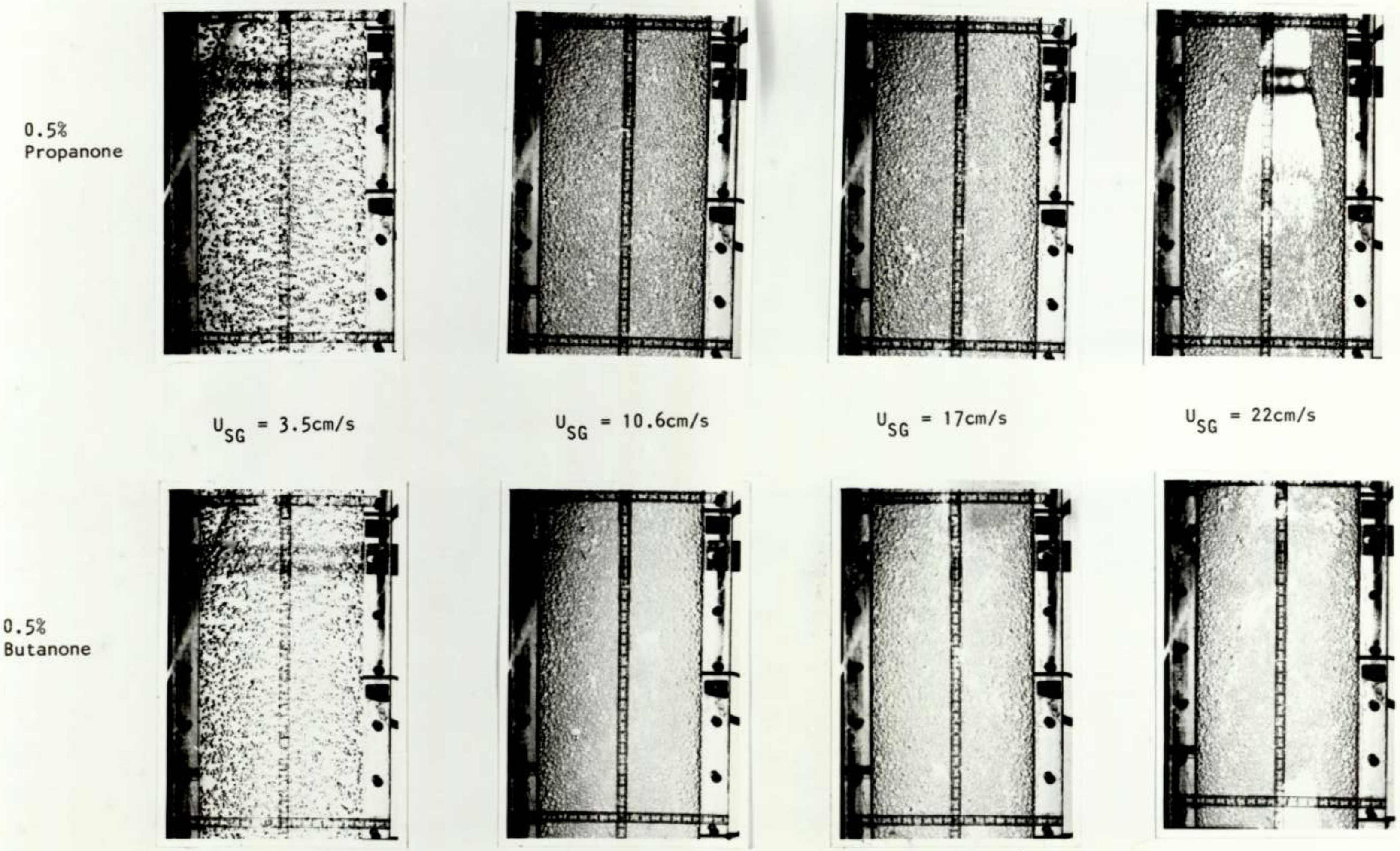
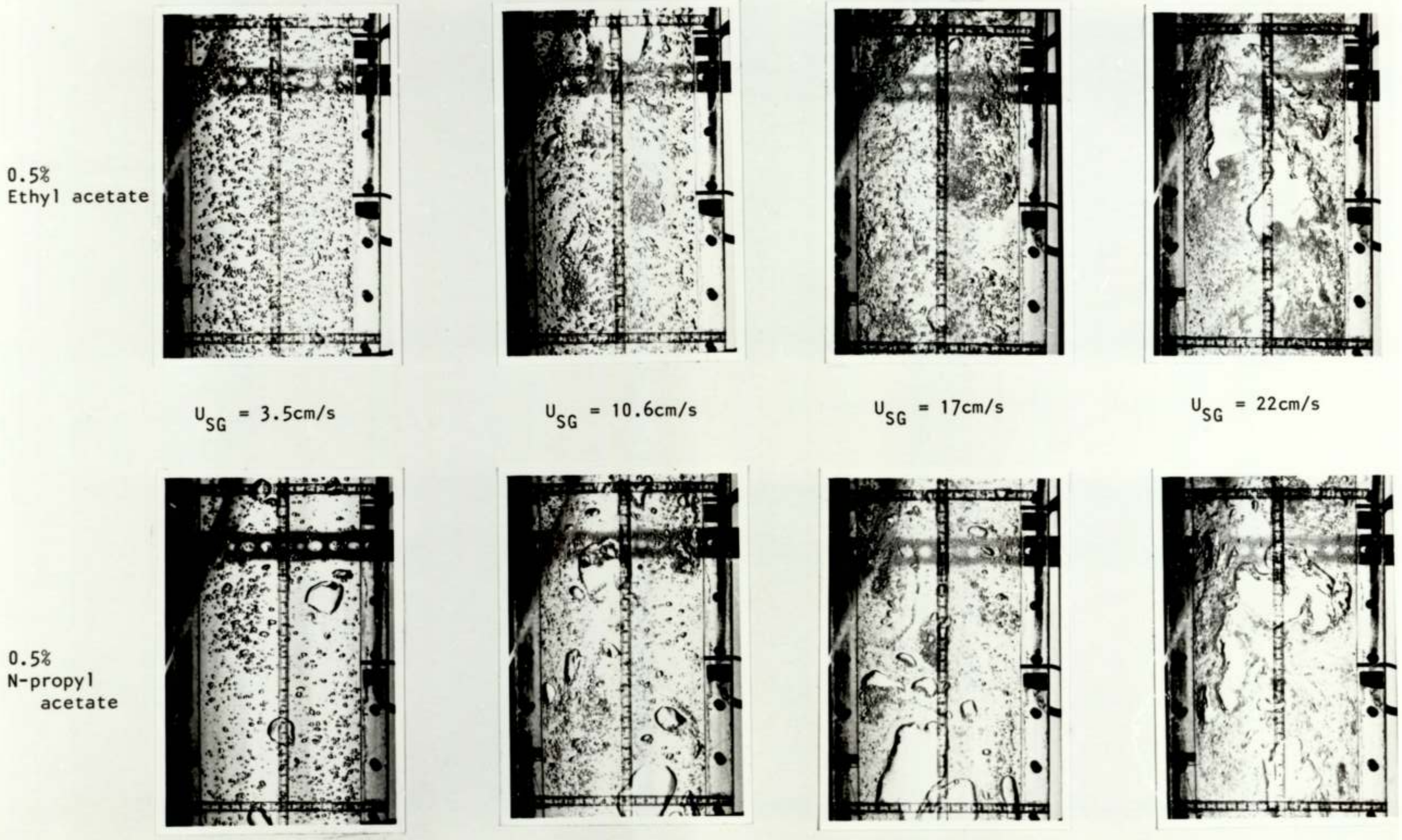


Figure 5.15 Effects of ethyl acetate and N-propl acetate of flow regimes and bubble dynamics (two-phase flow)



Since $U_S = \dfrac{USG}{\varepsilon_G} - \dfrac{USL}{(1 - \varepsilon_G)}$,

$$j_{GL} = U_{SG} (1 - \varepsilon_G) - U_{SL} \varepsilon_G \qquad \text{........ 5.7}$$

or

$$j_{GL} = U_{SG} - \varepsilon_G (U_{SG} + U_{SL}) \qquad \text{........ 5.8}$$

In principle, it should be possible to identify different flow regimes from plots of drift flux as a function of gas hold-up.

As shown in Figures 5.16 to 5.20, the clear transition in the slope corresponds to the change in flow regime. There is transition from bubbly flow to other flows. Figures 5.17 to 5.20 depict the effect of liquid additives on flow regime transition. Bubbly flow is maintained for a longer period while, in some cases, most of the data lie in the churn-turbulent flow regime. As the chain length of the organics increases, the transition occurs at higher gas velocities except for esters.

Liquid velocity has a rather small effect on flow regimes. It is noted that at low liquid velocities (< 1 cm/s) and for most of the systems with aqueous solutions, there is a similarity between the drift-flux vs gas hold-up plot and the gas hold-up vs superficial gas velocity plot. Thus, it is possible to predict flow regime transition from the gas hold-up vs superficial gas velocity plots, whereby a change in slope corresponds to a change in flow regime. The changes are more pronounced and apparent with drift-flux plots.

Figure 5·16 Effect of superficial gas and liquid velocities on flow regime transition

Two-phase system (Air-Water)

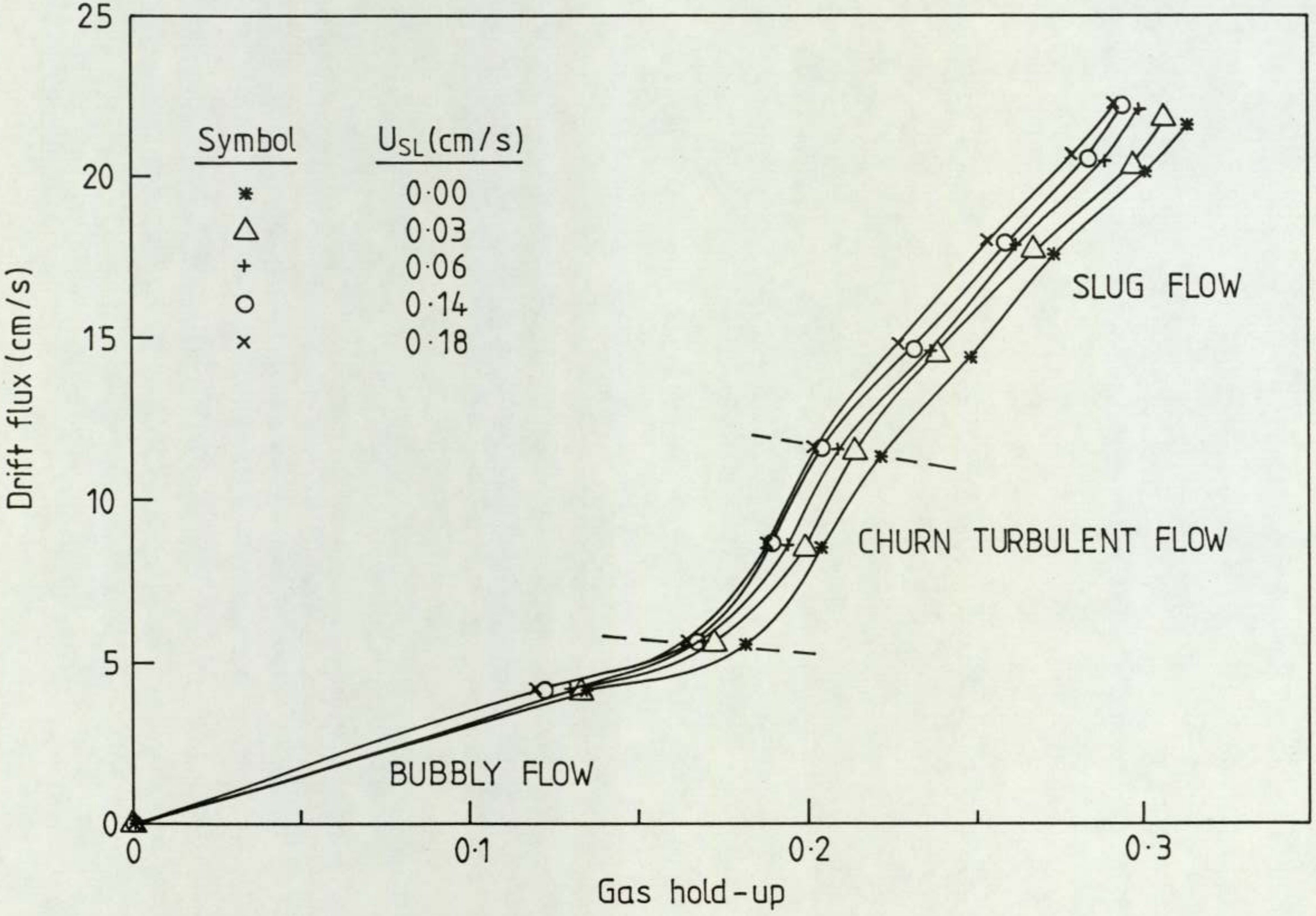


Figure 5.17 Effect of type of alcohol on flow regime transition.

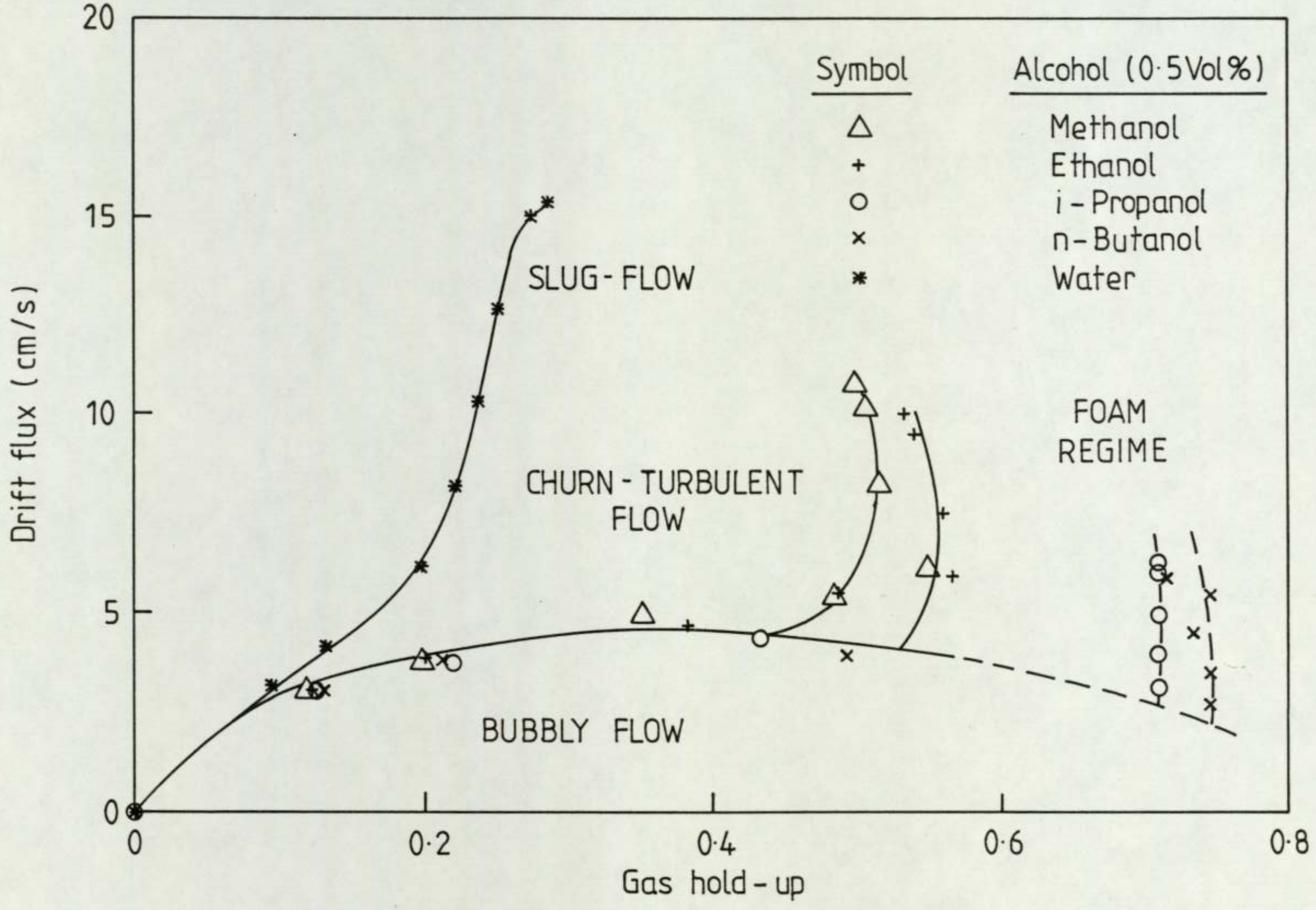


Figure 5.18 Effect of type of ketone on flow regime transition.

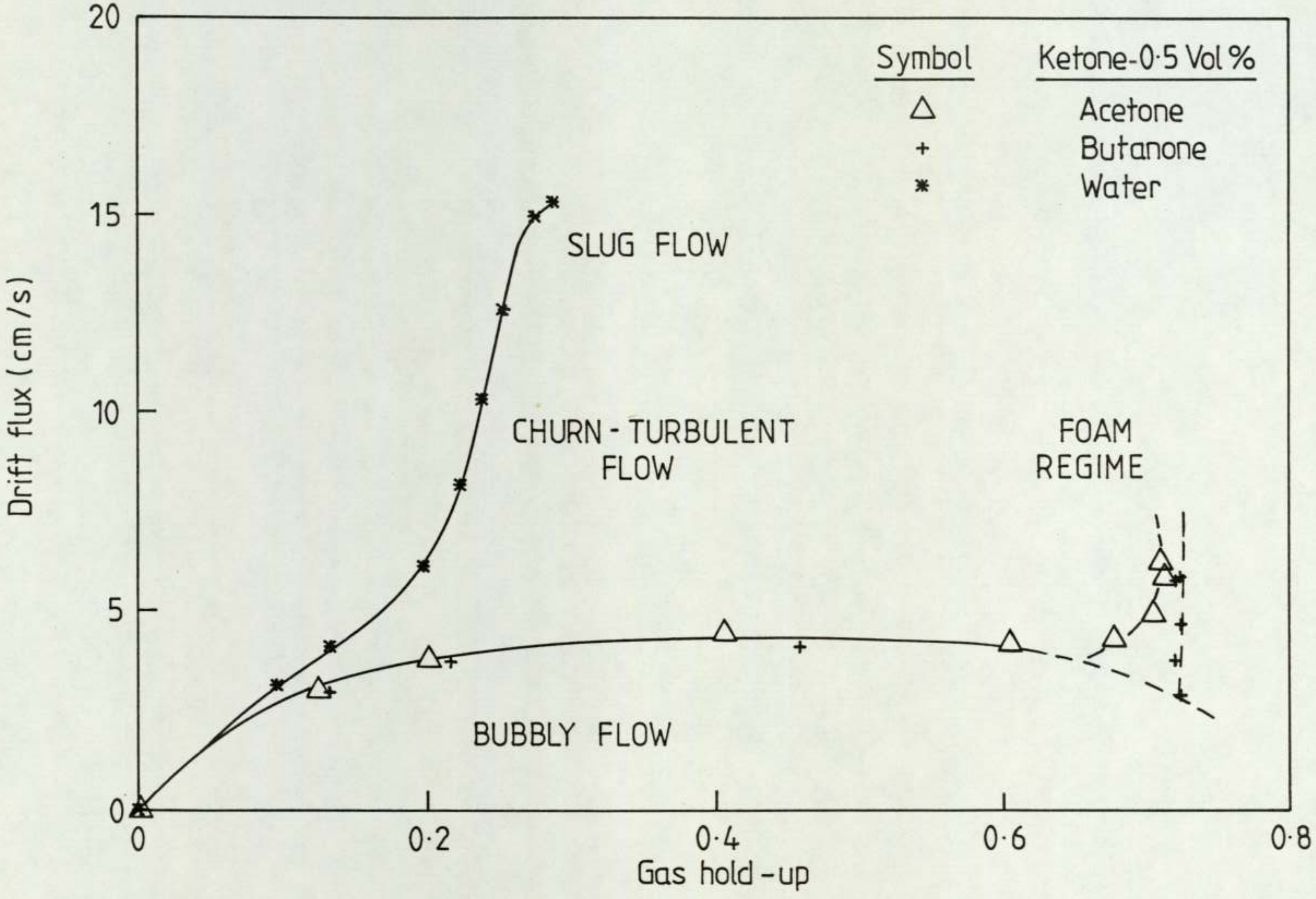


Figure 5.19 Effect of type of ester on flow regime transition

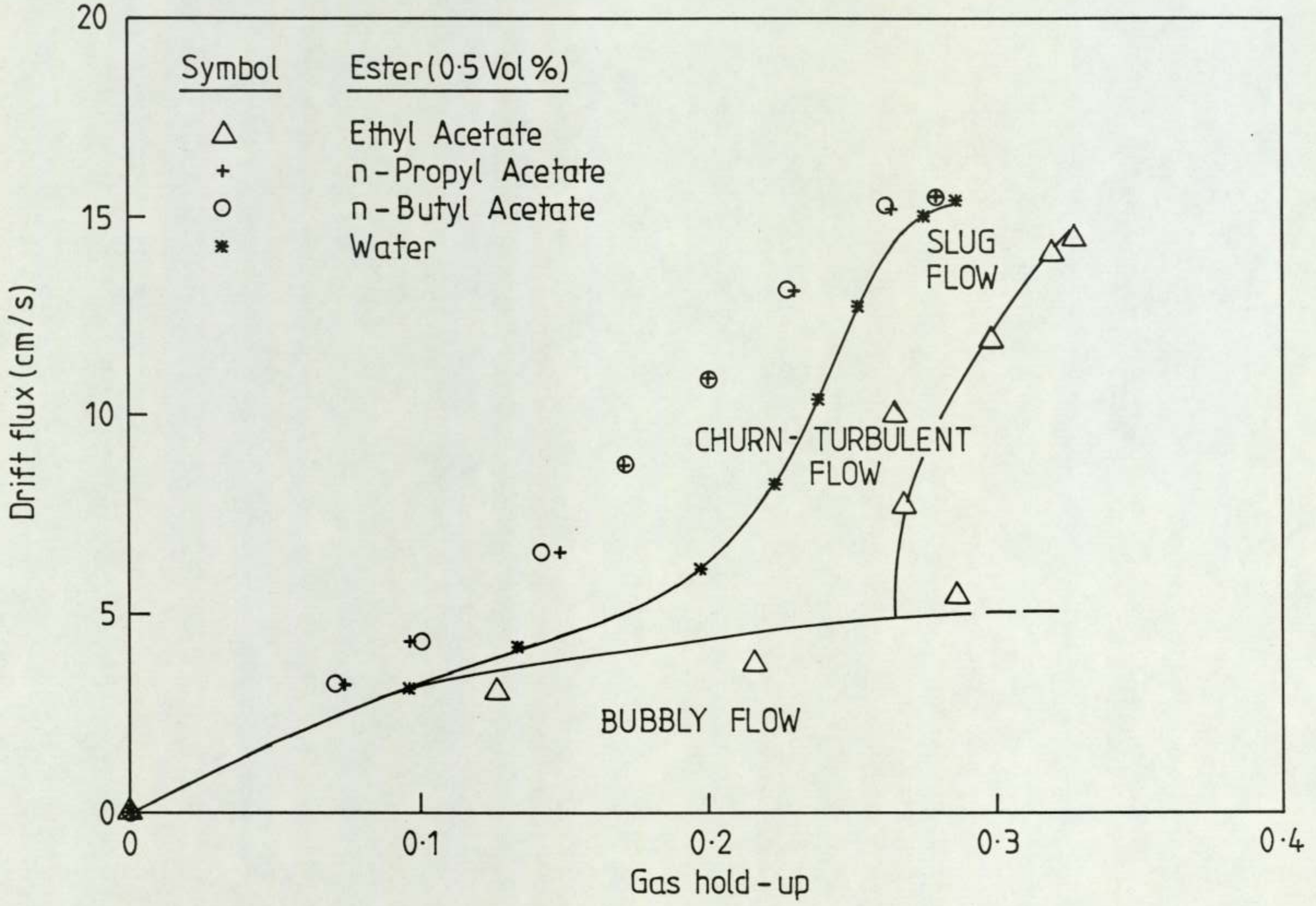
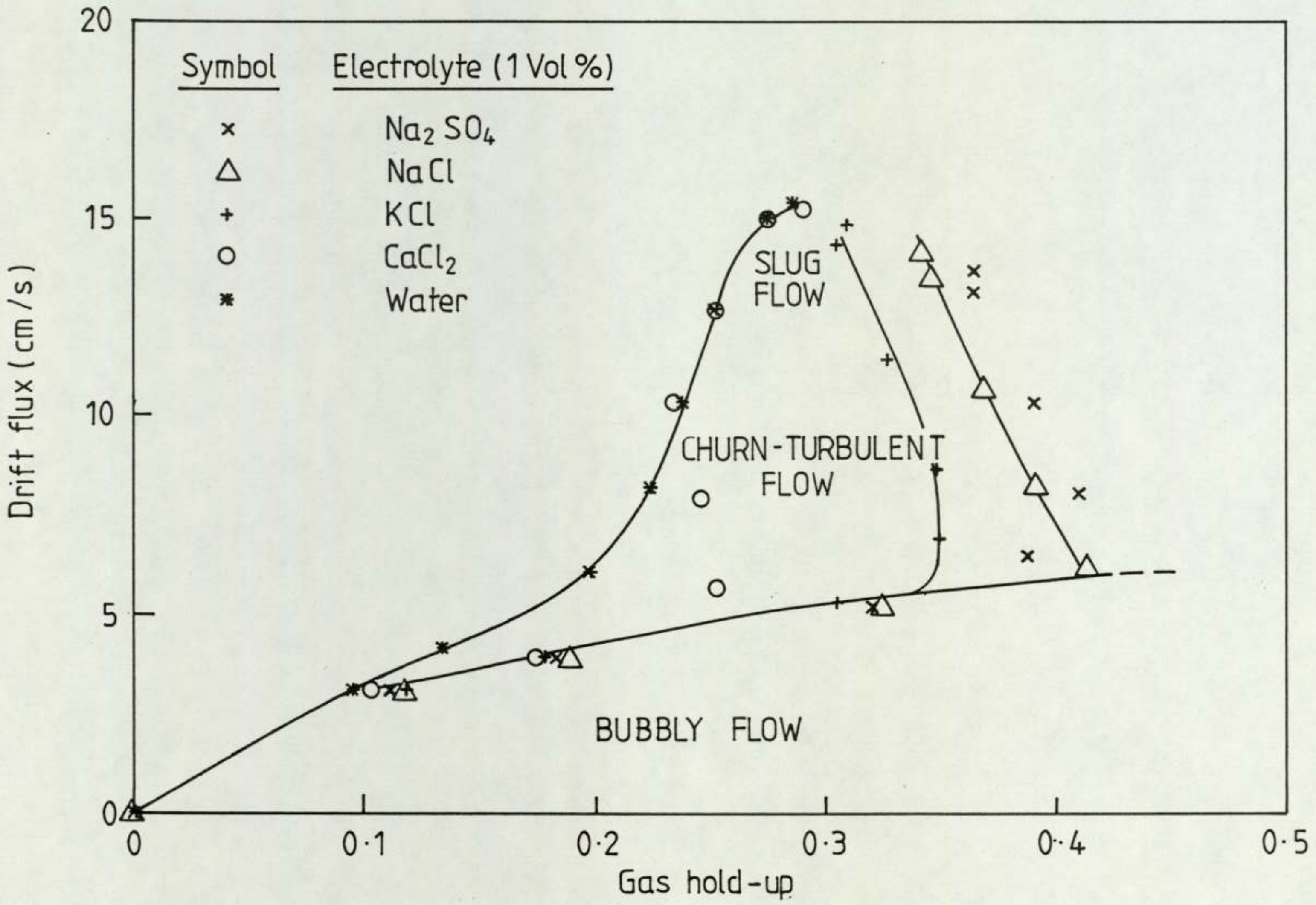


Figure 5.20 Effect of type of electrolyte on flow regime transition.



5.6 **Discussion**

5.6.1 Flow Regimes

The visual and photographic observations, and qualitative analysis made by the author have revealed five flow regimes.

Bubbly Flow Regime

In this regime, a swarm of bubbles rises uniformly within the column at low superficial gas velocities. The bubbles are relatively uniform in size and shape (see Figure 5.3): they are roughly elliptical in section with a major axis of about 5mm. Bubbly flow was observed at superficial gas velocities less than 5 cm/s in the 2-D column and about 3 cm/s in the 3-D column. The addition of organic compounds enhances bubbly flow.

Churn-Turbulent Regime

On increasing the gas velocity, bubbles coalesce to form larger bubbles. There is now a mixture of relatively small and large bubbles within the column. The regime is observable at velocities up to about 13 cm/s in the 2-D column and 10 cm/s in the 3-D column.

Slug-Flow Regime

With further increases of gas flow, bubbles get much bigger. The large, bullet-shaped bubbles characteristic of slug flow are then encountered. Here the gas elements are large enough to almost span the column. This was observed up to a gas velocity of about 17 cm/s in the 2-D column.

"Turbulent" Flow Regime

This regime occurs at high gas velocities and low initial liquid height (see Figure 5.4). The flow is much more irregular and disturbed as a result of the rapid breakdown of the gas bubbles. Some sources designate this flow as "wispy annular flow", because of oscillations between the slug-flow condition and annular-like flow, in which the central region of the flow is almost liquid free. Much liquid flows on the wall of the column; the gas is now concentrated at the centre and moves in a wavy form. For safety reasons, it was only possible to study this regime in the 2-D column.

Foam Regime

The presence of the higher alcohols and ketones in solution result in foaming especially at high gas velocities. The mechanism of foam formation and effect on column behaviour have been discussed in Chapter 4.

5.6.2 Bubble Dynamics

Generally, at low superficial gas velocities (approximately less than 5 cm/s), bubbles are small and uniform in size. Bubble coalescence rates are usually low and so bubbles can be uniformly distributed. The size and uniformity of bubbles depends on the design of the gas distributor and the column diameter. However, if the gas is uniformly distributed at the column inlet, HOMOGENEOUS flow is obtained.

At high superficial gas velocities (greater than 5 cm/s), the bubble coalescence rate increases significantly. The pseudohomogeneous bubble-in-liquid dispersion can no longer be maintained, and the flow thus becomes HETEROGENEOUS. The column contains a mixture of large and small bubbles. The fraction of gas occupied by large bubbles increases with the gas velocity. The size of the large bubbles is dependent on the nature of the gas distributor, column diameter and physical properties of the liquid. The large bubbles are non-spherical and though small in number, carry the bulk of the gas through the column.

Bubble-size distribution is quantified using the dynamic gas disengagement method. The large bubbles with high rise velocities disengage first before the small ones; changes from one flow regime to another can be seen. As shown in Figures 5.7 to 5.9, three bubble classes are present:

- the line with steepest slope corresponds to the disengagement of large bubbles - slug flow,
- the line with steeper slope corresponds to the disengagement of relatively large and small bubbles - churn-turbulent flow,
- while the line with least slope corresponds to the disengagement of small bubbles - bubbly flow.

The rise velocity of bubbles is very much dependent on the superficial gas velocity and nature of liquid phase.

5.6.3 Effect of Liquid Additives

It is clear that the addition of organic and inorganic compounds to the liquid phase alters the flow regime and affects the bubble dynamics considerably. Change from one regime to another was neither immediate nor very apparent. Bubble sizes were stabilized; bubble rise velocity and coalescence were reduced (see Figure 5.5). The strong effect of liquid additives is also shown by the drift-flux model (see Figures 5.17 to 5.20).

By means of the corpuscular analysis, the behaviour of such systems has been explained in Chapter 4.

5.7 Conclusions

The effects of operating parameters and the nature of the liquid phase on flow regime and bubble dynamics have been studied. The following conclusions can be drawn:

(1) Five regimes of flow can be identified, namely - bubbly flow, churn-turbulent flow, slug flow, turbulent flow and foam regimes.

(2) The motion of bubbles arises partly from buoyancy and partly from the motion caused by the entry of the two phases into the column.

(3) From the drift flux plot it can be seen that the appearance of large bubbles can be postponed or eliminated with the help of an additive in solution. Changes from one regime to another are clearly shown.

(4) The flow regimes and bubble dynamics in the column are strongly influenced by the gas velocity, initial liquid height, liquid properties and column design.

(5) The corpuscular theory can be used to interpret the effect of liquid properties.

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CHAPTER 6

GAS HOLD-UP IN THREE-PHASE FLUIDISATION

GAS HOLD-UP IN THREE-PHASE FLUIDISATION

GAS HOLD-UP IN THREE-PHASE FLUIDISATION

6.1 Introduction

Three-phase fluidisation is defined as an operation in which a bed of solid particles is suspended in gas and liquid media due to the net drag force of the gas and/or liquid flowing opposite to the net gravitational force or buoyancy force on the particles. Such an operation generates intimate contact between the gas, liquid and solid particles in these systems. This provides substantial advantages for applications in physical, chemical, biochemical and petrochemical processing involving gas, liquid and solid phases (see Figure 1.2). The gas-liquid-solid fluidised bed can be operated with a cocurrent or countercurrent flow of a gas and a liquid with liquid as the continuous phase. The various modes of operation are outlined in Chapter 2.

The successful design and operation of a gas-liquid-solid fluidised bed system depend on accurate prediction of the fundamental properties of the system, namely the hydrodynamics, mixing and transport properties. A comprehensive review of the literature has highlighted the need for further research on the effects of operating parameters and phases on these properties. In particular, reliable data on phase hold-up are few for systems involving aqueous organic solutions. Experimental, theoretical and modelling studies have therefore been carried out on three-phase systems, using a systematic approach to meet these challenges.

6.2 Review of the Literature

Three-phase fluidisation became a subject for fundamental research about two decades ago. Since then, considerable progress has been made on the understanding of the phenomena. Early reviews of the subject were made by Ostergaard (1969, 1971, 1977), Davidson et al (1977), Shah (1979) and Epstein (1981, 1983); more recent reviews have been presented by Wild et al (1984) and Muroyama et al (1985). Having established the three main factors that affect gas hold-up (see Chapter 4), available information in the literature is reviewed following this format.

6.2.1 Effects of Operating Parameters, Column Design and Nature of Phases on Gas Hold-up

The earlier studies of three phase systems made by Ostergaard (1964, 1965), Stewart and Davidson (1964), Adlington and Thompson (1965), Turner (1966) and Ostergaard and Theisen (1966) showed that the total hold-up of gas and liquid increased with increasing gas velocity.

Kato (1963) studied the effects of gas velocity, particle size and density, and the amount of solids and liquid in the bed on gas hold-up. The gas hold-up decreased with increasing particle size and amount of solids in the bed, and with a decrease in nominal gas velocity.

Adlington and Thompson (1965) reported results from experiments with 3in diameter beds of alumina particles of sizes from 0.3 to 2.8mm diameter fluidized by white spirit, and 10in diameter beds of sand particles of 0.3mm diameter fluidised by water. They found that the presence of solids had little influence on gas hold-up below nominal gas velocities of about 1.5 cm/s. At higher gas velocities the presence of solids caused a decrease of gas hold-up but was relatively independent of solids particle size.

Viswanathan et al (1965) measured gas hold-up in 2in diameter beds of quartz particles of 0.649 and 0.928mm diameter and glass beads of 4mm diameter, the liquid medium being water. The presence of quartz particles caused a decrease in gas hold-up, compared to solids-free gas-liquid system whereas the presence of glass beads caused an increase.

Sherrard (1966) obtained the gas hold-up by measuring x-ray transmission above the bed surface. In beds of large or heavy particles (beds of 1.6mm glass ballotini, 6mm acrylic spheres and 12-14 mesh lead shot) the gas hold-up decreased with increasing superficial liquid velocity; whereas in beds of small or light particles (beds of 12-14 mesh glass ballotini, 36-44 mesh glass ballotini), gas hold-up remained essentially independent of liquid velocity.

Schugerl (1967) reported that the presence of a small amount of 0.25mm particles in gas-liquid systems reduced the gas hold-up.

Ostergaard (1968) found that for very small sand particles (40 through 60 and 60 through 80 mesh), the gas (nitrogen) hold-up was independent of particle size and liquid (water) velocity, but increased linearly with nominal gas velocity. The gas hold-up in gas-liquid (solids-free) system was higher than that in a gas-liquid-solid fluidised bed.

Imafuku et al (1968) measured gas hold-up in a batch (ie. no liquid flow) three-phase fluidised bed column. The presence of solids caused significant coalescence of bubbles. Gas hold-up was correlated with the slip velocity between the gas and liquid. They also found that gas hold-up neither depends on the type of gas distributor nor the shape of the bottom of the column when solid particles are completely suspended.

Ostergaard and Michelsen (1968) measured gas hold-up in beds of 0.25, 1 and 6mm glass particles using a radioactive tracer technique. They found that

$$\varepsilon_G \propto U_{SG}^n \qquad \text{........ 6.1}$$

where U_{SG} is the superficial gas velocity and n took values of 0.88, 0.78 and 0.93, respectively for the three particle sizes. The solid-free bubble-column gave n = 1.05. They also found that, in the solid-free system and in beds of 6mm particles, gas hold-up decreased with increasing liquid flow rate; whereas in beds of 0.25 and 1mm particles, gas hold-up increased with increasing liquid flow rate.

Michelsen and Ostergaard (1970) extended the study of Ostergaard and Michelsen to a wider range of flow rates and other system conditions. Their data were obtained in a 15.2cm id and 11m tall concurrent upflow air-water-solid system. The results show that, while the liquid hold-up decreases with increasing gas velocity and decreasing liquid velocity for all particle sizes, both gas and solid hold-ups show different behaviour with different particle sizes. The gas hold-up data were explained on the basis of bubble coalescence in beds of small particles and the breakage of large bubbles in beds of large particles. The unusual effect of gas velocity on the solids hold-up with different particle sizes was explained on the basis of possible expansion and contraction of the beds under different conditions.

Kato et al (1972) found that the gas hold-up in an air-water-glass sphere system was less than that of the air-water system and that larger solid particles showed a smaller gas hold-up. Based on visual observations, they suggested that this is caused by the larger rising velocity of coalesced bubbles in the presence of solid particles.

Kim et al (1972) studied the effects of air and water velocity, and particle size ranging from 2.6 to 6.0mm on the gas and liquid hold-ups in a two-dimensional column. Based on their data, they reported dimensionless correlations for the gas and liquid hold-ups.

Kim et al (1975) extended their study to cover a wide range of liquid viscosities (1 to 70 cP) and surface tensions (40 to 73 dynes/cm). Solutions of sugar, aqueous carboxylmethyl cellulose of various concentrations, and acetone-water mixtures were investigated with particle sizes ranging from 1 to 6mm. The effect of liquid properties on the liquid hold-up and total gas plus liquid hold-up were correlated in terms of the dimensionless Weber number, We. The physical properties of the liquid affected the bubble characteristics and phase holdup: liquid hold-up increased with liquid viscosity for all particle sizes, whereas the dependence on surface tension of the liquid depended on both the size and nature of the solid particles.

Blum and Toman (1977) have reported extremely high gas hold-ups for a system comprising nitrogen/light mineral oil/coarse catalyst particles. The gas hold-up was observed to increase linearly up to 50% as the gas velocity increased to 14 cm/s in the system, foam formation was associated with the high hold-ups.

Kato et al (1972), Ying et al (1980) and Vasalos et al (1980) investigated the effect of solids concentration and concluded that an increase in solids concentration generally decreases the gas hold-up, although this effect becomes insignificant at high gas velocities (> 10 cm/s).

Axial solids concentration has been studied by Cova (1966), Imafuku et al (1968), Narayan et al (1969), Yamanaka et al (1970), and Govindarao (1975). Cova reported that high gas and liquid velocities and high viscosities tended to give more uniform solids distribution. Imafuku et al observed that the critical gas velocity for complete suspension mostly depends on the liquid flow near the gas distributor. Govindarao (1975) pointed out that particle diameter and bed volume have a strong influence on the axial distribution of solids.

Armstrong et al (1976) showed that the wettability of solids could significantly affect the gas and liquid hold-ups and the fluidized bed height. They made measurements in a large two-dimensional bed of 66cm x 2.54cm cross-section and 213cm height. 6mm glass beads and Teflon-coated glass beads were used as the wettable and non-wettable solids respectively. For all gas velocities, the gas hold-up was remarkably smaller with the Teflon-coated beads than with uncoated beads. Armstrong et al indicated that by rendering the solids nonwettable, gas -solid contact is improved and the liquid wake is inhibited.

6.2.2 Recent Studies

Three-phase fluidisation continues to receive attention owing to its increasing importance in industrial applications (see Figure 1.2). The results of recent studies on phase hold-up are summarised in Table 2.4.

Kara et al (1982) determined the effect of particle size, solids concentration, slurry velocity and gas velocity on gas, liquid and solid hold-up and flow regimes. With relatively small solid particles, significant effects of liquid and slurry velocities on gas hold-up were observed. Gas hold-up decreased with increasing slurry velocity, solids concentration and particle size. Based on their data, an empirical correlation for gas hold-up was presented.

Godbole et al (1983) studied the effects of gas velocity and the presence of solids on gas hold-up structure in viscous as well as low viscosity solutions. Gas hold-up was measured using the hydrostatic head technique and the dynamic gas disengagement method. Polystyrene, coal, sand and oil shale particles were employed as the solid phases. In viscous solutions, the addition of solids increased the fractional gas hold-up due to the small bubbles generated. High hold-ups in oil shale slurries were attributed to the effect of dissolution minerals. An attempt was made to explain the effect of solids on the basis of wettability characteristics.

Qualitative studies have been reported by Govindarao et al (1983) and Gidaspow et al (1983) on column operations. Hydrodynamic models were presented based on such studies.

Kelkar et al (1984) examined the effects of gas and slurry velocities, solid physical properties, particle sizes, solids concentration, interfacial tension and slurry viscosity on phase hold-ups. They concluded that phase hold-ups were not significantly affected by particle size, solids concentration and slurry velocities. However, solids wettability was found to enhance the coalescence tendencies in the liquid phase. Addition of surfactants increased gas hold-up; an increase in liquid viscosity reduced gas hold-up.

Koide et al (1984) studied the effects of column dimensions, gas velocity, properties of liquid and solids on gas hold-up. The presence of suspended solid particles reduced gas hold-up, the reduction being high in the transition regime and low in the heterogeneous regime. Based on these observations, they proposed empirical correlations for gas hold-up in transition and heterogeneous flows.

Sangnimnuan et al (1984) investigated gas hold-up under coal hydroliquefaction conditions in a continuous tubular bubble column reactor. Gas hold-up increased with increasing superficial gas velocity and temperature. The effects of gas molecular weight, pressure and superficial liquid velocity were negligible. The modified correlation of Hikita et al (1980) was found to satisfactorily predict gas hold-up under high temperature and pressure after correction for gas solubility and liquid vaporisation.

Capuder and Koloini (1984) measured gas hold-up during the carbonisation of lime suspensions with CO_2-air mixture. The influence of solids concentration was studied over a wide range of superficial gas velocities and power inputs. Gas hold-up decreased at all operating conditions with increasing solids concentration.

Smith and Ruether (1985) have presented new data and a theory for describing dispersed solids in slurry bubble columns. Axial solids concentration distributions were measured in a 10.8cm id slurry column operated at steady state conditions. The liquid-phase was either water or ethanol and the solid-phase consisted of narrow-sized fractions of glass spheres. With a one-dimensional sedimentation-dispersion model, the data were used to develop a method for predicting average solids loading and the axial distribution of solids. The new method predicts axial distribution of solids with improved accuracy over existing methods.

Kato et al (1985) studied the behaviour of gas and solid particles in a three-phase fluidised bed of 12cm diameter. The mean gas hold-ups in the dense and lean regions and the mean solid hold-up in the dense region were correlated by empirical equations. Glass spheres and porous alumina particles were used as the solid phases.

6.3 Experimental Programme and Procedure

Published quantitative data on hold-up in three-phase fluidisation is relatively scarce and covers too limited a range of process parameters for many design calculations. One of the objectives of this study is thus to provide data for a wide range of important process parameters. The literature also reveals contradictory conclusions. To overcome these anomalies, a systematic method of approach was developed and used in the study.

Experiments were conducted in the two- and three-dimensional bubble columns which have been described in Chapter 3. The columns were operated in a batch (semi-continuous) as well as continuous fashion. The fluid phases were air and water. A variety of solid particles and different liquid additives were used to simulate chemical, biochemical and petrochemical processes, eg. catalytic reactions, coal liquefaction, hydroprocessing and fermentations in particular the behaviour of microbial aggregates, immobilised enzymes and cells.

An outline of the experimental programme showing the factors studied is as follows:

(i) Operating Parameters

(a) Effect of superficial gas velocity, U_{SG}.

(b) Effect of superficial liquid velocity, U_{SL}.

(ii) Solid Phase

(a) Effect of solids concentration.

(b) Effect of solids size.

(c) Effect of solids density.

(d) Effect of solids wettability.

(iii) Liquid Phase

(a) Effect of alcohols on the three-phase system.

(b) Effect of ketones on the three-phase system.

(c) Effect of esters on the three-phase system.

(d) Effect of electrolytes on the three-phase system.

More discussion on the choice of factors and systems is given in Chapter 4.

Experimental Procedure

The physical properties of the solids were measured; these are given in Table 6.1. A 5% concentration (by volume) of the solid particles was introduced into the column; gas and liquid were then fed in at the required flow rates. After a steady concentration of the solids was established, gas hold-up was measured using the bed-expansion method. These figures were checked using the gas-disengagement method along with the photographic technique (see Chapter 5).

TABLE 6.1: PHYSICAL PROPERTIES OF SOLID PARTICLES

Reference for Solid	Chemical Composition	Particle Diameter $d_p \times 10^3$ (m)	Particle Density ρ_s (kg m^{-3})	Packed Bed Voidage ε_p
Styrocell R351	Polystyrene	1.25	1230	0.5
Styrocell R551X	Polystyrene	0.85	1030	0.45
Styrocell R551	Polystyrene	0.81	1150	0.47
Styrocell S	Polystyrene	0.60	1110	0.43
Ballotini	Glass	1.2	2900	0.5
Diakon I	Perspex	0.710-0.850	⩾1000	0.5
Diakon II	Perspex	0.600-0.710	⩾1000	0.5
Diakon III	Perspex	0.425-0.600	⩾1000	0.4
'Amberlite' resin IRA-400CI	Ion-exchange resin	0.55	⩾1000	0.4

After the experiments with different types of solid, additives were introduced into the liquid phase and the whole procedure repeated. The physical properties of each system are listed in Table 6.2.

6.4 Experimental Results

The results from experiments have all been tabulated and are presented in Appendix D. As in previous sections of the work, the graphical output was prepared using a computer graph plotter to achieve accuracy, neatness and clear presentation.

6.4.1 Effects of Operating Parameters

Effect of Superficial Gas Velocity, U_{SG}

The results of gas hold-up measurements as a function of superficial gas velocity, with superficial liquid velocity as a parameter, are shown in Figures 6.1 - 6.3. Styrocell "R551" and Amberlite "IRA-400CI" resin were the solid phases. The data used to plot these graphs are given in Tables D1, D2 and D3 of Appendix D. Results for other types of solids are given in subsequent sub-sections.

Gas hold-up increased with an increase in gas velocity in all cases. The relationship is almost linear in the bubbly flow regime.

Effect of Superficial Liquid Velocity, U_{SL}

Figures 6.1 - 6.3 also show the effect of liquid velocity on gas hold-up in three-phase fluidisation. Simultaneous flow of liquid with gas into the system caused a decrease in

TABLE 6.2: PHYSICAL PROPERTIES OF THREE-PHASE SYSTEMS WITH ADDITIVES IN THE LIQUID PHASE

Data were taken at 22°C. Solutions were made up in Birmingham mains water. Samples removed during tests with 5% Styrocell.

System (Vol %)	Density (kg m^{-3})	Surface Tension (N m^{-1})	Viscosity (Ns m^{-2} x 10^3)
Alcohols			
0.5% Methanol	1001	0.058	1.28
0.5% Ethanol	1001	0.048	1.35
Ketones			
0.5% Propanone	1001	0.052	1.41
0.5% Butanone	1001	0.051	1.48
Esters			
0.5% Ethyl Acetate	1000	0.053	1.22
0.5% n-propyl Acetate	1001	0.052	1.19
Electrolytes			
1.0% NaCl	1008	0.053	1.22
1.0% $CaCl_2$	1006	0.054	1.28

Figure 6.1 Effect of superficial gas and liquid velocities on gas hold-up in the two dimensional bubble column.

Three-phase system (Air-Water-Styrocell R551)

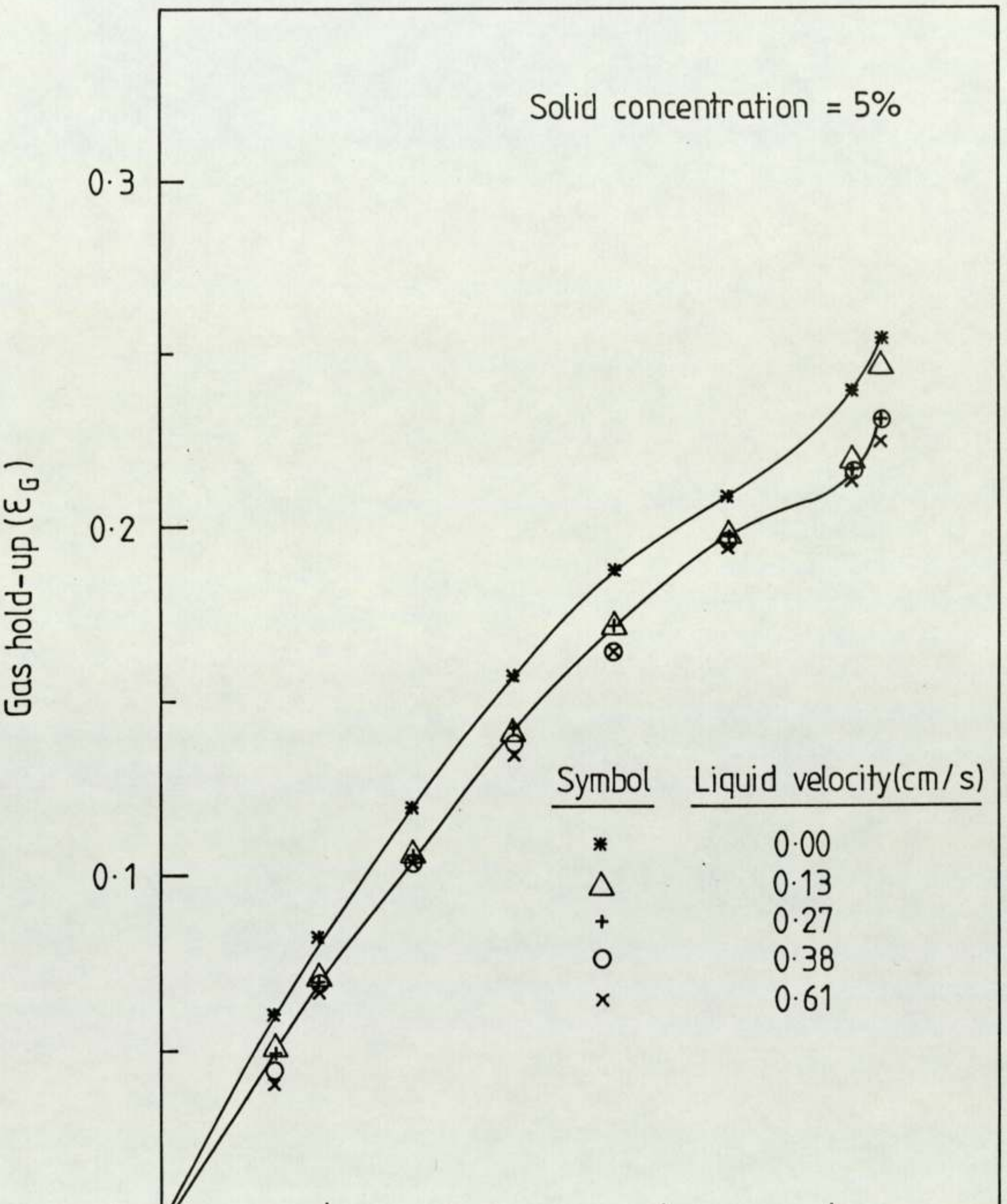


Figure 6.2 Effect of superficial gas and liquid velocities on gas hold-up in the three dimensional bubble column.

Three-phase system (Air - Water - Styrocell R551)

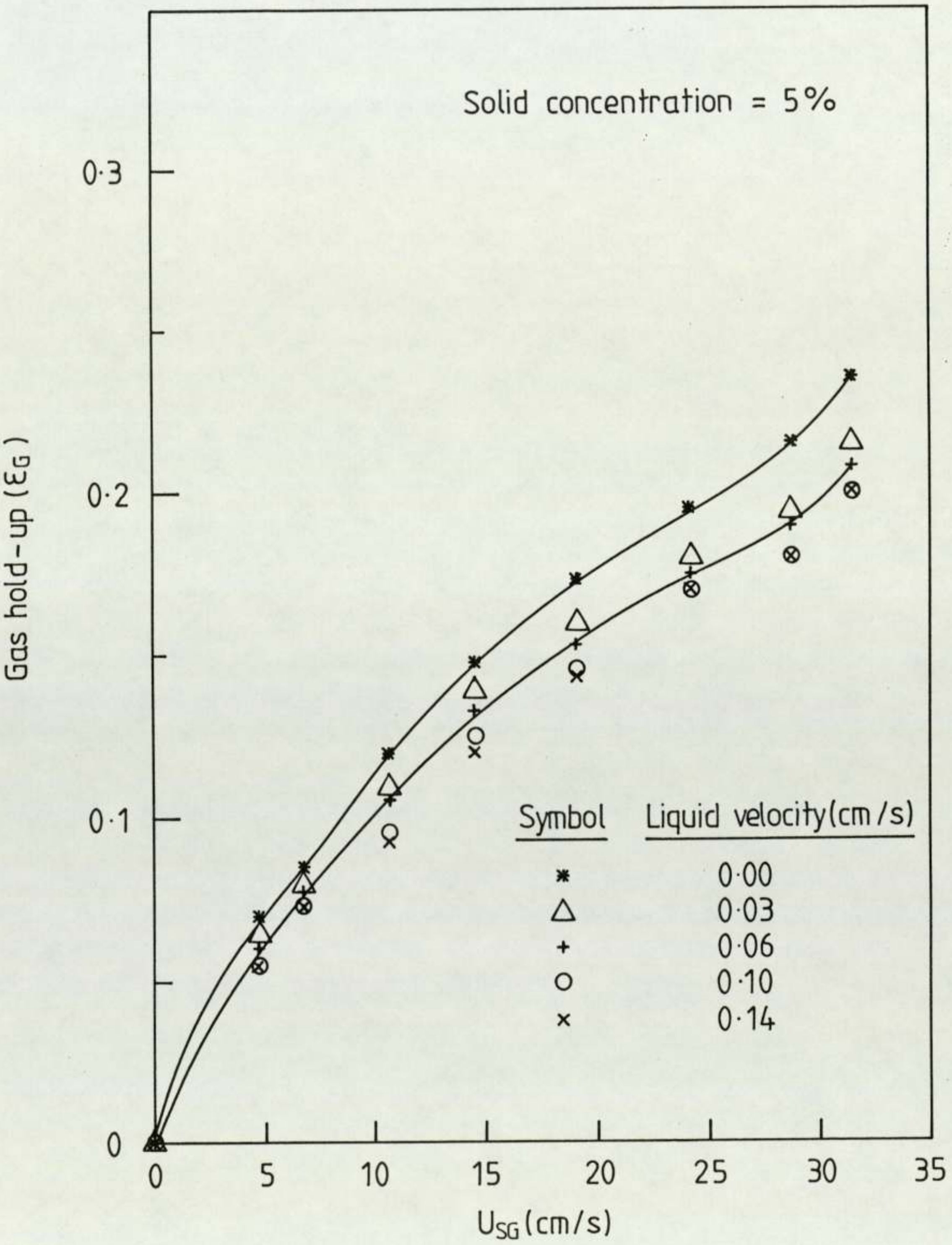
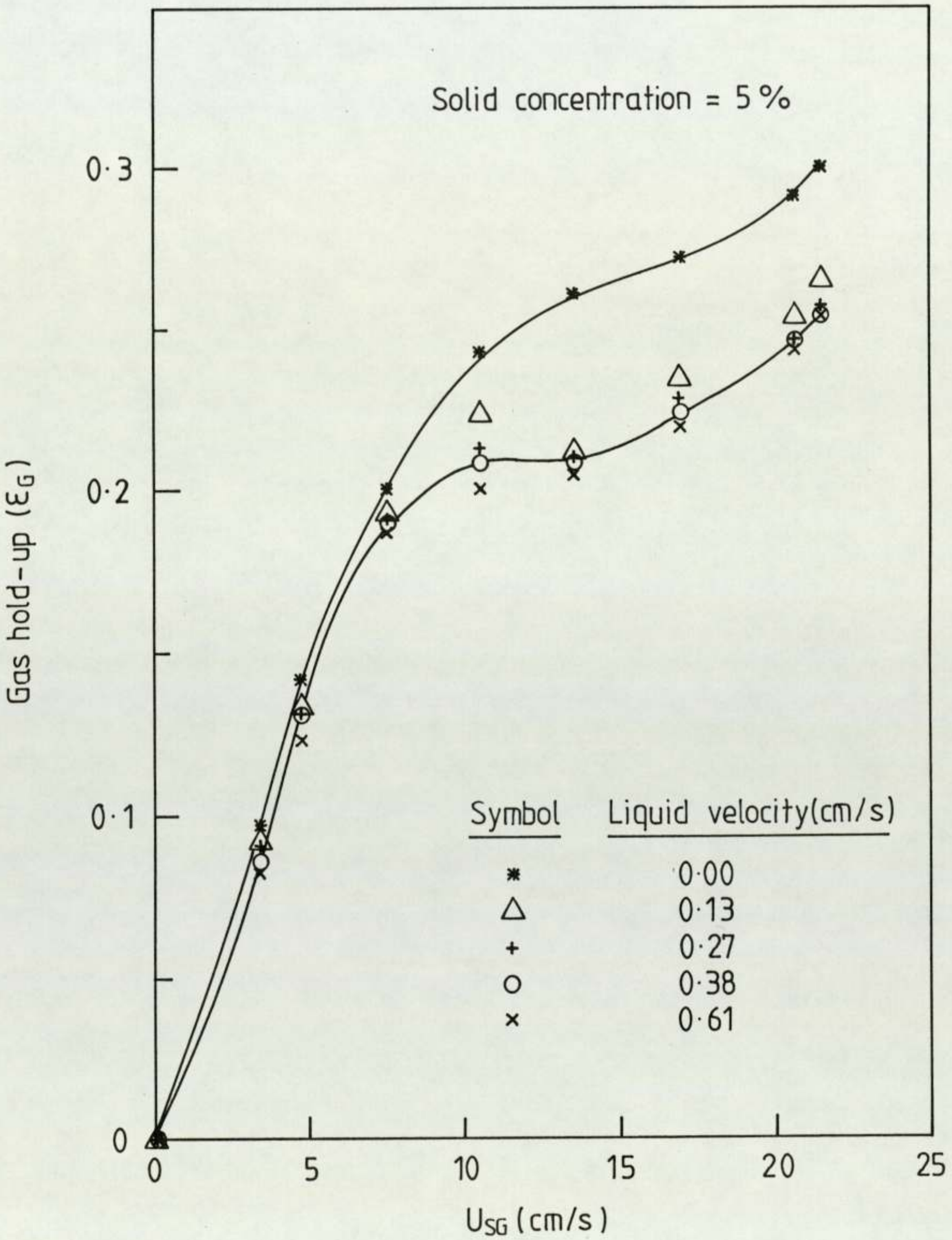


Figure 6.3 Effect of superficial gas and liquid velocities on gas hold-up in the two dimensional bubble column.

Three-phase system (Air-Water-Amberlite)



gas hold-up. Increase in the liquid flow rate gave a further, but very slight decrease in hold-up. With Amberlite resin, the decrease was more pronounced outside the bubbly flow regime of operation.

6.4.2 Effects of Solid-Phase Properties

Effect of Solids Concentration

The effect of solids concentration was studied by measuring the gas hold-up in systems with different amounts of solid (0-15% volume concentration). Figures 6.4 - 6.8 show the results for Styrocell "R351" and "R551", Ballotini and Amberlite resin particles, as the solid phases; the detailed data are given in Appendix D. A typical result with Diakon is compared in Figure 6.11.

Gas hold-up decreased with increasing solids concentration. The strong dependence at low solids concentration diminished as the concentration was increased.

Effect of Solids Size

Using the four different types of Styrocell particles the effect of solids size on gas hold-up was examined. This is shown in Figure 6.9 with the detailed data listed in Table D9 (Appendix D).

Within the size range considered, an increase in solids size decreased the gas hold-up. The effect of size was reduced for larger solid particles.

Figure 6.4 Effect of solid concentration on gas hold-up in the two dimensional bubble column.

Three-phase system (Air - Water - Styrocell R551)

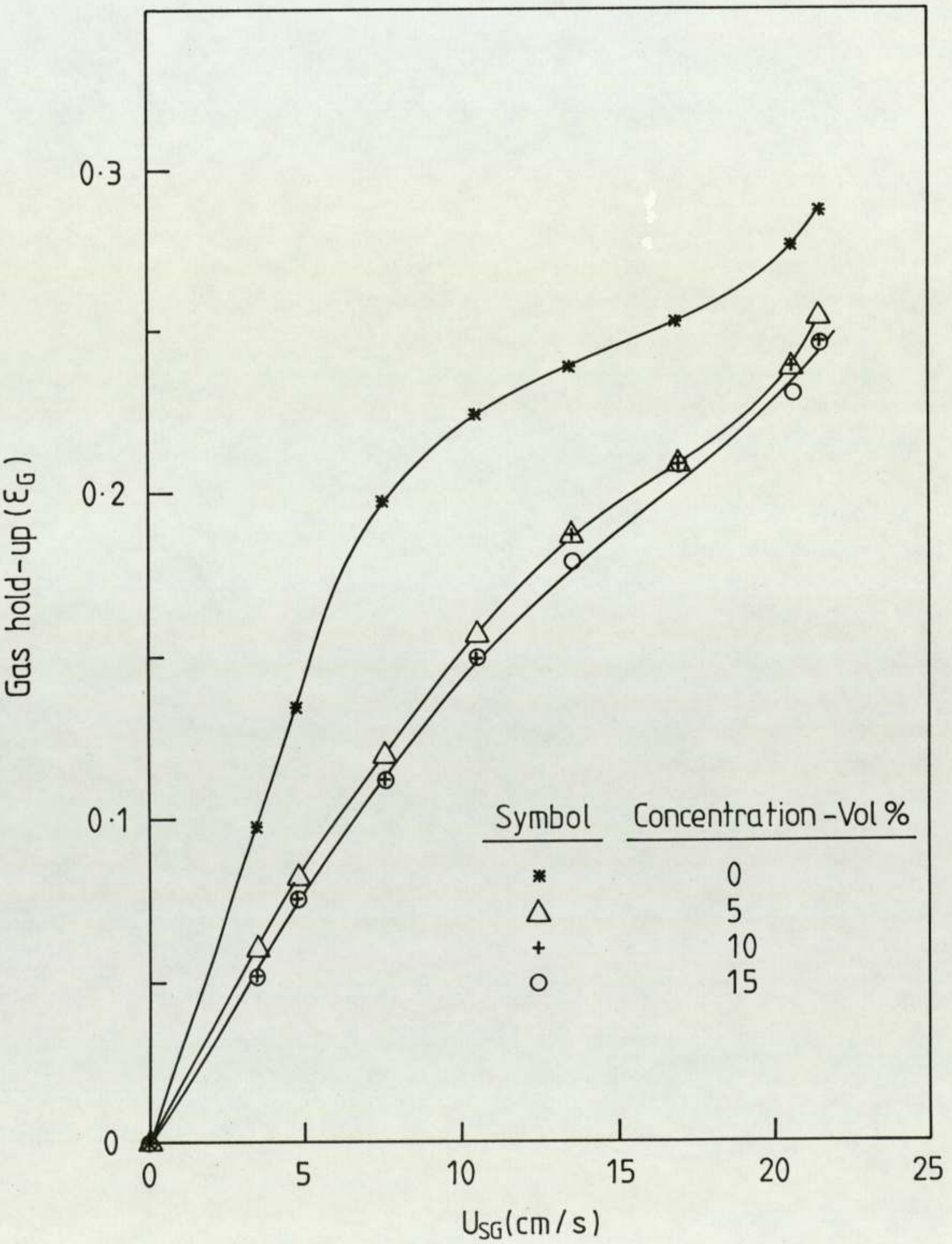


Figure 6.5 Effect of solid concentration on gas hold-up in the three dimensional bubble column.

Three-phase system (Air - Water - Styrocell R551)

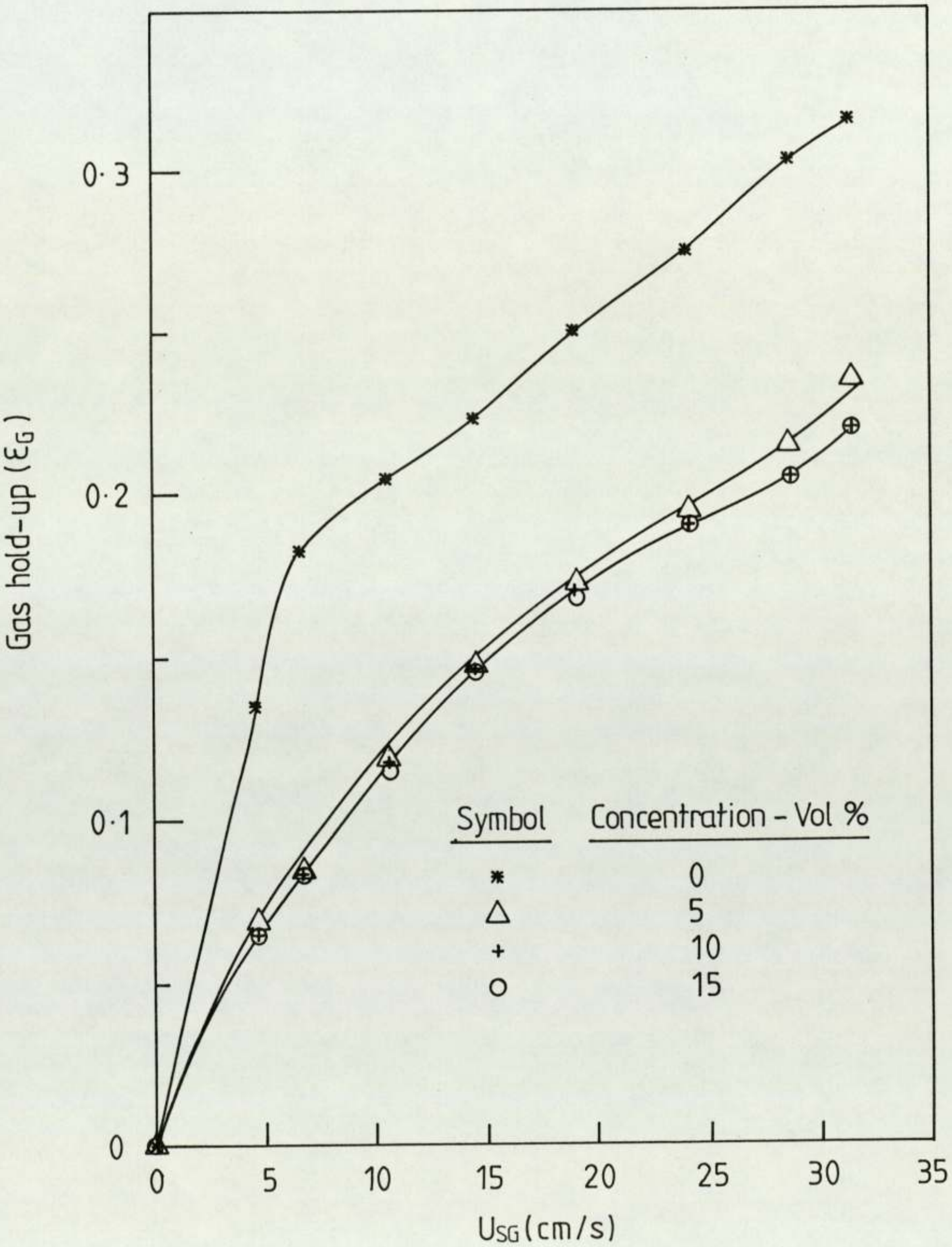


Figure 6.6 Effect of solid concentration on gas hold-up in the two dimensional bubble column.

Three-phase system (Air - Water - Styrocell R351)

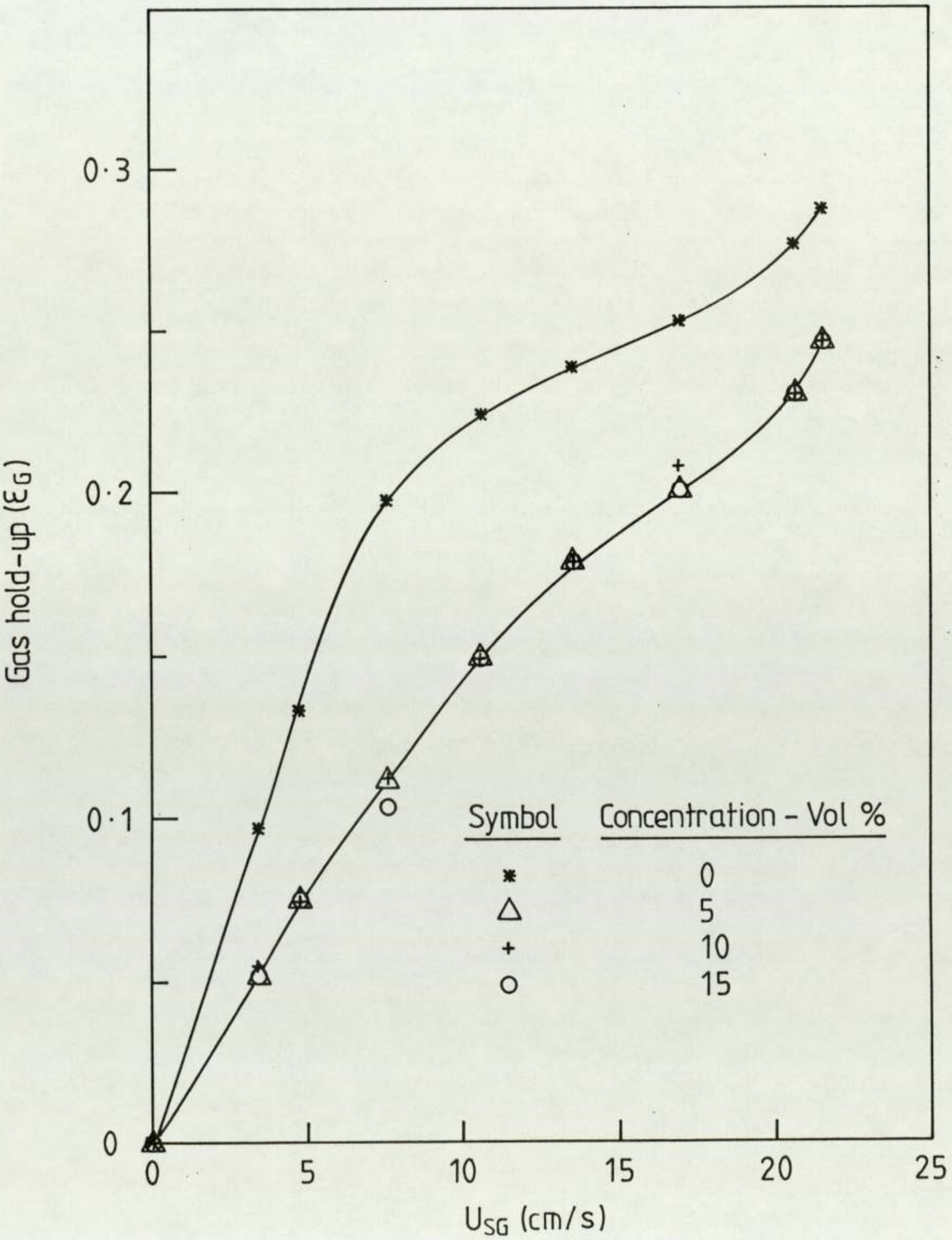


Figure 6.7 Effect of solid concentration on gas hold-up in the two dimensional bubble column.

Three-phase system (Air-Water-Ballotini)

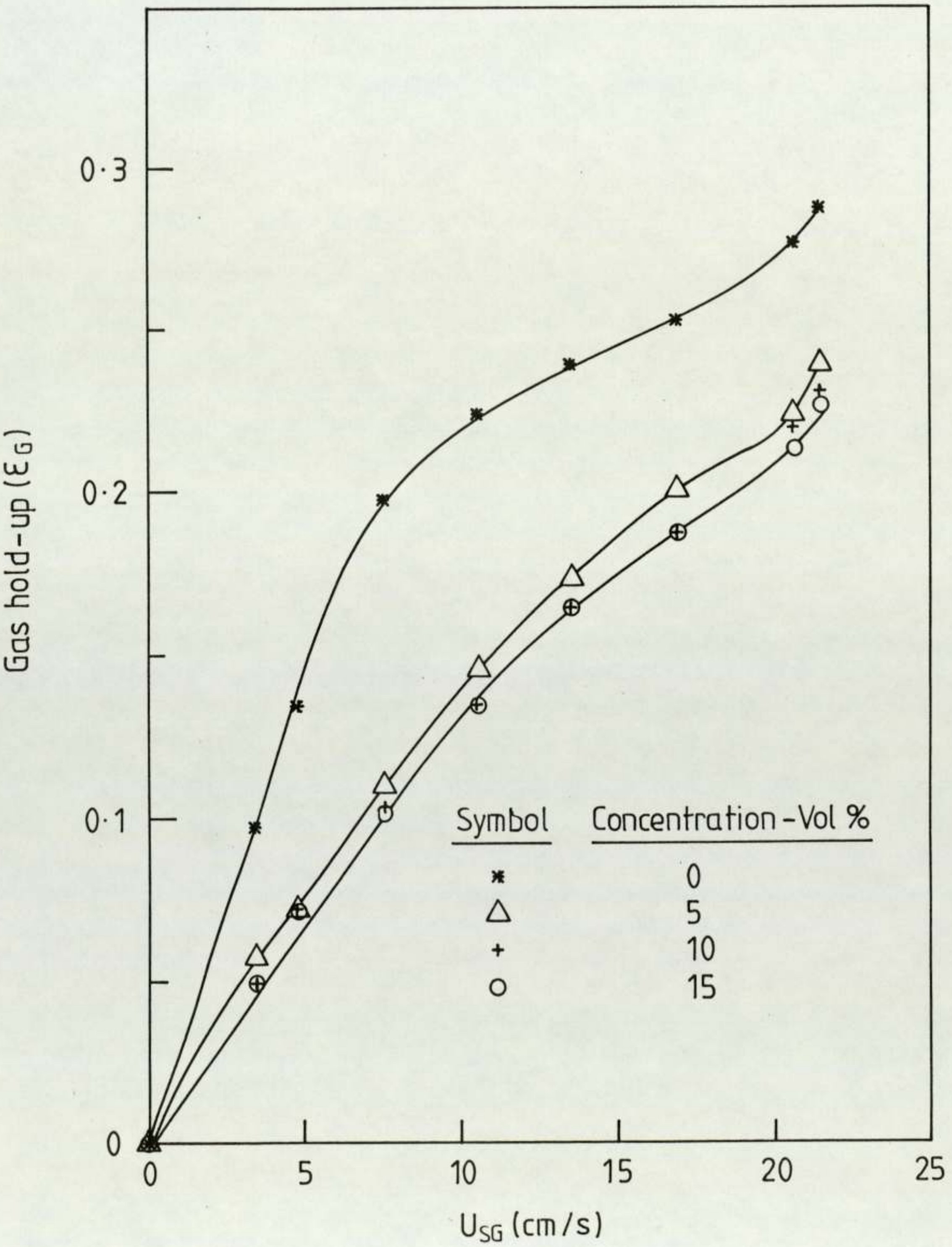


Figure 6.8 Effect of solid concentration on gas hold-up in the two dimensional bubble column.

Three-phase system (Air-Water-Amberlite)

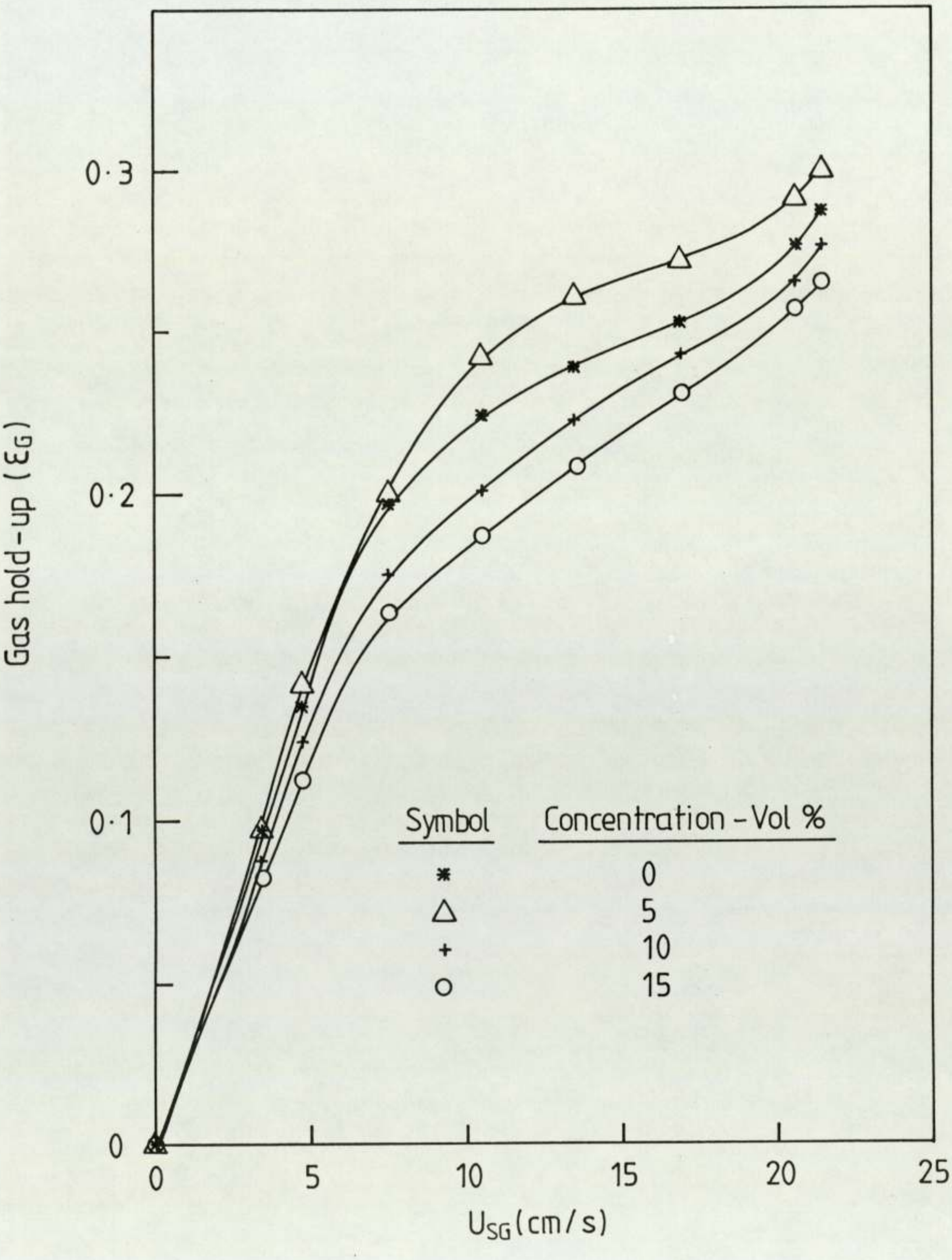
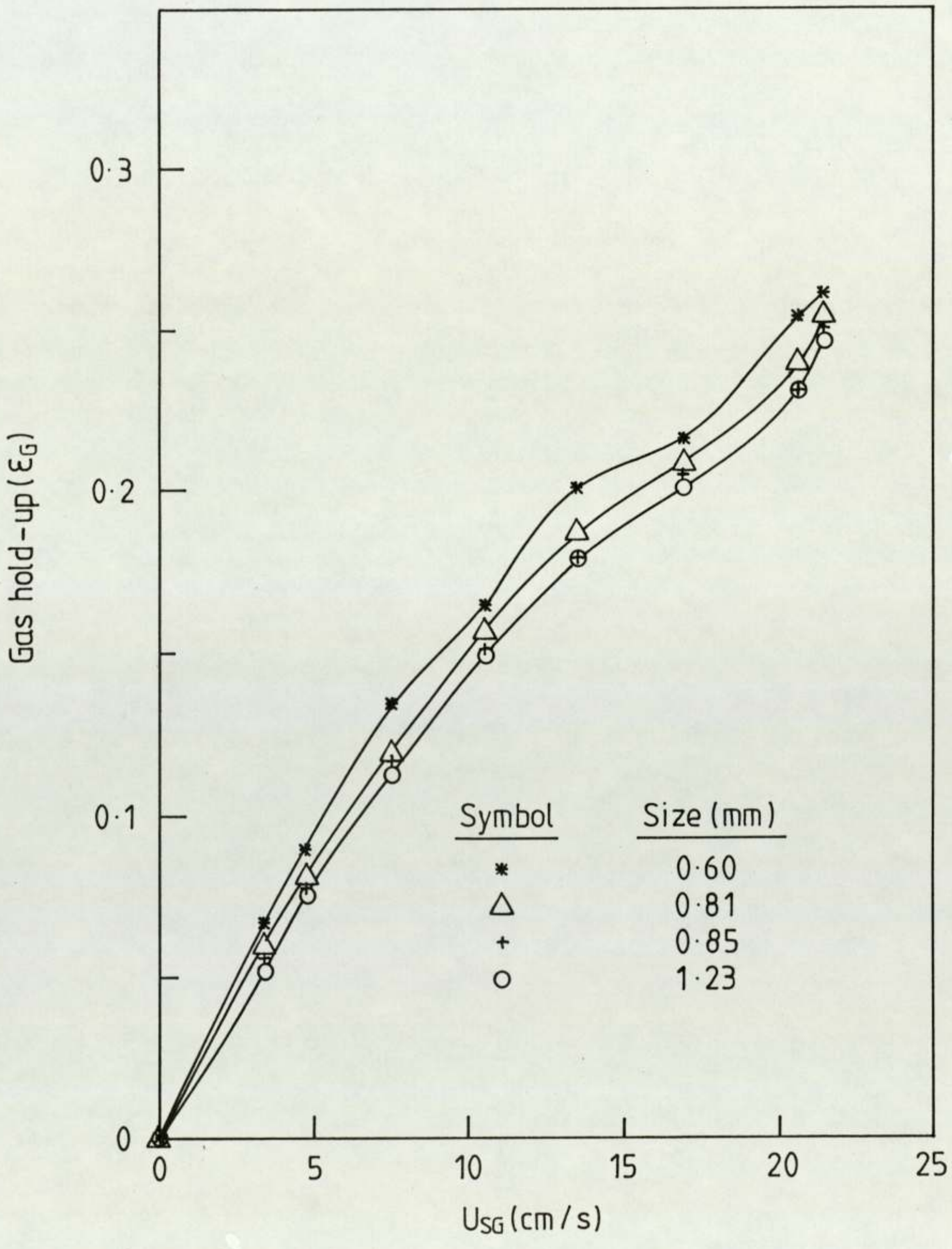


Figure 6.9 Effect of solid size on gas hold-up in the two dimensional bubble column.

Three-phase system (Air-Water-Styrocell)



Effect of Solid Density

From Table 6.1 and Figure 6.10, it can be deduced that increasing the solids density causes only a slight decrease in gas hold-up.

Effect of Solids Wettability

The surface property of solid particles proved to be a significant factor affecting gas hold-up. The behaviour of the four different types of solid particle were classed into two broad groups: non-wettable and wettable. Figure 6.10 shows the effect of non-wettable solids on gas hold-up in three-phase fluidisation. It should be noted that no attempt was made to clean the glass particles before use. The effect of wettable particles is shown in Figure 6.11. The detailed experimental data are to be found in Tables D10 and D11 in Appendix D.

6.4.3 Effects of Liquid Phase (Additives)

Most fermentation media consist of an aqueous solution and an organic liquid as the liquid phase. Liquid phase properties have been varied by the use of additives - both organic and inorganic - and the effects of these solutes on gas hold-up in the three-phase system have been determined. Alcohols, ketones, esters and electrolytes were used in the gas-liquid-solid systems; Styrocell "R551" particles was used as the solid phase.

Effect of Alcohols

The effects of methanol and ethanol on gas hold-up in the three-phase system are presented in Figures 6.12 and 6.13 respectively. The data are given in Tables D12 and D13 (Appendix D).

Figure 6.10 Effect of non-wettable solid particles on gas hold-up in the two dimensional bubble column.

Three-phase systems.

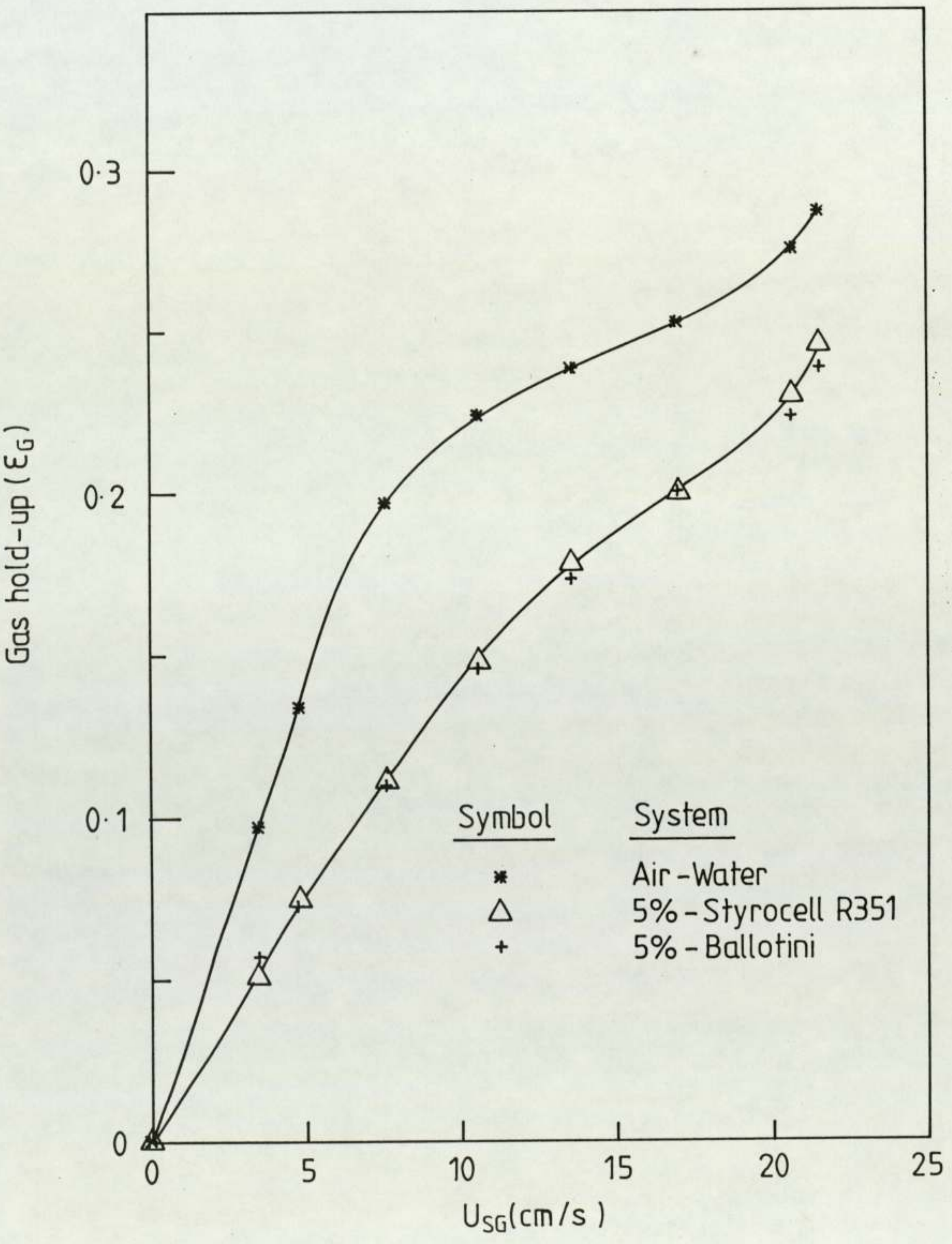


Figure 6.11 Effect of wettable solid particles on gas hold-up in the two dimensional bubble column.

Three-phase systems.

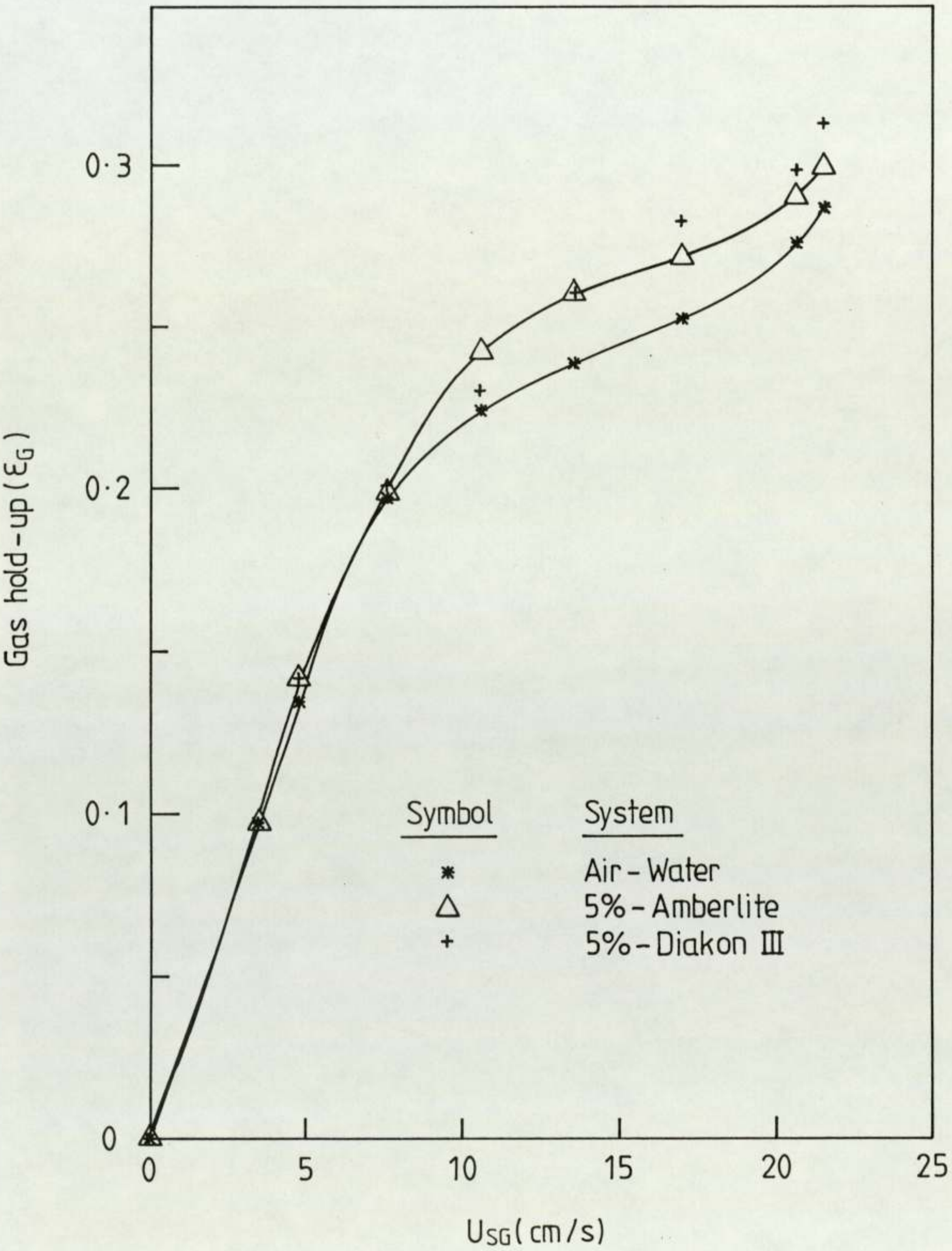


Figure 6.12 Effect of solid-phase and alcohol (Methanol) on gas hold-up in the two dimensional bubble column.

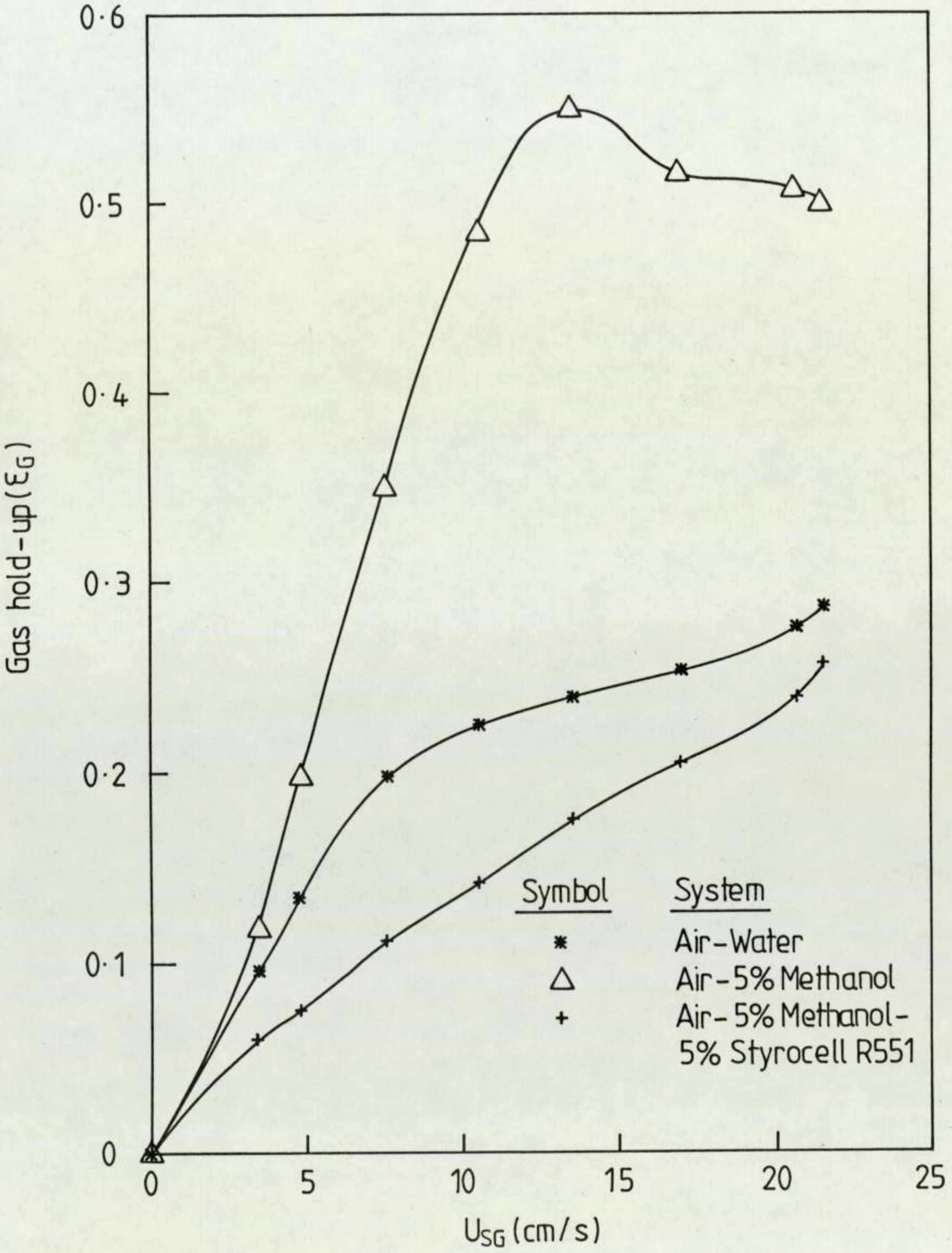
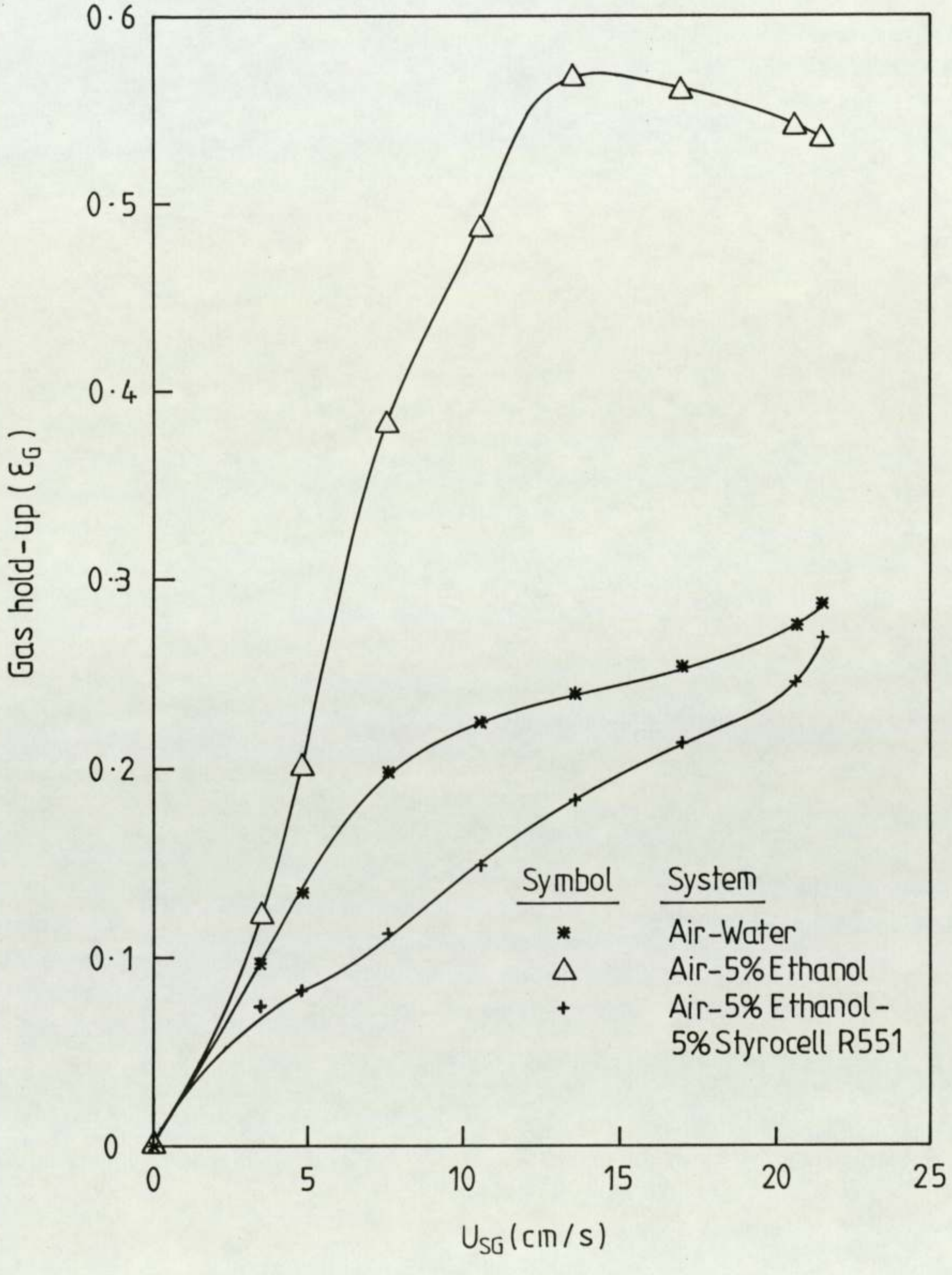


Figure 6.13 Effect of solid-phase and alcohol (Ethanol) on gas hold-up in the two dimensional bubble column.



There was a reduction in gas hold-up through the alcohols still caused a slight increase compared to systems without alcohols. The effect of solids was much more pronounced in reducing the hold-up values. The ethanol system gave a higher hold-up compared to the value when using methanol solution.

Effect of Ketones

The results showing the effects of acetone and butanone on gas hold-up are shown in Figures 6.14 and 6.15, with the data listed in Tables D14 and D15 in Appendix D.

Like alcohols, there was a reduction in hold-up though the ketones still caused a slight increase compared to that in systems without ketones. The effect of solids once again dominated system behaviour.

Butanone has a similar effect to that of acetone.

Effect of an Ester

The effect of ethyl acetate on gas hold-up in three-phase fluidisation is illustrated in Figure 6.16; the experimental data are listed in Table D16 in Appendix D.

The reduction in gas hold-up brought about by the addition of Styrocell particles followed a similar trend to that observed with the other organic solutes.

Effect of Electrolytes

The effects of using sodium chloride and calcium chloride on gas hold-up are shown in Figures 6.17 and 6.18 respectively. The full set of data is given in Tables D17 and

Figure 6.14 Effect of solid-phase and ketone (Acetone) on gas hold-up in the two dimensional bubble column.

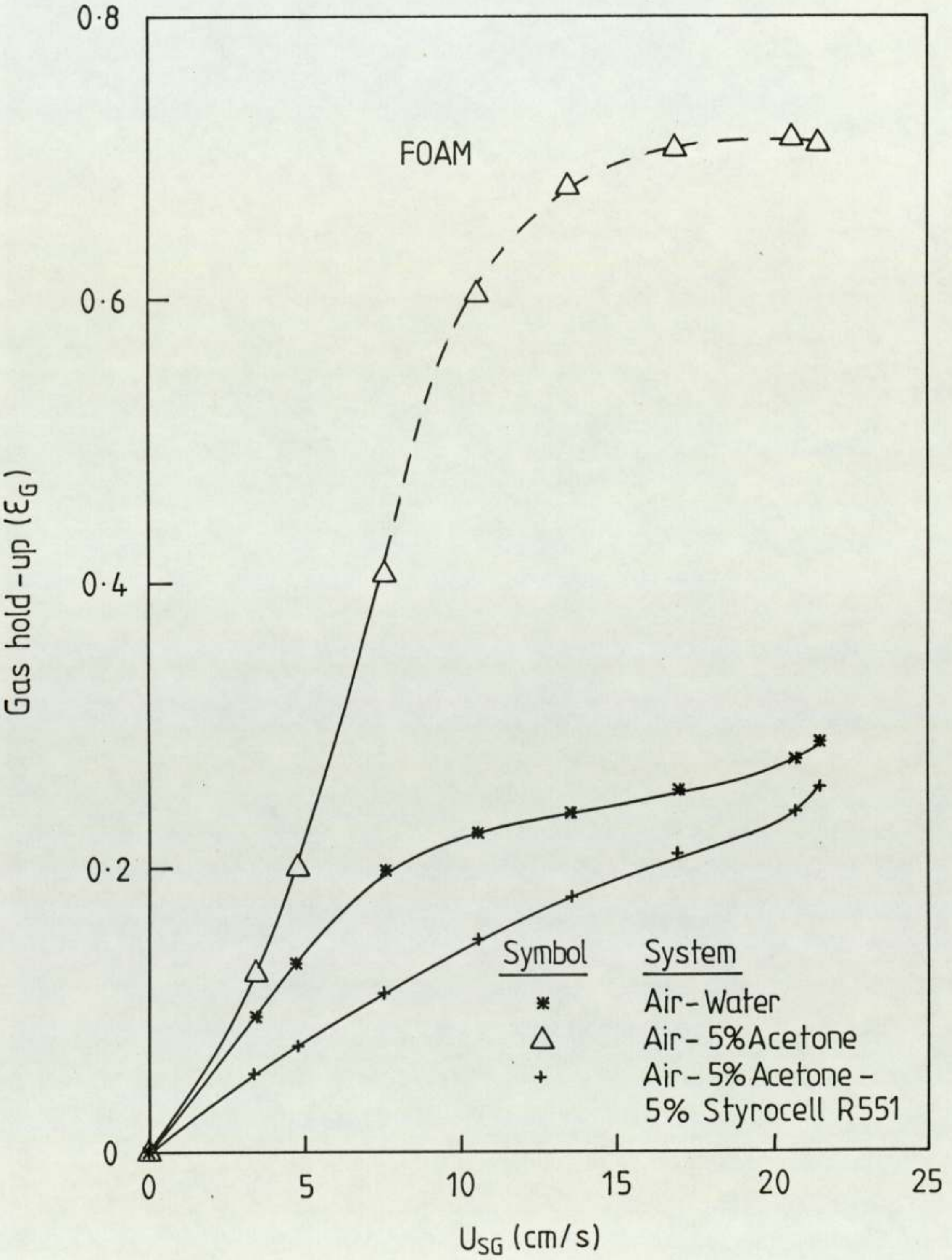


Figure 6.15 Effect of solid-phase and ketone (Butanone) on gas hold-up in the two dimensional bubble column.

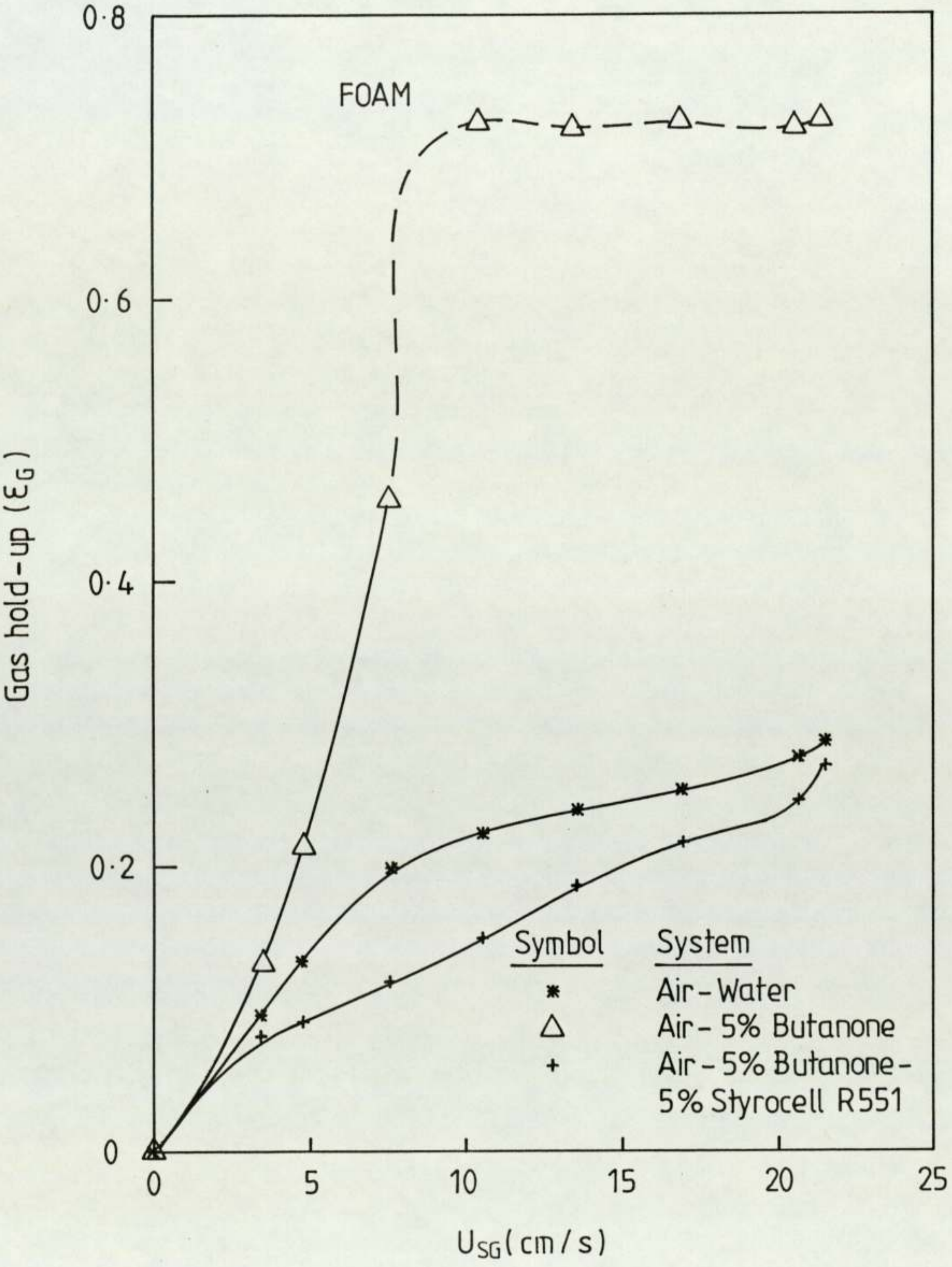


Figure 6.16 Effect of solid-phase and ester (Ethyl Acetate) on gas hold-up in the two dimensional bubble column.

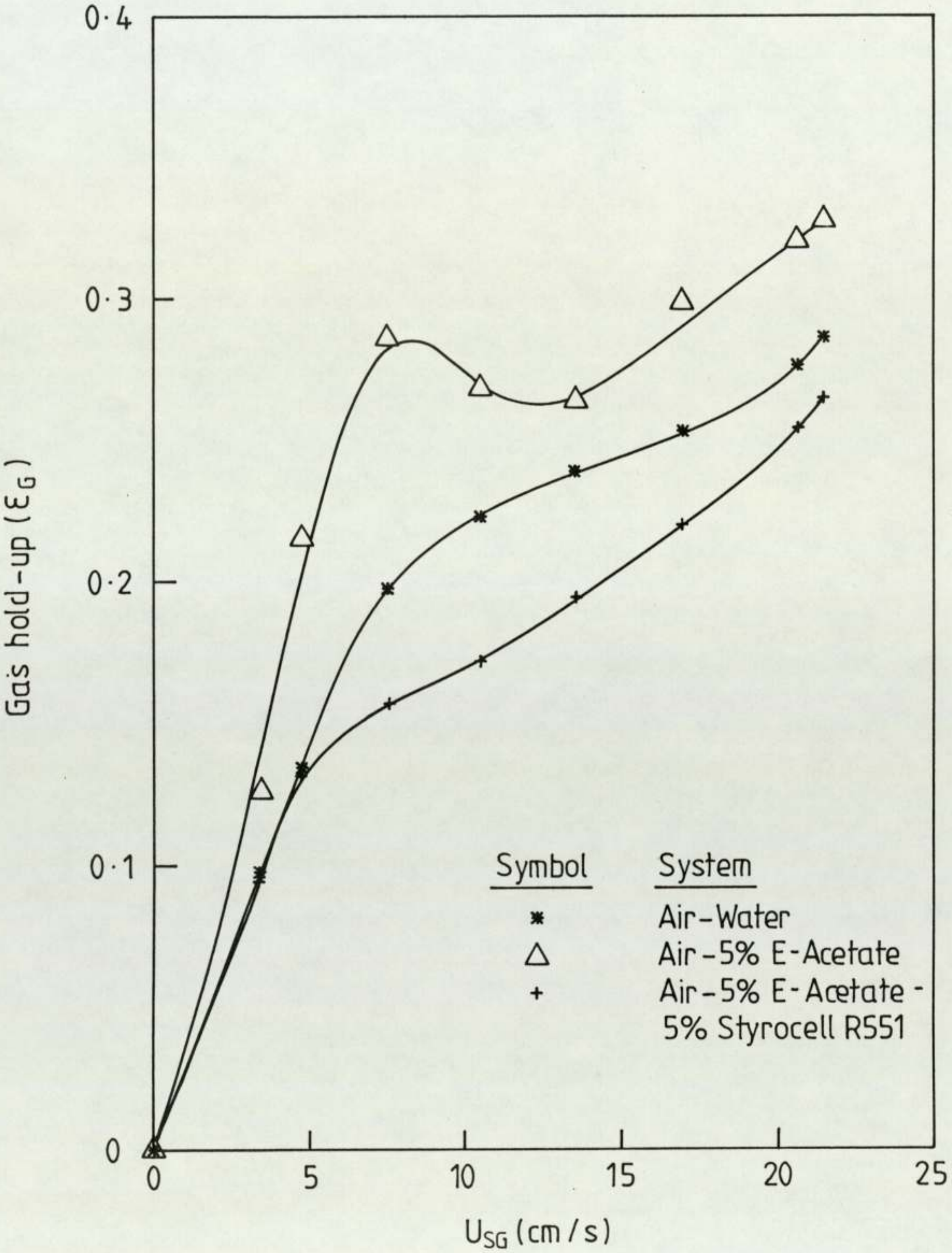


Figure 6.17 Effect of solid-phase and electrolyte (NaCl) on gas hold-up in the two dimensional bubble column

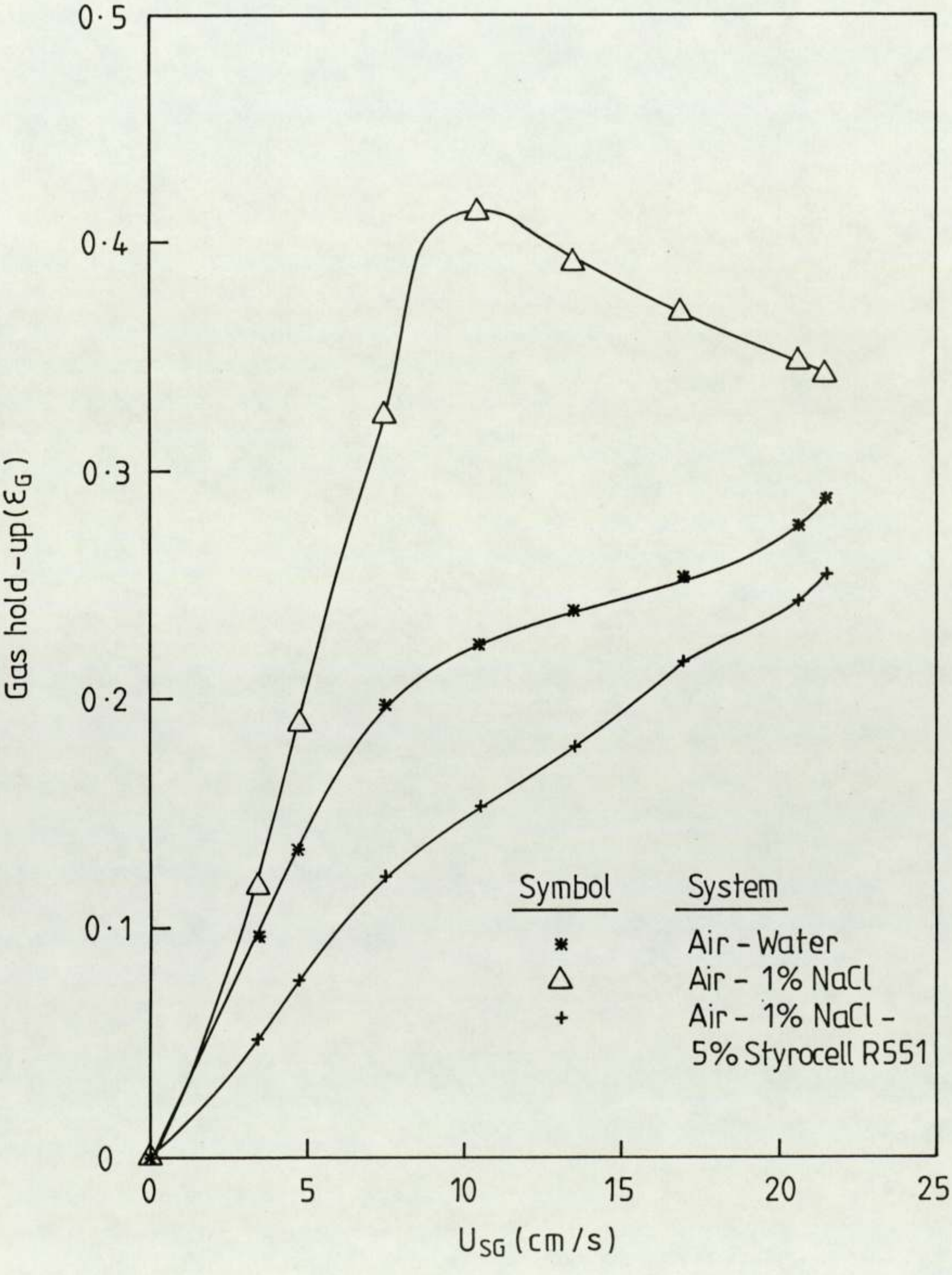
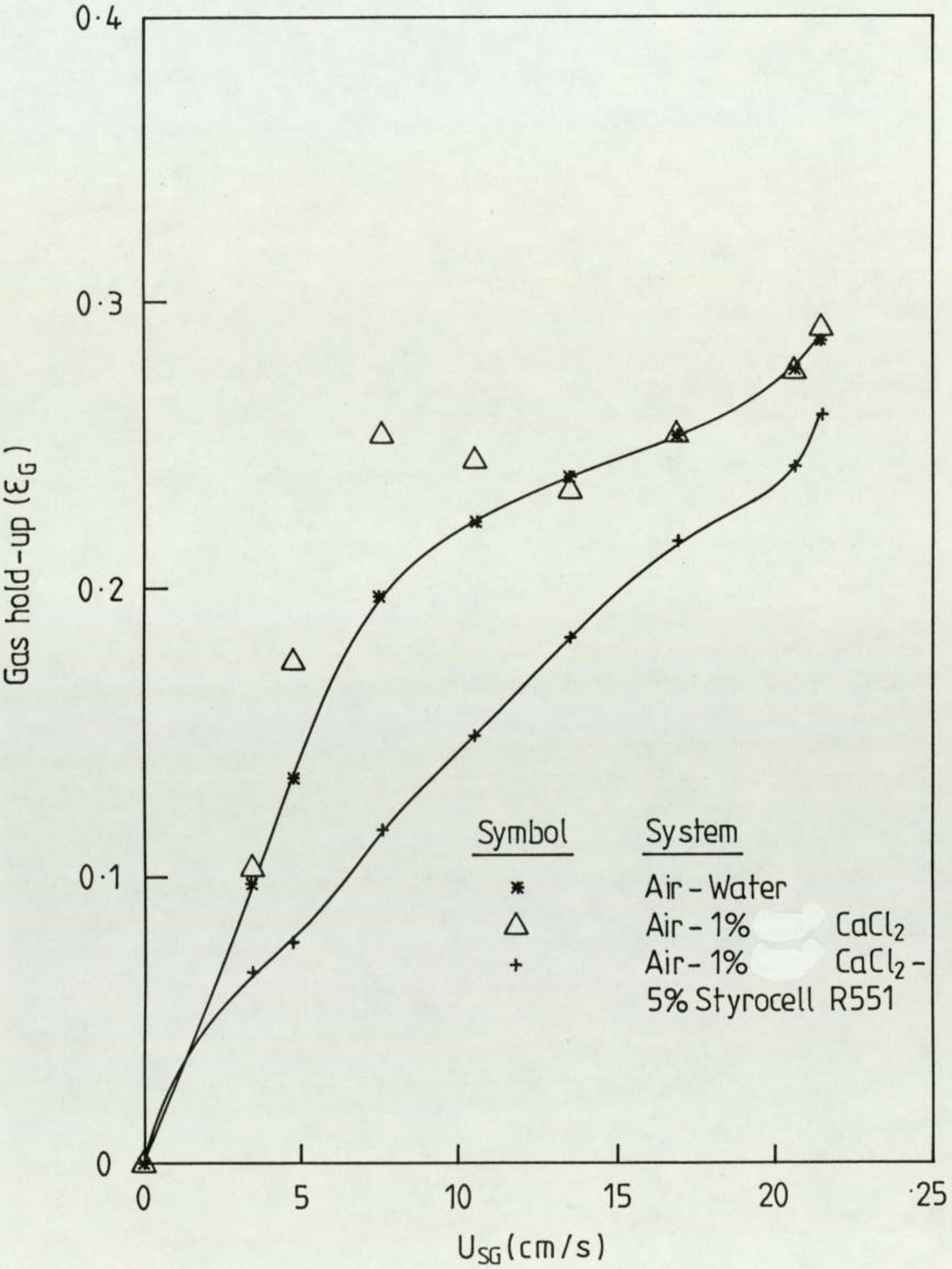


Figure 6.18 Effect of solid-phase and electrolyte ($CaCl_2$) on gas hold-up in the two dimensional bubble column.



D18 in Appendix D.

The expected reduction in gas hold-up on addition of solids is clearly illustrated by the plots.

6.5 **Discussion**

6.5.1 Effects of Operating Parameters

In three-phase fluidisation, a bed of solid particles is fluidised by upward co-current flow of gas and liquid. The liquid phase forms a continuous medium in which the solid particles are suspended and through which the gaseous phase flows as discrete bubbles. Momentum is transferred from the gas bubbles to the solids via the liquid medium. The behaviour of the bed is thus influenced by the operating parameters amongst other factors.

Effect of Superficial Gas Velocity

In general, an increase in gas velocity, increased the gas hold-up. Similar results have been reported by Ostergaard (1964), Viswanathan et al (1965), Kato et al (1972) and Kim et al (1972, 1975). At low gas velocities (< 10 cm/s), the bubbles rise separately in the liquid; the effect of gas velocity on the gas hold-up is quite pronounced. In this regime, an approximately linear relationship between gas hold-up and superficial gas velocity exisits. At higher gas velocities (10 to 20 cm/s), due to the significant coalescence of bubbles, a large fraction of the column is occupied by large bubbles. Under these conditions, an increase in gas velocity does not bring about such a significant increse in gas hold-up. As gas velocity is increased further (> 20 cm/s), there is a steep rise in gas hold-up.

Effect of Superficial Liquid Velocity

Introducing liquid into the column simultaneously with the gas causes a decrease in gas hold-up, although the effect is relatively small over the range of U_{SL} values (< 1 cm/s) explored in this study. This result is in agreement with those reported by Michelsen and Ostergaard (1970), Kim et al (1972), Dhanuka and Stepanek (1978), Begovich and Watson (1978), Kumar and Roy (1974). Increasing the liquid flow rate causes a quicker wash-out of the gas phase and thus reduces the gas hold-up.

6.5.2 Effects of Solid Phase Properties

In discussing the effects of solid phase properties, the effects of solids wettability, concentration, size and density are considered.

Effect of Solid Wettability

This study has revealed two different types of solids behaviour in gas-liquid-solid fluidisation. While the presence of some particles reduces hold-up compared to that in the solids-free system, the presence of others increases the gas hold-up. This effect due to the surface property of the solids is termed wettability. While the non-wettable solids caused a decrease, the wettable solids caused an increase at 5%.

Solids wettability can be characterized by the ability of a liquid to wet a solid surface. The work of adhesion, W, is given by Dupre's equation (Adamson, 1976):

$$W = \sigma_L (1 + \cos \theta), \qquad \text{........ 6.2}$$

where σ_L is the liquid surface tension and θ is the contact angle of the liquid on the given solid surface in the presence of the gas.

As the contact angle increases, W decreases and the solid shows pronounced non-wettability. Therefore, a decrease in contact angle will make gas-solid contact less likely. As the probability of gas-solid contact decreases, bubble breakup will be enhanced, and so the small bubble fraction will be increased. As the contact angle increases, the probability of gas-solid contact is more likely and hence the solids tend to induce bubble coalescence.

The wettable solids are hydrophilic in nature, while the non-wettable are hydrophobic. The latter tend to accumulate on top of the water surface and agglomerate with each other. The author's experiments indicate that, besides promoting bubble coalescence, the non-wettable nature of the particles also tends to reduce bubble break-up. An increase in the coalescence characteristics results in fast rising bubbles, which reduce gas hold-up. In the present study, Styrocell and Ballotini gave lower gas hold-ups than those obtained with Amberlite resin and Diakon particles. Styrocell and Ballotini are non-wettable solids and, because of the increase in gas-solid adherence, bubble break-up is reduced. Amberlite resin and Diakon particles are wettable by water. As a result, less coalescence is present and so high gas hold-up results. Kim et al (1972), Bhatia et al (1972), Armstrong et al (1976) and Godbole et al (1983) also noted similar behaviour between different wettable and non-wettable solids.

Effect of Solids Concentration

Increasing the solids concentration, generally causes a decrease in gas hold-up. Adding a small amount of solids is enough to break the resistance between bubbles and improve the possibility of coalescence even at low gas velocities. As more solids are added, there is more coalescence in the column. However, a critical point is reached beyond which further increase in concentration of solids has little impact on bubble behaviour. The effect of solids concentration is more pronounced with wettable solids than non-wettable solids. All the solids used are suspended during operation except Ballotini which is partially fluidised.

The dependence of gas hold-up on solids concentration can also be related to the liquid viscosity changing with solids concentration. Liquid viscosity is known to play an important role in determining the behaviour of bubbles. In general, low viscosity solutions exhibit bubble disintegrating behaviour. However, a trend towards bubble-coalescing behaviour is observed on increasing the viscosity of the liquid medium as shown by Calderbank (1967). Increase in liquid viscosity with the addition of solid particles has been observed by Vasalos et al (1980) and Javdani et al (1976); coalescence of bubbles then leads to reduced gas hold-up values.

Effect of Solids Size

An increase in solids size decreased the gas hold-up, the effect being less marked for larger particles. This observation is similar to the ones made by Kato et al (1973), Ying et al (1980), Kara et al (1982) and Kulkarni et al (1983). Greater coalescence of bubbles was found on increasing particle size, thus resulting in lower gas hold-up.

Effect of Solids Density

Increasing the solids density has been shown to reduce the gas hold-up, though slightly, a result in agreement with those from the investigation of Koide et al (1984).

In general, for particles to lower resistance to bubble coalescence, they should be able to mix well even at low superficial gas velocities and should have significant momentum. When particles densities are greater than that of water, at low superficial gas velocities, the bed only expands by a limited amount. Unlike particles whose densities are similar or lower to that of water, these heavier particles do not mix to any great extent. Consequently, heavier particles do not lower the resistance to bubble coalescence and there is a reduction in gas hold-up values.

Light particles whose densities compared with that of water are very low, will float at low superficial gas velocities. Even when the gas velocity is high enough for liquid circulation to drag them down, they do not have enough momentum to break the resistance between the bubbles. Hence, they lead to higher gas hold-up values compared with those associated with the use of heavy particles.

6.5.3 Effects of Liquid Phase (Additives)

Different additives, as discussed before in Chapter 4, have different degrees of polarity associated with their -OH group. Therefore, mixing water with different organic and inorganic solutes causes their fields of attraction to differ in intensity. The molecules which have greater fields of force tend to pass into the bulk aqueous phase and those with the smaller fields tend to remain at the air-liquid surface. Therefore, the surface layer will be more concentrated in the species with the smaller dipole moment. The polar group of short-chain organics confers solubility on the whole molecule whilst the long-chain organics spread out as a monomolecular film on the air-water surface. The lateral adhesion between the long alkyl chains and the polarity of their -OH groups are the main factors keeping the molecules together as a coherent film. The behaviour of soluble organic and inorganic solutes in water is discussed in detail in Chapter 4.

The Addition of a Non-Wettable Solid-Phase to Soluble Organic Solutions

The effects of adding a non-wettable solid to organic systems have been explored and it will be noted that gas hold-up was decreased even more than that for solids-free aqueous organic solutions. The solutes used have negligible non-polar ends, and the polarity of their -OH groups is almost the same as the polarity of water. When their concentration in water is low (about 0.5%) the system on the whole does not show much deviation from that of pure water. The hold-up was slightly higher than in water for some gas velocities suggesting that the solutes do have a small effect.

It would appear that the presence of solids has an over-riding effect on the gas hold-up. As discussed before, they promote bubble coalescence, giving rise to high bubble velocities and thus lower hold-up values. The effect of solids on the gas phase is thus stronger than that of the organic additives (in the liquid phase).

The Addition of a Non-Wettable Solid-Phase to Inorganic Solutions

The effect of adding non-wettable solids to electrolyte solutions is also to decrease gas hold-up though the figures are slightly higher than those for air-water-solid systems without electrolytes. When the concentration of electrolytes in water is low, they become distributed throughout the bulk of the system. The inter-molecular forces are electrostatic in nature and are much stronger than those in pure water. When non-wettable particles are added to such systems, their poor degree of wettability will tend to decrease gas hold-up or at least maintain it at values found in water. Therefore, the solid particles will act in the same way as in water.

Again, it will be noted that the presence of solids has an over-riding effect on gas hold-up. Looked at another way, this means that the effect of the electrolyte is significantly reduced by the action of the solids. Bubble coalescence is unhindered and lower gas hold-up values result.

6.6 Empirical Correlations

A computer program was developed which calculates gas hold-up from published correlations and compares them with the experimental values (see Appendix I). Four correlations were chosen for this purpose and are listed below:

Akita and Yoshida (1973)

$$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = A \left(\frac{g\,D_c^2\,\rho_L}{\sigma}\right)^{1/8} \left(\frac{g\,D_c^3}{\gamma_L^2}\right)^{1/12} \left(\frac{U_{SG}}{\sqrt{g\,D_c}}\right) \qquad \text{........ 6.3}$$

where A = 0.2 for pure liquids and non-electrolytes.
= 0.25 for electrolytes.

Kito et al (1976)

$$\frac{\varepsilon_G}{\{\varepsilon_G\,(1-\varepsilon_G)^2\}^{0.44}} = 0.5 \left(\frac{D_c\,U_{SG}\,\rho_L}{\sigma}\right)^{0.11} \left(\frac{U_{SG}}{\sqrt{g\,D_c}}\right)^{0.22} \qquad \text{........ 6.4}$$

Begovich and Watson (1978)

$$\varepsilon_G = 1.15\,U_{SG}^{0.692}\,d_p^{0.107}\,D_c^{-0.037} \qquad \text{........ 6.5}$$

Hikita et al (1980)

$$\varepsilon_G = 0.672 \left(\frac{U_{SG}\,\mu_L}{\sigma}\right)^{0.578} \left(\frac{\mu_L{}^4\,g}{\rho_L\,\sigma^3}\right)^{-0.131} \left(\frac{\rho_G}{\rho_L}\right)^{0.062} \left(\frac{\mu_G}{\mu_L}\right)^{0.107} \quad \text{........ 6.6}$$

The results are presented in Tables E1 to E8 in Appendix E. A plot of predicted gas hold-ups against experimentally obtained gas hold-ups for air-water-Styrocell systems is given in Figure 6.19.

The experimental values compare reasonably well with those based on the correlations. They are thus suitable and recommended for quick estimates of gas hold-up. The correlation of Begovich and Watson (1978) gave the closest values to the experimental data.

It is to be noted that some of these correlations: AKITA and YOSHIDA (1973) and HIKITA et al (1980) also apply to two-phase systems (see Chapter 4).

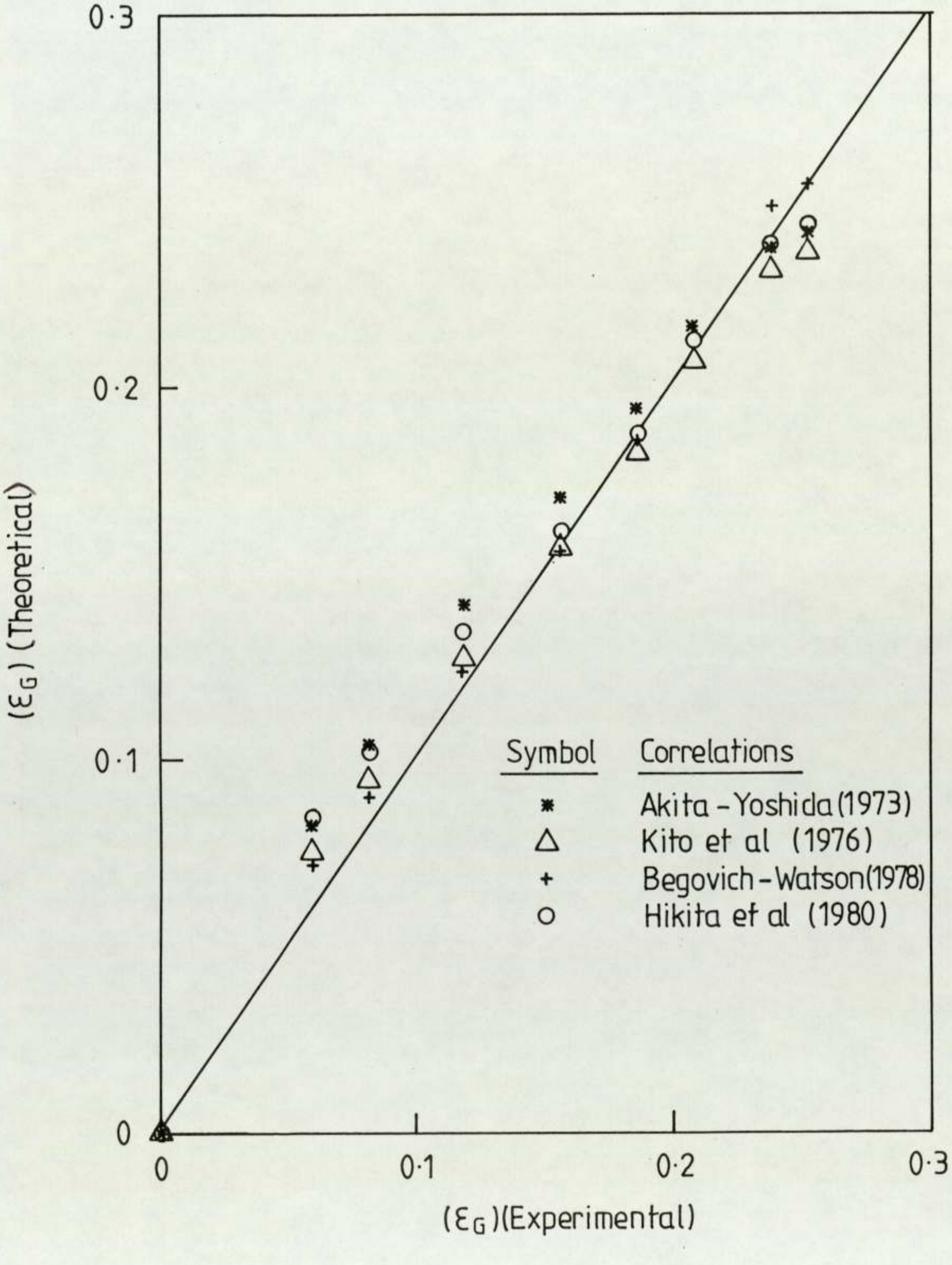
6.7 Conclusions

The effects of operating parameters and properties of solid and liquid phases on gas hold-up have been studied. The variables examined are superficial gas and liquid velocities, solids concentration, size, density and wettability. Key conclusions are as follows:

- gas hold-up increases with superficial gas velocity but decreases with liqiud velocity
- the type of solids influences gas hold-up significantly: with wettable solids, gas hold-up increases compared to that in a solids-free system, whereas, with non-wettable solids, gas hold-up decreases

Figure 6.19 Comparison of experimental gas hold-up with empirical correlations for the two dimensional bubble column

Three-phase system (Air - Water - Styrocell R551)



- gas hold-up decreases with an increase in solids concentration, the effect decreasing at higher concentrations
- an increase in solids size decreases the gas hold-up; the effect is reduced for larger particles
- increasing the solids density reduces the gas hold-up but only slightly
- the presence of non-wettable solids in organic and inorganic media decreases gas hold-up significantly; a change in the liquid phase property by using additives does not affect the gas hold-up in three phase systems
- empirical correlations can be used to provide estimates of gas hold-up and that due to Begowich-Watson (1978) is recommended.

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CHAPTER 7

FLOW REGIMES AND BUBBLE DYNAMICS IN THREE-PHASE FLUIDISED BEDS

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FLOW REGIMES AND BUBBLE DYNAMICS IN THREE-PHASE FLUIDISED BEDS

7.1 Introduction

The phenomena of bubbling and associated flows are striking and immediately obvious features of three-phase fluidised beds. Bubbles in fluidised beds are very important for they are responsible for most of the features that differentiate fluidised beds from other types of multiphase contractors. They dominate the flow of the gas phase through the system and cause particle movement which generally results in rapid and extensive mixing, and the very high heat transfer coefficients that can be obtained between the bed and immersed surfaces (including the containing walls).

An understanding of the phenomena of bubbling and the effects it produces is the key to understanding many of the relevant properties of fluidised beds. This coupled with the prevailing flow patterns provides the means for thorough analysis of operations in three-phase fluidised beds.

7.2 Review of the Literature

7.2.1 Bubble Dynamics

Previous studies of bubbles in fluidised beds can be divided into three categories:

(a) studies aimed at establishing the relationship between the size and rise velocity of single bubbles,

(b) studies of the characteristics of bubbles in freely bubbling beds, and

(c) studies of the mechanism of bubble coalescence and breakup.

It should be noted that this literature review covers cases where the solid phase is both partially and fully fluidised.

Characteristics of Single Bubbles

The flow behaviour of a single bubble was first investigated by Massimilla et al (1961) who measured the rise velocity of single bubbles of known volume injected into a 3.5in ID cylindrical, liquid fluidised bed of fine particles. Water and air were employed as the fluid phases, and the solids were 0.22mm silica sand, 0.79 and 1.09mm glass beads, and 0.2mm iron sand. The authors concluded that the relationship between the rise velocity and bubble size was essentially dependent upon bed expansion and was relatively insensitive to the particle size.

Henriksen and Ostergaard (1974), employing a cine photographic technique, measured the rise velocity of large single bubbles in a two-dimensional column. They used water, aqueous solutions of methanol and glycerol (as the fluidising liquids) to vary the surface tension and viscosity of the media between 22.7 - 72 dyne/cm and 1.0 - 170 cP respectively. The fluidised solids were 0.2, 1 and 3mm glass beads. They found that the relationship between the bubble rise velocity and the bubble size can be represented by an equation similar to that obtained by Davies and Taylor (1950) for spherical cap bubbles rising through inviscid liquid as:

$$U_B = \alpha \ (gR)^{\beta} \qquad \text{........ 7.1}$$

where g is acceleration due to gravity and R is the radius of circular cap bubble.

For spherical cap bubbles in gas-liquid systems, Davies and Taylor showed that the value of α is 2/3 and the exponent β to be 1/2. The values of β observed by Henriksen and Ostergaard in three phase beds ranged from 0.42 to 0.60. They concluded that surface tension had no effect on the rise velocity, but that the latter was somewhat reduced in beds of higher viscosity.

The rise velocity of single bubbles through a liquid fluidised bed was also studied by Darton and Harrison (1974). The equivalent diameter of the bubble, d_B, was found to vary from 5 to 25mm in a water fluidised bed of either 500μm or 1mm sand particles.

El-Temtamy and Epstein (1980) compared the bubble rise velocity predicted from the Davies and Taylor (1950) relationship with the experimental data of Darton and Harrison (1974). They showed that the predictions of the bubble rise velocity based on the Davies and Taylor equation is quite satisfactory provided the bed voidage is greater than about 0.5. The discrepancy at low bed voidage was attributed to the non-Newtonian behaviour of the concentrated solid suspension.

Characteristics of Bubble Swarms

Massimilla et al (1961) measured the size distribution of bubbles emerging from the top of freely bubbling three-phase beds. Their results showed that the mean bubble diameter increased with the gas flow rate, the distance from the distributor, and decreasing bed height. The latter parameter was found to be dominant. An empirical correlation was presented which related the bubble size to the distance above the distributor, the initial bubble size and the fractional expansion of the column. The results of Ostergaard (1966) qualitatively support these findings.

Rigby et al (1970) measured the local frequency, average size, size distribution and rise velocity of gas bubbles in three-phase fluidised beds of 10cm ID by electroresistivity probes. They used fine sand particles with diameters ranging from 0.12 to 0.775mm; air and water were the fluidising fluids. The average bubble size increased with increasing gas rate, decreasing liquid rate and increasing height above the distributor in a manner similar to that reported by Massimilla et al (1961). The distribution of the bubble sizes became wider with an increase of the axial distance or with a decrease of the liquid flow rate. They found that the bubble velocity obtained experimentally for the multibubble system is much higher than that predicted by the Davies and Taylor relationship for the single bubble system. An empirical correlation was developed for the bubble velocity involving the bed voidage and fluid superficial velocities.

Using a photographic technique, Page and Harrison (1972) measured the size distribution of bubbles leaving a three-phase fluidised bed of 0.5mm sand particles; air and water were used as the fluid phases. The bubble size was observed to decrease with increasing liquid velocity, and in contrast to Massimilla's results, with an increase in the gas rate at low liquid velocities. At the high liquid rates, it was observed to be independent of the gas flow and distributor design.

Darton and Harrison (1974) employed an impedance double-probe to study the bubble characteristics in water-fluidised beds of 0.55mm sand particles. They observed that the bubble size distributions were log-normal and the mean size increased with increasing gas rate and decreasing water rate.

Bruce and Revel-Chion (1974) investigated the average size of bubbles emerging at the surface of a fluidised bed of large glass particles ranging from 2 to 8mm in diameter.

They observed that the bubble diameter decreased with an increase of the particle size in the range of 2 to 4mm, reached a minimum at a particle size of 6mm and then increased at the size of 8mm. For the 6 and 8mm particles, the size of the bubbles decreased as the liquid velocity increased, reaching a minimum at a liquid velocity of 13.2 cm/s. A further increase in the liquid velocity increased the size of the bubbles. They also reported that the bubble diameter for a bed of 2mm particles was virtually independent of the gas flow rate, while for a bed of 6mm particles, it increased linearly with an increase in the gas flow rate.

Lee et al (1974) reported that the bubble size decreased with increasing particle diameter.

The characteristics of large two-dimensional bubbles in three-phase fluidised beds were studied by Kim et al (1977). The effects of surface tension and viscosity on the average bubble size and rise velocity were investigated by means of cine photography. Air, a variety of liquids and three sizes of particles (1.0mm glass beads, 2.6mm irregular gravel and 6mm glass beads) were employed as the three phases. Experiments were carried out at higher gas rates than have been previously studied to provide information in a regime of industrial importance. They indicated that the clouds of bubbles rise faster than a single isolated bubble expressed by the Davies and Taylor relationship. Both bubble size and rise velocity were found to increase with gas velocity but were relatively insensitive to the liquid velocity, viscosity and surface tension. At the high gas rates employed, the bubble characteristics were independent of particle size. Correlations were presented for calculating bubble size and rise velocity as follows:

$$d_B = 13.4\, U_{SL}^{0.052}\, U_{SG}^{0.248}\, \gamma^{0.008}\, \sigma^{0.034} \qquad \text{........ 7.2}$$

$$U_B = 83.1\, U_{SL}^{0.065}\, U_{SG}^{0.339}\, \gamma^{0.025}\, \sigma^{0.179} \qquad \text{........ 7.3}$$

or $U_B = 18.0\, d_B^{0.989}$ 7.4

where d_B is in mm, U_{SL} and U_{SG} in mm/s, σ in dyne/cm and γ in Ns/m^2.

It should be noted that wall effects may have influenced these correlations due to the fact that the experimental column had a rectangular cross section with a rather narrow width.

Bubble Coalescence and Breakup

The behaviour of a dispersion of gas in a fluidised bed is of particular interest, as coalescence and breakup of the bubbles in the dispersion occur simultaneously. Some of the published studies on these mechanisms are introduced below.

Massimilla et al (1961) reported that the bubble diameter increased considerably with the distance above the gas distributor in beds of small particles with diameters ranging from 0.22 to 1.1mm: this was attributed to the fact that the bubbles coalesced as they moved up the bed. Adlington and Thompson (1965) also observed considerable bubble coalescence taking place in a three-phase fluidised bed containing distributed sizes of alumina particles. Ostergaard (1964) observed that the bubble size at low liquid flow rates was larger than that at high liquid flow rates. Sherrard (1966) carried out a large number of observations on bubble size, for varying particle size, density, bed height and bed porosity. The high rate of bubble coalescence observed in beds of small particles of relatively low density was explained by reference to the relatively high viscosity of such beds.

For beds of small particles, Ostergaard (1966) measured the bubble frequency from a single orifice and the bubble frequency at the bed surface. It was noted that the rate of coalescence decreased as the bed porosity increased and the change in bubble frequency took place within a relatively short distance from the orifices.

On the contrary, Lee (1965) reported that break-up of bubbles takes place in a three-phase fluidised bed when the size of the particle is comparable to that of the bubble. Utilizing 6mm glass beads and a gas distributing device which provided rather large bubbles, he observed that the bubble diameter decreased with an increase in the bed height. The measurement of the bubble diameter was made immediately above the bed, which was maintained at a constant expansion ratio. In such a bed of coarse particles, the bubbles emerging from the surface of the bed appear to be of uniform size (Ostergaard, 1969).

Lee, Sherrard and Buckley (1974) postulated that bubbles are split when they are penetrated by a solid particle. In the absence of viscous forces, this was shown to occur if the Weber number exceeded three.

ie. We > 3

$$\frac{\rho_s U_B^2 d_p}{\sigma} > 3 \qquad \text{........ 7.5}$$

Photographic evidence was presented which the authors claimed supported their contention.

Henriksen and Ostergaard (1974) studied the breakup of 2cm bubbles in beds of water and methanol. Three solids (5mm steel spheres; 3 and 6mm glass spheres) were used. A bubble was held stationary by a downward flow of liquid and a particle was allowed to fall through it. In no case was the bubble observed to disintegrate. They concluded that the bubbles are broken up as a result of Taylor instability (1950) of their roofs.

Since the minimum sized particle capable of splitting a bubble in air-water beds was shown to be about 8.5mm, it was concluded that the instability was generated by fingers of liquid projecting down through the roof of the bubble.

7.2.2 Flow Regimes

The flow behaviour in a three-phase fluidised bed varies depending on the particle size, the degree of bed expansion, the liquid velocity and the gas velocity. Three flow regimes have received attention:

(a) the ideal bubbly flow regime (Darton and Harrison, 1975) or the bubble break-up regime (Michelsen and Ostergaard, 1970), or the bubble disintegration regime (Kim et al, 1972)

(b) the churn-turbulent regime (Darton and Harrison, 1975) or the bubble coalescing regime (Kim et al, 1972)

(c) the slug flow regime (Michelsen and Ostergaard, 1970; Kim et al, 1972).

Kim et al (1975) reported the existence of a critical particle size which separates the bubble disintegrating regime and bubble coalescing regime. The critical size for particles with a density similar to that of glass was reported to be about 2.5mm in diameter for an air-water system. However, the studies by Muroyama et al (1981) and Fan et al (1982) indicate that in addition to the particle size, flow regimes also strongly depend on the liquid and gas velocities used in the system.

7.3 Experimental Programme and Procedure

Experiments were carried out in the two-dimensional column described previously (see Chapter 3). The solids used were Styrocell and Amberlite resin particles. Water with various aqueous additives was employed as the fluidising liquid and air was used as the gas phase throughout the study.

At the start of an experiment, gas and liquid were introduced into the column at the desired velocities. When steady state was reached, bubble swarms in the two and three phase systems were photographed using a Canon AEI programmable camera. The column was lit from behind so that the bubbles were clearly visible through the front face. At varying velocities, photographs of different sections of the column were taken to obtain representative samples of bubbles and to observe any variation of bubble size in the axial direction. The negatives were examined by projecting stills onto a grid using a Motion Analyser. The mean bubble size was taken as the diameter of the circle having the same area as the average bubble. At higher gas velocities, about ten of the larger bubbles were measured to characterise bubble size; very small bubbles were ignored.

A JVC (GX-N70E) Colour video camera was used to film the experimental set-up during the operation. With the gas-flow suddenly cut-off, a recording of the fall of aerated height was also made for each gas velocity; the results were analysed on a TV screen. From the video recordings, analysis of bubble behaviour in terms of coalescence and breakup was possible although precise measurements of bubble size could not be obtained easily. As a consequence, the video recordings were only used to check results obtained by still photography.

The effects of superficial gas and liquid velocities, solids concentration, size and wettability on flow regimes and bubble dynamics were studied. The experimental programme is summarised below:

- studies with air-water-solid particles
- effects of liquid additives on the three-phase system

The additives used were alcohols, ketones, esters and electrolytes. The operating conditions were similar to those described in previous chapters.

7.4 Experimental Results

7.4.1 Effects of Operating Parameters

Effect of Superficial Gas Velocity

Typical plots of bubble size against superficial gas velocity for the air-water system and three-phase beds of Styrocell particles are shown in Figures 7.1 and 7.2. The data are given in Tables F1 and F2 (Appendix F). Figure 7.3 shows the effect of superficial gas velocity on flow regimes and bubble dynamics. Bubble sizes increased with gas velocity in all cases. The prevailing flow regime is a strong function of gas velocity.

Effect of Superficial Liquid Velocity

The effect of superficial liquid velocity is also portrayed in Figure 7.3 for the case of the Amberlite particles. Increasing the liquid velocity has little or no noticeable effect on bubble dynamics and flow regimes.

7.4.2 Effects of Solid Properties

The effects of solids concentration, size, density and wettability are reported in this section.

Figure 7.1 Variation of bubble diameter with superficial gas velocity in the two dimensional bubble column.

Two-phase system (Air - Water)

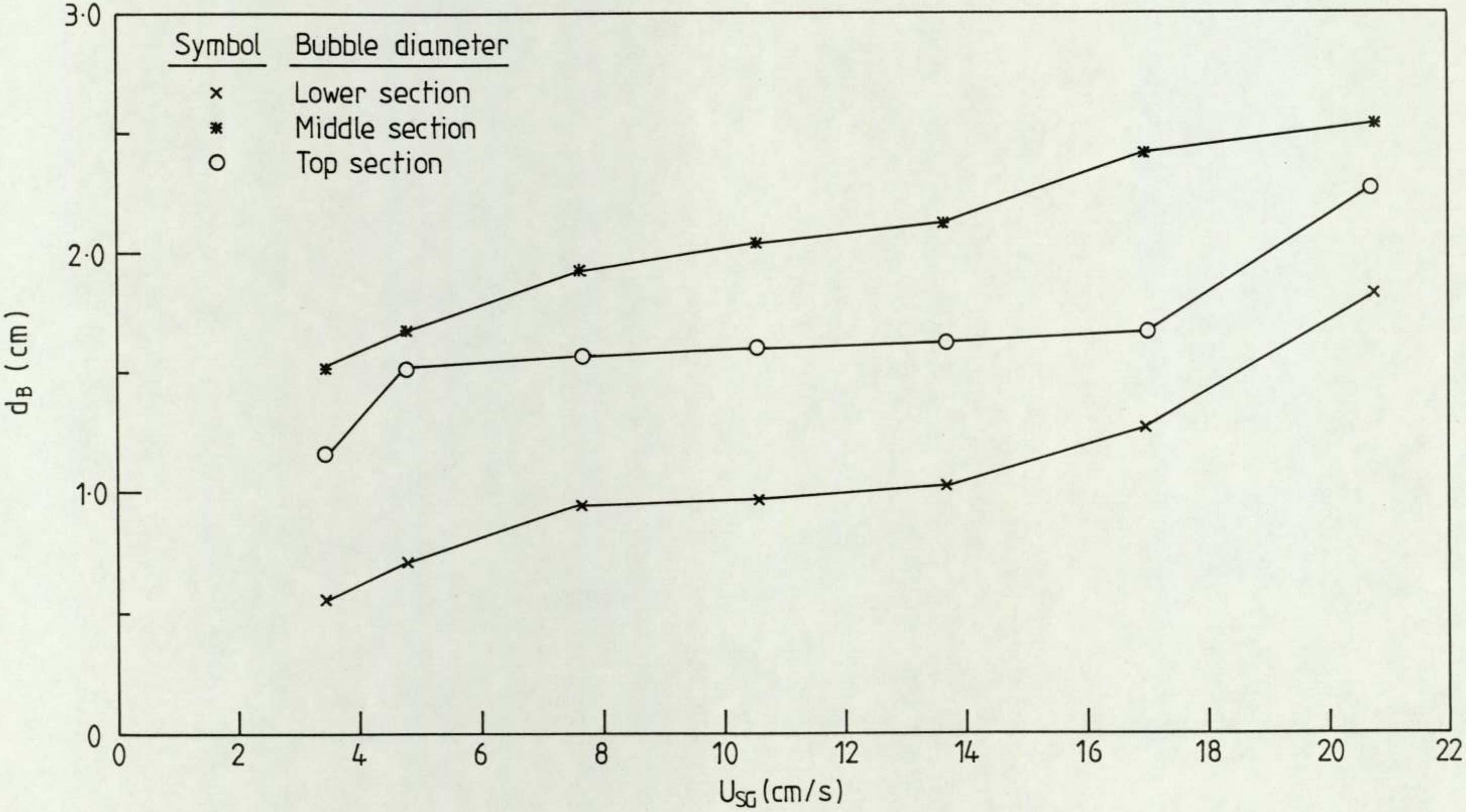


Figure 7.2 Variation of bubble diameter with superficial gas velocity in the two dimensional bubble column.

Three-phase system (Air-Water-Styrocell R551)

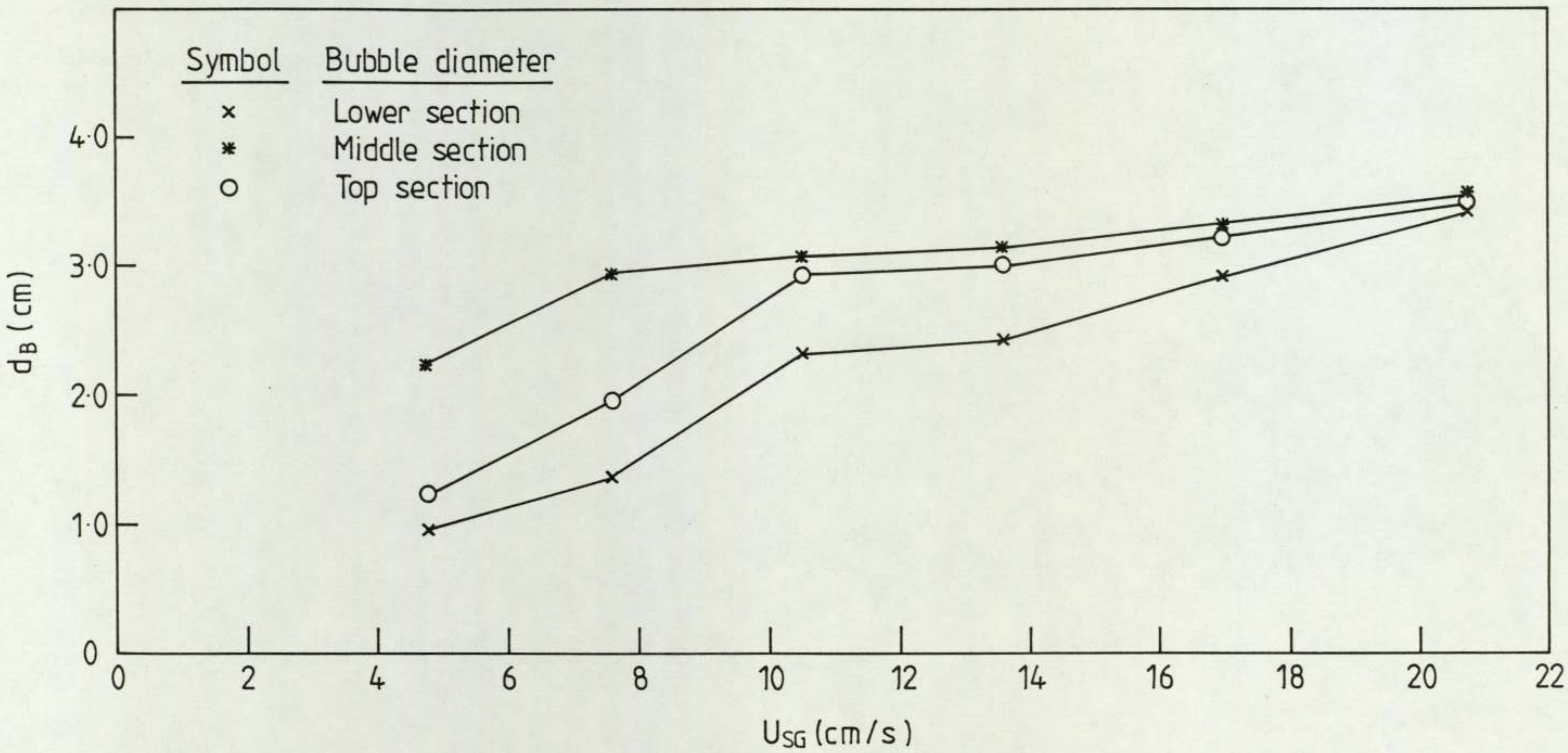


Figure 7.3 Effects of superficial gas and liquid velocities on flow regimes and bubble dynamics (three-phase fluidisation)

<u>5% vol Amberlite Resin</u>

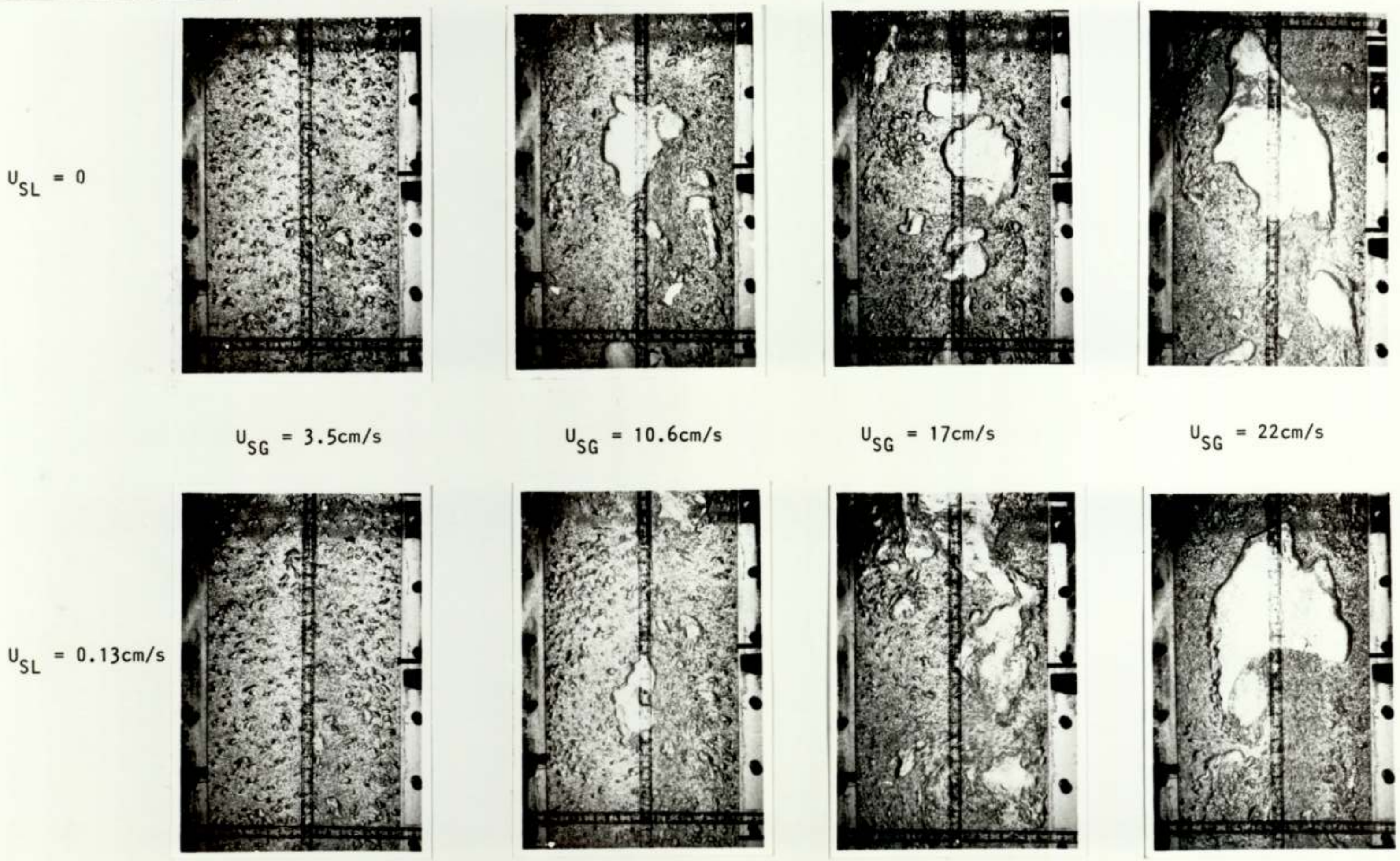
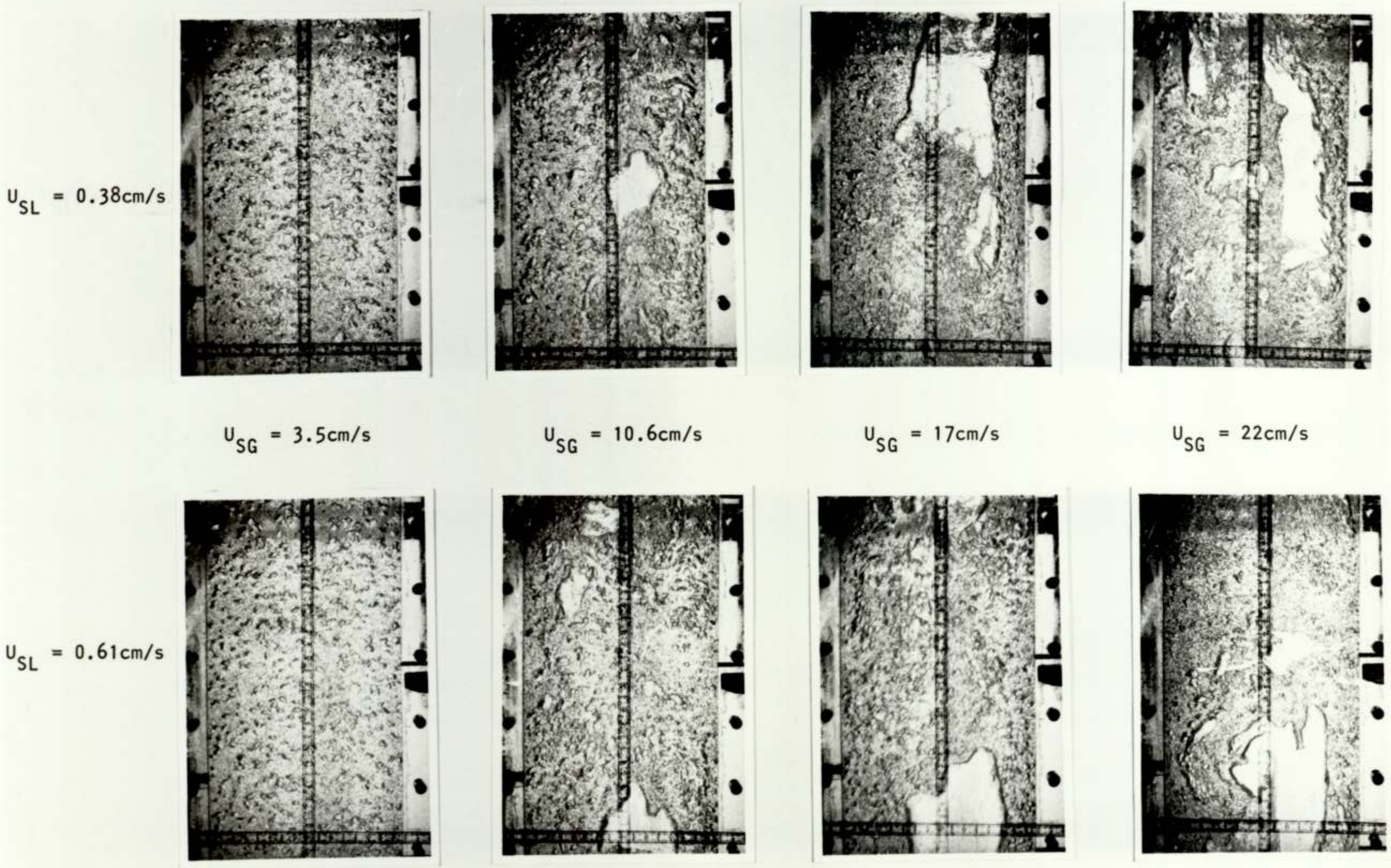


Figure 7.3 (continued)



Effect of Solids Concentration

Figure 7.4 shows the effect of solids concentration on bubble dynamics and flow regime with gas velocity as a parameter. There is no significant effect of concentration. This figure also illustrates how the presence of non-wettable Styrocell solids enhances coalescence of bubbles.

Effect of Solids Size and Density

Different sizes of Styrocell were used in the three-phase fluidised bed. Visual inspection of Figure 7.5 does not reveal any significant effect.

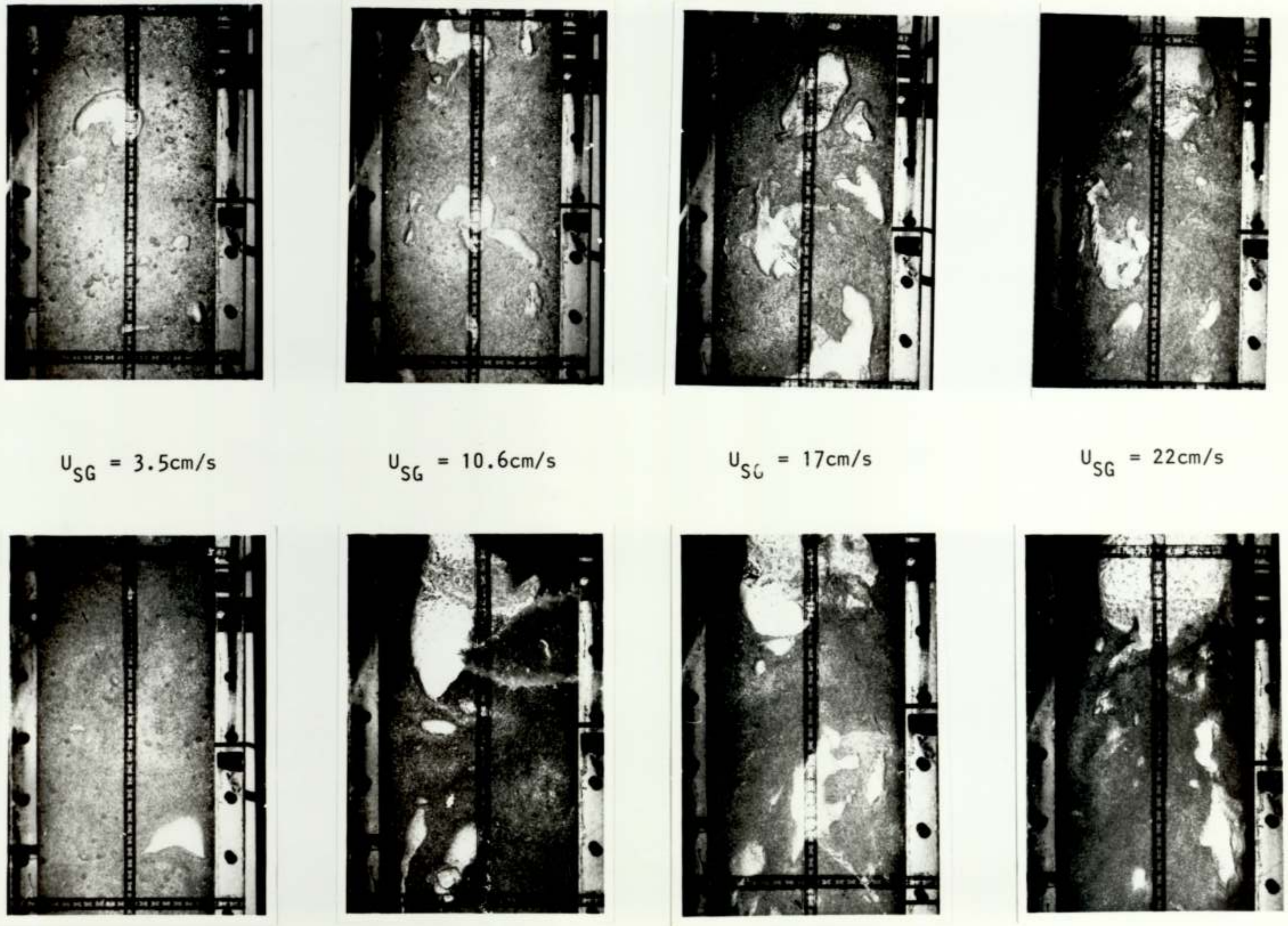
Effect of Solids Wettability

Amberlite resin and Styrocell particles were used as the wettable and non-wettable solids respectively. As shown in Figures 7.3 and 7.4, bubbles were smaller and coalescence was reduced with Amberlite resin especially at low gas velocities. With Styrocell particles, the bubbles were bigger due to coalescence.

7.4.3 Effects of Liquid Additives

Different types of additives (alcohols, ketones, esters and electrolytes) were used with suspensions of Styrocell particles. Figures 7.6 to 7.9 show the effects of organic and inorganic additives on bubble dynamics and flow regimes. Whereas in two-phase systems the effect of additives was very strong, in three-phase systems the solids had an overriding effect. Thus the use of additives to alter the properties of the liquid phase had little effect on the behaviour of the three-phase fluidised beds; indeed, the normal coalescence of bubbles persisted in all cases.

Figure 7.4 (continued)



10% vol
Styrocell

15% vol
Styrocell

U_{SG} = 3.5cm/s

U_{SG} = 10.6cm/s

U_{SG} = 17cm/s

U_{SG} = 22cm/s

Figure 7.5 Effects of solids size and density on flow regimes and bubble dynamics

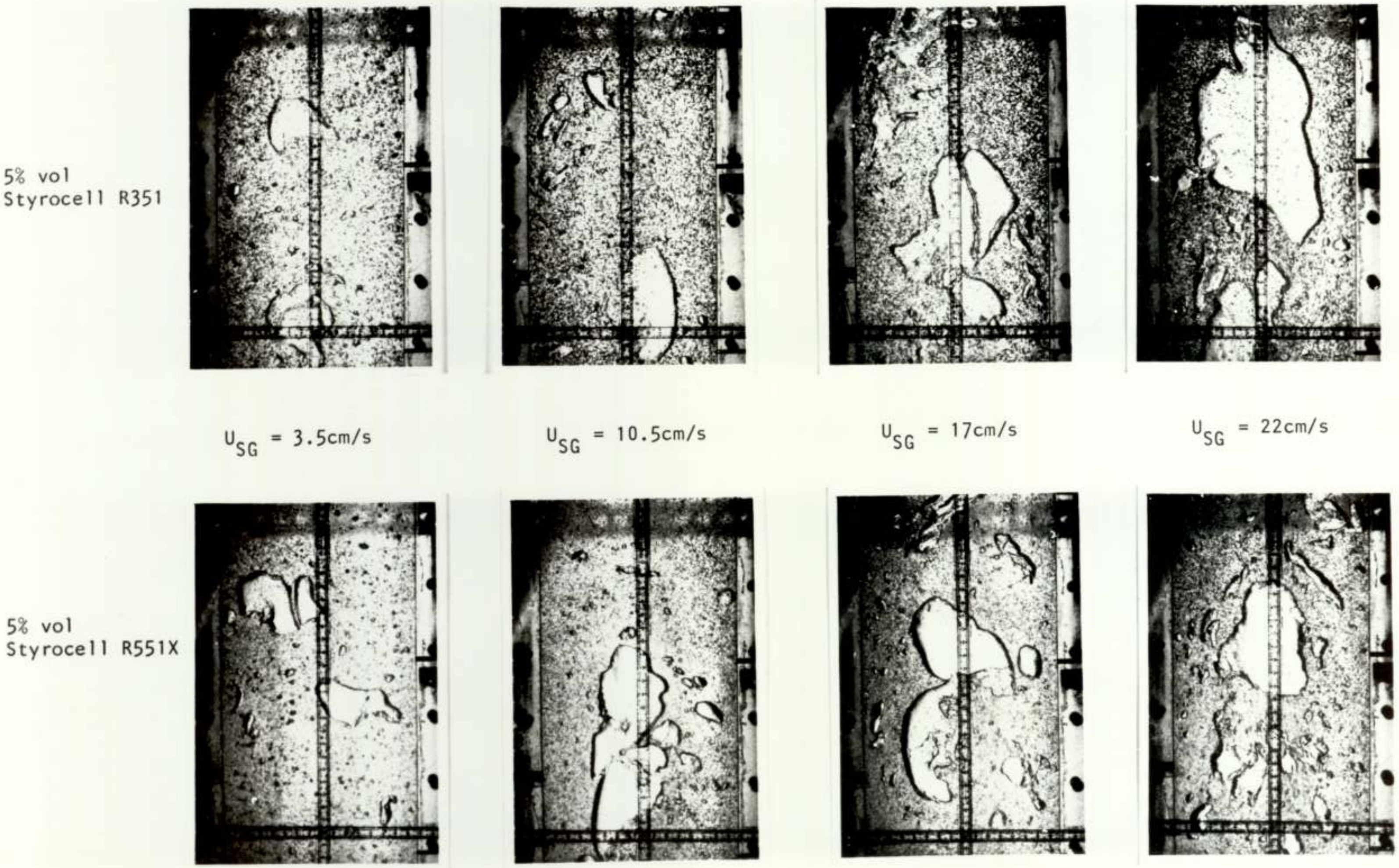


Figure 7.5 (continued)

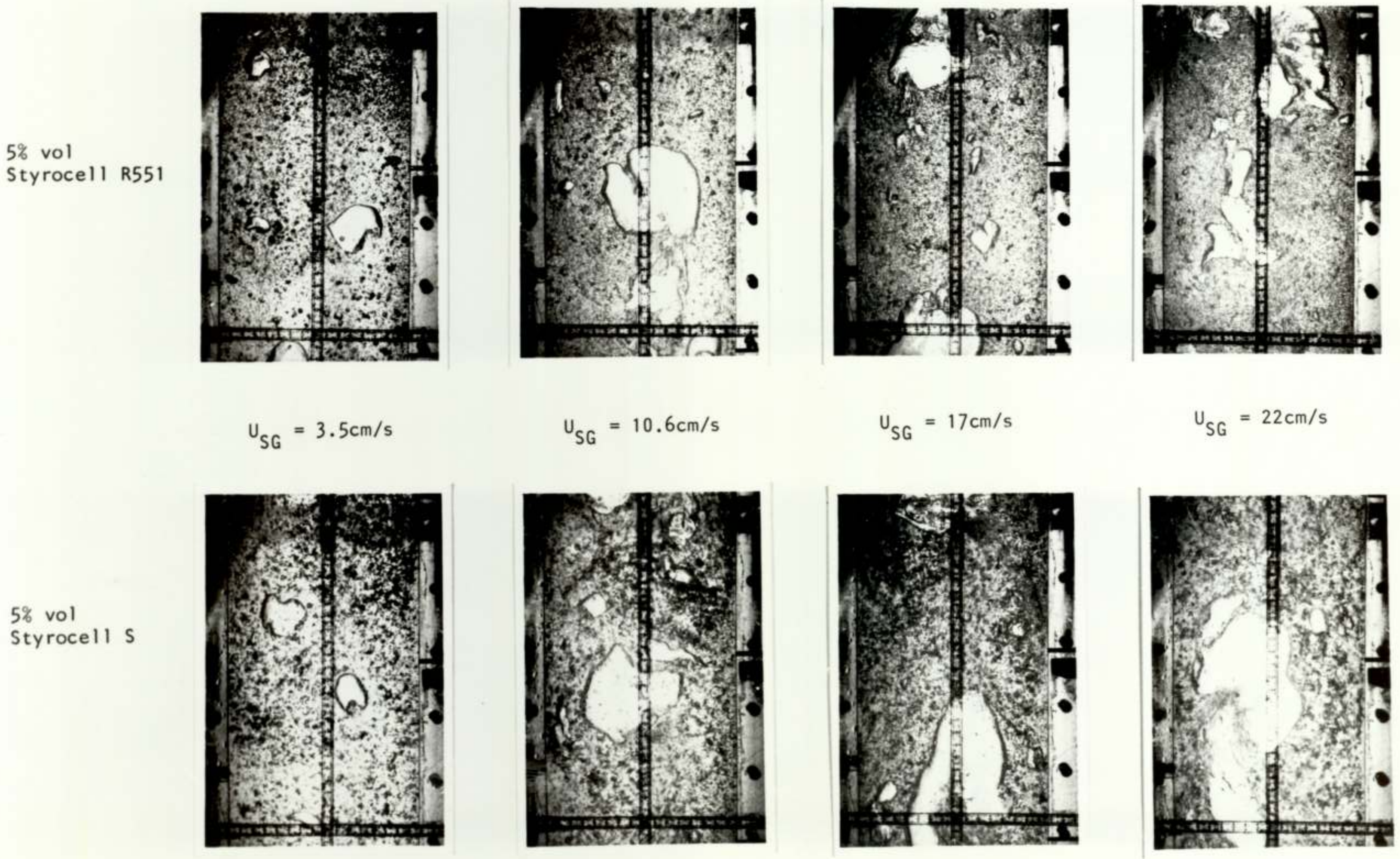


Figure 7.6 Effects of methanol and ethanol on flow regimes and bubble dynamics (three-phase fluidisation)

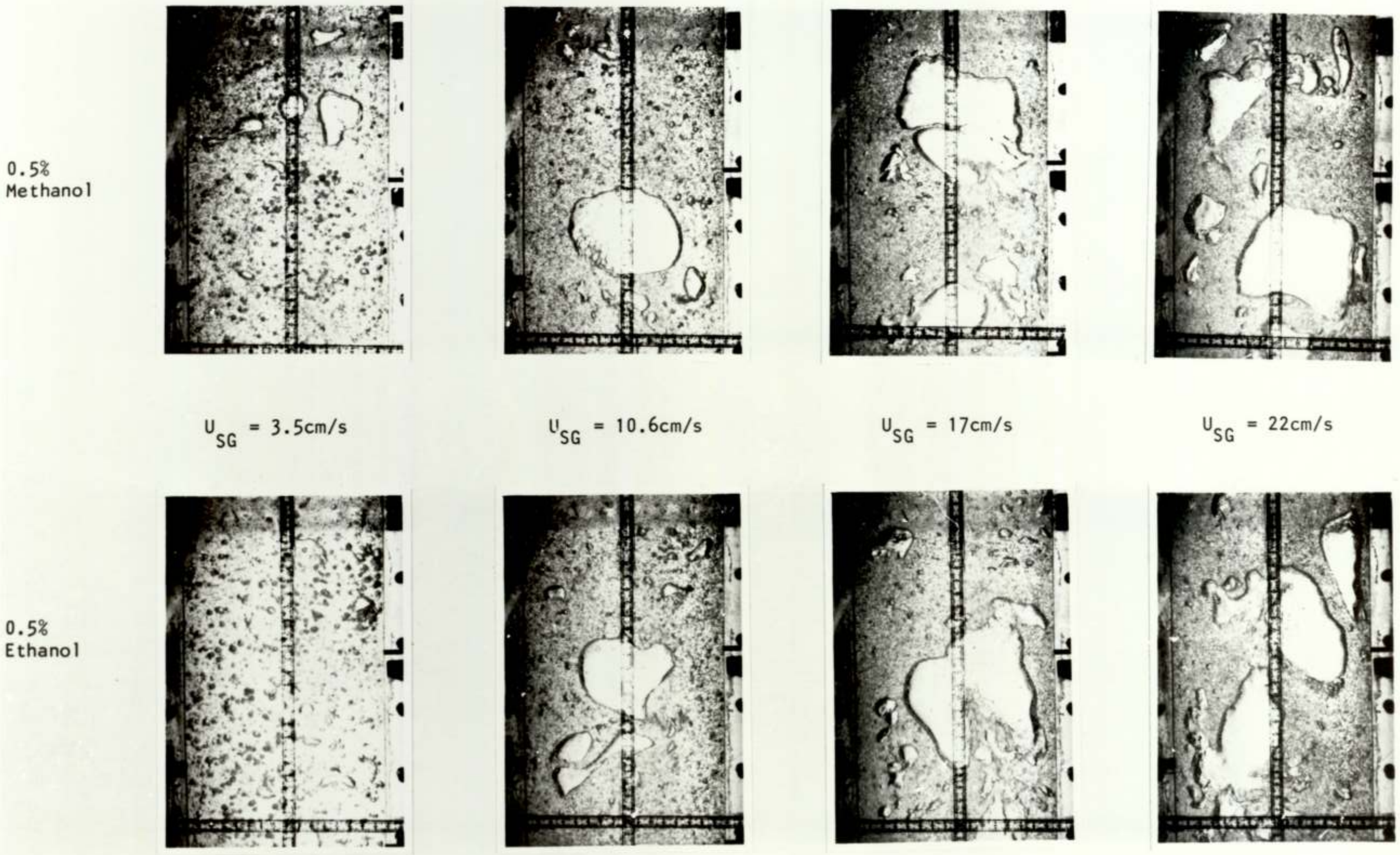


Figure 7.7 Effects of propanone and butanone on flow regimes and bubble dynamics (three-phase fluidisation)

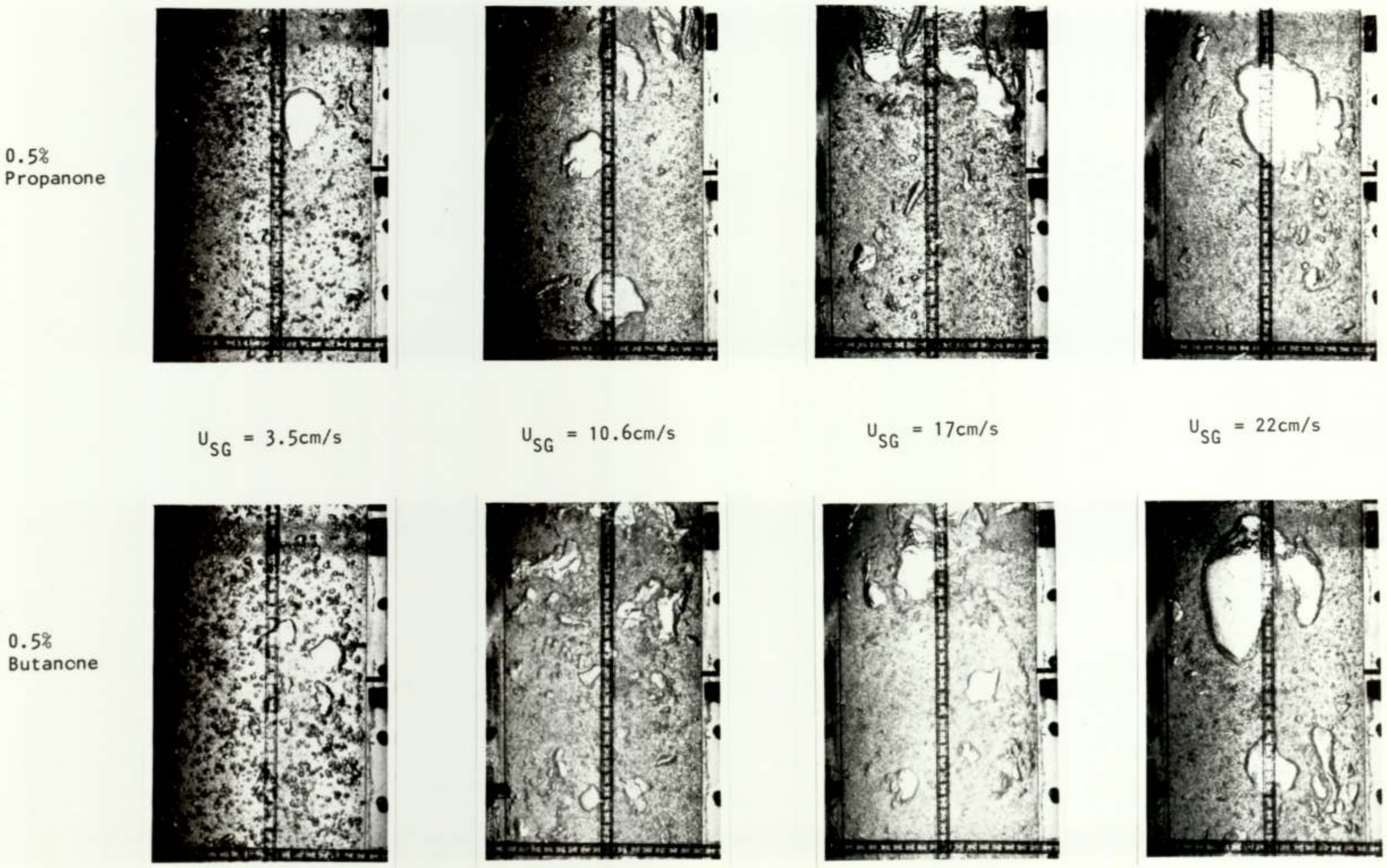


Figure 7.8 Effects of ethyl acetate and N-propyl acetate on flow regimes and bubble dynamics (three-phase fluidisation)

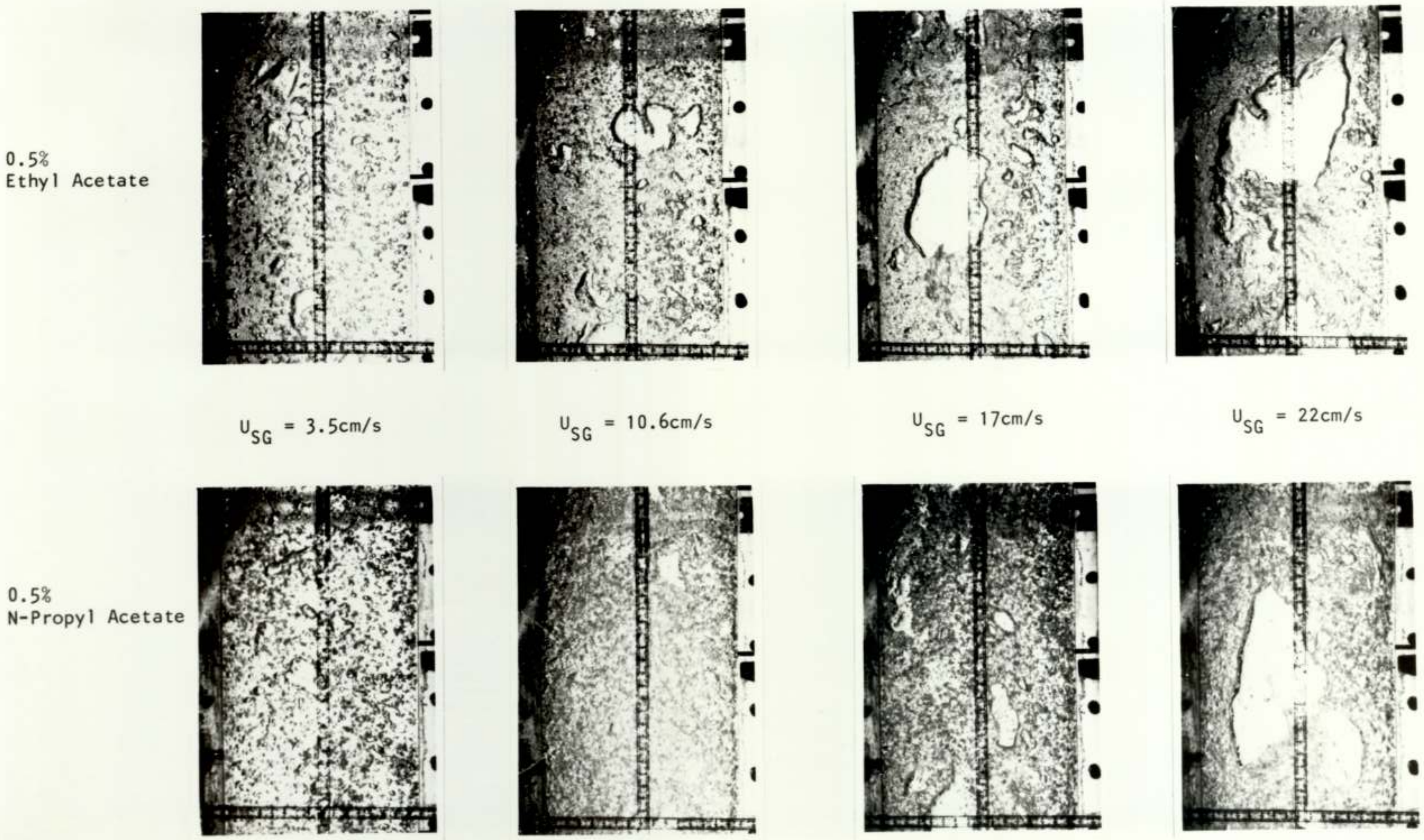
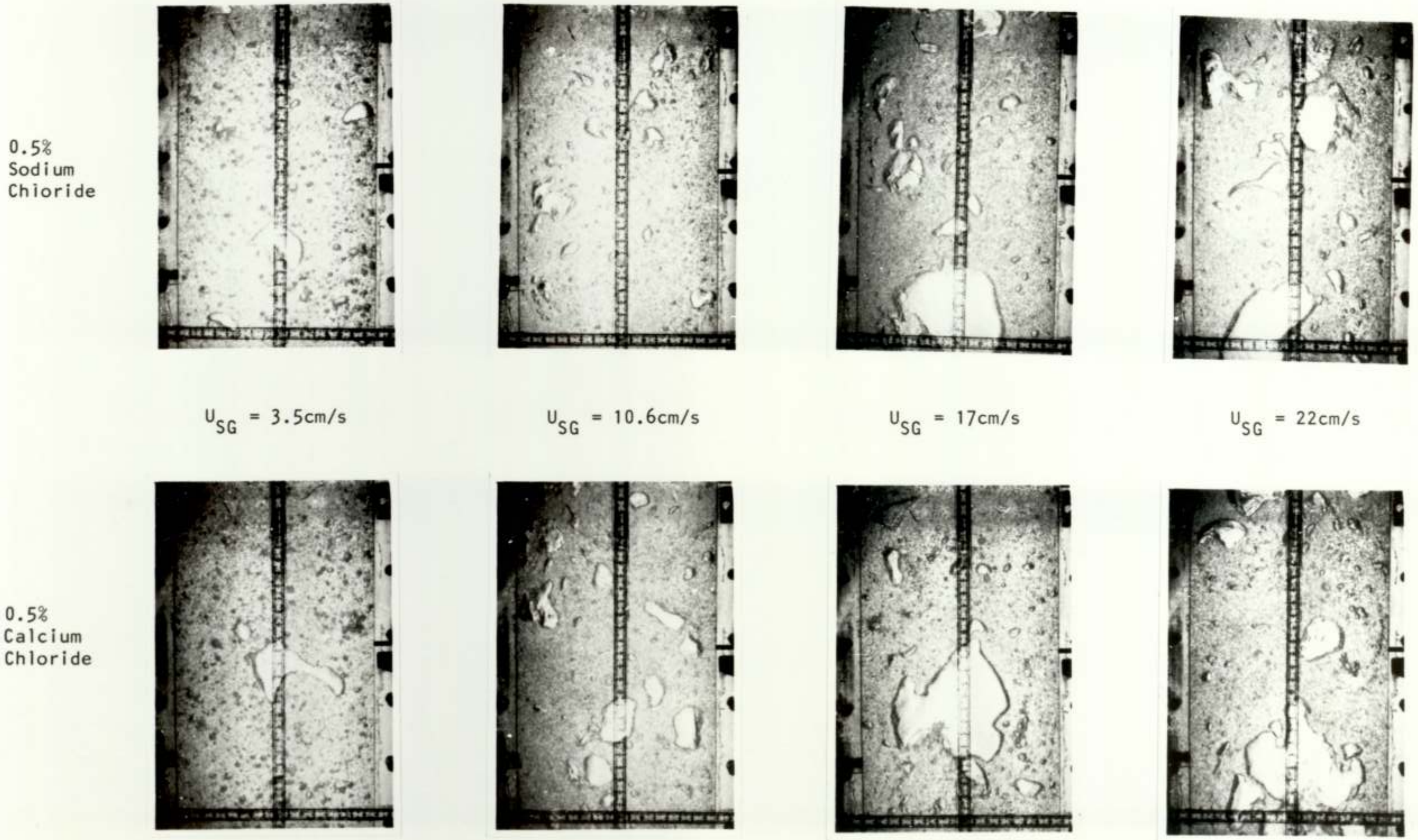


Figure 7.9 Effects of sodium chloride and calcium chloride on flow regimes and bubble dynamics (three-phase fluidisation)



7.5 The Drift-Flux Model and Flow Regimes

Using the drift-flux model described in Chapters 4 and 5, the effects of superficial gas and liquid velocities and solids concentration, size, density and wettability on flow regime transition were enumerated. The results of this analysis are shown in Figures 7.10 to 7.14. Styrocell and Amberlite resin were used as the solid phases with air and water as the fluid phases. The data are presented in Tables F3-F7 of Appendix F.

The transition in the slope of drift flux velocity as a function of gas hold-up corresponds to the change in flow regime. Three flow regimes could thus be distinguished, namely:

- bubbly flow regime,
- churn turbulent flow regime, and
- slug flow regime.

with air-water-Amberlite systems.

In the case of Styrocell particles, true bubbly flow is never achieved and so only the latter two regimes are to be found.

7.6 Discussion

7.6.1 Operating Parameters

In all cases, bubble diameters and, consequently, bubble rise velocities increased with increasing gas flow rate. Similar trends have been observed by most investigators. In the cases reported, relatively low superficial gas velocities were used. In contrast, experiments have been performed in the present study at gas rates as high as about 25 cm/s.

Figure 7.10 Effect of superficial gas and liquid velocities on flow regime transition in the two dimensional bubble column. Three-phase system (Air-Water-Styrocell R551)

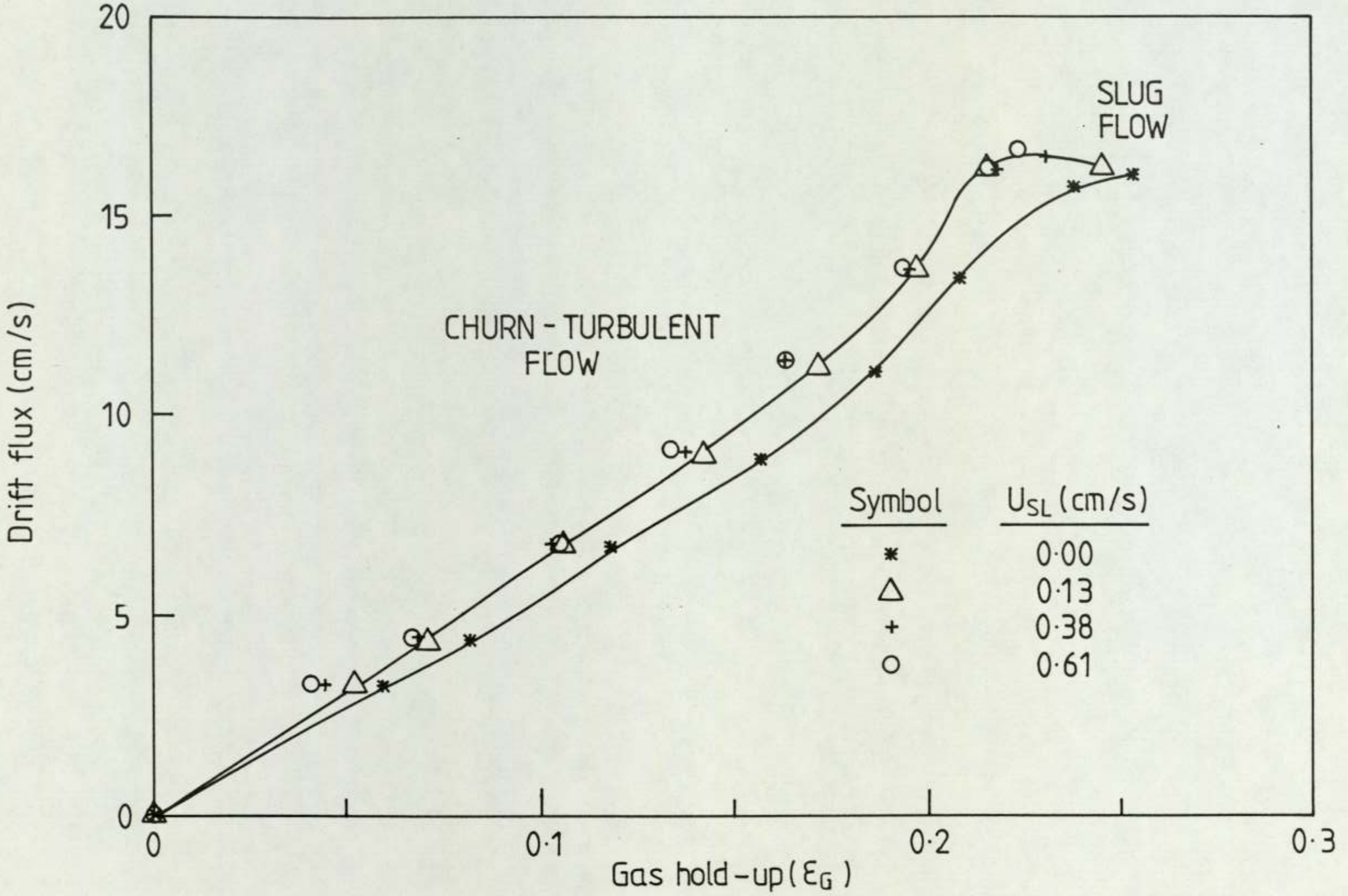


Figure 7.11 Effect of superficial gas and liquid velocities on flow regime transition in the two dimensional bubble column. Three-phase system (Air - Water - Amberlite)

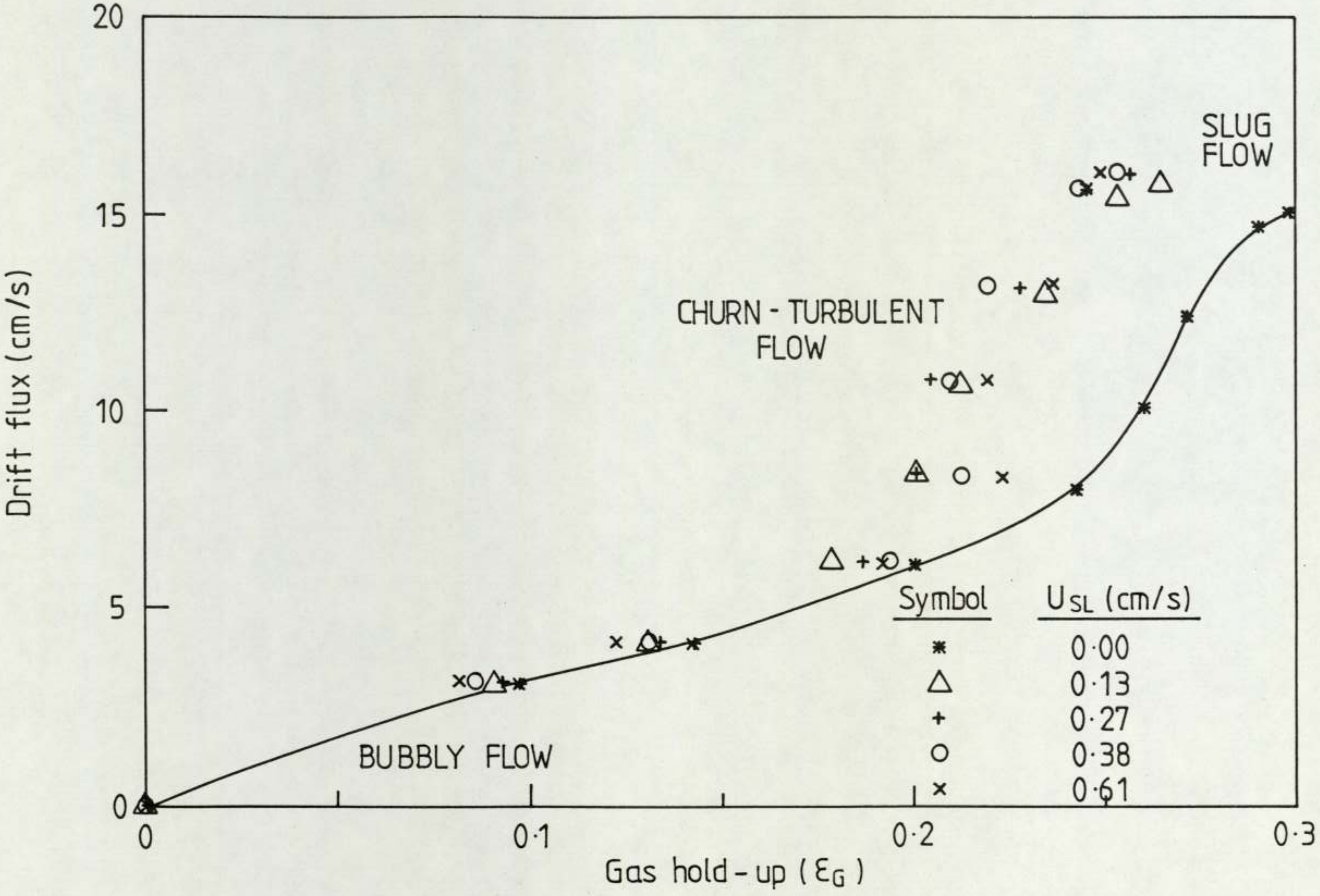


Figure 7.13 Effect of solids concentration on flow regime transition in the two dimensional bubble column

Three-phase system (Air - Water - Amberlite)

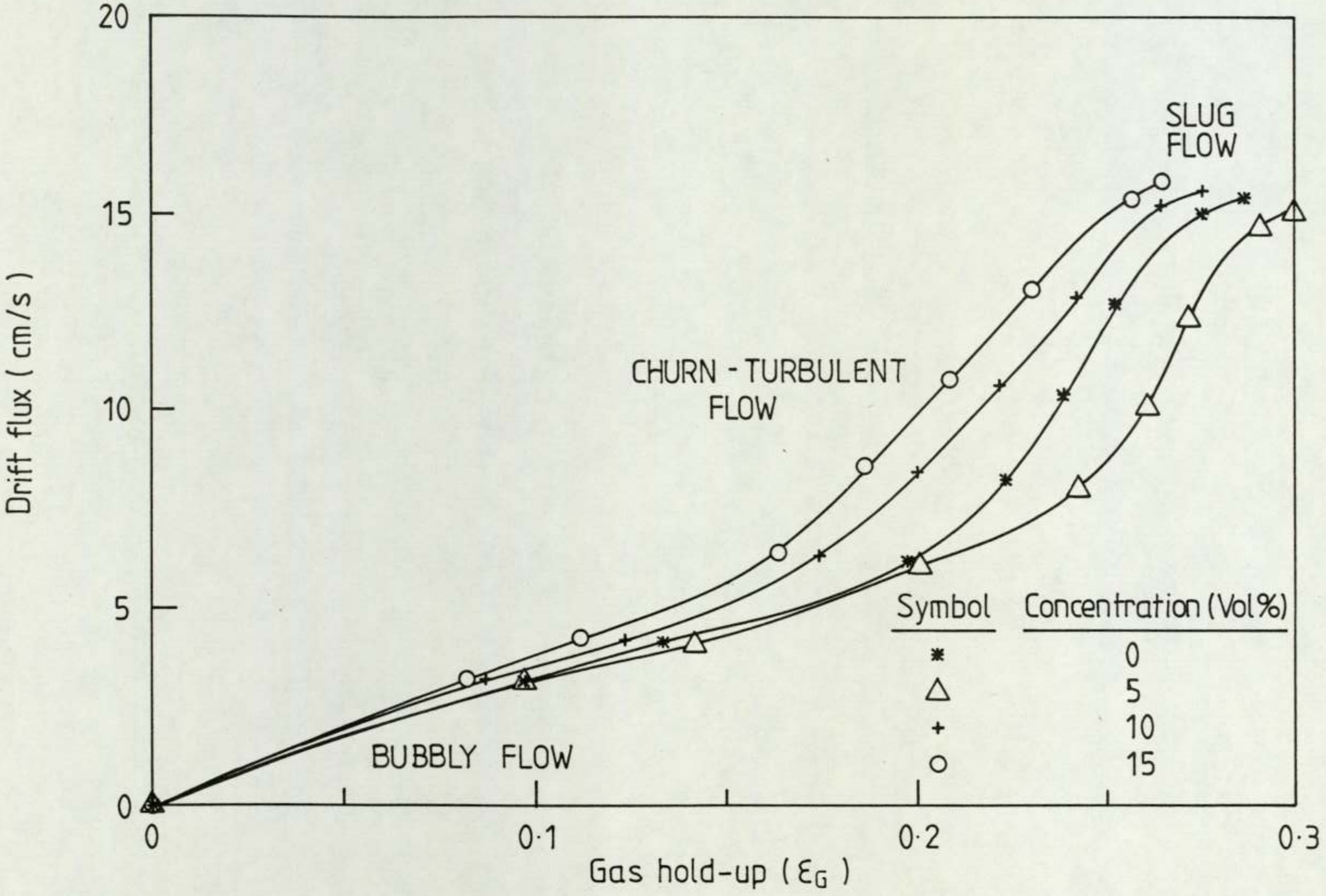
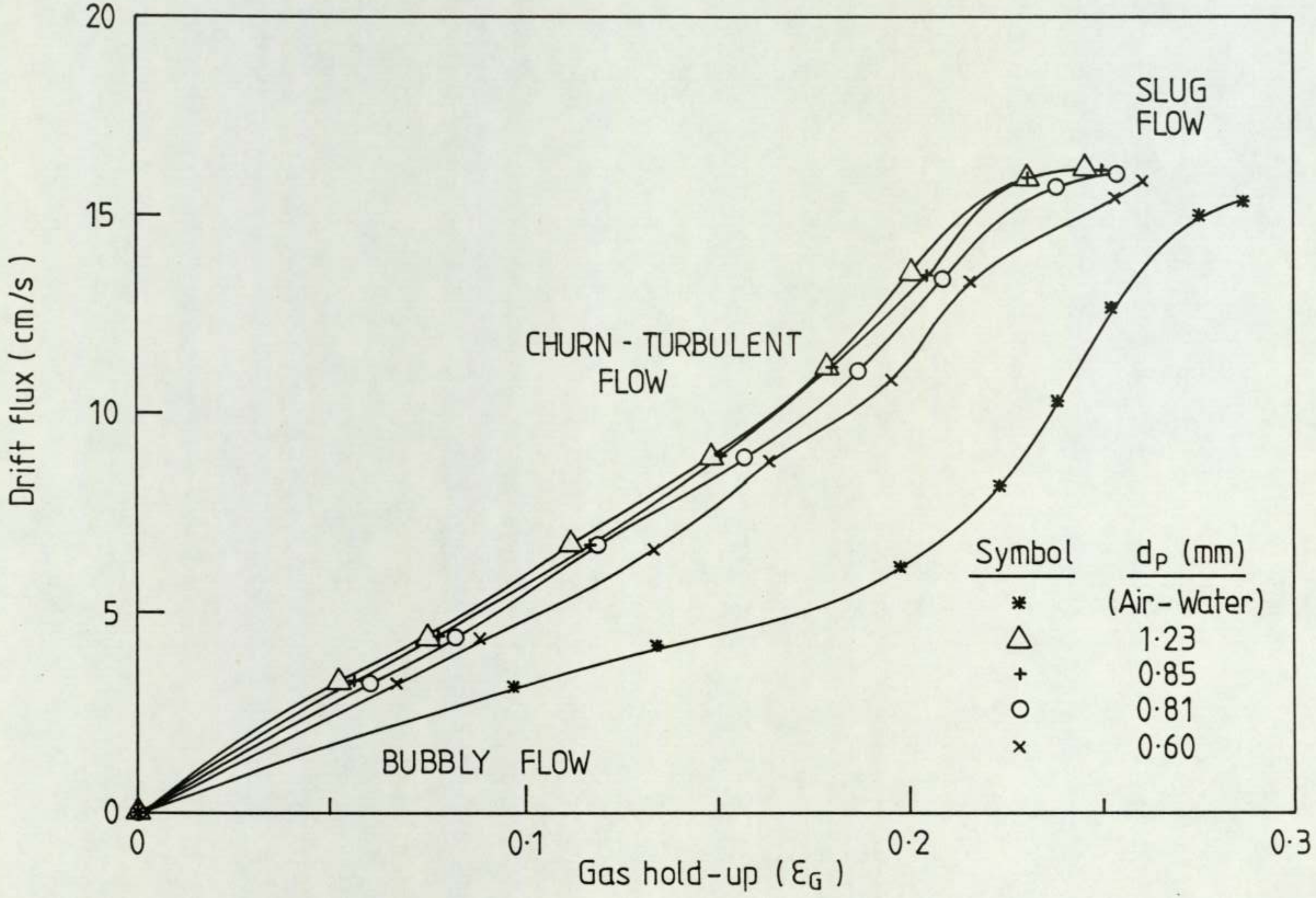


Figure 7.14 Effect of solid size on flow regime transition in the two dimensional bubble column.

Three-phase system (Air-Water-Styrocell)



At low gas rates, in the case of air-water systems containing wettable particles the bubbles are generally small, coalescence is not too marked and the fluidisation is smooth. In contrast, at higher gas rates, considerable coalescence resulting in much larger bubbles prevails. The beds are then far more turbulent. True bubbly flow is never observed with beds containing non-wettable solids.

Liquid velocity was found to have only a relatively minor effect on the bubble size and hence rise velocity. The effect fell off at higher liquid velocities.

7.6.2 Solid Properties

Bubble size was increased by the presence of non-wettable solids and to a much lesser extent by wettable particles. Increasing the solids concentration, size and density lead to no significant change in the bubble size. The effect of wettability was more pronounced. At low gas rates, the nature of the solid influences the size of bubbles in the bed. At high gas rates, bubble characteristics and flow regimes are independent of the size and density of the particles. This is of importance since it indicates that in this regime of potential industrial importance, the gas-liquid interfacial area is unlikely to be influenced significantly by the characteristics of the solids.

When water, which is highly polar, is brought into contact with solid surfaces (eg. polystyrene), the water cannot "adhere" to them. The tension at the interfaces between solid particles and water will thus be very high. Because the attraction between water-water molecules is much higher than between water and solid particles, the solid will tend to migrate towards bubble interfaces. As a result the concentration of solids at the bubble interface tends to be greater than in the bulk. Also, the motion of solids in the bulk will lead to some lowering of the resistance to bubble coalescence. Therefore, the tendency for bubbles to coalesce will be higher even at low gas velocities in the presence of small and large particles.

However, wettable solids having strong polar groups will remain in the bulk liquid phase. This is due to their compatibility with the polar water molecules. This thus reduces the bubble rise velocity and bubble coalescence.

7.6.3 Liquid Properties

The presence of solids has an overriding effect on the behaviour of the bed, so much so that the effect of soluble additives is largely destroyed. Thus, the nature of the liquid phase has little effect on bubble dynamics and flow regimes.

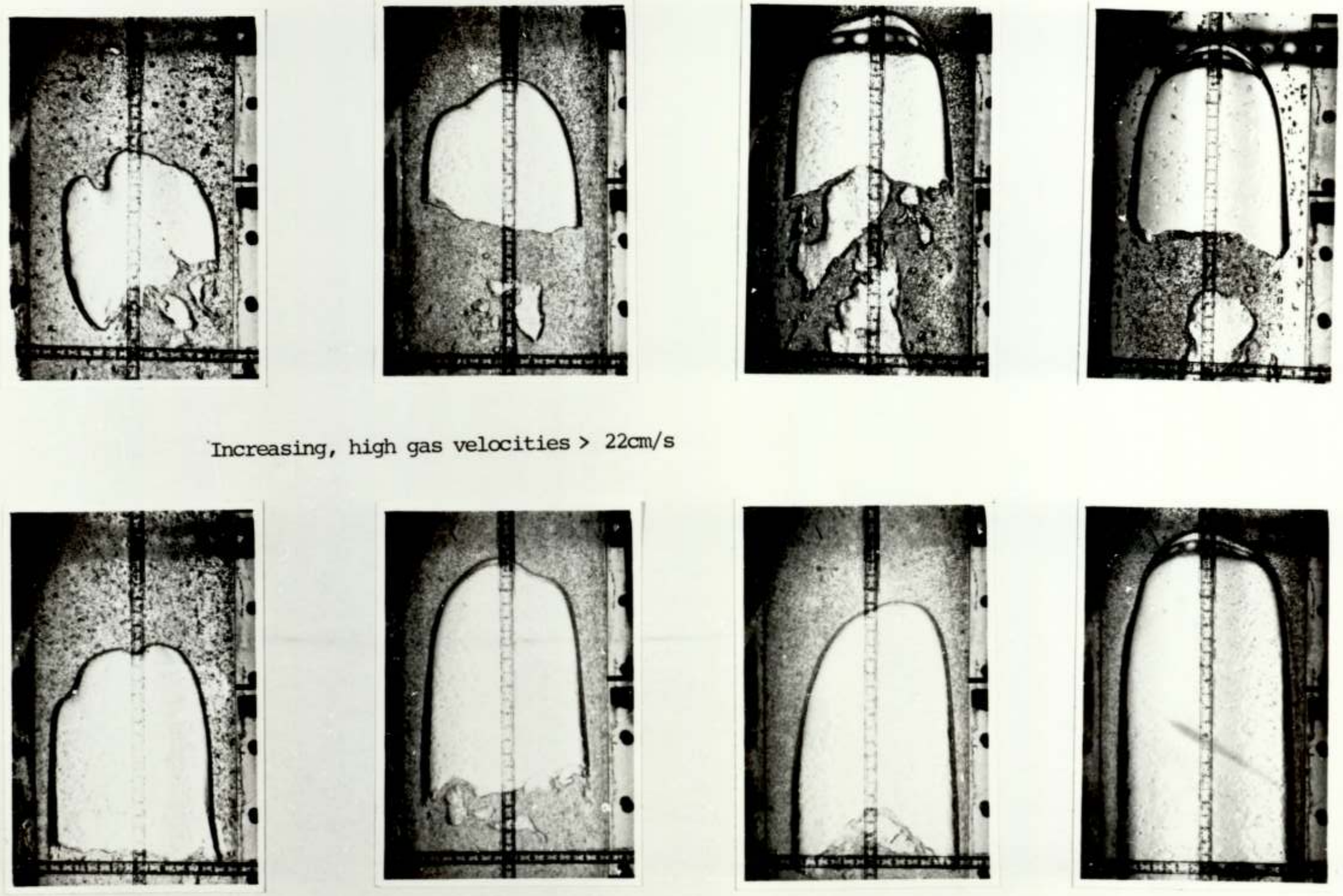
7.6.4 Flow Regime Transition

The visual, photographic and qualitative studies have revealed three flow regimes:

- the bubbly flow regime,
- the churn-turbulent flow regime, and
- the slug flow regime.

In the bubbly flow regime, bubbles are relatively uniform though there is a tendency for bubble clusters to form. In the churn-turbulent regime, bubble coalescence occurs; this results in large bubbles existing simultaneously with small ones. The slug flow regime is characterised by slug formation in the column (see Figure 7.15). This usually occurs at very high gas velocities. The large slugs almost span the entire section of the column.

Figure 7.15 Slugs in three-phase fluidised beds (air-water-styrocell R551)



Increasing, high gas velocities > 22cm/s

7.6.5 Bubble Coalescence and Break-up

From this study, the behaviour of three-phase fluidised beds can be divided into three different regions:

- the formation zone,
- zone of active bubble coalescence and break-up,
- the break-up and ejection zone.

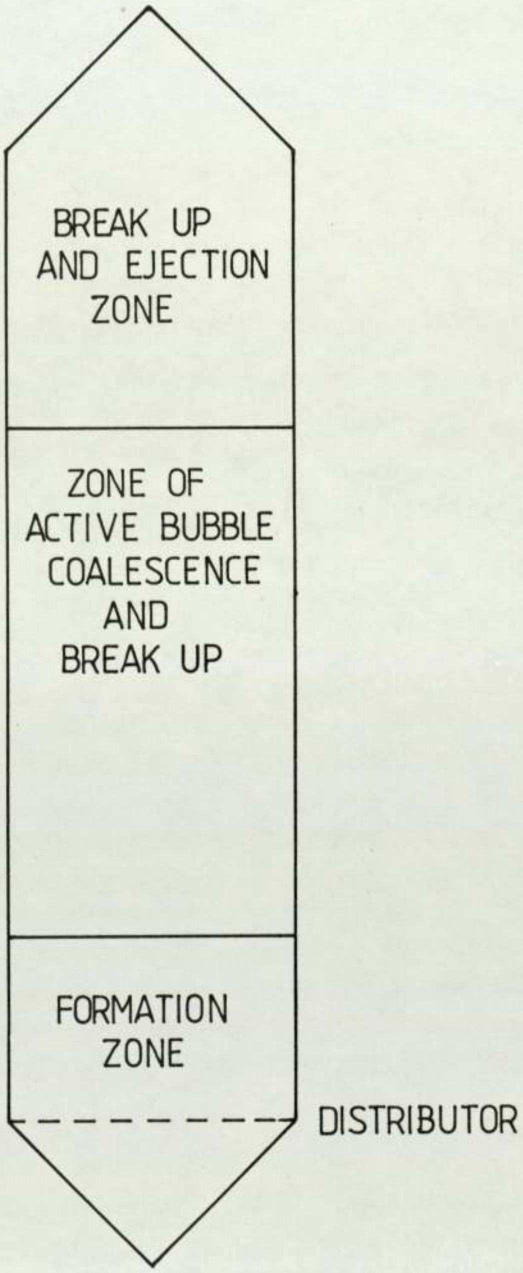
In the formation zone, (see Figure 7.16) just above the gas distributor, bubbles are formed. This leads to the dynamic coalescence and break-up of bubbles which covers the major part of the column; it is this zone that has been studied by the author in detail. Bubble break-up and subsequent ejection finally ends the journey of the bubbles through the column.

The data show that, under the conditions used in this study, the size of bubbles is not highly sensitive to the nature of the solids. This would tend to eliminate the bubble break-up mechanism proposed by Lee, Sherrard and Buckley (1974) involving the perforation of bubbles by the passage of particles through them. On the other hand, the photographic evidence to a certain extent supports Ostergaard's (1974) contention that Taylor instability may be responsible for bubble break-up. This has been clearly shown by "fingering" and roof instability of bubbles (eg. Figure 7.5).

7.7 **Conclusions**

The characteristics of bubble swarms, their coalescence, break-up and flow behaviour have been studied over a wide range of experimental conditions. The following conclusions are drawn:

Figure 7.16 Regions in three-phase fluidised bed.



1 Bubble dynamic and flow regimes are primarily the function of gas velocity, the liquid velocity having little or no effect.

2 Liquid properties have only limited influence on these properties in the presence of solids.

3 At high gas rates, bubble dynamics and flow regimes are not greatly affected by solids properties.

4 Three regimes of operation can be readily distinguished.

5 Future analysis of bubble behaviour must embrace consideration of bubble formation, coalescence and break-up in the bed, and break-up at the surface of the bed.

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CHAPTER 8

INTRODUCTORY STUDIES ON MIXING IN TWO- AND THREE-PHASE FLUIDISATION

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8.1 Introduction

Mixing, a widely practiced operation occurs whenever fluids are moved in the conduits and vessels of laboratory and industrial-processing equipment. It usually results in the dispersion of one component in another and also provides a means for the promotion of heat transfer, mass transfer, solid suspension and reaction.

8.1.1 The Concept of Mixing

Mixing is applied to operations which tend to reduce non-uniformities or gradients in composition, properties, or temperature of material in bulk. It is accomplished by movement of material between various parts of the whole mass. For fluids, the movement occurs by a combination of the following mechanisms:

- bulk flow in both laminar and turbulent regimes
- eddy and molecular diffusion.

The mixing process is conventionally divided into two parts (Shah et al, 1978):

macromixing - giving information about the retention time of the elementary volumes,
micromixing - describing the communication between the elementary volumes.

8.1.2 Applications

The performance of process equipment depends not only on the relevant intrinsic kinetics but also on the physical processes occurring in such columns. These physical processes such as interphase, interparticle and intraparticle mass and heat transfer depend very significantly upon the mixing characteristics of the various phases involved.

Mixing enhances the movement or transfer of materials to or from the surface of particles or phases as in chemical reactions, dissolution, leaching, gas absorption and liquid-liquid extraction processes. It also enables the production of more uniform mixture of components.

Mixing can also have significant effects on reaction rates and product selectivity. This is the case with backmixing in the various phases in a multiphase reactor or column. For isothermal systems, backmixing has a detrimental effect on the reaction rate in many cases because of the reduction in the effective concentration of the species leading to a reduction in the effective driving force. However, in non-isothermal systems, the backmixing of heat with the fluid compensates for the reduction in the reactant concentration due to the effect of temperature on the reaction rate constant.

8.2 Literature Review

The importance of mixing to the chemical process industries has attracted many investigators. Pertinent studies on mixing in gas-liquid and gas-liquid-solid systems are reviewed in this section.

8.2.1 Mixing in Gas-Liquid Dispersion

Early reviews on this subject are given by Mashelkar (1970), Bischoff (1966), Pavlica and Olson (1970). A more recent review is that of Shah et al (1978). Some of the previous studies are summarized in Table 8.1.

TABLE 8.1: SOME PREVIOUS STUDIES ON LIQUID PHASE MIXING

REFERENCE	SYSTEM	FLOW RATE (cm/s)	TRACER SYSTEM	COLUMN CHARACTERISTICS (cm)	DISPERSION STUDIED
Ohki & Inoue (1970)	Air-Water	U_{SL} = Batch U_{SG} = 0-25	4N aqueous KCl pulse Pt electrical conductivity cell	D_c = 4, 8, 16 L = 20, 25, 30	Axial
Eissa et al (1971)	Air-Water Cocurrent	U_{SL} = 0.75-1.3 U_{SG} = 1.4-7.8	Salt solution Steady state point source Conductivity meter	D_c = 5 L = 42	Axial & Radial
Kato & Nishiwaki (1972)	Air-Water Cocurrent	U_{SL} = 0.5-1.5 U_{SG} = 0-30	1-1.5N KCl pulse Pt Electrical conductivity cell	D_c= 6.6, 12.2, 21.4 L = 200, 201, 405	Axial
Towell & Ackerman (1972)	Air-Water Cocurrent & Counter current	U_{SL} = 0.24-1.4 U_{SG} = 0-15.2	Methylene blue dye Step and pulse spectrophotometer	D_c = 40.6, 106.7, L = 274.3, 518.2	Axial
Deckwer et al (1974)	Air-Water	U_{SL} = Batch U_{SG} = 0-5	Dye, heat, electrolytes Dirac pulse	D_c = 15, 20 L = 440, 723	Axial
Hikita & Kikukawa (1974)	Air-Water Air-Aq Me OH solutions	U_{SL} = Batch U_{SG} = 4.3-33.8	Aq KCl pulse Pt electrical conductivity cell	D_c = 10, 19 L = 150, 240	Axial
Cova (1974)	N_2 - Water N_2 - Acetone, CCl_4	U_{SL} = 0.2-0.87 U_{SG} = 0.51-17.3	Heat, steady state Thermocouples	D_c = 1.91, 4.57 L = 116, 122	Axial

Eissa & Schugerl (1975)	Air-Water Cocurrent	U_{SL} = 0.35-1.4 U_{SG} = 0.2-6	NaCl solution Steady state injection Conductivity meter	D_c = 15.9 L = 390	Axial
Alexander & Shah (1976)	Air-aq glycerol solns	U_{SL} = Batch U_{SG} = 0-7	$H_2 SO_4$ pulse Electrical conductivity	D_c = 603, 15.24 7.62 x 22.9 rectangular L = 124, 308, 112	Axial
Stiegel & Shah (1977)	Air-Water	Mass Velocities U_{SL} = 7.32-73.2 (kg/s m^2) U_{SG} = 0-0.47 (kg/s m^2)	$H_2 SO_4$, pulse Electrical conductivity	16.8 x 2.06 x 122	Axial
Smith et al (1977)	Air-Water	U_{SL} = 0.5-2.5 U_{SG} = 0.5-12.0	KCl solns Steady state Stimulus-response technique	D_c = 7.6, 15.2, 30.5, 61 L = 300	Axial & Radial
Field & Davidson (1980)	Air-Water	U_{SL} = 3-3.5 U_{SG} = 4-5.5	Radioactive tracer Detectors	D_c = 320 L = 1890	Axial
Kelkar et al (1983a, b)	Air-Alcohol solns - Electrolytes	U_{SL} = 0-15 U_{SG} = 1-30	Heat, steady state Thermocouples	D_c = 15.4, 30 L = 335, 244	Axial
This study	Air-Water Air-Water (solids)	U_{SL} = Batch U_{SG} = 0-35	Methylene blue dye Stimulus-response technique	D_c = 15.2 L = 350	Axial

As long as the sparged vessel has a sufficiently large length to diameter ratio, the gas phase is usually assumed to move in plug flow. Earlier studies indicate that the mixing in the liquid phase depends on the column diameter, gas velocity and nature of the gas distributor. Although there are some discrepancies in the literature, it appears that the axial dispersion coefficient in a vertical column is essentially independent of liquid velocity and the liquid properties.

8.2.2 Mixing in Gas-Liquid-Solid Fluidisation

Liquid Mixing

The effects of suspended solid particles on the liquid phase mixing have been studied by Schugerl (1967), Vail et al (1968), Ostergaard and Michelsen (1969), Michelsen and Ostergaard (1970) and Kato et al (1972). The data show that, in a three-phase column, the liquid phase mixing still increases with gas rate. In contrast to what happens in a gas-liquid bubble column, the liquid phase mixing coefficient is dependent on the liquid velocity. Vail et al showed that the effect of liquid velocity diminishes with increasing liquid flow rates. The nature of the effect is also dependent on the gas rate and solids particle size. Vail et al also showed that the presence of solids reduces the liquid phase mixing. El-Temtamy et al (1979a) measured the axial liquid-phase dispersion coefficient in three-phase fluidised beds using the steady-state tracer injection technique as employed by Vail et al.

Kim and Kim (1983) examined the effects of liquid viscosity and surface tension on the axial dispersion coefficient of the liquid in a three-phase fluidised bed. The axial dispersion coefficient slightly increased with the liquid viscosity for a bed of large particles, ie. 6mm glass beads, but it decreased with viscosity at high gas flow rates in a bed of small particles, ie. 1.6 and 3mm particles. The axial dispersion coefficient increases with an increase in the surface tension of the liquid.

An experimental study was conducted on radial liquid mixing for cocurrent three-phase fluidisation by El-Temtamy et al (1979b). The radial dispersion coefficient was found to be much less than the axial dispersion coefficient. Radial liquid dispersion coefficients were also measured by Muroyama et al (1981) who obtained the coefficient from the effective thermal conductivity data in heat transfer experiments. The Peclet number for radial liquid mixing appreciably decreases with an increase of the gas flow rate.

A unified correlation of the axial liquid mixing in gas-liquid two-phase columns and three-phase fluidised beds has been proposed by Joshi (1980), who developed an energy balance model for mixing. The correlation is:

$$D_L = 0.29\,(U_{SL} + U_c)\,D_c \qquad \text{........ 8.1}$$

Where D_L is in m^2/s, D_c in m, U_{SL} and U_c in m/s.

Kara et al (1982) studied the mixing properties in a stainless steel column with three phases (air, water and coal and dried mineral ash) in cocurrent upflow. The effect of solids, slurry velocity and gas velocity on axial heat dispersion coefficients were determined. The heat dispersion model represented the observed axial temperature profiles. The heat dispersion coefficients were found to increase with gas velocity, be independent of solids concentration, and decrease with both increasing slurry velocity and increasing solid size. The dispersion coefficients for both heat and mass were judged to be equivalent. The experimental data were correlated empirically and the results were compared with other published data.

Kelkar et al (1984) also studied mixing properties with three-phases flowing cocurrently upward in a column. The effects of fluid velocities, solid properties, interfacial tension and slurry viscosity on axial dispersion coefficients were examined. The axial dispersion coefficients were not significantly affected by particle size, solids concentration and slurry velocities. However, the presence of solids was found to enhance the coalescence tendencies in the liquid phase and hence led to a decrease in the axial dispersion coefficient compared with values for air-water systems. Addition of surfactants decreased the axial dispersion coefficient. An increase in liquid viscosity was found to reduce the dispersion coefficient.

Gas Mixing

The gas phase residence time distribution (RTD) in a three-phase fluidised bed has been measured by Schugerl (1967), Michelsen and Ostergaard (1970), and Ostergaard and Michelsen (1969). Schugerl reported that, just as in a bubble column, the intensity of mixing in both the gas and liquid phases decreases from the top to the bottom of the column. At low liquid velocities, the gas phase Peclet number increases with the gas rate, but at high liquid velocities, the Peclet number shows a maximum with respect to gas rate. Ostergaard and Michelsen used argon-41 as a gaseous tracer and measured the responses of the tracer pulse in the gas phase indirectly from the outside of the column. The values of HMUg (height of mixing unit of the gas) were found to be comparable to those in gas-liquid bubbling columns.

Gas phase mixing in a cocurrent slurry flow system containing water, air and 250 µm quartz particles in suspension was investigated by Afshar and Schugerl (1968). The degree of gas mixing in a bubble column slurry reactor has been assessed using the tanks-in-series mixing model (Kolbel et al, 1962).

Solids Mixing

Cova (1966) has measured the steady state concentration distribution of solids in a three-phase reactor, and Farkas and Leblond (1969) used their data to calculate the mixing coefficient of solids. The behaviour of suspended solid particles and their mixing has also been studied by Imafuku et al (1968) and Kato et al (1972). Both batch and flow systems were examined; various types of solid particles including glass spheres, copper powder, ion exchange resin were examined. The column diameters ranged from 0.05 to 0.22m, and the effects of solids concentration up to 20% was studied. They concluded that, for small particles in small diameter columns, the longitudinal dispersion coefficient for the solids is the same as that of the liquid.

A qualitative investigation of solids mixing in cocurrent three-phase fluidised beds containing a binary mixture of glass beads was conducted by Fan et al (1982). They noted that the complete mixing state occurs largely in the coalesced bubble regime and the slugging regime.

Quantitative studies of solid concentration profiles and of the degree of mixing and segregation of a binary mixture of particles were recently conducted by Fan et al (1984). They found that the degree of particle segregation decreases sharply with an increase in the gas velocity. The degree of solids mixing was found to be lower than that of liquid mixing as observed by Al-Dibouni and Garside (1979).

8.3 Strategy for Mixing Studies

8.3.1 Measurement of Residence Time Distribution (RTD)

The mixing characteristics of various phases can be evaluated from the residence time distribution of a tracer injected into the phase of interest. These tracer techniques usually involve the injection of a tracer at one or more locations in the system and detection of its concentration as a function of time at one or more downstream positions. This is the "stimulus-response" technique. Various types of tracer input, such as a step or pulse, may be used.

The selection of the proper tracer for a given system is important and must be made so that the RTD is characteristic of the flowing phase and not influenced by the tracer itself. The basic requirements for a satisfactory tracer experiment are:

- The tracer should be miscible in and have physical properties similar to the fluid phase of interest, and it should not be transferrable to the other phase or phases in the system.
- The tracer should be accurately detectable at low concentrations so that only a small quantity need be injected into the system, thus minimizing disturbances in the established flow patterns.
- The tracer should be non-reacting so that the analysis of the RTD is kept simple.
- The tracer detection device should cause as little disturbance as possible in the flow patterns.
- When using tracers in fast moving phases, the detection and recording equipment must have a high sensitivity and short response time.

The nature of the tracer selected usually dictates the detection system. For the liquid phase, quite often the tracer (eg. sodium chloride) is such that the detection probe can be inserted directly into the column and tracer concentration monitored continuously.

For the solid phase, a magnetic tracer can sometimes be used. The concentration of a solid phase tracer can also be measured by a capacitance probe if the tracer material has a different dielectric constant than the solid phase. In the case of both solid and gas phases, radioactive tracers have their attraction. Since the tracer detection devices can be placed externally, no disturbances in the established flow patterns are created.

8.3.2 Modelling

Once the data for the RTD are obtained, the mixing characteristics for each phase can be quantitatively evaluated by fitting appropriate models to these data. The models for mixing can be largely divided into two categories:

- differential models, and
- stagewise models.

The differential models result in differential equations, whereas under steady state conditions, stagewise models result in algebraic equations.

Several models have been used to characterise mixing effects and non-ideal flow patterns in process vessels. Among these, the axially dispersed plug-flow model is perhaps the most widely applied.

The Axially Dispersed Plug-Flow Model

This model characterizes mixing by a simple, one-dimensional dispersion equation based on an analogy with Fick's law of diffusion. The constant of proportionality in this equation is known as the axial dispersion coefficient.

Derivation

Levenspiel and Bischoff, when reviewing the patterns of flow in chemical process vessels, gave the following differential equation for the general dispersion model including chemical reaction and source terms:

$$\frac{\delta C}{\delta t} + U \nabla C = \nabla \cdot (D \cdot \nabla C) + S + r_c \qquad \text{........ 8.2}$$

Because of the difficulties of specifying velocity profiles and limitation in experimental methods, the above equation is often simplified by assuming that:

- bulk flow occurs in the axial direction only,
- the dispersion coefficient is independent of position,
- fluid flows at the mean velocity (plug-flow), and
- there is no variation in properties in the radial direction.

The equation can then be written as:

$$\frac{\delta C}{\delta t} + U \frac{\delta C}{\delta X} = D_1 \frac{\delta^2 C}{\delta X^2} + S + r_c \qquad \text{....... 8.3}$$

The usual method of estimating dispersion coefficients is to inject a suitable tracer at a point or plane in the system and monitor the changes in tracer concentration at one or more points. The dispersion coefficient is then found from an analysis of the concentration data. In such stimulus-response experiments, chemical reactions do not occur and $r_c = O$. When the injection point for the tracer and the measuring points are sufficiently far apart and when there is no flow of liquid through the column, equation 8.3 reduces to:

Leads to:

$$\ln \frac{C}{Co} = \frac{-U}{D_1} X \qquad \text{........ 8.7}$$

$$\text{Where } U = \frac{U_{SL}}{1 - \varepsilon_G} \qquad \text{........ 8.8}$$

If U/D_1 is constant, a plot of ln (C/Co) against X gives a straight line of slope - U/D_1 from which D_1 may be calculated.

8.4 Experimental Procedure

The experimental procedure consisted of three parts:

- the setting of flow conditions,
- injection of tracer, and
- concentration measurement.

The unsteady-state tracer technique was used to evaluate liquid phase axial mixing in the three-dimensional bubble column. Air and water were used as the gas and liquid phases respectively.

For each experiment, after setting the appropriate gas flow rate, an automatic timer was switched on simultaneously with the introduction of liquid tracer (methylene blue solution) at the top of the column. Samples were then taken near the bottom of the column and the times at which samples were taken were recorded. A spectrophotometer (see Figure 8.1) was then used to measure the dye concentration in the solution. The calibration graph of the spectrophotometer is given in Figure G1 (Appendix G). Since it has been shown that the liquid flow rate has an insignificant effect on dispersion coefficient, all the data reported in this study were obtained from a batch system.

The whole procedure was repeated with a three-phase system using Styrocell particles as the solid phase.

8.5 Experimental Results

The axially dispersed plug flow model was used to fit the experimental data. A graphical solution of the model, taking X/L as a parameter (equation 8.5), is presented in Figure 8.2.

For each gas flow rate, the axial dispersion coefficient was calculated for the two- and three-phase systems. Tables G1 to G14 in Appendix G give the detailed results.

8.5.1 Two-Phase Axial Dispersion Coefficient

The effect of superficial gas velocity on the liquid-phase axial dispersion coefficient is shown in Figure 8.3. The dispersion coefficient increased with increasing gas velocity reaching a maximum at a superficial gas velocity of 15 cm/s. At the higher gas rates, the coefficient increased further.

8.5.2 Three-Phase Axial Dispersion Coefficient

Also shown in Figure 8.3 is the effect of solids on the liquid-phase dispersion coefficient. A decrease will be observed although the effect is not very significant especially at the mid-range of gas velocities.

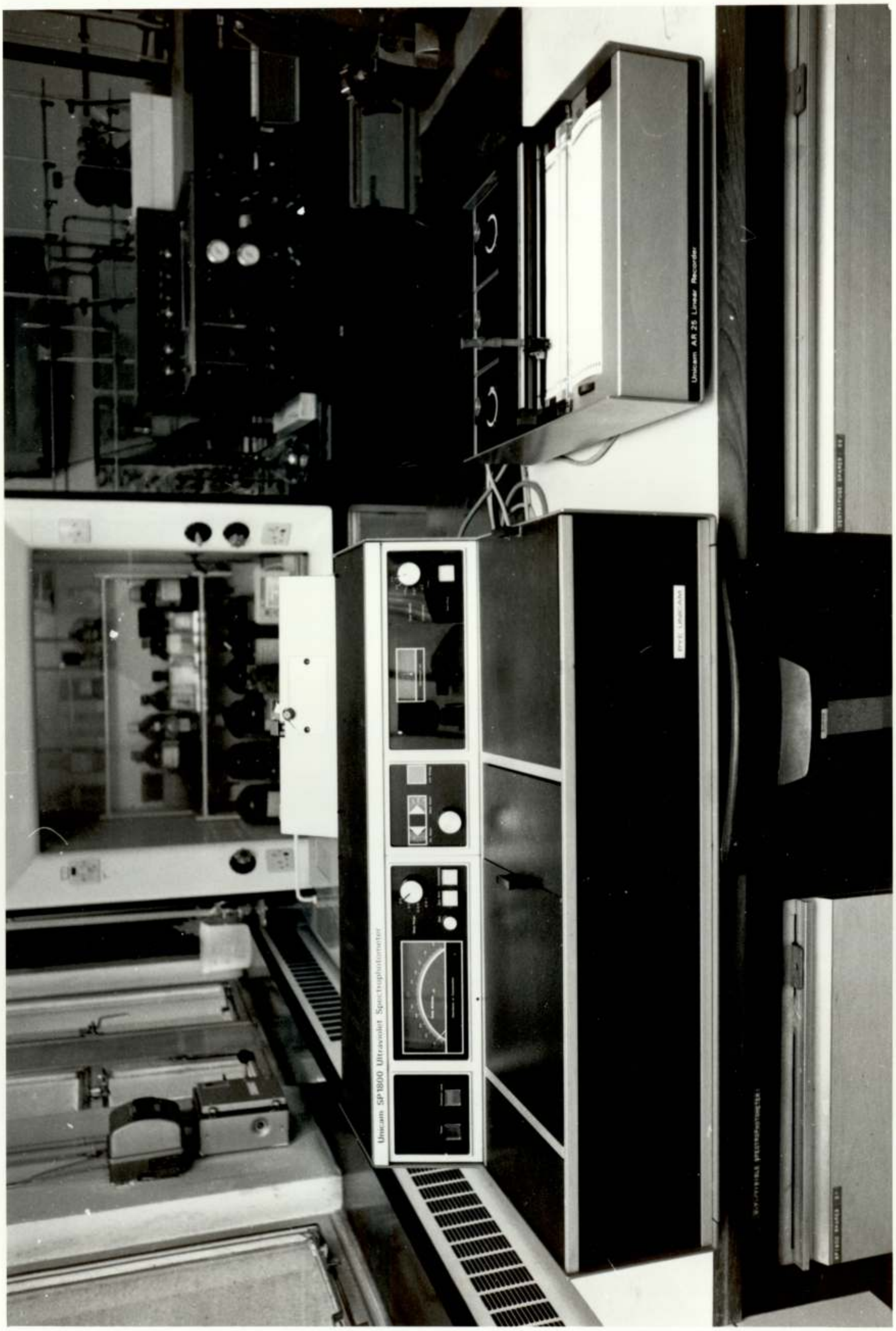


Figure 8.1 The Unicam SP1800 Ultraviolet Spectrophotometer

Figure 8.2 Graphical solution to the diffusion equation

$x/L = 0{\cdot}88$ (the actual measurement position)

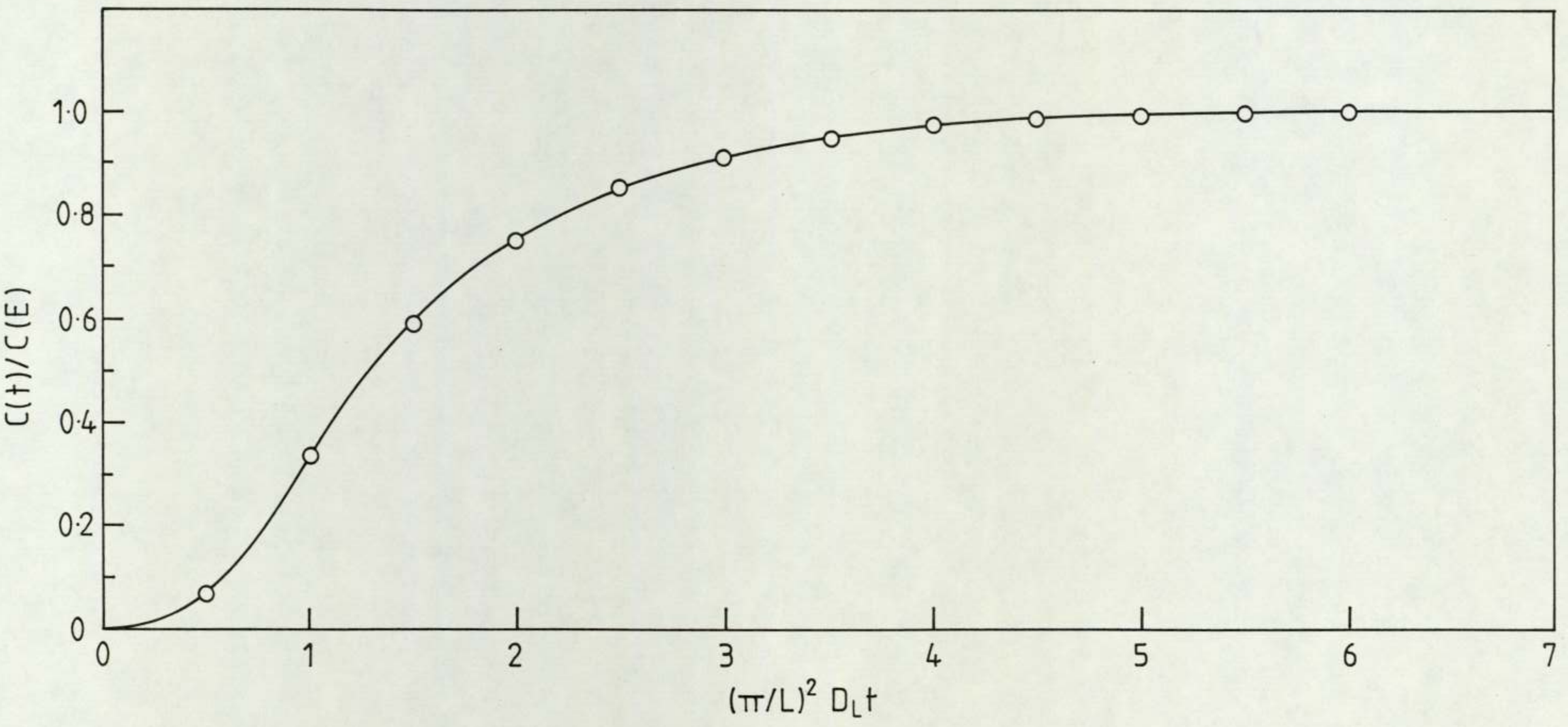
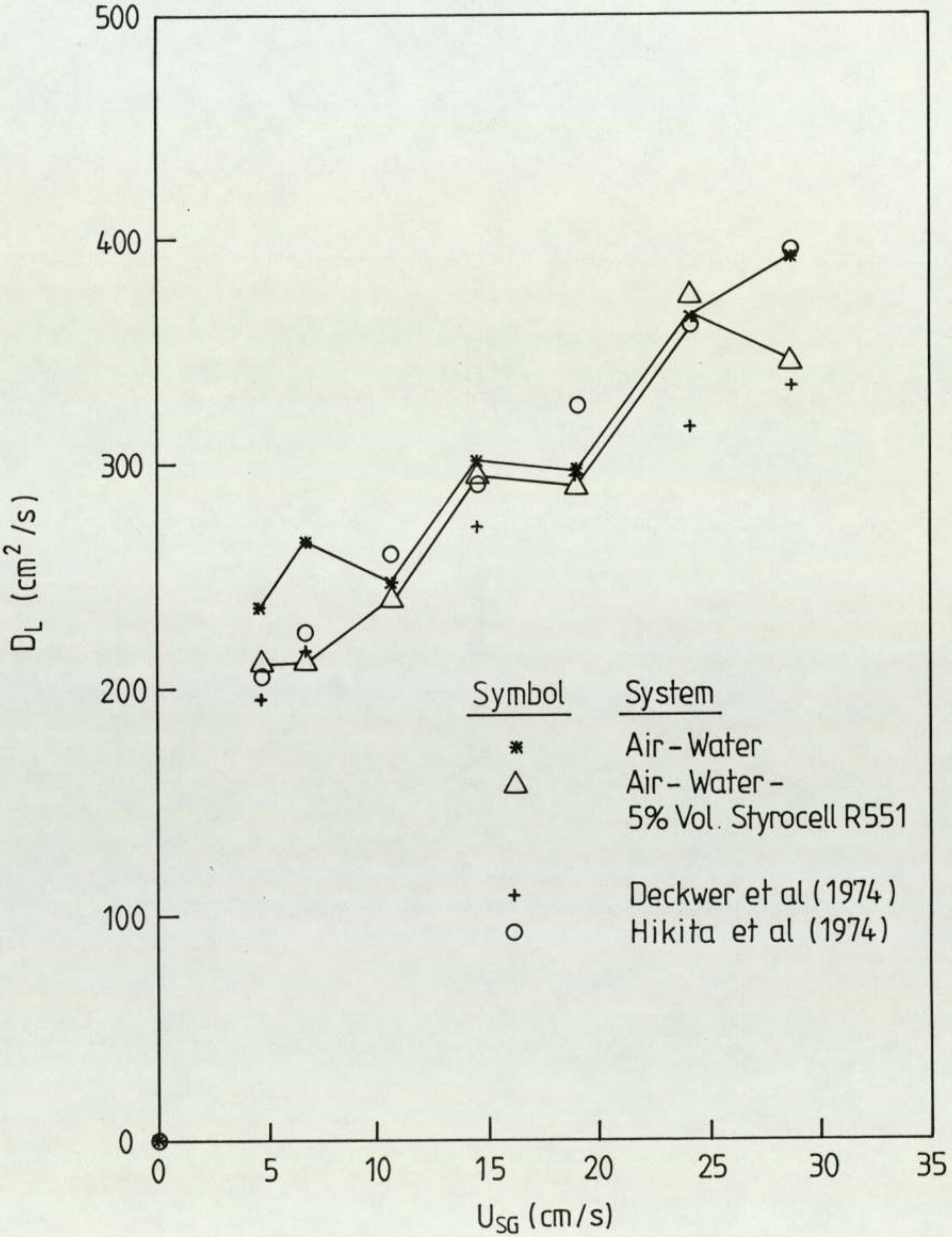


Figure 8.3 Effect of superficial gas velocity on dispersion coefficient in the three dimensional bubble column. Two and three-phase systems.



8.6 Empirical Correlations

A computer program (see Appendix I) was written to calculate the values of the dispersion coefficient from published correlations. These have been compared with the experimental data, the results being given in Table 8.2. There is a good fit between the calculated and experimental dispersion coefficients. (See Figure 8.3 also).

The correlations used are:

Deckwer et al (1974)

$$D_L = 0.678\, D_C^{1.4}\ U_{SG}^{0.3} \qquad \text{........ 8.9}$$

Hikita and Kikukawa (1974)

$$D_L = (0.15 + 0.69\, U_{SG}^{0.77}) D_C^{1.25}\ (1/\mu_L)^{0.12} \qquad \text{........ 8.10}$$

Baird and Rice (1975)

$$D_L = 0.35\, D_C^{1.333}\ (U_{SG}\, g)^{0.333} \qquad \text{........ 8.11}$$

Riquarts (1981)

$$D_L = 0.068\, D_C \sqrt{D_C\, g}\ (U_{SG}^{3}/\nu g)^{0.125} \qquad \text{........ 8.12}$$

TABLE 8.2: Comparison of Experimental Dispersion Coefficient Data with Empirical Correlations

Two- and Three-phase Systems

	Experimental					
U_{SG} (m/s)	D_L-2Phase (m^2/s)	D_L-3Phase (m^2/s)	Deckwer (1974)	Baird-Rice (1975)	Hikita et al (1974)	Riquarts (1981)
0.0480	0.0236	0.0212	0.0195	0.0221	0.0206	0.0171
0.0681	0.0265	0.0213	0.0217	0.0248	0.0225	0.0195
0.1070	0.0247	0.0241	0.0248	0.0289	0.0260	0.0231
0.1461	0.0301	0.0295	0.0272	0.0320	0.0291	0.0259
0.1918	0.0297	0.0291	0.0296	0.0351	0.0326	0.0287
0.2426	0.0365	0.0375	0.0317	0.0379	0.0362	0.0314
0.2881	0.0392	0.0345	0.0334	0.0402	0.0394	0.0335

8.7 Discussion

8.7.1 Effects of Operating Conditions

Rising gas bubbles cause rapid liquid phase circulation which has a dominating effect on mixing in the column. This is believed to be caused by a combination of the following phenomena:

- Differences in densities due to the existence of a phase rich in bubbles near the centre of the column and a phase relatively lean in bubbles near the wall of the column.
- Downward liquid flow compensating for the liquid transported upwards in the bubble wakes.
- Liquid displacement due to the rise of bubbles.

The contribution of each of these factors to the liquid circulation varies with the operating conditions.

The main factors that may cause axial mixing in the liquid phase of a fluidised bed may also be seen as:

- the presence of bubbles and particles,
- the motion of bubbles and particles,
- the continuous formation and separation of bubble and particle wake,
- the radial velocity profile of the gas phase.

When the bubbles are in motion, they are trailed by turbulent wakes which recurrently form and separate during their rise. These faster moving liquid elements will simultaneously cause axial mixing in the direction of flow, and a downflow of the displaced liquid phase in other parts of the bed. The extent to which such downflow takes place depends upon the bubble relative velocity (U_{SG} - U_{SL}) and bubble and wake sizes.

Fluidised particles are essentially in a state of continuous oscillation, and their amplitude of oscillation in the vertical direction increases with increasing gas velocity. Ruckenstein (1964) stated that in addition to the turbulence induced by the splitting and recombination of the fluid streams, this mode of particle motion is also responsible for fluid mixing in fluidised beds. Evidence for the existence of particle wakes and their importance in effecting axial mixing has been presented by Letan and Elgin (1972). The wake associated with the particle was assumed to embody highly mixed zones which multiply with an increase in either particle size or liquid rate. Axial mixing in the direction of flow was claimed to be mainly generated by the continuous formation and detachment of these wakes.

In view of the above description, it is to be expected that increasing the gas flow rate, ie. increasing the gas hold-up and bubble size, would under all conditions lead to an increase in the mixing parameter.

An increase in liquid flow rate will normally cause finer bubble dispersion (with no great effect on gas hold-up), lower particle concentration (in the entire system) and increased particle oscillations. The overall effect of changing liquid flow rate is difficult to predict but it may be expected that mixing will depend on particle characteristics.

The decrease of dispersion coefficient in three-phase fluidised beds over those measured in solid free systems is a result of the additional mechanisms offered by the presence of the solid particle phase. Non-wettable solids enhance bubble coalescence, decrease the gas hold-up and thus lead to a reduction though slight in the dispersion coefficient. This is supported by the findings of Ying et al (1980) and Kara et al (1982).

8.7.2 Effects of Liquid-Phase Properties

Liquid-phase properties - surface tension, viscosity and density - have been found to have an insignificant effect on axial dispersion coefficient (Aoyama et al (1968), Cova (1974), Alexander and Shah (1976)).

Ohki and Inoue (1970) explained their data in the bubbly flow regime using a velocity distribution model, which assumes a homogeneous bubble column with a uniform velocity independent of the column diameter. Baird and Rice (1975), on the other hand, proposed an isotropic turbulence model based on the Kolmogoroff theory in a homogeneous medium. Both these models also assumed that large scale interactions (ie. large eddies or diffusion on a large scale compared with the size of gas bubbles) play an important role in the dispersion process. Therefore, fluid viscosity, which partly determines the minimum eddy size, and fluid surface tension, which determines the shape and size of the bubbles, should have little effect on the magnitude of the dispersion coefficient. Both the velocity distribution model and the isotropic turbulence model predict strong effects of gas velocity and column diameter on the dispersion coefficient because the size of a large eddy should be dependent on these parameters.

8.8 Conclusions

Axial dispersion coefficients in the liquid phase of both gas-liquid and gas-liquid-solid fluidised beds have been determined from unsteady state tracer concentration measurements downstream of the injection plane.

On the basis of the limited amount of work carried out by the author, the following conclusions are drawn:

1. The axial dispersion coefficient increased with increasing gas flow rate.
2. The coefficients were lower for three-phase fluidised beds than those of the corresponding two-phase, particle free, systems.
3. Published correlations can be used to give preliminary estimates of dispersion coefficients.

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CHAPTER 9

OVERVIEW

OVERVIEW

A systematic study of the effects of operating parameters and nature of the phases on bubble column behaviour has been carried out. Two- and three-phases have been examined using two- and three-dimensional columns.

The comprehensive literature survey was helpful in defining the scope of the research, leading to the research strategy summarised in Table 9.1

Bubble Formation, Coalescence and Break-up

At formation, the initial bubble size at the gas distributor is controlled by the buoyancy and interfacial forces at low gas flow rates, by the instability of the gas-liquid interface of the gas jet at intermediate gas flow rates and by the ratio of the dynamic pressure force to the interfacial force at high gas flow rates. This initial bubble size is not necessarily preserved in the entire column. The ascending bubbles coalesce, if the initial bubble size is smaller than the equilibrium bubble size in the column and the coalescence is not suppressed. The bubbles break-up if the initial bubble size is larger than the equilibrium bubble size. In systems with hindered coalescence, the bubbles formed at the gas distributor can be preserved and, therefore, the size of the bubbles can be smaller than the equilibrium size.

The rising velocity of bubbles is influenced by the properties of the phases, bubble size, bubble density and the coalescence rate. At high bubble densities, the velocity of the bubble swarm is low due to the mutual hinderance of the bubbles and the back-flow of liquid. Under many practical conditions, coalescence results in growth of bubbles leading to an increase in their rising velocity.

Table 9.1: The Research Strategy

(1) LITERATURE REVIEW

(2) EXPERIMENTATION

- A DESIGN PARAMETERS
  - Gas Hold-up
  - Flow Regimes
  - Bubble Dynamics
  - Dispersion Coefficient

- B VARIABLES
  - Operating Parameters - Superficial gas and liquid velocities, initial liquid height, temperature.
  - Phase properties - Gas, liquid, solid
  - (Column Design - Column dia, height, type, distributor design)

- C SYSTEMS
  - Two-phase (Air-Water)
  - Two-phase with liquid additives (complex two-phase)
  - Three-phase (Gas-Liquid-Solid)
  - Three-phase with liquid additives (complex three-phase)

- D COLUMNS
  - Two-dimensional bubble column
  - Three-dimensional bubble column

- E MEASUREMENTS
  - Visual Observation
  - Bed Expansion Technique
  - Manometric Technique
  - Gas-Disengagement Technique
  - Photographic Technique
  - Tracer Injection Technique

(3) QUALITATIVE ANALYSIS

(4) DISCUSSION

(5) CONCLUSIONS

Complex Two-phase Systems

It should be noted that the most important difference between water and aqueous solutions of alcohols, ketones, esters and electrolytes is that in water the bubble coalescence rate is relatively high while in the others, the coalescence rate is low. The Corpuscular analysis has been developed to explain such behaviour. Organic and inorganic additives stabilize the bubble size by formation of an ionic or polar double layer at the interface which suppresses the coalescence rate. The rise velocity is also reduced. The effect of each additive is summarised below.

Alcohols - possess a surface tension lower than that of water. The surface tension of an aqueous solution is thus between the surface tension of pure water and alcohol. The alcohol molecule orientates itself with the hydrocarbon groups away from water. At equilibrium, the surface region contains an excess of orientated alcohol molecules which diminish the surface tension. The initial bubble size as well as the equilibrium diameter is also reduced (equation 5.1). The presence of the dipole orientated molecules at the surface suppresses the coalescence rate of bubbles. With increasing length of the hydrocarbon "tail" the degree of enrichment of orientation of the molecules increases: lower surface tension and reduced coalescence rate result.

Ketones - are surface active; therefore they are adsorbed at the gas-liquid interface. They also reduce the surface tension in solution. They thus behave fairly in a similar way to alcohols.

Esters - are also surface active to a certain extent and lower the surface tension. However, the effect of chain length is more pronounced than in alcohols and ketones. While the lower esters behave similarly, the higher esters show quite a different behaviour: coalescence seems less affected and lower hold-up results.

Electrolytes - increase the surface tension. These salts make the whole solution more cohesive or strongly bound together due to the prevailing electrostatic forces. Finely dispersed bubbles are well retained. The formation of electrical double layer at the interface suppresses bubble coalescence.

Foam Formation - the higher alcohols and ketones result in foaming, especialy at high gas velocities. This behaviour has a significant effect on column characteristics.

Three-phase Systems

In three-phase systems, the effects of gas and liquid velocities, solid physical properties on gas hold-up, flow regimes, bubble dynamics and dispersion coefficient were examined. It was found that the presence of solids exerts a significant influence on these parameters compared to two-phase systems. However, increasing the particle size, concentration and liquid velocity did not exert any significant effect on the parameters.

The effect of solids is closely linked to the surface property of WETTABILITY. Solids wettability suppresses the coalescence tendencies in the liquid phase, affecting column behaviour. Styrocell and Ballotini were observed to exhibit non-wetting characteristics, while Diakon and Amberlite exhibit wetting characteristics.

Complex Three-phase Systems

The presence of organic and inorganic additives in the liquid phase did not have any significant influence on the properties of the three-phase system. It would appear that any potential effect is largely counter-balanced by the presence of solids. In other words, the presence of solids has an over-riding effect on the additives.

Modelling and Empirical Correlations

Most of the experimental data were compared with empirical correlations available in the literature. It was found that the experimental values were a reasonable match with those predicted by such correlations. However, these correlations should be treated with caution while using them for systems containing surfactants.

The slip velocity, drift flux and Zuber-Findlay models were helpful in analysing column behaviour under different conditions.

CHAPTER 10

RECOMMENDATIONS FOR FURTHER STUDY

Additional Scope

Many industrial reactions are either exothermic or endothermic. Heat transfer studies using internals and their possible effect on fluid dynamics are relatively scarce in the literature and should be the subjects of further research.

Having done extensive studies with single stage, two- and three-dimensional bubble columns, work on multi-stage columns should now be initiated and developed.

APPENDIX A

Gas Hold-up in Two-Phase Systems

(Tables of Experimental Results)

TABLE A1: Effect of Superficial Gas and Liquid Velocities on Gas Hold-up in the 2-D Bubble Column

Two-phase System (Air-Water)

	U_{SL} (cm/s)				
U_{SG} (cm/s)	0	0.13	0.27	0.38	0.73
3.48	0.097	0.093	0.093	0.090	0.089
4.80	0.134	0.134	0.133	0.131	0.127
7.63	0.198	0.197	0.194	0.194	0.190
10.57	0.224	0.220	0.216	0.213	0.213
13.62	0.239	0.235	0.231	0.231	0.224
17.00	0.253	0.250	0.246	0.243	0.239
20.70	0.276	0.269	0.261	0.250	0.246
21.51	0.287	0.280	0.276	0.265	0.264

TABLE A2: Effect of Superficial Gas and Liquid Velocities on Gas Hold-up in the 3-D Bubble Column

Two-phase System (Air-Water)

	U_{SL} (cm/s)				
U_{SG} (cm/s)	0	0.03	0.06	0.14	0.18
4.80	0.135	0.133	0.130	0.122	0.120
6.81	0.183	0.173	0.170	0.168	0.165
10.70	0.205	0.200	0.195	0.190	0.189
14.61	0.223	0.215	0.210	0.205	0.203
19.18	0.250	0.243	0.238	0.233	0.228
24.26	0.275	0.268	0.263	0.260	0.255
28.81	0.303	0.298	0.290	0.285	0.280
31.56	0.315	0.308	0.300	0.295	0.293

TABLE A3: Effect of Alcohols on Gas Hold-up in the 2-D Bubble Column

Two-phase System (Gas-Liquid)

		Alcohols			
U_{SG} (cm/s)	Air-Water	0.5 Vol % Methanol	0.5 Vol % Ethanol	0.5 Vol % Propanol	0.5 Vol % Butanol
3.48	0.097	0.119	0.123	0.127	0.131
4.80	0.134	0.198	0.201	0.220	0.213
7.63	0.198	0.351	0.384	0.433	0.493
10.57	0.224	0.485	0.487	0.709	0.746
13.62	0.239	0.549	0.567	0.710	0.746
17.00	0.253	0.515	0.560	0.710	0.735
20.70	0.276	0.507	0.541	0.709	0.716
21.51	0.287	0.500	0.534	0.709	0.746

TABLE A4: Effect of Ketones on Gas Hold-up in the 2-D Bubble Column

Two-phase System (Gas-Liquid)

		Ketone	
U_{SG} (cm/s)	Air-Water	0.5 Vol % Acetone	0.5 Vol % Butanone
3.48	0.097	0.127	0.134
4.80	0.134	0.201	0.216
7.63	0.198	0.407	0.459
10.57	0.224	0.604	0.724
13.62	0.239	0.679	0.720
17.00	0.253	0.705	0.724
20.70	0.276	0.713	0.720
21.51	0.287	0.710	0.725

TABLE A5: Effect of Esters on Gas Hold-up in the 2-D Bubble Column

Two-phase System (Gas-Liquid)

		Ester		
U_{SG} (cm/s)	Air-Water	0.5 Vol % Ethyl Acetate	0.5 Vol % N-propyl Acetate	0.5 Vol % N-Butyl Acetate
3.48	0.097	0.127	0.075	0.071
4.80	0.134	0.216	0.097	0.101
7.63	0.198	0.287	0.149	0.142
10.57	0.224	0.269	0.172	0.172
13.62	0.239	0.265	0.201	0.201
17.00	0.253	0.299	0.231	0.228
20.70	0.276	0.321	0.265	0.263
21.51	0.287	0.328	0.281	0.280

TABLE A6: Effect of Electrolytes on Gas Hold-up in the 2-D Bubble Column

Two-phase System (Gas-Liquid)

		Electrolytes			
U_{SG} (cm/s)	Air-Water	1% NaCl	1% KCl	1% $CaCl_2$	1% Na_2SO_4
3.48	0.097	0.119	0.119	0.104	0.112
4.80	0.134	0.190	0.179	0.175	0.183
7.63	0.198	0.325	0.306	0.254	0.321
10.57	0.224	0.414	0.351	0.246	0.388
13.62	0.239	0.392	0.349	0.235	0.410
17.00	0.253	0.369	0.328	0.254	0.392
20.70	0.276	0.347	0.306	0.276	0.366
21.51	0.287	0.343	0.310	0.291	0.366

APPENDIX B

Comparison of Experimental Gas Hold-up Data with Empirical Correlations

(Two-phase Systems)

TABLE B2: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-0.5 Vol % Methanol)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.119	0.083	0.135	0.426	0.084
4.80	0.198	0.105	0.175	0.460	0.101
7.63	0.351	0.141	0.246	0.507	0.132
10.57	0.485	0.170	0.296	0.539	0.160
13.62	0.549	0.195	0.327	0.563	0.185
17.00	0.515	0.217	0.343	0.582	0.210
20.70	0.507	0.238	0.342	0.600	0.236
21.51	0.500	0.242	0.340	0.603	0.241

TABLE B3: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-0.5 Vol % Ethanol)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.123	0.084	0.138	0.427	0.086
4.80	0.201	0.105	0.179	0.461	0.103
7.63	0.384	0.142	0.249	0.509	0.135
10.57	0.487	0.171	0.299	0.540	0.163
13.62	0.567	0.196	0.329	0.564	0.188
17.00	0.560	0.218	0.343	0.584	0.214
20.70	0.541	0.239	0.341	0.601	0.240
21.51	0.534	0.243	0.338	0.604	0.245

TABLE B4: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-0.5 Vol % I-Propanol)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.127	0.085	0.141	0.428	0.087
4.80	0.220	0.106	0.183	0.462	0.105
7.63	0.433	0.143	0.254	0.510	0.137
10.57	0.709	0.173	0.303	0.541	0.166
13.62	0.710	0.197	0.332	0.565	0.192
17.00	0.710	0.220	0.344	0.584	0.218
20.70	0.709	0.240	0.339	0.602	0.245
21.51	0.709	0.244	0.336	0.605	0.250

TABLE B5: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-0.5 Vol % N-Butanol)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.131	0.085	0.142	0.429	0.088
4.80	0.213	0.107	0.184	0.462	0.106
7.63	0.493	0.144	0.255	0.510	0.138
10.57	0.746	0.173	0.304	0.541	0.167
13.62	0.746	0.198	0.333	0.565	0.193
17.00	0.735	0.220	0.344	0.585	0.219
20.70	0.716	0.241	0.338	0.602	0.246
21.51	0.746	0.245	0.335	0.605	0.252

TABLE B7: Comparison of Experimental Gas Hold-up with Empirical Correlations

Two-phase System (Air-0.5 Vol % Butanone)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.134	0.084	0.138	0.428	0.086
4.80	0.216	0.106	0.179	0.462	0.103
7.63	0.459	0.142	0.250	0.509	0.135
10.57	0.724	0.171	0.300	0.540	0.163
13.62	0.720	0.196	0.330	0.564	0.189
17.00	0.724	0.218	0.343	0.584	0.215
20.70	0.720	0.239	0.340	0.601	0.241
21.51	0.725	0.243	0.338	0.604	0.246

TABLE B8: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-0.5 Vol % Ethyl Acetate)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.127	0.085	0.140	0.428	0.086
4.80	0.216	0.106	0.181	0.461	0.104
7.63	0.287	0.143	0.252	0.509	0.136
10.57	0.269	0.172	0.301	0.540	0.164
13.62	0.265	0.196	0.331	0.564	0.190
17.00	0.299	0.219	0.344	0.584	0.216
20.70	0.321	0.240	0.340	0.601	0.242
21.51	0.328	0.244	0.337	0.604	0.247

TABLE B9: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-0.5 Vol % N-Propyl Acetate)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.075	0.085	0.142	0.429	0.088
4.80	0.097	0.107	0.183	0.463	0.105
7.63	0.149	0.143	0.254	0.510	0.138
10.57	0.172	0.173	0.304	0.541	0.166
13.62	0.201	0.197	0.332	0.565	0.193
17.00	0.231	0.220	0.344	0.585	0.219
20.70	0.265	0.240	0.339	0.602	0.245
21.51	0.281	0.245	0.336	0.605	0.251

TABLE B10: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-0.5 Vol % N-Butyl Acetate)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.071	0.086	0.143	0.432	0.089
4.80	0.101	0.107	0.185	0.465	0.107
7.63	0.142	0.144	0.257	0.513	0.140
10.57	0.172	0.173	0.305	0.544	0.169
13.62	0.201	0.198	0.334	0.567	0.196
17.00	0.228	0.220	0.344	0.587	0.223
20.70	0.263	0.241	0.338	0.604	0.250
21.51	0.280	0.245	0.335	0.608	0.255

TABLE B11: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-1% NaCl)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.119	0.085	0.141	0.428	0.087
4.80	0.190	0.106	0.182	0.462	0.105
7.63	0.325	0.143	0.253	0.509	0.137
10.57	0.414	0.172	0.302	0.541	0.165
13.62	0.392	0.197	0.332	0.565	0.191
17.00	0.369	0.219	0.344	0.584	0.218
20.70	0.347	0.240	0.339	0.602	0.244
21.51	0.343	0.244	0.337	0.605	0.249

TABLE B12: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-1% KCl)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.119	0.084	0.138	0.429	0.086
4.80	0.179	0.105	0.179	0.463	0.103
7.63	0.306	0.142	0.249	0.510	0.135
10.57	0.351	0.171	0.299	0.542	0.163
13.62	0.349	0.196	0.329	0.565	0.189
17.00	0.328	0.218	0.343	0.585	0.215
20.70	0.306	0.239	0.341	0.602	0.241
21.51	0.310	0.243	0.338	0.605	0.246

TABLE B13: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-1% $CaCl_2$)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.104	0.085	0.141	0.428	0.087
4.80	0.175	0.106	0.183	0.461	0.105
7.63	0.254	0.143	0.254	0.509	0.137
10.57	0.246	0.173	0.303	0.540	0.165
13.62	0.235	0.197	0.332	0.564	0.192
17.00	0.254	0.220	0.344	0.584	0.218
20.70	0.276	0.240	0.339	0.601	0.244
21.51	0.291	0.244	0.336	0.605	0.249

TABLE B14: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Two-phase System (Air-1% Na_2So_4)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kumar et al (1976)	Mersmann (1978)	Hikita et al (1980)
3.48	0.112	0.085	0.140	0.428	0.086
4.80	0.183	0.106	0.181	0.461	0.104
7.63	0.321	0.143	0.252	0.509	0.136
10.57	0.388	0.172	0.301	0.540	0.164
13.62	0.410	0.196	0.331	0.564	0.190
17.00	0.392	0.219	0.344	0.584	0.216
20.70	0.366	0.240	0.340	0.601	0.242
21.51	0.366	0.244	0.337	0.604	0.247

TABLE C1: Variation of Bubble Velocity with Superficial Gas Velocity

	U_B (cm/s)				
U_{SG} (cm/s)	Water	0.5 Vol % Ethanol	0.5 Vol % Acetone	0.5 Vol % Ethyl Acetate	0.5 Vol % NaCl
3.48	35.88	28.29	27.40	27.40	29.24
4.80	35.75	23.83	23.83	22.18	25.21
7.63	38.83	19.87	18.75	26.59	23.48
10.57	47.19	21.70	17.50	39.29	25.53
13.62	56.99	24.02	20.06	51.40	34.74
17.00	67.15	30.34	24.10	56.82	46.04
20.70	75.00	38.26	29.03	64.49	59.65
21.51	75.00	40.28	30.30	65.58	62.71

TABLE C2: Variation of Buoyancy Velocity with Superficial Gas Velocity

Two-phase System (Air-Water)

U_{SG} (cm/s)	U_o (cm/s)
3.48	18.52
4.79	17.86
6.15	17.24
7.63	16.56
8.98	15.72
10.57	15.15
12.09	14.88
13.62	15.63
15.25	14.75
17.00	14.29

TABLE C3: Variation of Dynamic Gas Hold-up with Time in the 3-D Bubble Column (Li = 90 cm)

Two-phase System (Air-Water)

	U_{SG} (cm/s)				
Time (s)	1.23	3.78	5.00	6.03	7.35
0	0.043	0.119	0.144	0.168	0.184
1	0.022	0.101	0.136	0.160	0.176
2	0.011	0.073	0.101	0.152	0.168
3	0.006	0.043	0.063	0.119	0.136
4	-	0.022	0.033	0.083	0.101
5		-	0.011	0.043	0.073
6			-	0.011	0.033
7				-	0.011
8					-

TABLE C4: Variation of Dynamic Gas Hold-up with Time in the 3-D Bubble Column (Li = 100 cm)

Two-phase System (Air-Water)

	U_{SG} (cm/s)					
Time (s)	1.23	2.17	4.31	6.03	6.62	7.35
0	0.039	0.066	0.124	0.161	0.175	0.182
1	0.029	0.057	0.100	0.132	0.147	0.168
2	0.015	0.029	0.066	0.092	0.116	0.124
3	0.005	0.020	0.039	0.066	0.075	0.092
4	-	0.010	0.020	0.039	0.048	0.057
5		-	0.005	0.020	0.029	0.039
6			-	0.010	0.020	0.030
7				-	0.010	0.020
8					-	0.010
9						-

TABLE C6: Typical Fall of Liquid Height with Time (Li = 100 cm) in the 3-D Bubble Column

Two-phase (Air-Water)

	U_{SG} (cm/s)					
Time (s)	1.23	2.17	4.31	6.03	6.62	7.35
0	103	016	113	118	120	121
1	102	105	110	114	116	119
2	100.5	102	106	109	112	113
3	99.5	101	103	106	107	109
4	-	100	101	103	104	105
5	-	-	99.5	101	102	103
6		-	-	100	101	102
7			-	-	100	101
8				-	-	100
9					-	-

APPENDIX D

Gas Hold-up in Three-phase Fluidisation

(Tables of Experimental Results)

TABLE D1: Effect of Superficial Gas and Liquid Velocities on Gas Hold-up in the 2-D Bubble Column

Three-phase System (Air-Water-5 Vol % Styrocell R551)

	U_{SL} (cms)				
U_{SG} (cm/s)	0	0.13	0.27	0.38	0.61
3.48	0.060	0.052	0.049	0.045	0.041
4.80	0.082	0.071	0.070	0.069	0.067
7.63	0.119	0.106	0.105	0.104	0.104
10.57	0.157	0.142	0.142	0.138	0.134
13.62	0.187	0.172	0.172	0.164	0.164
17.00	0.209	0.198	0.198	0.196	0.194
20.70	0.239	0.219	0.216	0.216	0.213
21.51	0.254	0.246	0.231	0.231	0.224

TABLE D2: Effect of Superficial Gas and Liquid Velocities on Gas Hold-up in the 3-D Bubble Column

Three-phase System (Air-Water-5 Vol % Styrocell R551)

	U_{SL} (cms)				
U_{SG} (cm/s)	0	0.03	0.06	0.10	0.14
4.80	0.070	0.065	0.060	0.055	0.055
6.81	0.085	0.080	0.078	0.073	0.073
10.70	0.120	0.110	0.105	0.095	0.093
14.16	0.148	0.140	0.133	0.125	0.120
19.18	0.173	0.160	0.153	0.145	0.143
24.26	0.195	0.180	0.175	0.170	0.170
28.81	0.215	0.195	0.190	0.180	0.180
31.56	0.235	0.215	0.208	0.200	0.200

TABLE D3: Effect of Superficial Gas and Liquid Velocities on Gas Hold-up in the 2-D Bubble Column

Three-phase System (Air-Water-5 Vol % Amberlite)

	U_{SL} (cms)				
U_{SG} (cm/s)	0	0.13	0.27	0.38	0.61
3.48	0.097	0.093	0.091	0.086	0.082
4.80	0.142	0.134	0.131	0.131	0.123
7.63	0.201	0.194	0.192	0.190	0.187
10.57	0.243	0.224	0.213	0.209	0.201
13.62	0.261	0.213	0.210	0.209	0.205
17.00	0.272	0.235	0.229	0.224	0.220
20.70	0.291	0.254	0.246	0.246	0.244
21.51	0.300	0.265	0.257	0.254	0.254

TABLE D4: Effect of Solid Concentration on Gas Hold-up in the 2-D Bubble Column

Three-phase System (Air-Water-Styrocell R551)

	Concn (Vol %)			
U_{SG} (cm/s)	0	5%	10%	15%
3.48	0.097	0.060	0.052	0.052
4.80	0.134	0.082	0.075	0.075
7.63	0.198	0.119	0.112	0.112
10.57	0.224	0.157	0.149	0.149
13.62	0.239	0.187	0.187	0.179
17.00	0.253	0.209	0.209	0.209
20.70	0.276	0.239	0.239	0.231
21.51	0.287	0.254	0.246	0.246

TABLE D5: Effect of Solid Particle Concentration on Gas Hold-up in the 3-D Bubble Column

Three-phase System (Air-Water-Styrocell R551)

	Concn (Vol %)			
U_{SG} (cm/s)	0	5%	10%	15%
4.80	0.135	0.070	0.065	0.065
6.81	0.183	0.085	0.083	0.083
10.70	0.205	0.120	0.118	0.115
14.61	0.223	0.148	0.145	0.145
19.18	0.250	0.173	0.170	0.168
24.26	0.275	0.195	0.190	0.190
28.81	0.303	0.215	0.205	0.205
31.56	0.315	0.235	0.220	0.220

TABLE D6: Effect of Solid Concentration on Gas Hold-up in the 2-D Bubble Column

Three-phase System (Air-Water-Styrocell R351)

	Concn (Vol %)			
U_{SG} (cm/s)	0	5%	10%	15%
3.48	0.097	0.052	0.055	0.052
4.80	0.134	0.075	0.075	0.075
7.63	0.198	0.112	0.112	0.104
10.57	0.224	0.149	0.149	0.149
13.62	0.239	0.179	0.179	0.179
17.00	0.253	0.201	0.208	0.201
20.70	0.276	0.231	0.231	0.231
21.51	0.287	0.246	0.246	0.246

TABLE D7: Effect of Solid Concentration on Gas Hold-up in the 2-D Bubble Column

Three-phase System (Air-Water-Ballotini)

	Concn (Vol %)			
U_{SG} (cm/s)	0	5%	10%	15%
3.48	0.097	0.058	0.049	0.049
4.80	0.134	0.073	0.073	0.071
7.63	0.198	0.110	0.104	0.101
10.57	0.224	0.146	0.134	0.134
13.62	0.239	0.174	0.164	0.164
17.00	0.253	0.201	0.187	0.187
20.70	0.276	0.224	0.220	0.213
21.51	0.287	0.239	0.231	0.227

TABLE D8: Effect of Solid Concentration on Gas Hold-up in the 2-D Bubble Column

Three-phase System (Air-Water-Amberlite)

	Concn (Vol %)			
U_{SG} (cm/s)	0	5%	10%	15%
3.48	0.097	0.097	0.087	0.082
4.80	0.134	0.142	0.124	0.112
7.63	0.198	0.201	0.175	0.164
10.57	0.224	0.243	0.201	0.187
13.62	0.239	0.261	0.222	0.209
17.00	0.253	0.272	0.243	0.231
20.70	0.276	0.291	0.265	0.257
21.51	0.287	0.300	0.276	0.265

TABLE D9: Effect of Solid Size on Gas Hold-up in the 2-D Bubble Column

Three-phase System (Air-Water-5 Vol % Styrocell)

	d_p (mm)			
U_{SG} (cm/s)	0.60	0.81	0.85	1.23
3.48	0.067	0.060	0.056	0.052
4.80	0.089	0.082	0.078	0.075
7.63	0.134	0.119	0.117	0.112
10.57	0.164	0.157	0.151	0.149
13.62	0.201	0.187	0.179	0.179
17.00	0.216	0.209	0.205	0.201
20.70	0.254	0.239	0.231	0.231
21.51	0.261	0.254	0.250	0.246

TABLE D10: Effect of Non-Wettable Solid Particles on Gas Hold-up in the 2-D Bubble Column

Three-phase Systems (Air-Water-Styrocell)

(Air-Water-Ballitoni)

	SYSTEM		
U_{SG} (cm/s)	Air-Water	5 Vol % Styrocell R351	5 Vol % Ballotini
3.48	0.097	0.052	0.058
4.80	0.134	0.075	0.073
7.63	0.198	0.112	0.110
10.57	0.224	0.149	0.146
13.62	0.239	0.179	0.174
17.00	0.253	0.201	0.201
20.70	0.276	0.231	0.224
21.51	0.287	0.246	0.239

TABLE D11: Effect of Wettable Solid Particles on Gas Hold-up in the 2-D Bubble Column

Three-phase Systems (Air-Water-Amberlite)

(Air-Water-Diakon)

	SYSTEM		
U_{SG} (cm/s)	Air-Water	5 Vol % Amberlite Resin	5 Vol % Diakon III
3.48	0.097	0.097	0.097
4.80	0.134	0.142	0.142
7.63	0.198	0.201	0.201
10.57	0.224	0.243	0.231
13.62	0.239	0.261	0.261
17.00	0.253	0.272	0.283
20.70	0.276	0.291	0.299
21.51	0.287	0.300	0.313

TABLE D12: Effect of Solid-phase and Alcohol (Methanol) on Gas Hold-up in the 2-D Bubble Column

	SYSTEM		
U_{SG} (cm/s)	Air-Water	Air - 0.5% Methanol	Air - 0.5% Methanol 5% Styrocell
3.48	0.097	0.119	0.060
4.80	0.134	0.198	0.075
7.63	0.198	0.351	0.112
10.57	0.224	0.485	0.142
13.62	0.239	0.549	0.175
17.00	0.253	0.515	0.205
20.70	0.276	0.507	0.239
21.51	0.287	0.500	0.257

TABLE D13: EDffect of Solid-phase and Alcohol (Ethanol) on Gas Hold-up in the 2-D Bubble Column

		SYSTEM	
U_{SG} (cm/s)	Air-Water	Air - 0.5% Ethanol	Air - 0.5% Ethanol 5% Styrocell
3.48	0.097	0.123	0.075
4.80	1.134	0.201	0.082
7.63	0.198	0.384	0.112
10.57	0.224	0.487	0.149
13.62	0.239	0.567	0.183
17.00	0.253	0.560	0.213
20.70	0.276	0.541	0.246
21.51	0.287	0.534	0.269

TABLE D14: Effect of Solid-phase and Ketones (Acetone) on Gas Hold-up in the 2-D Bubble Column

	SYSTEM		
U_{SG} (cm/s)	Air-Water	Air - 0.5% Acetone	Air - 0.5% Acetone 5% Styrocell
3.48	0.097	0.127	0.056
4.80	0.134	0.201	0.075
7.63	0.198	0.407	0.113
10.57	0.224	0.604	0.149
13.62	0.239	0.679	0.179
17.00	0.253	0.705	0.209
20.70	0.276	0.713	0.239
21.51	0.287	0.710	0.257

TABLE D15: Effect of Solid-phase and Ketone (Butanone) on Gas Hold-up in the 2-D Bubble Column

		SYSTEM	
U_{SG} (cm/s)	Air-Water	Air - 0.5% Butanone	Air - 0.5% Butanone 5% Styrocell
3.48	0.097	0.134	0.082
4.80	0.134	0.216	0.090
7.63	0.198	0.459	0.119
10.57	0.224	0.724	0.149
13.62	0.239	0.720	0.187
17.00	0.253	0.724	0.216
20.70	0.276	0.720	0.246
21.51	0.287	0.725	0.269

TABLE D16: Effect of Solid-phase and Ester (Ethyl Acetate) on Gas Hold-up in the 2-D Bubble Column

	SYSTEM		
U_{SG} (cm/s)	Air-Water	Air - 0.5% Ethyl Acetate	Air - 0.5% Ethyl Acetate 5% Styrocell
3.48	0.097	0.127	0.094
4.80	0.134	0.216	0.132
7.63	0.198	0.287	0.157
10.57	0.224	0.269	0.172
13.62	0.239	0.265	0.194
17.00	0.253	0.299	0.220
20.70	0.276	0.321	0.254
21.51	0.287	0.328	0.265

TABLE D17: Effect of Solid-phase and Electrolyte (NaCl) on Gas Hold-up in the 2-D Bubble Column

	SYSTEM		
U_{SG} (cm/s)	Air-Water	Air - 1.0 % NaCl	Air - 1.0% NaCl 5% Styrocell
3.48	0.097	0.119	0.052
4.80	0.134	0.190	0.078
7.63	0.198	0.325	0.123
10.57	0.224	0.414	0.153
13.62	0.239	0.392	0.179
17.00	0.253	0.369	0.216
20.70	0.276	0.347	0.243
21.51	0.287	0.343	0.254

TABLE D18: Effect of Solid-phase and Electrolyte ($CaCl_2$) on Gas Hold-up in the 2-D Bubble Column

	SYSTEM		
U_{SG} (cm/s)	Air-Water	Air - 1.0 % $CaCl_2$	Air - 1.0% $CaCl_2$ 5% Styrocell
3.48	0.097	0.104	0.067
4.80	0.134	0.175	0.078
7.63	0.198	0.254	0.116
10.57	0.224	0.246	0.149
13.62	0.239	0.235	0.183
17.00	0.253	0.254	0.216
20.70	0.276	0.276	0.243
21.51	0.287	0.291	0.261

APPENDIX E

Comparison of Experimental Gas Hold-up Data with Empirical Correlations

(Three-phase Systems)

TABLE E1: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Three-phase System (Air-Water-Styrocell R551)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kito et al (1976)	Begovich-Watson (1978)	Hikita et al (1980)
3.48	0.060	0.083	0.076	0.072	0.085
4.80	0.082	0.104	0.095	0.090	0.102
7.63	0.119	0.141	0.0128	0.124	0.134
10.57	0.157	0.170	0.157	0.156	0.161
13.62	0.187	0.194	0.183	0.185	0.187
17.00	0.209	0.216	0.207	0.216	0.212
20.70	0.239	0.237	0.231	0.248	0.238
21.51	0.254	0.241	0.236	0.254	0.243

TABLE E2: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Three-phase System (Air-0.5% Methanol-5% Styrocell R551)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kito et al (1976)	Begovich-Watson (1978)	Hikita et al (1980)
3.48	0.060	0.085	0.079	0.072	0.087
4.80	0.075	0.106	0.098	0.090	0.105
7.63	0.112	0.143	0.133	0.124	0.138
10.57	0.142	0.172	0.163	0.156	0.166
13.62	0.175	0.197	0.189	0.185	0.192
17.00	0.205	0.219	0.214	0.216	0.218
20.70	0.239	0.240	0.238	0.248	0.245
21.51	0.257	0.244	0.243	0.254	0.250

TABLE E3: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Three-phase System (Air-0.5% Ethanol-5% Styrocell R551)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kito et al (1976)	Begovich-Watson (1978)	Hikita et al (1980)
3.48	0.075	0.086	0.082	0.072	0.090
4.80	0.082	0.108	0.101	0.090	0.109
7.63	0.112	0.145	0.137	0.124	0.142
10.57	0.149	0.175	0.168	0.156	0.171
13.62	0.183	0.199	0.194	0.185	0.199
17.00	0.213	0.222	0.220	0.216	0.226
20.70	0.246	0.243	0.244	0.248	0.253
21.51	0.269	0.247	0.249	0.254	0.259

TABLE E4: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Three-phase System (Air-0.5% Acetone-5% Styrocell R551)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kito et al (1976)	Begovich-Watson (1978)	Hikita et al (1980)
3.48	0.056	0.086	0.081	0.072	0.089
4.80	0.075	0.107	0.100	0.090	0.107
7.63	0.113	0.144	0.135	0.124	0.140
10.57	0.149	0.174	0.166	0.156	0.169
13.62	0.179	0.198	0.192	0.185	0.195
17.00	0.209	0.221	0.217	0.216	0.222
20.70	0.239	0.241	0.242	0.248	0.249
21.51	0.257	0.246	0.246	0.254	0.254

TABLE E5: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Three-phase System (Air-0.5% Butanone-5% Styrocell R551)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kito et al (1976)	Begovich-Watson (1978)	Hikita et al (1980)
3.48	0.082	0.086	0.081	0.072	0.089
4.80	0.090	0.107	0.100	0.090	0.107
7.63	0.119	0.144	0.136	0.124	0.140
10.57	0.149	0.174	0.166	0.156	0.169
13.62	0.187	0.199	0.193	0.185	0.195
17.00	0.216	0.221	0.218	0.216	0.222
20.70	0.246	0.242	0.242	0.248	0.249
21.51	0.269	0.246	0.247	0.254	0.254

TABLE E6: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Three-phase System (Air-0.5% Ethyl Acetate-5% Styrocell R551)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kito et al (1976)	Begovich-Watson (1978)	Hikita et al (1980)
3.48	0.094	0.086	0.080	0.072	0.089
4.80	0.132	0.107	0.100	0.090	0.107
7.63	0.157	0.144	0.135	0.124	0.140
10.57	0.172	0.173	0.165	0.156	0.169
13.62	0.194	0.198	0.191	0.185	0.196
17.00	0.220	0.220	0.217	0.216	0.223
20.70	0.254	0.241	0.241	0.248	0.250
21.51	0.265	0.245	0.246	0.254	0.255

TABLE E7: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Three-phase System (Air-1% NaCl-5% Styrocell R551)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kito et al (1976)	Begovich-Watson (1978)	Hikita et al (1980)
3.48	0.052	0.086	0.080	0.072	0.089
4.80	0.078	0.107	0.100	0.090	0.107
7.63	0.123	0.144	0.135	0.124	0.140
10.57	0.153	0.174	0.165	0.156	0.170
13.62	0.179	0.198	0.192	0.185	0.196
17.00	0.216	0.221	0.217	0.216	0.223
20.70	0.243	0.241	0.241	0.248	0.250
21.51	0.254	0.245	0.246	0.254	0.256

TABLE E8: Comparison of Experimental Gas Hold-up Data with Empirical Correlations

Three-phase System (Air-1% $CaCl_2$-5% Styrocell R551)

U_{SG} (cm/s)	EXPTL ε_G	Akita-Yoshida (1973)	Kito et al (1976)	Begovich-Watson (1978)	Hikita et al (1980)
3.48	0.067	0.085	0.080	0.072	0.089
4.80	0.078	0.107	0.99	0.090	0.106
7.63	0.116	0.144	0.135	0.124	0.139
10.57	0.149	0.173	0.165	0.156	0.168
13.62	0.183	0.198	0.191	0.185	0.195
17.00	0.216	0.220	0.216	0.216	0.221
20.70	0.243	0.241	0.240	0.248	0.248
21.51	0.261	0.245	0.245	0.254	0.254

APPENDIX F

Flow Regimes and Bubble Dynamics in Three-Phase Fluidised Beds

(Tables of Experimental Results)

TABLE F2: Variation of Bubble Diameter with Superficial Gas Velocity in the 2-D Bubble Column

Three-phase System (Air-Water-Styrocell R551)

	d_B(cm)		
U_{SG} (cm/s)	d_B(L)	d_B(M)	d_B(T)
4.80	1.00	2.23	1.28
7.63	1.32	2.16	1.97
10.57	2.31	3.07	2.96
13.62	2.44	3.15	3.04
17.00	2.93	3.31	3.26
20.77	3.44	3.54	3.45

TABLE F3: Effect of Superficial Gas and Liquid Velocities on Flow Regime Transition in the 2-D Bubble Column

Three-phase System (Air-Water-Styrocell R551)

	U_{SL} (cm/s)			
U_{SG} (cm/s)	0	0.13	0.38	0.61
3.48	3.27	3.30	3.32	3.34
4.80	4.40	4.45	4.46	4.47
7.63	6.72	6.82	6.84	6.83
10.57	8.91	9.07	9.11	9.15
13.62	11.07	11.28	11.39	11.39
17.00	13.44	13.63	13.66	13.69
20.70	15.75	16.23	16.17	16.23
21.51	16.05	16.22	16.54	16.69

TABLE F4: Effect of Superficial Gas and Liquid Velocities on Flow Regime Transition in the 2-D Bubble Column

Three-phase System (Air-Water-Amberlite)

	U_{SL} (cm/s)				
U_{SG} (cm/s)	0	0.13	0.27	0.38	0.61
3.48	3.14	3.16	3.16	3.20	3.18
4.80	4.11	4.16	4.15	4.20	4.16
7.63	6.10	6.26	6.20	6.18	6.15
10.57	8.00	8.45	8.45	8.36	8.32
13.62	10.07	10.72	10.83	10.77	10.76
17.00	12.37	13.00	13.10	13.18	13.25
20.70	14.68	15.44	15.61	15.61	16.65
21.51	15.06	15.81	15.98	16.05	16.05

TABLE F5: Effect of Solids Concentration on Flow Regime Transition in the 2-D Bubble Column

Three-phase System (Air-Water-Styrocell R551)

	Conc (Vol %)			
U_{SG} (cm/s)	0	5	10	15
3.48	3.14	3.27	3.30	3.30
4.80	4.15	4.40	4.43	4.43
7.63	6.12	6.72	6.78	6.78
10.57	8.20	8.91	9.00	9.00
13.62	10.36	11.07	11.07	11.18
17.00	12.70	13.44	13.44	13.44
20.70	15.00	15.75	15.75	15.92
21.51	15.34	16.05	16.22	16.22

TABLE F6: Effect of Solids Concentration on Flow Regime Transition in the 2-D Bubble Column

Three-phase System (Air-Water-Amberlite)

	Conc (Vol %)			
U_{SG} (cm/s)	0	5	10	15
3.48	3.14	3.14	3.18	3.20
4.80	4.15	4.11	4.20	4.25
7.63	6.12	6.10	6.30	6.38
10.57	8.20	8.00	8.45	8.59
13.62	10.36	10.07	10.60	10.77
17.00	12.70	12.37	12.86	13.07
20.70	15.00	14.68	15.22	15.38
21.51	15.34	15.06	15.57	15.81

TABLE F7: Effect of Solid Size on Flow Regime Transition in the 2-D Bubble Column

Three-phase System (Air-Water-Styrocell)

	d_p (mm)			
U_{SG} (cm/s)	1.23	0.85	0.81	0.60
3.48	3.30	3.29	3.27	3.25
4.80	4.43	4.42	4.40	4.36
7.63	6.78	6.74	6.72	6.61
10.57	9.00	8.97	8.91	8.84
13.62	11.18	11.17	11.07	10.88
17.00	13.58	13.50	13.44	13.32
20.70	16.00	15.96	15.75	15.44
21.51	16.22	16.13	16.05	15.90

APPENDIX G

Introductory Studies on Mixing in Two- and Three-Phase Fluidisation

(Tables of Experimental Results)

Figure G1 Calibration of the spectrophotometer

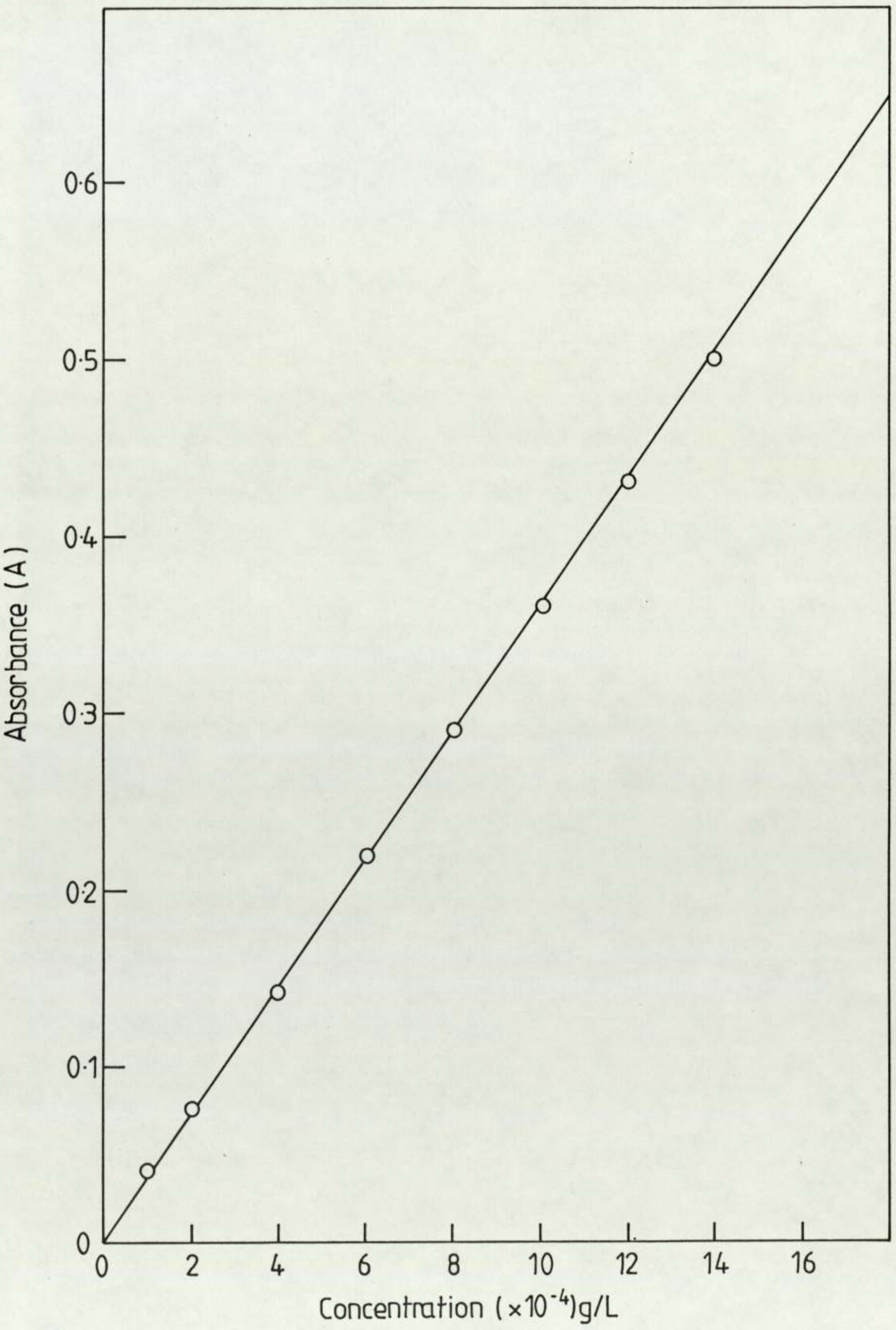

TABLE G1: Liquid Phase Mixing in the 3-D Bubble Column

Two-phase System (Air-Water)

U_{SG} = 4.80 cm/s X = 176.5 cm L = 200 cm X/L = 0.88

t sec	A	C (t) g/l	C (t)/C(E)	$(\pi/L)^2 D_L t$	D_L (cm^2/s)
30	0.102	2.8×10^{-4}	0.7	1.8	243.17
40	0.124	3.4	0.85	2.5	253.30
50	0.130	3.6	0.9	2.9	235.07
60	0.138	3.8	0.95	3.5	236.42
180	0.144	4.0	1.0	8.5	-

(D_L)av = 242 cm^2/s

TABLE G2:

$U_{SG} = 6.81$ cm/s

t (sec)	A	C (t) g/l	C (t)/C(E)	$(\pi/L)^2 D_L t$	D_L (cm²/s)
10	0.024	0.60×10^{-4}	0.154	0.70	283.70
20	0.064	1.70	0.436	1.2	243.17
30	0.104	2.9	0.744	1.95	263.44
50	0.134	3.7	0.949	3.5	283.70
60	0.136	3.75	0.962	3.7	250.0
120	0.142	3.9	1.0	8.5	-

$(D_L)_{av} = 265$ cm²/s

TABLE G3

U_{SG} = 10.7 cm/s

t sec	A	C (t) g/l	C (t)/C(E)	$(\pi/L^2)\, D_L\, t$	D_L (cm^2/s)
15	0.040	1.1×10^{-4}	0.262	0.9	243.17
20	0.070	1.95	0.464	1.2	243.17
40	0.132	3.61	0.859	2.55	258.37
50	0.138	3.8	0.905	3.0	243.17
120	0.150	4.2	1.0	8.5	-

$(D_L)_{av}$ = 247 cm^2/s

TABLE G4

U_{SG} = 14.61 cm/s

t sec	A	C (t) g/l	C (t)/C (E)	$(\pi/L^2)\,D_L t$	D_L (cm^2/s)
15	0.045	1.2×10^{-4}	0.27	0.92	248.6
20	0.092	2.5	0.566	1.44	291.8
40	0.138	3.8	0.860	2.55	258.37
50	0.152	4.2	0.950	3.5	283.7
120	0.161	4.42	1.0	8.5	-

$(D_L)_{av}$ = 270.62 cm^2/s

TABLE G5

U_{SG} = 19.18 cm/s

t sec	A	C (t) g/l	C (t)/C (E)	$(\pi/L)^2 D_L t$	D_L (cm^2/s)
15	0.068	1.9×10^{-14}	0.404	1.1	297.21
20	0.098	2.7	0.574	1.45	293.83
30	0.138	3.65	0.777	2.07	279.65
40	0.156	4.3	0.915	3.05	309.03
60	0.165	4.6	0.979	4.5	303.96
120	0.168	4.7	1.0	8.5	-

$(D_L)_{av}$ = 296.74 cm^2/s

TABLE G6

$U_{SG} = 24.26$ cm/s

t sec	A	C (t) g/l	C (t)/C (E)	$(\pi/L^2)\,D_L t$	D_L (cm^2/s)
15	0.089	2.5×10^{-4}	0.543	1.38	372.86
20	0.116	3.2	0.696	1.8	364.76
30	0.144	4.0	0.870	2.65	358.00
40	0.158	4.4	0.957	3.6	364.76
120	0.162	4.6	1.0	8.5	-

$(D_L)_{av} = 365.1$ cm^2/s

TABLE G7

U_{SG} = 28.81 cm/s

t sec	A	C (t) g/l	C (t)/C (E)	$(\pi/L)^2 D_L t$	D_L (cm^2/s)
15	0.087	2.4×10^{-4}	0.608	1.55	418.8
20	0.104	2.9	0.734	1.93	391.1
30	0.128	3.6	0.911	3.0	405.28
50	0.141	3.9	0.987	5.0	405.28
60	0.143	3.95	1.0	8.5	-

$(D_L)_{av}$ = 405.12 cm^2/s

TABLE G8: Liquid Phase Mixing in the 3-D Bubble Column - Effect of Solid Particles

Three-phase System (Air-Water-5% Styrocell R551)

U_{SG} = 4.80 cm/s

X = 176.5 cm L = 200cm X/L = 0.88

t sec	A	C (t) g/l	C (t)/C(E)	$(\pi/L)^2 D_L t$	$D_L (cm^2/s)$
20	0.036	1.0×10^{-4}	0.294	0.95	192.51
30	0.072	2.0	0.588	1.50	202.64
40	0.098	2.7	0.794	2.18	220.88
50	0.107	3.0	0.882	2.75	222.91
60	0.114	3.15	0.926	3.25	219.53
70	0.122	3.4	1.0	8.50	-

$(D_L)_{av}$ = 211.69 cm²/s

TABLE G9

$U_{SG} = 6.81 \text{cm/s}$

t sec	A	C (t) g/l	C (t)/C(E)	$(\pi/L)^2 D_L t$	D_L (cm^2/s)
25	0.057	1.6	0.485	1.28	207.5
30	0.068	1.9	0.576	1.46	197.2
50	0.102	2.8	0.848	2.50	202.64
60	0.111	3.1	0.939	3.6	243.17
120	0.120	3.3	1.0	8.5	-

$(D_L)_{av} = 212.63 \text{ cm}^2/\text{s}$

TABLE G10

U_{SG} = 10.7 cm/s

t sec	A	C (t) g/l	C (t)/C (E)	$(\pi/L)^2 D_L t$	D_L (cm^2/s)
20	0.048	1.3×10^{-4}	0.361	1.05	212.77
30	0.083	2.3	0.639	1.63	220.20
40	0.113	3.1	0.861	2.58	261.40
50	0.118	3.3	0.917	3.1	251.28
70	0.127	3.5	0.972	4.5	260.54
120	0.131	3.6	1.0	8.50	-

$(D_L)_{av}$ = 241.24 cm^2/s

TABLE G11

$U_{SG} = 14.61$ cm/s

t sec	A	C (t) g/l	C (t)/C (E)	$(\pi/L)^2 D_L t$	D_L (cm^2/s)
20	0.067	1.85×10^{-4}	0.529	1.35	273.57
30	0.104	2.85	0.814	2.28	308.02
40	0.116	3.2	0.914	3.0	303.96
60	0.123	3.4	0.971	4.50	303.96
120	0.128	3.5	1.0	8.50	287.08

$(D_L)_{av} = 295.32$ cm^2/s

TABLE G12

U_{SG} = 19.18 cm/s

t sec	A	C (t)g/l	C (t)/C (E)	$(\pi/L)^2 D_L t$	D_L (cm^2/s)
20	0.082	2.25 x 10-4	0.592	1.52	303.96
30	0.114	3.15	0.829	2.38	321.53
40	0.122	3.4	0.895	2.90	293.83
50	0.128	3.5	0.921	3.1	251.28
60	0.135	3.7	0.974	4.2	283.70
70	0.136	3.8	1.0	8.50	-

$(D_L)_{av}$ = 290.86 cm^2/s

TABLE G13

$U_{SG} = 24.26$ cm/s

t sec	A	C (t) g/l	C (t)/C (E)	$(\pi/L)^2 D_L t$	D_L (cm^2/s)
20	0.090	2.5×10^{-4}	0.676	1.75	354.62
40	0.130	3.6	0.973	4.0	405.28
50	0.132	3.65	0.986	5.0	405.28
60	0.132	3.65	0.986	5.0	337.74
70	0.134	3.7	1.0	8.5	-

$(D_L)_{av} = 375.73$ cm^2/s

TABLE G14

$U_{SG} = 28.81$ cm/s

t sec	A	C (t) g/l	C (t)/C (E)	$(\pi/L)^2 D_L t$	D_L (cm^2/s)
20	0.100	2.75×10^{-4}	0.679	1.75	354.62
30	0.126	3.5	0.864	2.58	348.54
35	0.134	3.7	0.914	3.0	347.39
40	0.135	3.75	0.926	3.25	329.30
70	0.146	4.05	1.0	8.5	-

$(D_L)_{av} = 344.96$ cm^2/s

APPENDIX H

Pressure Correction for Gas Flow Rate Measurement

PRESSURE CORRECTION FOR GAS FLOW RATE MEASUREMENT

SUPERFICIAL GAS VELOCITY, U_{SG}, is the gas volumetric flow rate at a given cross section of the column divided by the cross-sectional area.

$$U_{SG} = \frac{Q}{A} \qquad \text{........ H1}$$

Where Q is the gas flow rate, A = local cross-sectional area.

The gas flow rate was measured by taking readings from a rotameter incorporated into the air line. A rotameter chart was then used to estimate the flow rate and as the mains pressure was different from atmospheric pressure a correction was made.

Correction for Gas Flow-rate Measurement

Mass flow rate is given as (Coulson and Richardson, 1977):

$$G = C_D A_2 \sqrt{\frac{2g V_f \rho (\rho_f - \rho)}{A_f \{1 - (A_2/A_1)^2\}}} \text{ for a variable area meter.} \qquad \text{........ H2}$$

ie

$$G \propto \sqrt{\rho (\rho_f - \rho)} \qquad \text{........ H3}$$

with s - standard or calibration conditions, a - actual condition.

$$G_s \propto \sqrt{\rho_s (\rho_f - \rho_s)} \qquad \text{........ H4}$$

$$G_a \propto \sqrt{\rho_a (\rho_f - \rho_a)} \qquad \text{........ H5}$$

But,

$G = Q\rho$ H6

Hence

$Q_s \propto 1/\rho_s \sqrt{\rho_s(\rho_f - \rho_s)}$ H7

and

$Q_a \propto 1/\rho_a \sqrt{\rho_a(\rho_f - \rho_a)}$ H8

Assume

$\rho_f >> \rho_a$ and ρ_s.

Therefore $Q_s \propto \sqrt{1/\rho_s}$ H9

$Q_a \propto \sqrt{1/\rho_a}$ H10

Dividing gives:

$Q_a/Q_s = \sqrt{\rho_s/\rho_a}$ H11

But

$$\frac{\rho_s}{\rho_a} = \frac{P_a}{T_a} \cdot \frac{T_s}{P_s}$$ H12

Therefore

$$Q_a = Q_s \sqrt{\frac{T_s}{T_a} \cdot \frac{P_a}{P_s}}$$ H13

Since $T_a \doteqdot T_s$

$Q_a = Q_s \sqrt{P_a/P_s}$ H14

Hence the correction formula adopted is:

$$Q_{corrected} = Q_{graph} \sqrt{\frac{\text{Actual Air Pressure (Mains Pressure)}}{\text{Atmospheric pressure}}} \quad \text{........ H15}$$

Q corrected is thus used in equation H1 in calculating the superficial gas velocity, U_{SG}.

For the TWO-DIMENSIONAL COLUMN

$$Q_{corrected} = Q_{graph} \sqrt{\frac{10.0 + 14.7}{14.7}}$$

$$= Q_{graph} \times 1.3$$

For the THREE-DIMENSIONAL COLUMN

$$Q_{corrected} = Q_{graph} \sqrt{\frac{32 + 14.7}{14.7}}$$

$$= Q_{graph} \times 1.78$$

Any variation in line pressure is accounted for accordingly.

REFERENCE

Coulson JM and Richardson JF: "Chemical Engineering", Vol 1, 3rd Ed, Pergamon Press, Oxford (1977)

SYMBOLS

A_1	Area at the height of float, cm^2
A_2	Area of the annulus, cm^2
A_f	Area of rotameter float, cm^2
C_D	Coefficient of discharge (-)
g	Acceleration due to gravity cm/s^2
G	Mass flow rate, g/s
Q	Volumetric flow rate, cm^3/s
V_f	Volume of rotameter float, cm^3
ρ	Density of fluid, g/cm^3
ρ_f	Density of rotameter float, g/cm^3

APPENDIX I

Listing of Computer Programs

PROGRAM 1

This program compares experimental gas hold-up data with empirical correlations for two-phase systems

NOMENCLATURE

G	Gravitational acceleration
DC	Column diameter
RHOL	Liquid density
RHOG	Gas density
SIGMA	Surface tension
KVIS	Kinematic viscosity
DVISL	Liq dynamic viscosity
DVISG	Gas dynamic viscosity
USG	Superficial gas velocity
EG	Experimental gas hold-up

```
REAL KVIS
COMMON/BLOCK1/USG, RHOL, SIGMA, G
COMMON/BLOCK2/DC, KVIS
COMMON/BLOCK3/DVISL, RHOG
COMMON/BLOCK4/DVISG

READING IN OF DATA
------------------------------------------------------------------------------------------------

DATA G, DC, RHOL, RHOG, SIGMA, KVIS, DVISL, DVISG/981.0, 15.2, 1.0,
1.21E-3, 72.8, 0.01, 0.0101, 0.000181/
PRINT 10
PRINT 20
PRINT 25
PRINT 30
DO 50 I=1, 8
READ *, USG, EG

CALLING THE CORRELATIONS
------------------------------------------------------------------------------------------------

CALL AKITA (EGA)
CALL KUMAR (EGK)
CALL MERSM (EGM)
CALL HIKITA (EGH)

PRINTING FORMAT
------------------------------------------------------------------------------------------------

PRINT 40, USG, EG, EGA, EGK, EGM, EGH

10FORMAT ( / 40X, "TABLE    "/)
20FORMAT ( /5X, "COMPARISON OF EXPERIMENTAL GAS HOLD-UP DATA
WITH LITERATURE CORRELATIONS. "/)
25FORMAT ( /10X," : TWO-PHASE SYSTEMS (Air-Water). "/)
30FORMAT ( /5X, "USG (cm/s)", 3X, "EXPTL, EG" 3X, "AKITA-YOSHIDA), 3X,
"KUMAR ET AL", 3X, "MERSMANN", 3X, "HIKITA ET AL "/)
40FORMAT ( /6X, F5.2, 7X, F5.3, 8X, F5.3, 10X, F5.3, 9X, F5.3, 6X, F5.3/)
```

```
50CONTINUE
STOP
END


THE CORRELATION OF AKITA AND YOSHIDA (1973)
----------------------------------------------------------------------------------------------------

PROGRAM SOLVES FOR EGA USING THE CORRELATION

SUBROUTINE AKITA (EGA)
REAL KVIS
COMMON/BLOCK1/USG, RHOL, SIGMA, G
COMMON/BLOCK2/DC, KVIS
A1=(G*DC**2*RHOL/SIGMA)**0.125
A2=(G*DC**3/KVIS**2)**0.0833
A3=USG/SQRT(G*DC)
EGA = 0.25
1FUNC = (EGA/(1.0-EGA)**4)-0.2*A1*A2*A3
DIFF-(1.0+3.0*EGA)/(1.0-EGA)**5
EGA1=EGA-FUNC/DIFF
IF (ABS((EGA-EGA1) LT 1 0E-6) GO TO 2
EGA=EGA2
GO TO 1
2CONTINUE
RETURN
END


THE CORRELATION OF KUMAR ET AL (1976)
----------------------------------------------------------------------------------------------------

PROGRAM SOLVES FOR EGK USING THE CORRELATION

SUBROUTINE KUMAR (EGK)
COMMON/BLOCK1/USG, RHOL, SIGMA, G
U=USG*(RHOL/(SIGMA*G))**0.25
EGK=0.728*U-(0.485*U**2)+(0.0975*U**3)
RETURN
END


THE CORRELATIONS OF MERSMANN (1978)
----------------------------------------------------------------------------------------------------

PROGRAM SOLVES FOR EGM USING THE CORRELATION

SUBROUTINE MERSM (EGM)
REAL M1, M2, M3
COMMON/BLOCK1/USG, RHOL, SIGMA, G
COMMON/BLOCK3/DVISL, RHOG
M1=(RHOL/SIGMA*G)**0.25
M2=(RHOL*SIGMA**3/(DVISL**4*G))**0.0417
M3=(RHOL/RHOG)**0.06944
EGM=0.95
3FUNC=(EGM.(1.0-EGM)**4)-0.14*USG*M1*M2*M3
DIFF=(1.0+3.0*EGM)/(1.0-EGM)**5
EGM1=EGM-FUNC/DIFF
```

```
IF (ABS((EGM-EGM1)/EGM). LT. 1.0E-6) GO TO 4
EMG-EMG1
GO TO 3
4CONTINUE
RETURN
END


THE CORRELATION OF HIKITA ET AL (1980)
----------------------------------------------------------------------------------------------------

PROGRAM SOLVES FOR EGH USING THE CORRELATION

SUBROUTINE HIKITA (EGH)
COMMON/BLOCK1/USG, RHOL, SIGMA, G
COMMON/BLOCK3/DVISL, RHOG
COMMON/BLOCK4/DVISG
H1= (USG*DVISL/SIGMA)**0.578
H2=(DVISL**4*G/(RHOL)**0.062
H4=(DVISG/DVISL)**0.107
EGH=0.672*H1*H3*H4/H2
RETURN
END
----------------------------------------------------------------------------------------------------
```

PROGRAM 2

This program compares experimental gas hold-up data with empirical correlations for three-phase systems

NOMENCLATURE

G	Gravitational acceleration
DC	Column diameter
RHOL	Liquid density
RHOG	Gas density
SIGMA	Surface tension
KVIS	Kinematic viscosity
DVISL	Liquid dynamic viscosity
DVISG	Gas dynamic viscosity
USG	Superficial gas velocity
EG	Experimental gas hold-up

```
Real KVIS
COMMON/BLOCK1/USG
COMMON/BLOCK2/RHOL, SIGMA, G
COMMON/BLOCK3/KVIS, DC
COMMON/BLOCK4/RHOG, DVISL, DVISG

READING IN OF DATA
------------------------------------------------------------------------------------------------------
Data G, DC, RHOL, RHOG, SIGMA, KVIS,  DVISL,  DVISG/981. 0, 15.2, 1.0,
1.21E-3, 72.8, 0.01, 0.0101, 0.000181/

PRINT 10
PRINT 20
PRINT 25
PRINT 30
DO 50 I=1, 8
READ *, USG, EG

CALLING THE CORRELATIONS
------------------------------------------------------------------------------------------------------

CALL AKITA (EGA)
CALL KITO (EGK)
CALL BEGWATS (EGB)
CALL HIKITA (EGH)

PRINTING FORMAT
------------------------------------------------------------------------------------------------------

PRINT 40, USG, EG, EGA, EGK, EGB, EGH

10FORMAT ( /40X,"TABLE       "/)

20FORMAT ( /40X, "COMPARISON OF EXPERIMENTAL GAS HOLD-UP DATA
WITH LITERATURE CORRELATIONS"/)
25FORMAT ( /10X, ":THREE-PHASE SYSTEM (Air-Water-Styrocell)."/)
30FORMAT (/5X, "USG (cm/s)", 3X, "EXPTL. EG", 3X, "AKITA-YOSHIDA", 3X
"KITO ET AL", 3X, "BEGOWATS", 3X, "HIKITA ET AL"/)
40FORMAT ( /6X, F5.2, 7X, F5.3, 8X, F5.3, 10X, F5.3, 9X, F5.3, 6X, F5.3/)
```

```
50CONTINUE
STOP
END
```

THE CORRELATION OF AKITA AND YOSHIDA (1973)

PROGRAM SOLVES FOR EGA USING THE CORRELATION

```
SUBROUTINE AKITA (EGA)
REAL KVIS
COMMON/BLOCK1/USG
COMMON/BLOCK2/RHOL, SIGMA, G
COMMON/BLOCK3/KVIS, DC
A1= (G*DC**2*RHOL/SIGMA)**0.125
A2=(G*DC**3/KVIS**2)**0.0833
A3=USG/SQRT(G*DC)
EGA = 0.25
1FUNC= (EGA/(1.0-EGA)**4)-0.2*A1*A2*A3
DIFF=(1.0+3 0*EGA)/(1.0-EGA)**5
EGA1=EGA-FUNC/DIFF
IF (ABS( (EGA-EGA1)/EGA). LT. 1.0E-6) GO TO 2
EGA=EGA1
GO TO 1
2CONTINUE
RETURN
END
```

THE CORRELATION OF KITO ET AL (1976)

PROGRAM SOLVES FOR EGK USING THE CORRELATION

```
SUBROUTINE KITO (EGK)
REAL K1
COMMON/BLOCK1/USG
COMMON/BLOCK2/RHOL, SIGMA, G
K1=((USG**4*RHOL)/(G*SIGMA))**0.11
EGK=0.95
3FUNC=(EGK/EGK*(1.0-EGK)**2)**0.44)-0.5*K1
DIFF=(0.56+0.32*EGK)/(1.0-EGK)/(EGK*(1.0-EGK)**2)**0.44
EGK1=EGK-FUNC/DIFF
IF (ABS((EGK-EGK1)/EGK). LT. 1.0E-6) GO TO 4
EGK=EGK1
GO TO 3
4CONTINUE
RETURN
END
```

THE CORRELATION OF BEGOVICH AND WATSON (1978)

```
PROGRAM SOLVES FOR EGB USING THE CORRELATION.

SUBROUTINE BEGWATS (EGB)
COMMON/BLOCK1/USG
EGB=1.15*((USG/100.0)**.692)*(.0081**.107)*(.152**(-.037))
RETURN
END

THE CORRELATION OF HIKITA ET AL (1980)
------------------------------------------------------------------------------------------------

PROGRAM SOLVES FOR EGH USING THE CORRELATION

SUBROUTINE HIKITA (EGH)
COMMON/BLOCK1/USG
COMMON/BLOCK2/RHOL, SIGMA, G
COMMON/BLOCK4/RHOG, DVISL, DVISG
H1=(USG*DVISL/SIGMA)**0.578
H2=(DVISL**4*G/(RHOL*SIGMA**3))**0.131
H3=(RHOG/RHOL)**0.062
H4=(DVISG/DVISL)**0.107
EGH=0.672*H1*H3*H4/H2
RETURN
END
------------------------------------------------------------------------------------------------
```

PROGRAM 3

This program compares experimental dispersion coefficient data with literature correlations for 2- and 3-phase systems

NOMENCLATURE

G	Gravitational acceleration
DC	Column diameter
RHOL	Liquid density
RHOG	Gas density
SIGMA	Surface tension
KVIS	Kinematic viscosity
DVISL	Liq dynamic viscosity
DVISG	Gas dynamic viscosity
USG	Superficial gas velocity
DL	Axial liq dispersion coeff

```
REAL KVIS
COMMON/BLOCK1/USG,DC
COMMON/BLOCK2/G
COMMON/BLOCK3/DVISL
COMMON/BLOCK4/KVIS

READING IN OF DATA
----------------------------------------------------------------------------------------------------

DATA G, DC, DVISL, KVIS/9.810,0,152,1.0,0.01/
PRINT 10
PRINT 20
PRINT 30
PRINT 40
PRINT 50
PRINT 60
PRINT 70
DO 80 I=1,7
READ *, USG, DL2, DL3

CALLING THE CORRELATIONS
----------------------------------------------------------------------------------------------------

CALL DECKWER (DLD)
CALL BAIRD (DLB)
CALL HIKITA (DLH)
CALL RIQUARTS (DLR)

PRINTING FORMAT
----------------------------------------------------------------------------------------------------

PRINT 70, USG, DL2, DL3, DLD, DLB, DLH, DLR

10FORMAT ( /40X, "TABLE   "/)
20FORMAT (/5X, "COMPARISON OF LIQ DISPERSION COEFFICIENT DATA
WITH LITERATURE CORRELATIONS  "/)
30FORMAT ( /10X, "  : TWO AND THREE PHASE SYSTEMS.  "/)
40FORMAT ( /17X, "EXPERIMENTAL")
50FORMAT ( /5X, "USG (m/s)", 3X, "DL-2P", 4X, "DL-3P", 4X, "DECKWER",
4X,
```

```
"BAIRD-RICE", 3X, "HIKITA ET AL", 4X, "RIQUARTS"/)
60FORMAT (15X, "m**2/s", 4X, "m**2/s", "m**2/s", 6X, "m**2/s", 8X,
"m**2/s"/)
70FORMAT ( /6X, F6.4, 3X, F6.4, 4X, F6.4, 4X, F6.4, 8X, F6.4, 8X, F6.4/)

80CONTINUE
STOP
END


THE CORRELATION OF DECKWER (1974)
-----------------------------------------------------------------------------------------------

PROGRAM SOLVES FORDLD USING THE CORRELATION

SUBROUTINE DECKWER (DLD)
COMMON/BLOCK1/USG, DC
DLD=0.678*DC**1.4*USG**0.3
RETURN
END


THE CORRELATION OF HIKITA ET AL (1974)
-----------------------------------------------------------------------------------------------

PROGRAM SOLVES FOR DLH USING THE CORRELATION

SUBROUTINE HIKITA (DLH)
COMMON/BLOCK1/USG, DC
COMMON/BLOCK2/DVISL
DLH=(0.15+0.69*USG**0.77)*DC**1.25*(1.0/DVISL)**0.12
RETURN
END


THE CORRELATION OF BAIRD AND RICE (1975)
-----------------------------------------------------------------------------------------------

PROGRAM SOLVES FOR DLB USING THE CORRELATION

SUBROUTINE BAIRD (DLB)
COMMON/BLOCK1/USG, DC
COMMON/BLOCK2/G
DLB=0.35*DC**1.333*(USG*G)**0.333
RETURN
END


THE CORRELATION OF RIQUARTS (1981)
-----------------------------------------------------------------------------------------------

PROGRAM SOLVES FOR DLR USING THE CORRELATION

SUBROUTINE RIQUARTS (DLR)
REAL KVIS
COMMON/BLOCK1/USG, DC
COMMON/BLOCK2/G
COMMON/BLOCK4/KVIS
```

```
R1=SQRT(DC*G*10000.0)
R2=((USG*100.0)**3/(KVIS*G*100.0))**0.125
DLR=(0.068*DC*R1*R2)/100.0
RETURN
END
------------------------------------------------------------------------------------------------
```